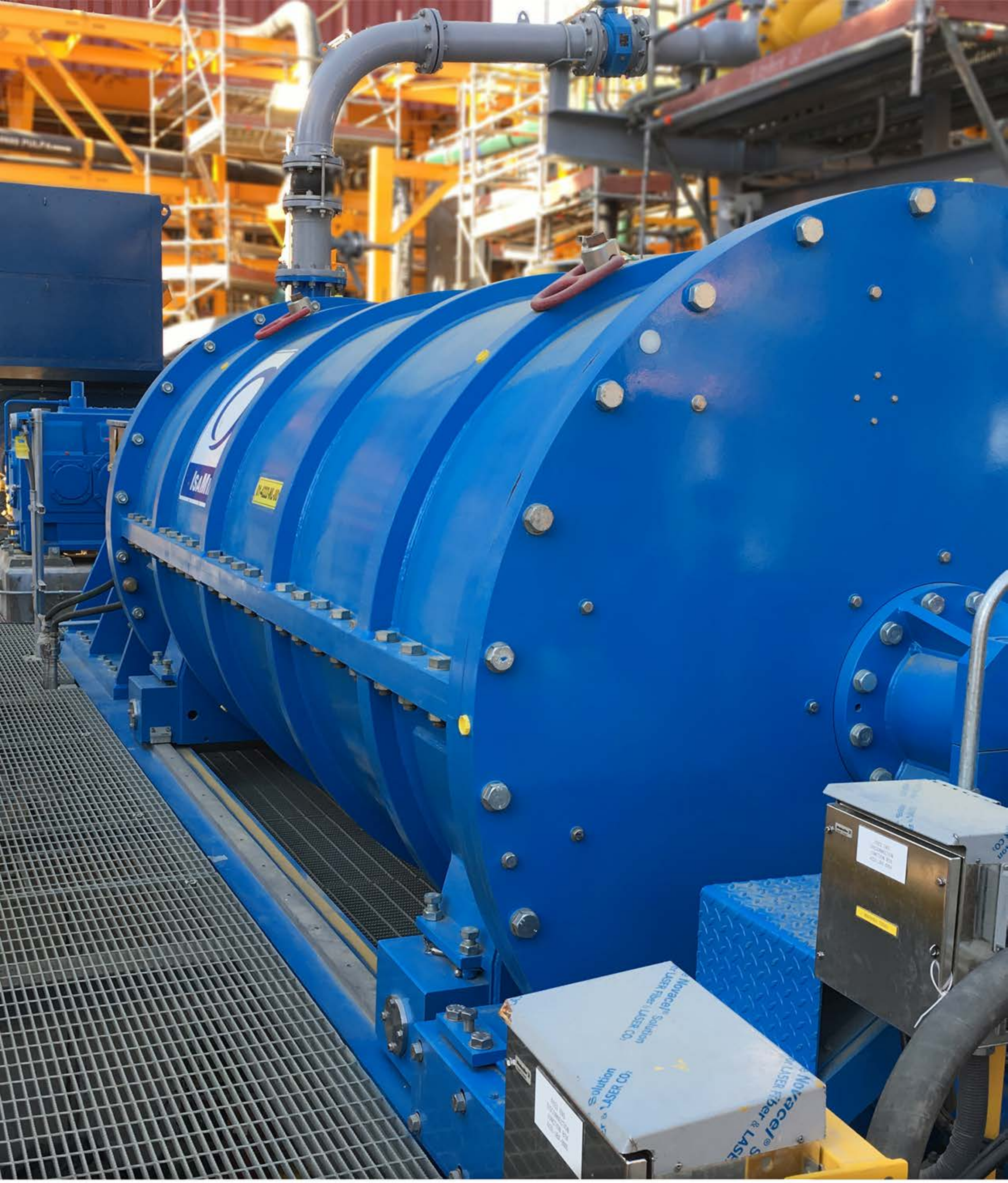
GLENCORE TECHNOLOGY



ISAMILL™ 25 YEARS

A HISTORY OF ISAMILL PROGRESS AT THE TECK RED DOG MINE

*Michael Larson[1], Brigitte Lacouture[2], & Greg Anderson[3]

[1]Molycop USA
8116 Wilson Road, Kansas City, MO 64125, USA

[2]Teck, Red Dog Mine
3105 Lakeshore Drive, Suite A101, Anchorage, AK 99517, USA

[3]Glencore Technology
Level 10, 160 Ann St, Brisbane, QLD 4000, Australia

(*Corresponding author: mike.larson@molycop.com)

Abstract

In December 2011, Teck's Red Dog Operation commissioned two 1.5 MW M3000 IsaMills as part of a project to improve their zinc metallurgy. This paper examines the history, including initial performance, characterization of a feed that has managed to be both abrasive and viscous at the same time, reviews improvements to the mill flexibility through an operating vessel size upgrade and the optimization of the internal component configuration for improved wear life. Red Dog has also completed a program for grinding media optimization. Recently, Red Dog finalized testwork and design on a value improvement project (VIP#2) that will install the world's first M15000 IsaMill in 2019 into their grinding circuit to ensure throughput and grind size targets are maintained as harder ores are processed in the near future.

Keywords

IsaMill, regrind

Introduction

The Teck Red Dog Operation in Alaska has been running two IsaMills in the zinc circuit since 2011. These two IsaMills replaced seven smaller vertical tower mills. The circuit is shown in Figure 1. One IsaMill treats the zinc rougher concentrate while the other treats the feed to the zinc retreat circuit.

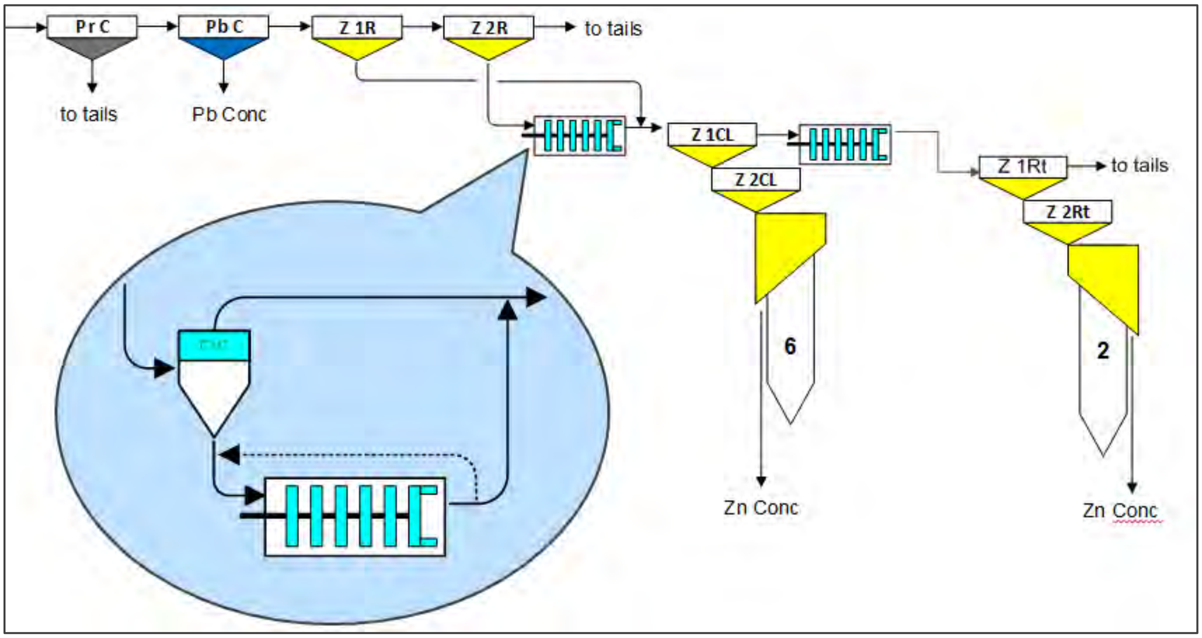


Figure 1 – Teck Red Dog flotation and regrind flowsheet (Lacouture, 2013)

The following will attempt to explain the methods involved in optimizing the wear and reliability of the two Red Dog IsaMills. The focus of this was not to optimize the grinding performance, but to improve the runtime, stability, and internal component wear of the two mills. This was a more demanding project, in that in a typical IsaMill operational challenge, either abrasive wear or viscosity will be an issue. Red Dog has both. The general operating conditions of the two Red Dog IsaMills are shown in Table 1.

This optimization program turned out to be a complex process involving flow, mixing, viscosity, grinding media, internal components, and the entire shell liner. Thermal imaging and regular viscosity measurements would prove to be critical tools in this process.

Table 1 – Teck Red Dog IsaMill feed and operating parameters (Lacouture, 2013)

Parameter	Zn Rougher	Zn Retreat
Cyclone U/F (µm – Malvern)	90 – 120	50 – 70
Discharge (µm Malvern)	45 – 65	3 – 50
Recombine (µm Malvern)	28 – 37	20 – 30
Power (kW)	1000 – 1200	900 – 1100

Parameter	Zn Rougher	Zn Retreat
Power (kWh/t)	11 – 13	14 – 17
Feed Flowrate (gpm/m^3/h)	500/115	400/90
Feed Density (%)	53 – 60	47 – 55
Feed Viscosity (cp)	20 – 40	20 – 40
Feed Silica (%)	17 – 27	35 – 55
Feed Barite (%)	7 – 12	8 – 15

Discussion

DISC SIZE AND THERMAL IMAGING

From January 2012 (one month after startup) to June 2012 Glencore Technology and Red Dog experimented with different disc configurations in the rougher IsaMill to minimize disc and shell liner wear. This was aided with the use of a thermal imaging camera to track media movement and compression in the mill. The thermal imagery has proved to be a useful indication of the initiation and development of wear areas within the mill, particularly on the shell lining.

Initially the rougher mill was operated at around 1250 kW with excessive wear found towards the feed end of the mill on both the discs and shell liner. The thermal imaging in Figure 2 (Anderson, 2012) shows the concentration of media towards the feed end of the mill resulting in higher heat generation and wear on the shell liner.

As an initial step to address this, smaller diameter discs (SDD) were installed into the mill. Initially the maximum of seven were installed from the feed end, leaving a single normal-diameter disc in front of the rotor. The SDDs rotate at the same shaft speed but at a lower tip speed due to the reduced diameter. In addition, the smaller diameter creates a larger gap between the disc tip and the shell that allows for reduced compression and a more fluidized media pattern at the shell liner, as well as a reduced media impact speed on the shell liner. Upon startup under this configuration, the mill was limited to a power draw of 650 kW. Thermal imaging of the shell in Figure 3 (Anderson, 2012) showed the media had shifted towards the discharge end of the mill. The reduction in power was a direct result of the smaller disc diameter drawing less power per disc and also drawing the media further down the mill overcoming some of the pumping action of the rotor.

To improve the pumping action and distribution of the media within the mill, the rotor was reconfigured with a rhomboidal finger design in place of the standard square design. Adjustment of the rotor finger configuration allows for changes in the pumping volume of the rotor, the volume pumped by the rotor when operated at a fixed rotational speed. The disc configuration was left as seven SDDs. The resultant thermographic image is shown in Figure 4 (Anderson, 2012), clearly illustrating that the change in the rotor configuration had allowed the media to be pushed back towards the feed end of the mill. In addition, the overall power drawn by the mill was able to be increased to a maximum of 900 kW under these conditions. This configuration stabilized the wear on both the discs and shell liner in the mill. The disc wear can be seen from both the feed and discharge end in Figure 5.

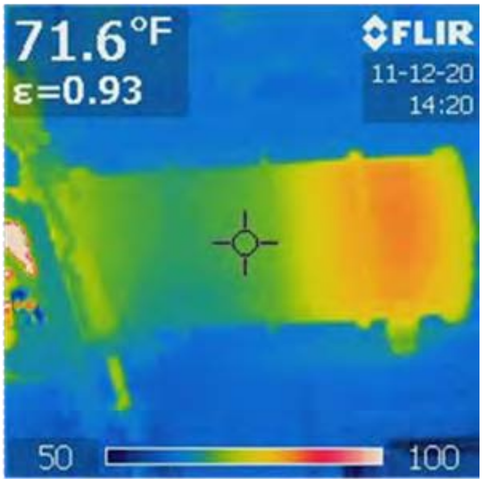


Figure 2 – IsaMill thermal image – gouging

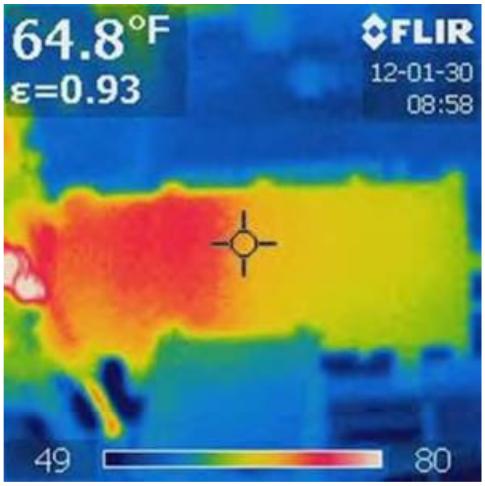


Figure 3 – IsaMill thermal image – media at discharge (left side)

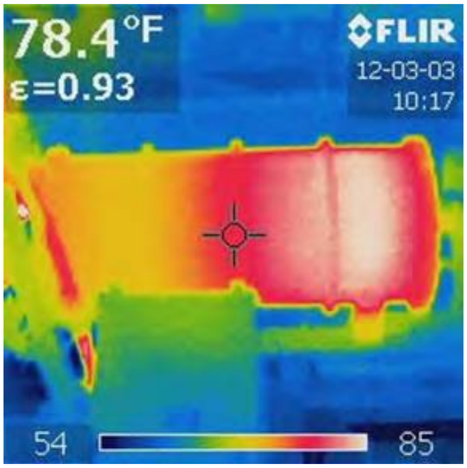


Figure 4 – IsaMill thermal image – media at feed end (right side)

Figure 5 – IsaMill small diameter disc installation from discharge (left picture) and feed (right picture) ends

SHELL LINER

In order to increase the power draw on the mill under stable wear conditions it became apparent that the Red Dog IsaMills could benefit from a retrofit in shell sizing. Though the original IsaMills in operation at Mount Isa and McArthur River Mine were 3,000 litre (L) shells, a 5,000 L shell had recently been designed for non ultra-fine grinding duties with larger ceramic media. This shell would allow for a larger gap between the shell wall and the grinding discs, thereby reducing wear while allowing for a higher power draw. The disc-to-shell ratio of the M5000 with the normal diameter discs would replicate the use of the SDDs in the M3000. This extra space allows the media to properly fluidize and mix between the discs and the shell wall, rather than packing and gouging the rubber. In September of 2012 Glencore Technology converted both mill shells from 3,000 L to 5,000 L. This allowed the use of full-size discs and a return to +1000 kW operation without the previous wear issues.

GRINDING MEDIA

When the Red Dog IsaMills were first started, the retreat mill utilized 2 mm ceramic media and the rougher mill 3.5 mm ceramic media. The original IsaMill grinding theory was to use the smallest media possible. This was valid when the mills were first introduced in the industry given the approximately 7-micron (µm) target regrind sizes from an already fine feed, and limited selection of grinding media at the time. However, with improvements in grinding media and a better understanding of the IsaMill grinding process it became clear that the original theory was flawed, and that in many cases larger media would be beneficial to both the grind and internal wear properties of the process. By utilizing larger media, the coarser particles present in the flotation regrind feeds can be broken down quicker, so they do not accumulate at the front of the mill and prematurely wear those rubber components at the feed end of the mill.

To prove this point, Red Dog contracted ALS Metallurgical in Kamloops, BC, to run a media comparison trial on the zinc retreat feed in March of 2012. ALS tested graded charges of three different media top sizes, 2, 2.8, and 3.5 mm. These energy signature plot results are shown in Figure 6.

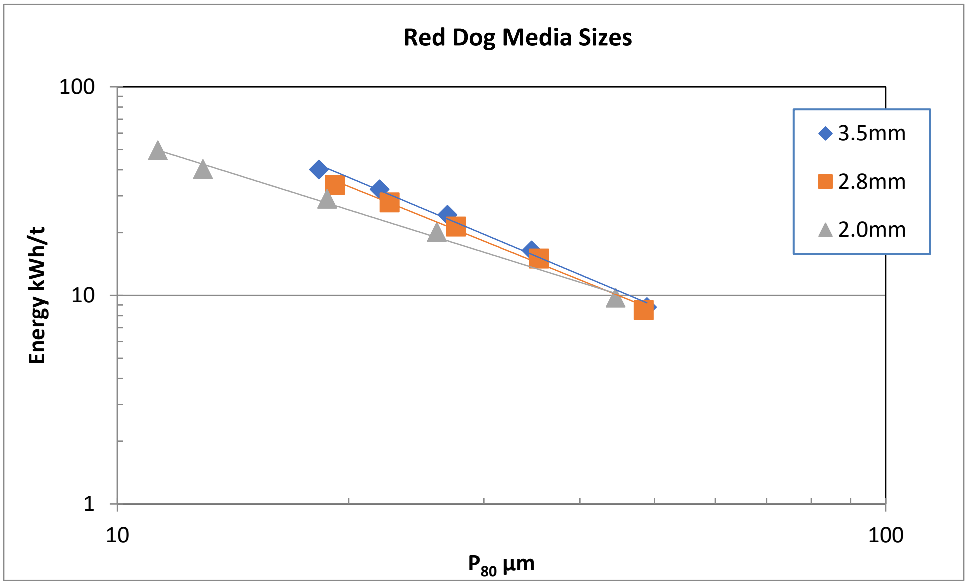


Figure 6 – IsaMill laboratory media sizing comparisons, retreat feed (Mehrfert, 2012)

The 2 mm results would indicate better performance, but what Figure 6 does not show is that the 2 mm test was holding in the coarse material, indicated by a higher motor power-draw — that insight is in the full test report.

As a result of this testwork, grinding media size was adjusted from 2 to 3 mm in the retreat. At the same time the rougher mill was switched from 3.5 to 4 – 5 mm. These changes also coincided with a change in supplier. This was brought on by availability of sizes, wear performance, and shape of worn media. The newer media was found to maintain a round shape longer than the original media, which would also contribute to improved component wear. As an added benefit it is thought that the relatively larger media with bigger gaps between packed pieces is impacted less by viscosity.

SEPARATOR

The discharge end of an IsaMill contains a centripetal separator that provides internal classification for the mill. This also provides the backflow necessary to keep grinding media and coarse particles in the mill. Typically, the fingers on this separator are a square or hexagonal shape. The Retreat IsaMill was experiencing higher than normal wear at the feed end flange, indicating a concentration of material in this area. To remedy the problem, some of the separator fingers were rounded (Figure 7) to make it less aggressive and reduce back flow.

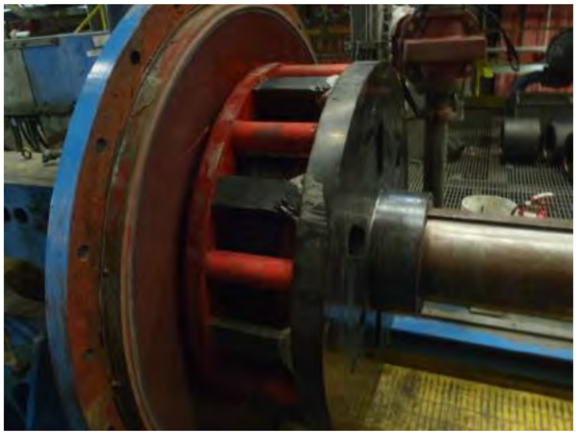


Figure 7 – IsaMill separator retrofitted with round fingers (Lacouture, 2013)

VISCOSITY

If viscosity were regularly measured at all IsaMill installations, it would be likely that Red Dog would rank at the top of challenging rheology. In this case, over the normal operating density range of 50% to 55% solids for the rougher mill, the viscosity increases by over 50% from the low to high point (Figure 8).

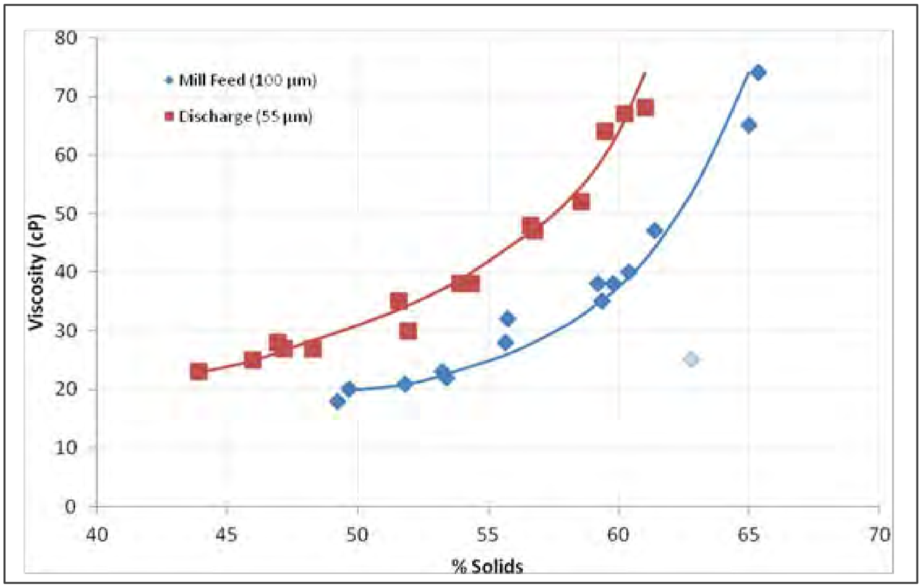


Figure 8 – Rougher IsaMill feed and discharge viscosity at variable densities (Merritt, 2012)

The mineralogy of barite, sphalerite, and silicates ground fine has resulted in some unique challenges to mill operations. The general power trend of this variable viscosity feed is shown in Figure 9. This is a good example that had been previously theorized for IsaMill operations but rarely demonstrated with actual plant data. As the feed solids and viscosity increase, the power draw increases, eventually past a point of optimum operation. As the viscosity continues to rise, the power draw will eventually drop off as the mixing of the mill decreases and the discs begin to rotate freely without agitating the charge adequately.

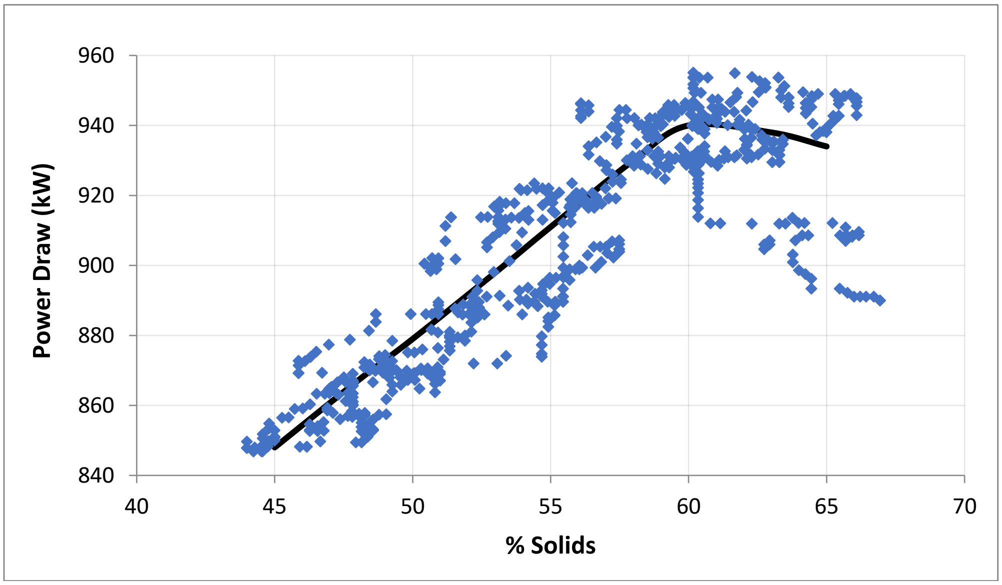


Figure 9 – Effect of density on power draw (Merritt, 2012)

One option to reduce viscosity was to reduce the amount of recycle directed back to the mill feed. The recycle is primarily used to maintain constant flow into the mill, but there was room to reduce this. This was preferred over reducing density, as it has historically been shown that operation at dilute density also increases the component wear in the mill. All else being normal, it is usually recommended to operate an IsaMill at about twenty percent solids by volume. Across sites this has been a good starting guide to regulate surface area present.

One interesting product of this work is the beginnings of a rheological model based on surface area of the solids in slurry. A series of samples were taken in the Red Dog plant and from laboratory grinding. The plant samples included feed and product from both IsaMills and the laboratory samples were ground for varying lengths of time in a laboratory rod mill. These were analyzed in a Malvern laser sizer for a full-size distribution. Each individual particle size was assumed to be a sphere for simplicity, and then assigned its given surface area. Given the solid specific gravity (SG) and slurry density the total number of each particle size and then the total surface area per unit of volume was calculated. This was plotted against the viscosity of each sample as measured in the Red Dog laboratory (Figure 10). In general, there will be some differences based on method of grind (and resulting particle shape) and slurry stream mineralogy, but overall this looks like it could potentially be a start to a guide to viscosity characteristics if ever necessary.

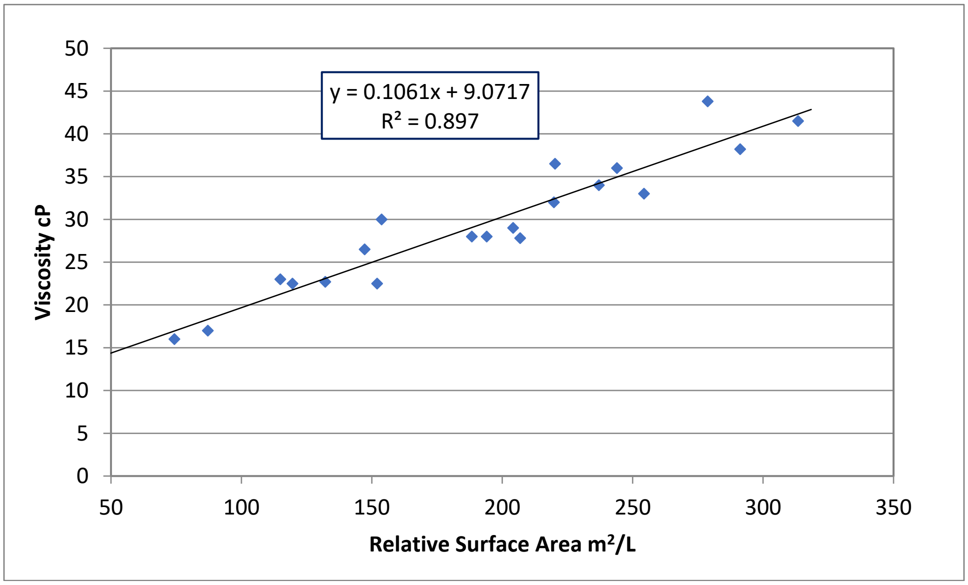


Figure 10 – Surface area vs. viscosity (Larson, 2012)

IsaMill Grind

A series of surveys was completed in June of 2013 to review the two IsaMills' performance. Figure 11 and Figure 12 show typical examples of mill breakage performance for the zinc retreat and rougher mills. In this case the rougher IsaMill ground from an F_{80} of 105 µm to a P_{80} of 42 µm at 10 kWh/t. The retreat IsaMill ground from an F_{80} of 45 µm to a P_{80} of 30 µm at 12 kWh/t.

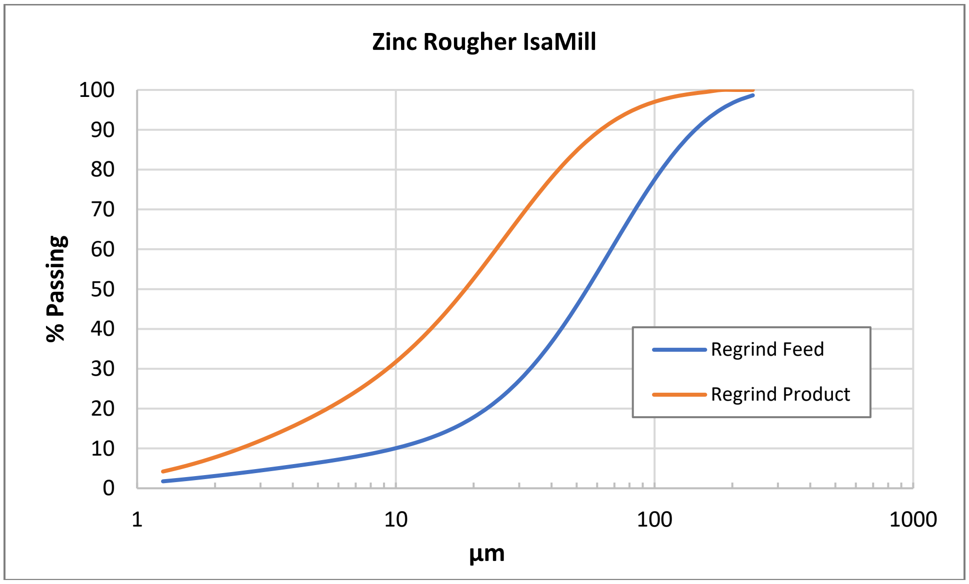


Figure 11 – Rougher IsaMill performance at 10 kWh/t (Larson, 2012)

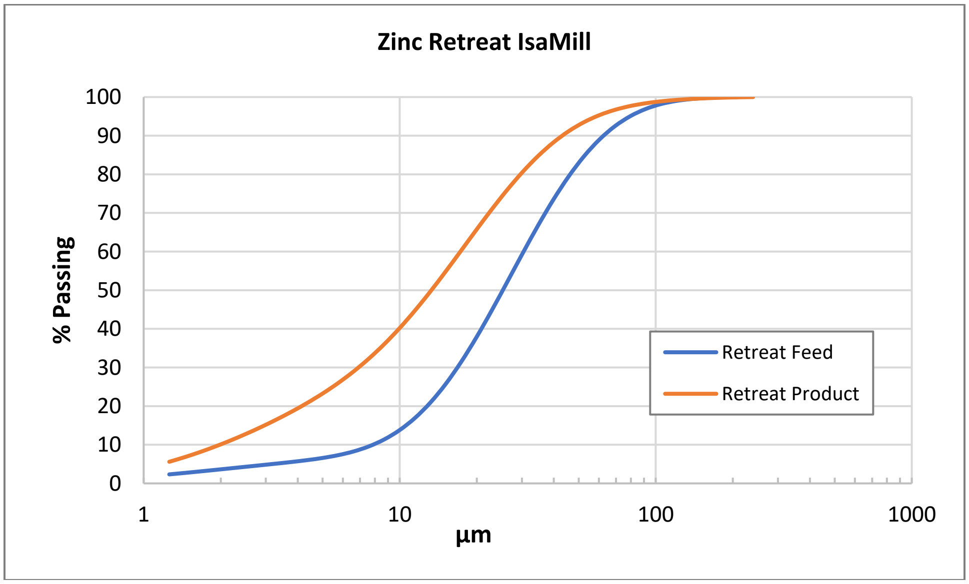


Figure 12 – Retreat IsaMill performance at 12 kWh/t (Larson, 2012)

M15000

In 2019 Red Dog will commission a new M15000 IsaMill as part of Red Dog's latest mill upgrade. The mill will treat the prefloat tailings, grinding the material from a feed of 150 µm down to a product of 65 µm to feed the lead flotation circuit. The M15000 was chosen over the M10000 due to the larger gap between the discs and shell liner, along with a capability to process a higher feed flow than the M10000.

Conclusions

Given the complexity of the issues around the Red Dog Zinc IsaMills and the relative quickness with which they were addressed, this program should be considered a success. Less than eight months after commissioning, the mill had been adjusted both in internal components and media sizing to reduce wear and internal instability. Within one-and-a-half years Red Dog gained an understanding of the effect of viscosity on their mills that should be the envy of any number of researchers. This has resulted in more than seven years of reliable operation of the first two IsaMills at Red Dog and has given the team the confidence to install the first ever M15000 IsaMill to maintain mill throughput with harder ores.

Acknowledgements

The authors wish to thank the Teck Red Dog Mine for permission to publish this work.

References

Anderson, G. (2012). *Red Dog Disc Configurations*. Brisbane, Australia: Glencore Technology.

Lacouture, B. (2013). *IsaMill Conference Red Dog Overview*. Cape Town, South Africa: IsaMill Users Conference.

Larson, M. (2012). *Red Dog Site Report*. Red Dog Mine, AK: Glencore Technology.

Mehrfert, P. (2012). *Red Dog IsaMill Grinding Test Report*. Kamloops, BC: ALS Laboratories.

Merritt, A. (2012). *Rougher IsaMill Power Curve Observations*. Red Dog Mine, AK: Teck Red Dog.

Comminution 2018

Improving IsaMill™ Energy Efficiency Through Shaft Spacer Design

G.S. Anderson
IsaMill™ Technology Manager, Glencore Technology
Level 10, 160 Ann Street, Brisbane, Queensland 4000
Greg.Anderson@glencore.com.au

P.A. Bandarian
Contract Metallurgist, Glencore Technology
Level 10, 160 Ann Street, Brisbane, Queensland 4000
Paul.Bandarian@glencore.com.au

Abstract

The IsaMill™ was specifically developed to address the energy efficiency issues associated with grinding to fine and ultrafine sizes. It has been doing this successfully for more than 20 years in over 120 installations.

Recent efforts by Glencore Technology to further improve the energy efficiency of the IsaMill™ have resulted in the development of a new IsaMill™ shaft spacer design. The spacer was developed through laboratory, pilot and full scale testing prior to being commercialised. Full-scale results indicate that the new configuration was able to reduce the specific energy consumption by 13-17% in an M1000 gold application grinding to a P_{80} of 18 microns. The spacers were also shown to have no adverse impact on the shape of the product particle size distribution.

This paper discusses the development process of the new spacers through lab, pilot and full scale trials, operating results and processing implications.

Key Words

IsaMill™, fine grinding, energy efficiency, inert media

Introduction

It is well known that comminution energy consumption is a major concentrator operating cost. As a result there has always been a strong focus at mine sites to reduce energy consumption where possible to minimise operating costs – this has become increasingly important in the current economic environment. Increasingly, there are also political and social pressures to reduce overall energy consumption and associated carbon footprint. However, as orebodies become more complex, finer liberation sizes are required in order to produce saleable concentrates, driving up the specific energy required to produce each tonne of saleable concentrate.

The IsaMill™ was developed in the 1990's by Mount Isa Mines (MIM, now Glencore) and Netzsch Feinmahltechnik, to specifically address the requirement for fine grinding down to a P_{80} of 7µm for the further development of MIM's McArthur River deposit and Mount Isa Pb/Zn ore bodies. The IsaMill™ addressed the two major issues of energy efficiency and the resultant downstream metallurgy, primarily through the use of small stirred, inert grinding media. The first 1.1 MW IsaMill™ using inert media was commissioned at Mt Isa in 1994 and four 1.1 MW IsaMills™ became the enabling technology for the McArthur River project in 1995. The early development and implementation of the IsaMill™ is well described by a number of authors including (Enderle, Woodall, Duffy, & Johnson, 1997), (Harbort, Murphy, Vargas, & Young Michael, 1999), (Johnson, Gao, Young, & Cronin, 1998) and (Pease, Curry, Barns, Young, & Rule, 2006).

Since its inception, the IsaMill™ has been successfully used to grind metalliferous concentrates to P_{80} product sizes from 7µm to 60µm for over 20 years in over 120 installations and proven to be significantly more energy and metallurgically efficient than conventional ball mills and tower mills (Larson, Young, & Morrison, 2008) and (Larson M. , Anderson, Morrison, & Young, 2011).

Ongoing efforts by Glencore Technology to further improve the energy efficiency of the IsaMill™ have resulted in a new IsaMill™ shaft spacer design. The new spacer was developed through laboratory, pilot and full scale testing prior to being commercialised and has shown operational specific energy reductions of 13-17% in a production scale gold concentrate regrind application. This paper discusses the development, plant performance and processing implications of the new spacer design.

Grinding Mechanism and Conventional Spacer

A conventional IsaMill™ configuration uses a cylindrical spacer to separate the grinding discs at the required distance. Figure 1 shows the conventional spacer on the shaft of an M1000 IsaMill™

Figure 1 - Conventional Shaft Spacer between two Grinding Discs for an M1000 IsaMill™

Figure 2 illustrates the grinding mechanism within the IsaMill™. The IsaMill™ typically operates around 70% media filling volume. As the shaft rotates, the grinding discs agitate the media such that it is drawn out along the face of the discs towards the shell liner. As it reaches the shell liner, the media is turned around and directed back towards the mill shaft area. This happens on the face of each disc, where there is sufficient media present and sets up a chamber of agitated grinding media between each of the grinding discs. Slurry to be ground enters opposite the shaft end cap at one end of the IsaMill™ and must pass through each of the agitated grinding chambers in series before it can exit, making it virtually impossible for any material to short circuit the IsaMill™. At the discharge end of the IsaMill™ is the patented product separator which makes use of a closer spacing between the final disc and the rotor to centrifuge any coarse particles and media towards the shell. The rotor, which acts like a pump, then returns this material to the grinding zones. This mechanism allows ground product to flow through and exit the IsaMill™ whilst retaining the grinding media inside, all without the use of fine screens.

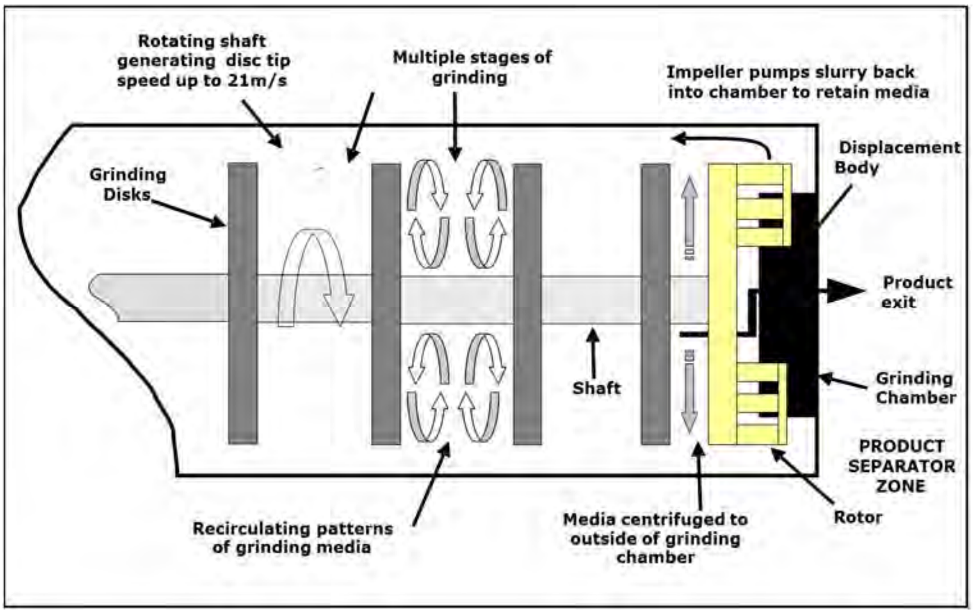


Figure 2 - Simplified IsaMill™ Grinding Mechanism

Laboratory Testwork

In 2010, a large IsaMill™ development testwork program was undertaken at the Netzsch Feinmahltechnik laboratory in Germany. The program utilised an M20 IsaMill™ (20 litre volume), with a clear shell, to investigate and understand the impact on IsaMill™ power draw and axial media distribution of different designs and configurations of the discs, shaft spacers and rotor in conjunction with different operating shaft speeds and media loadings. The use of conical shaft spacers was investigated as part of this work.

The clear shell M20 is the same as a standard M20 IsaMill™ used for pilot scale test work other than being fitted with a clear plastic shell to allow the inner workings of the IsaMill™ to be observed under different operating conditions. It was built specifically for investigative work and demonstration purposes. To allow observation of the inner workings and preserve the integrity of the clear shell, only water and glass beads are used in this mill.

Prior to any data recording, the M20 was operated for 20-30 minutes to allow the bearings to warm up, so as not to impact on the power draw readings. Two shaft configurations were tested to isolate the impact of changing from standard cylindrical to conical spacers. Firstly, standard grinding discs and standard cylindrical spacers were used to generate the baseline data. Following this, the standard cylindrical spacers were replaced with conical spacers – the disc spacing and disc position along the shaft remained the same. In each of the two cases, the M20 was filled with 13.6 litres of 2mm glass beads and operated at a steady 16.3 litres/min of water throughput while the shaft speed was varied in stages from 600 to 1400 rpm. Typical operating speed for the M20 is 1200-1400rpm. The media distribution inside the mill was visually observed and the overall mill power draw recorded for each case. Figure 3 shows the conical spacer arrangement inside the glass shell M20 IsaMill™, with the feed end on the left and the product separator on the right.

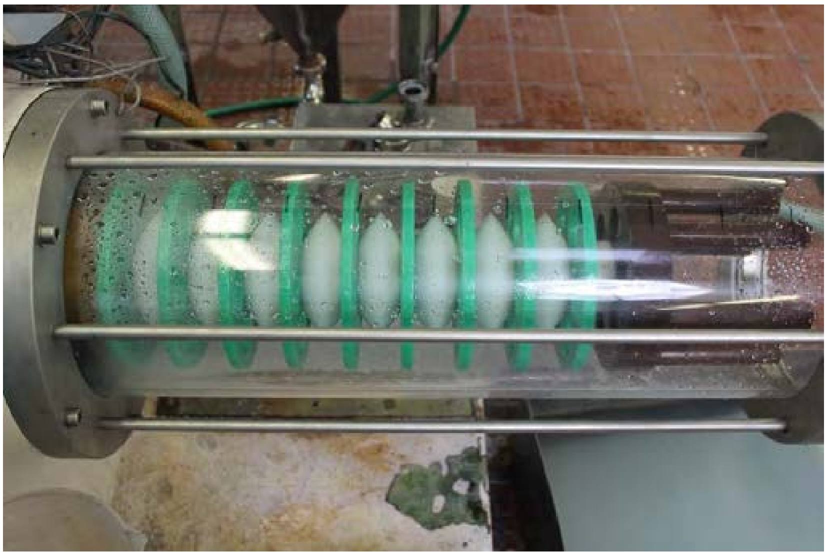


Figure 3 - M20 Clear-Shell IsaMill™, fitted with Conical Spacer Arrangement

As a note, the laboratory M20 IsaMill™ used was configured with a 9-disc arrangement and therefore a narrower shaft spacing compared to the standard 7-disc setup used for pilot scale metalliferous grinding. This could not be altered as all shaft parts were sized to fit the 9 disc arrangement. Other than that, the configuration was the same as that conventionally used with the 7-disc M20 arrangement.

Figure 4 illustrates the power draw as a function of the M20 IsaMill™ shaft speed and the shaft spacer type – which was the only parameter difference between the two curves. For a given shaft speed, there was a clear reduction in overall mill power draw when the conical spacers were used. A reduction of 20%-30% was evident at the typical M20 operating speeds of 1200-1400rpm. This was a direct result of using the conical spacers. As the shaft speed decreased, the difference between the two spacer designs decreased to zero at 800rpm and at 600rpm the conical spacer actually drew more power than the standard cylindrical spacer.

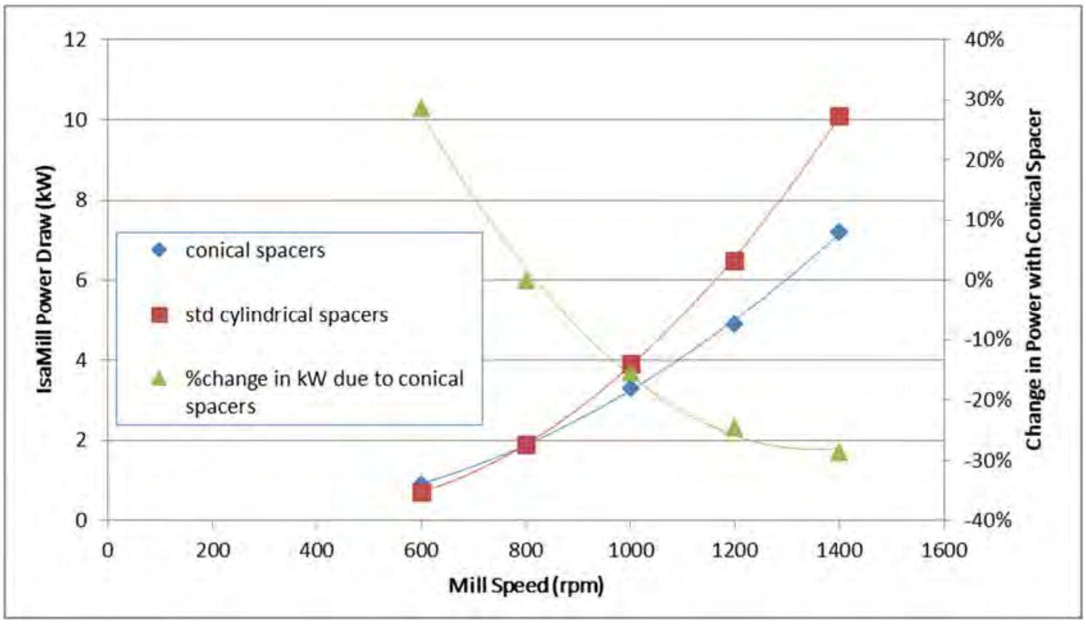


Figure 4 - Relationship between Shaft Speed and Power Draw for Cylindrical and Conical Spacers

The conical spacers allowed the same volume of media to be agitated at the same mill shaft speed, but at a lower drawn power, than when the standard cylindrical spacers were used. Given that the agitated media does the grinding, the impact of the reduced power draw on grinding efficiency – which could not be determined in the current testing program – was questioned. A separate pilot plant testwork program was designed and executed to answer this question.

Pilot Scale Testwork

An opportunity was identified during an onsite pilot campaign in January 2012, utilising an M20 IsaMill™, to conduct a grinding efficiency comparison between the standard cylindrical spacer and conical spacer configurations. This was a standard M20 pilot plant mill configured for metalliferous testing, with 7 discs rather than the 9 discs of the Netzsch laboratory M20. The spacing between each disc was therefore wider, which resulted in an increased included angle at the peak of the conical spacers. The spacers used in this testwork were specifically designed and built for the 7-disc arrangement and all tests conducted at the pilot plant utilised the 7-disc configuration.

Stage 1 of the pilot tests was designed to replicate the water and media testing carried out at Netzsch. Again, the only changed parameter between the two tests was replacement of the standard cylindrical shaft spacers with the conical spacers. The disc spacing and position along the shaft remained the same for both cases. At the selected pilot plant operating speed of 1390rpm and using 13 litres of 2.5mm ceramic grinding media at a water flowrate of 17.6 litres/min, there was a 15% reduction in the drawn power when the new conical spacer design was used. The overall power reduction attributed to the change in spacer design was less than that achieved at Netzsch but was still significant. Differences between the pilot tests and the Netzsch tests could likely be attributed to some of the key parameter differences including the number of discs, resultant changes in the conical spacer geometry, the shell material (rubber vs perspex), the media type (ceramic vs glass) and media volume. Figure 5 shows the cylindrical and conical spacer shaft configurations.

Figure 5 - M20 Pilot Plant IsaMill™ Shaft Configuration with Cylindrical (left) and Conical (right) Spacers

Stage 2 of the pilot tests involved conducting standard IsaMill™ signature plots using each configuration. The feed material was a coal and was managed such that the feed type and size distribution was as similar as possible between the two tests. Figure 6 illustrates the signature plot results for the two sets of spacer configurations.

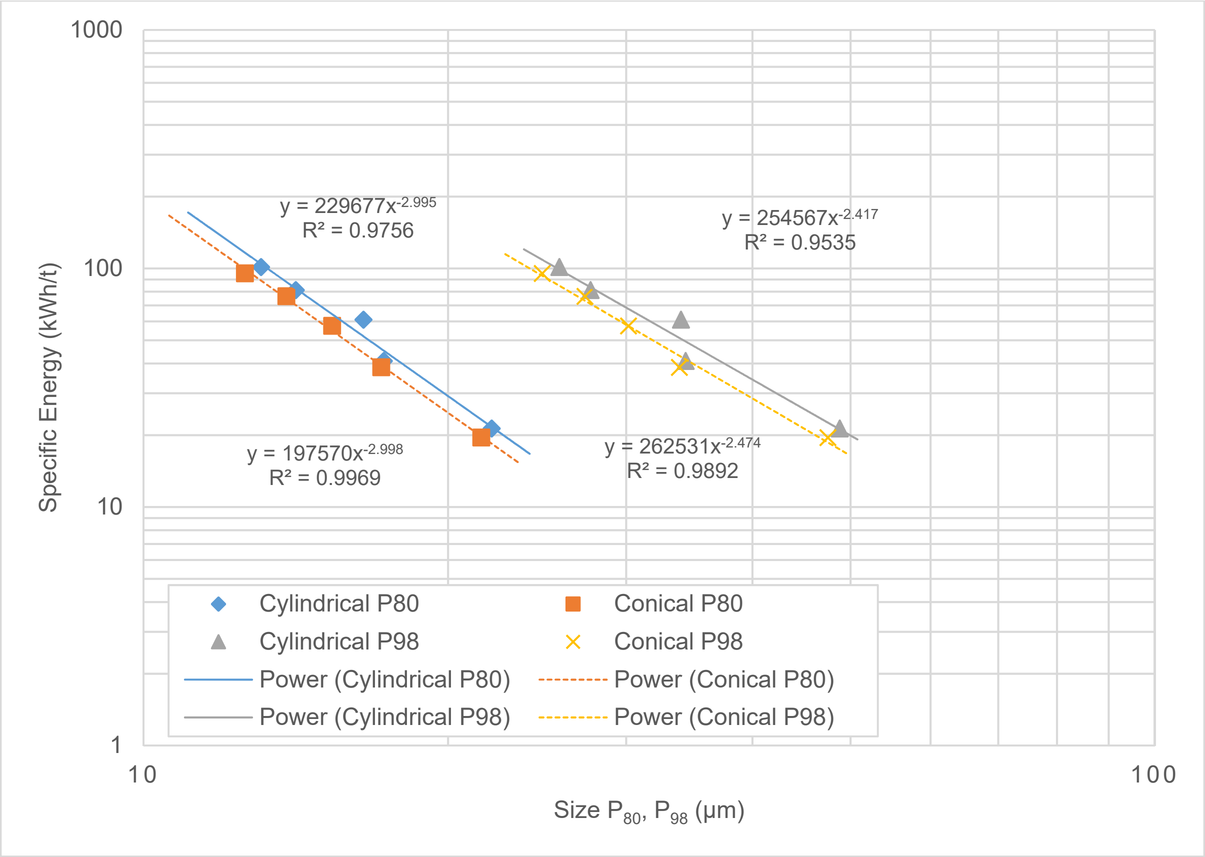


Figure 6 - Signature plot comparison between cylindrical and conical spacer configurations

There was a clear distinction between the two sets of curves. Average net power draw treating the slurry for the conical spacers of 7.5kW was 6% lower than the standard cylindrical spacers at 8.0kW. At the grind target P_{80} of 15 microns, the conical spacer configuration used 15% less specific energy than the standard cylindrical spacer configuration. This was the first evidence to support the contention that the conical spacers not only resulted in a reduced power draw for the same volume of agitated media, but that the resultant grinding action was more energy efficient.

Unfortunately due to the pilot plant requirements and schedule there was no further opportunity to complete any duplicate or further investigative work. However, it was considered that the result obtained in the single comparison, and the potential benefits it offered, warranted progression to full scale testing.

IsaMill™ M1000 Spacer Design

The initial production tests were carried out on an M1000 (500kW) IsaMill™. A series of design parameters for the M1000 spacer and for other full scale IsaMills™ were agreed upon based on the IsaMill™ M20 spacer design. In order to conduct the testwork as efficiently as possible, the test spacers were manufactured from polyurethane rather than the standard rubber lined method.

IsaMill™ M1000 Stage 1 Testwork

An initial, basic testwork program was designed to determine whether the observations from the pilot scale M20 IsaMill™ translated to the production scale M1000 IsaMill™. A site with a large surge tank ahead of the IsaMill™ circuit was identified to minimise variations in feed mineralogy, flow and size distribution between the tests as much as possible. The trial plan consisted of three stages with the aim of having the spacers as the only changed variable. The tests were conducted over a 4-day period in July 2013 and the results summarised in Table 1 below. All particle size analysis was done on a Malvern laser sizer.

Table 1 - M1000 Stage 1 testwork results

Survey	**Test Comment**	**Spacer Type**	**Power (kW)**	**Feed (tph)**	**kWh/t**	**F_{80} (µm)**	**P_{80} (µm)**
1	Std conditions	Cylindrical	430	7.5	54.7	129	21.1
2	Std conditions	Cylindrical	427	7.4	55.0	129	21.4
3	Std conditions	Conical	333	6.7	46.8	127	21.5
4	Lower tph	Conical	337	6.2	51.4	141	20.4
5	Std conditions	Cylindrical	428	7.9	52.1	138	21.3
6	Std conditions	Cylindrical	422	8.0	50.6	139	21.2

The first stage involved surveying the IsaMill™ circuit under the existing standard IsaMill™ configuration using the cylindrical spacers (survey 1, 2). Under standard operating conditions in survey 1, the IsaMill™ drew a gross power of 430kW for a specific energy of 54.7kWh/t, producing a product P_{80} of 21.1µm. In survey 2, the IsaMill™ drew a gross power of 427kW for a specific energy of 55kWh/t producing a product P_{80} of 21.4µm.

For the second stage (survey 3, 4), the IsaMill™ was reconfigured by swapping the standard cylindrical spacers with the conical spacers. As far as possible, the same volume of grinding media that was removed from the IsaMill™ after the completion of the first stage was returned to it. Figure 7 shows the conical spacers installed in the M1000 IsaMill™.

Figure 7 - M1000 IsaMill™ configured with conical spacers

After the initial startup of each stage and stabilisation, top up grinding media was added in the usual way to maintain the initial power draw for each stage. This ensured, as much as possible, that the grinding media volume in the mill remained constant throughout the testwork period.

In survey 3, the IsaMill™ drew a gross power draw of 333kW - a reduction of 22% compared to the standard cylindrical spacer configuration. The IsaMill™ operated at 46.8 kWh/t – a 15% reduction compared to the cylindrical spacer configuration – and produced a product P_{80} of 21.5µm, similar to that produced from the cylindrical spacer configuration survey 2 and slightly higher than survey 1. This first comparative result suggested that the introduction of the conical spacer design had the potential for a 15% reduction in the required specific energy to produce the same product size, similar to what was observed in the pilot scale work.

The aim of survey 4 was to increase the specific energy, using the conical spacers, towards that of survey 1 and 2. By reducing the tonnage through the mill, the specific energy was increased to

51.4kWh/t. The resultant product size decreased to a P_{80} of 20.4μm. Note that survey 4 onwards was subject to a slightly coarser feed size due to an upstream ore change, which coarsened the feed into the surge tank ahead of the IsaMill™. Although the specific energy was increased in survey 4, it was still less than that of survey 1 and 2. Despite this, and the coarser feed size of survey 4, a finer product size was produced from survey 4. This again suggested a grinding efficiency benefit from the conical spacers.

At the completion of survey 4, the IsaMill™ was returned to the original configuration using standard cylindrical spacers for the third stage of testwork. Again, as far as possible, the same volume of grinding media that was removed from the IsaMill™ at the completion of the second stage was returned to it. Two more surveys (surveys 5, 6) were completed to compare against the original standard configuration surveys and the conical spacer surveys.

In survey 5, the IsaMill™ gross power draw returned to 429kW, essentially the same power that was drawn under the same spacer configuration in the first two surveys. This confirmed that essentially the same volume of media was present in the IsaMill™ as had been present under the original conditions in stage 1. The IsaMill™ operated at 52.1kWh/t for a product size P_{80} of 21.3μm.

The specific energy in survey 5 with the cylindrical spacers had increased slightly from 51.4 to 52.1 kWh/t (in comparison to Survey 4, which had similar feed size distribution but with the conical spacers installed) but the product sizing had also increased from P_{80} of 20.4 to 21.3μm.

Survey 6 confirmed a similar result, consuming 50.6kWh/t to a P_{80} of 21.3μm. When compared to surveys 3 and 4, surveys 5 and 6 both suggested a grinding energy efficiency benefit for the conical spacers over the standard cylindrical spacers.

Overall, the results of the initial on site testwork were very encouraging and supported the findings from the pilot plant work. The M1000 data suggested that the same product size could be produced about 15% more efficiently by using the conical spacer configuration. Given the exponential increase in specific energy requirements when grinding to finer sizes, this was a significant finding. Based on this success, a further, more detailed technical program was proposed to better quantify and confirm the advantages of the conical spacers.

IsaMill™ M1000 Stage 2 Testwork – Standard Mill Operating Speed / Varied Throughput

The second stage of testwork was conducted at the same site in September 2013 and again involved trials of the two different spacer configurations. A standard IsaMill™ signature plot, shown in Figure 8, was generated for each configuration by varying the throughput, within the operational limitations of the site. As per the initial stage of M1000 site testwork, the IsaMill™ was operated with nominally the same media load in each case to isolate the impact of the change in spacer design.

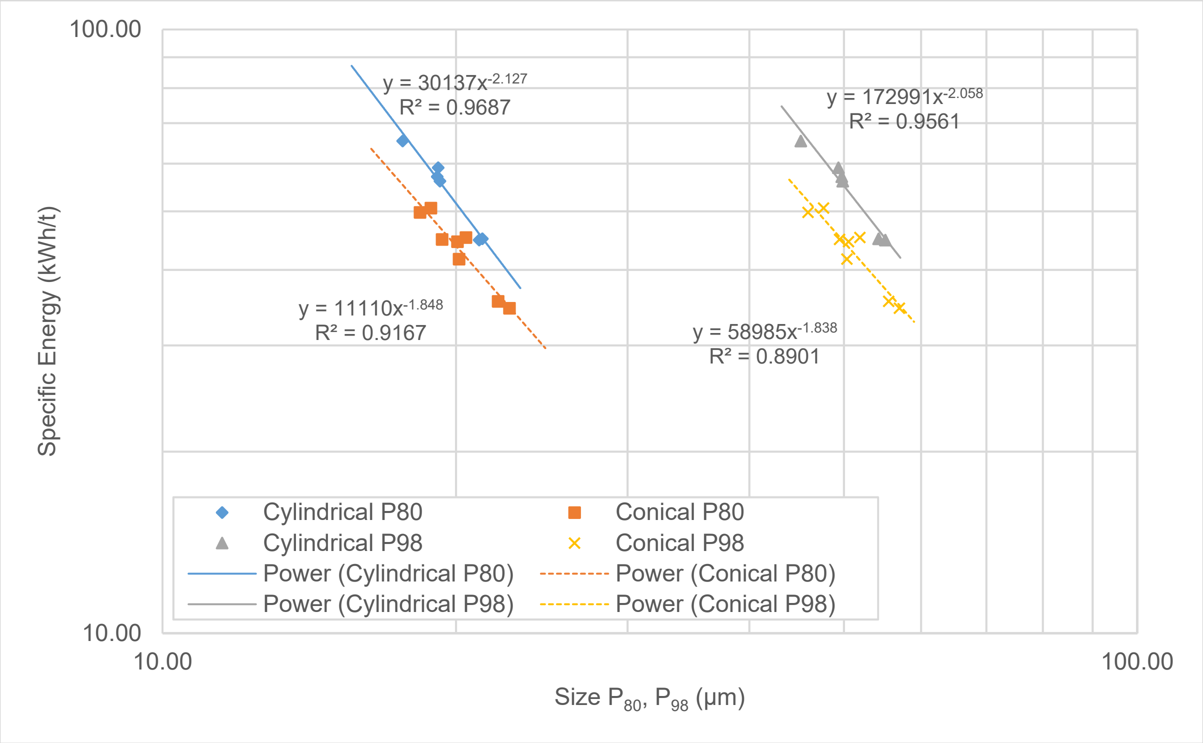


Figure 8 - M1000 IsaMill™ Signature Plot for Conical and Cylindrical Spacers - by varying Throughput

The signature plot equations from Figure 8 were used to construct Figure 9, which illustrates the reduction in specific energy requirement to a given target product P_{80} and P_{98} sizing. Table 2 summarises the data (within the range of product sizes produced during the testwork, i.e. 18μm < P_{80} < 22μm). The data indicated that for a P_{80} target of 22μm, the conical spacers require 36.8kWh/t. This was 12.7% more efficient than the cylindrical spacers which require 42.1kWh/t. The P_{98} predicted for the conical spacer configuration was 55.5μm at 36.8kWh/t, 17.4% more efficient than the cylindrical spacers, which would require 44.5kWh/t to produce the same P_{98}. Similarly, at a P_{80} target of 18μm, the conical spacers were predicted to consume 53.3kWh/t. This was 17.4% more efficient than the cylindrical spacers which would consume 64.5kWh/t. The P_{98} predicted for the conical spacer configuration was 45.3μm at 53.3kWh/t, 21.1% more efficient than the cylindrical spacers, which would require 67.4kWh/t to produce the same P_{98}.

An additional comparison to a target P_{80} of 15μm was made by using the signature plot equations (signature plots have been extensively proven to exhibit a linear relationship on a log-log plot (Larson M. , Anderson, Barns, & Villadolid, 2012)). Analysis of Figure 9 shows that the predicted efficiency improvement of the conical spacers increased further (21.5% in P_{80} terms and 24.1% in terms of the energy equivalent P_{98} of 37.8μm) as the desired grind size was reduced.

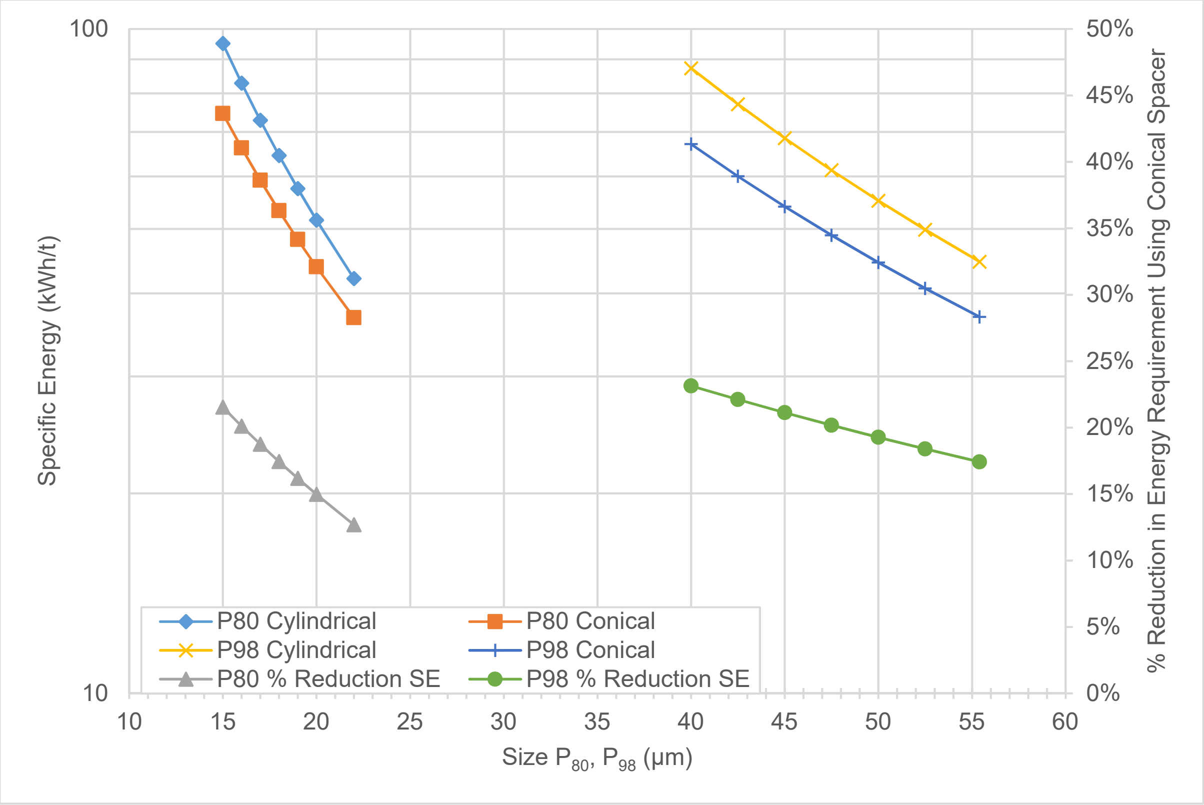


Figure 9 - Reduction in specific energy to target product P_{80} and P_{98} sizing (varied throughput)

Table 2 - Energy Efficiency Improvement Presented by Conical Spacers during Varied Throughput Trial

	P_{80} = 22µm	P_{80} = 18µm	P_{80} = 15µm	P_{98} = 55.5µm	P_{98} = 45.3µm	P_{98} = 37.8µm
Cylindrical (kWh/t)	42.1	64.5	95.1	44.5	67.4	98.3
Conical (kWh/t)	36.8	53.3	74.6	36.8	53.3	74.6
Efficiency Improvement	12.7%	17.4%	21.5%	17.4%	21.1%	24.1%

This data suggested that the impact of the conical spacers on grinding efficiency became more significant at finer grind targets and diminished at coarser targets. Although it was not proven whether the impact at coarser targets was neutral, or even negative.

It is well known that the shape of product size distributions from different grinding devices or circuits can vary such that the P_{80} values may be the same for two very differently shaped curves (Gao, Reemeyer, Obeng, & Holmes, 2007) and (Larson M. , Anderson, Morrison, & Young, 2011). Further, the downstream metallurgical performance can be influenced by the shape of the coarse end of the size distribution curve. In this case, the P_{98} of each product size distribution was also measured and compared to the corresponding P_{80} value as a method of quantifying changes in the shape of the coarse end of the size distribution – both between the two spacer designs and also as the specific energy input changed. The smaller the P_{98}/P_{80} ratio, the tighter the distribution and therefore a lower proportion of particles at coarser sizes for a given P_{80}.

Figure 10 illustrates the P_{98}/P_{80} ratios generated from the Figure 8 signature plot equations. It highlights the reduction in P_{80} and P_{98} values for a given specific energy as a result of using the conical spacers compared to the cylindrical spacers. For example, at a specific energy of 40kWh/t, the P_{80} was reduced by around 7% and the P_{98} by 9%. This corresponded to a reduction in the P_{98}/P_{80} ratio from 2.59 to 2.52 – indicating a slightly tighter size distribution produced by the conical spacers, at the same specific energy input. Figure 10 indicates that larger percentage reductions in P_{80} and P_{98} product sizings occurred at higher specific energies but the improvement in the P_{98}/P_{80} ratio between the spacer designs also diminished but was still in favour of the conical design.

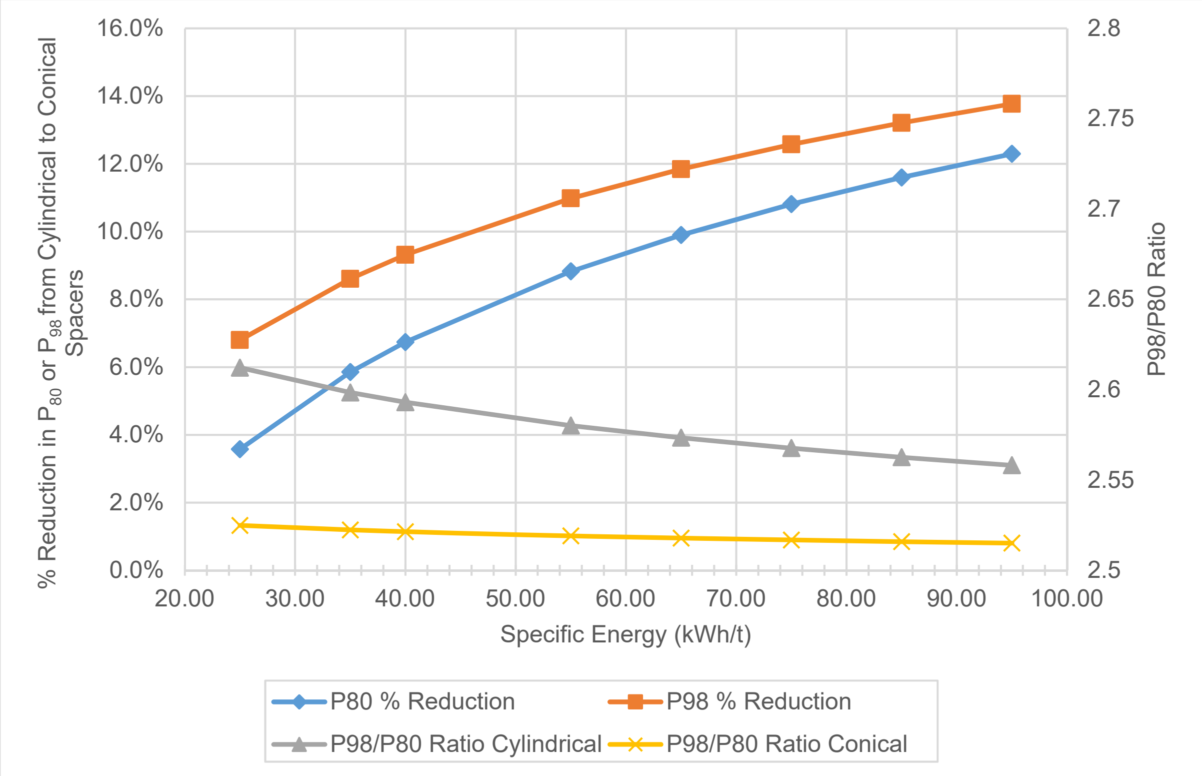


Figure 10 - Effect of Conical Spacers on P_{80}, P_{98} and P_{98}/P_{80} Ratio

In reality, any efficiency gains will likely be realised as reduced specific energy consumption to the target product size, rather than reduced product sizing at the same specific energy target. Table 3 summarises the impact on the P_{98}/P_{80} ratio at target sizes of 22, 18 and 15µm for the M1000 trials. Clearly, there was only a nil to small improvement to the P_{98}/P_{80} ratio once the same target P_{80} sizing was considered for both spacer designs; however, the important fact was to confirm that it had not increased. The P_{80}/P_{50} and P_{80}/P_{20} ratios are also included for completeness. The ratios were largely consistent between the two spacer types, indicating that the conical spacers did not adversely impact the shape of the size distribution curves (at equivalent P_{80} target sizing).

Table 3 - P_{98}/P_{80} ratio for cylindrical ■ and conical ⬢ spacers at same target P_{80} product sizing

	Specific Energy (kWh/t)			**P_{98}/P_{80}**		**P_{80}/P_{50}**		**P_{80}/P_{20}**	
Target P_{80}	■	⬢	**Saving**	■	⬢	■	⬢	■	⬢
15µm	95.1	74.6	21.5%	2.56	2.52	2.17	2.17	4.77	4.78
18µm	64.5	53.3	17.4%	2.57	2.52	2.22	2.20	5.16	5.13
22µm	42.1	36.8	12.7%	2.60	2.52	2.28	2.24	5.62	5.54

IsaMill™ M1000 Stage 2 Testwork – Varied Mill Speed

An additional set of tests to construct signature plots for both spacer designs was conducted where the mill speed was varied, rather than the mill throughput, to adjust the drawn mill power. Figure 11 illustrates the signature plots and indicates similar relationships to those developed when the mill throughput was varied.

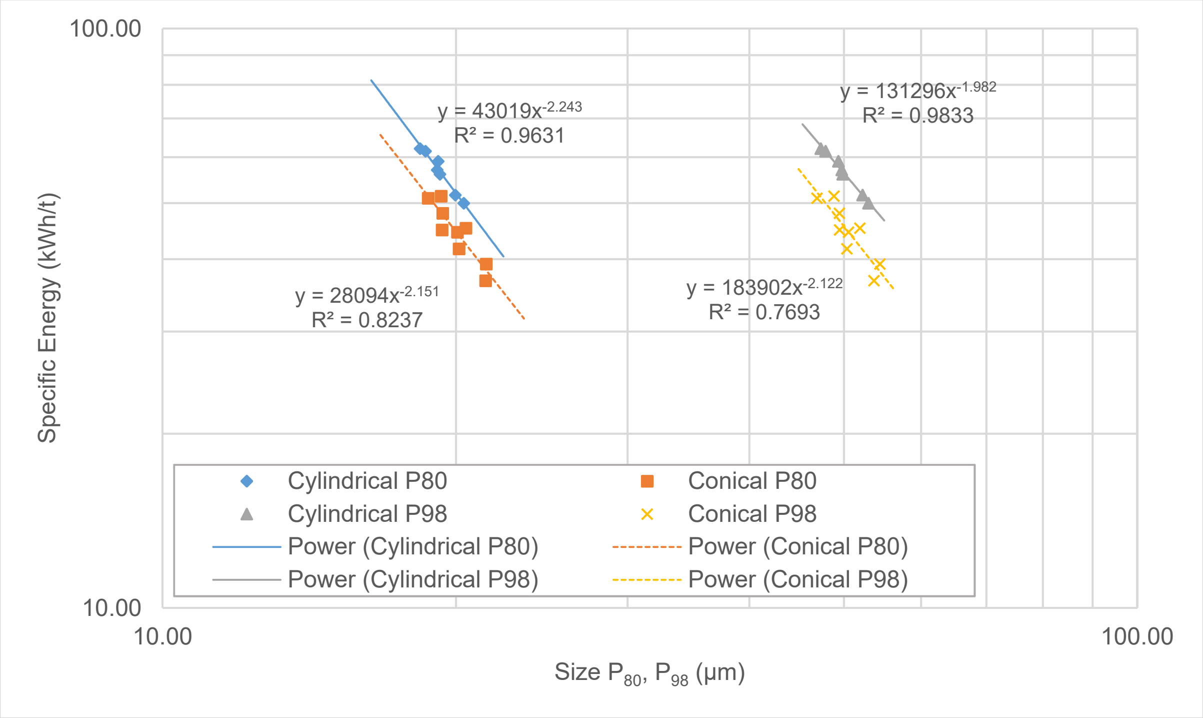


Figure 11 - M1000 IsaMill™ Signature Plot for Conical and Cylindrical Spacers – by varying Mill Speed

Figure 12 illustrates the reduction in specific energy requirement to a given target product P_{80} and P_{98}, based on the signature plot data equations from Figure 11, and shows similar to trends to those observed for Figure 9 where the IsaMill™ throughput was varied. Table 4 summarises some of the data. To achieve a P_{80} target of 22um, the conical spacers required 36.4kWh/t. This was 13.4% more efficient than the cylindrical spacers, which would consume 42kWh/t. The P_{98} predicted for the conical spacer configuration was 55.6µm at 36.4kWh/t, 20.3% more efficient than the cylindrical spacers, which would require 45.6kWh/t to produce the same P_{98}. Similarly, at a P_{80} target of 18um, the conical spacers would require 56kWh/t. This was 15% more efficient than the cylindrical spacers, which would require 65.9kWh/t. The P_{98} predicted for the conical spacer configuration was 45.4µm at 56kWh/t, 18% more efficient than the cylindrical spacers, which would require 68.2kWh/t to produce the same P_{98}.

An additional comparison to a target P_{80} of 15µm was made by using the signature plot equations above. Analysis of Figure 12 shows that the conical spacers required 82.9kWh/t for a P_{80} of 15µm, 16.4% more efficient than the cylindrical spacers, which required 99.1kWh/t.

These results are consistent with the magnitude of reductions observed by varying the mill throughput.

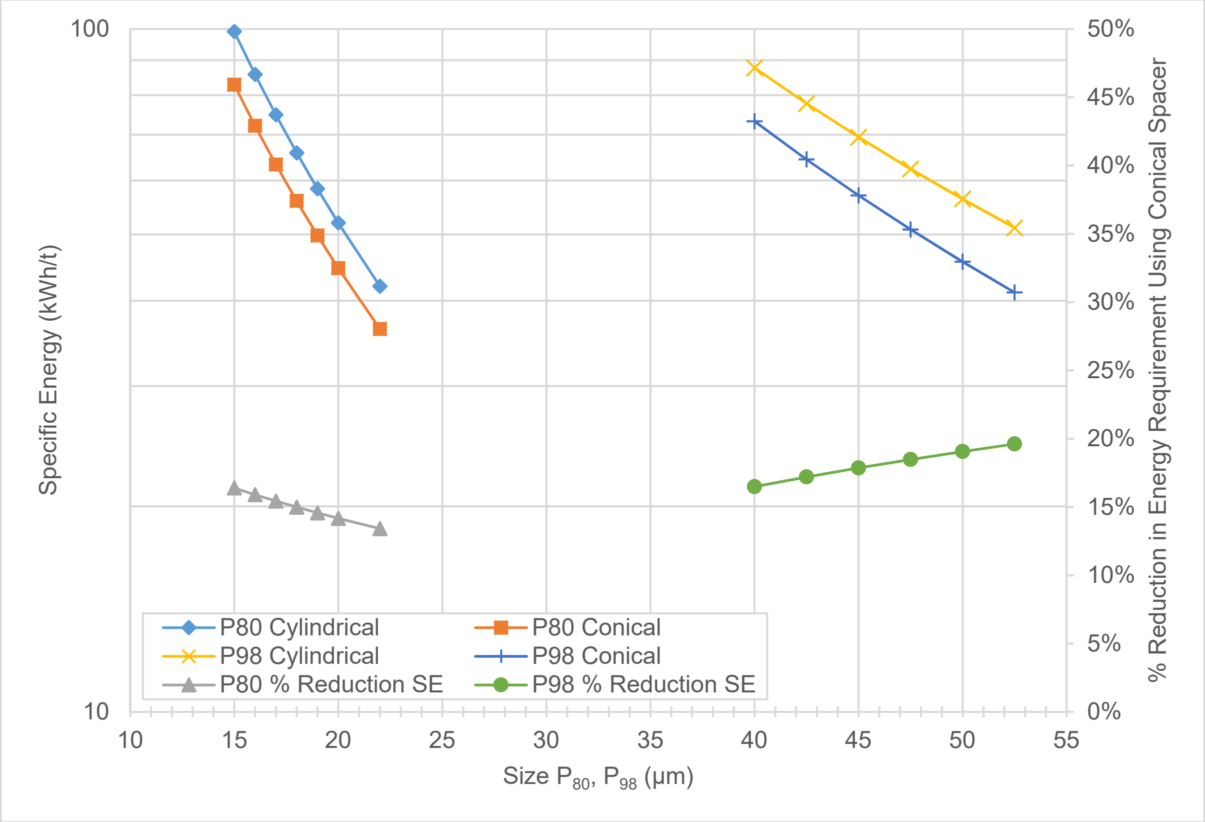


Figure 12 - Reduction in specific energy to target product P_{80} and P_{98} sizing (varied mill speed)

Table 4 - Energy Efficiency Improvement Presented by Conical Spacers during Varied Mill Speed Trial

	P_{80} = 22.0µm	**P_{80} = 18.0µm**	**P_{80} = 15.0µm**	**P_{98} = 55.6µm**	**P_{98} = 45.4µm**	**P_{98} = 37.7µm**
Cylindrical (kWh/t)	42.0	65.9	99.1	45.6	68.2	98.4
Conical (kWh/t)	36.4	56.0	82.9	36.4	56.0	82.9
Efficiency Improvement	13.4%	15.0%	16.4%	20.3%	18.0%	15.8%

Continuous Operation

Based on the results achieved during the testwork campaigns, the test site opted to install the spacers for continuous operation. A set of rubber-lined spacers was designed and manufactured for this purpose. The site was able to operate the IsaMill™ at reduced power draw and produce their required grind size, in line with the testwork results. The spacers were expected to become more of a wear item in the mill due to their contribution to the grinding process. Figure 13 shows the spacers after 375 hours operation.

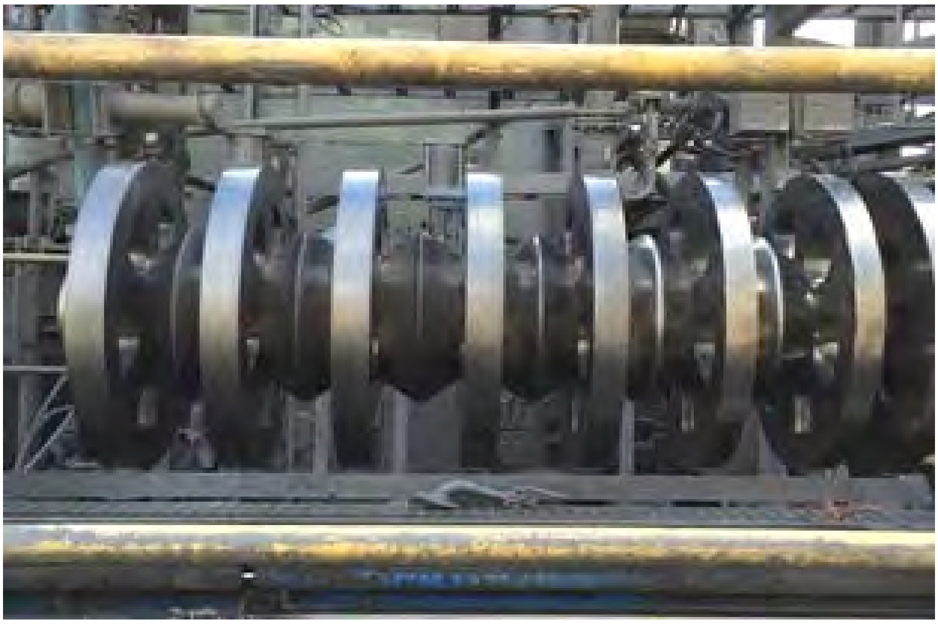


Figure 13 - M1000 IsaMill™ with Conical Spacers after 375 hours of Operation

Process Implications and Considerations

There are a number of potential benefits and considerations for IsaMills™ operating with the new conical spacers due to the improved grinding efficiency.

- Reduced specific energy consumption for the same product size target or a finer product at the same specific energy consumption.
- Small improvement in the top end of the size distribution (P_{98}/P_{80} ratio) and associated downstream benefits, particularly if the same specific energy input is maintained.
- Reduced operating temperature for the same product size target or improved product size at the current operating temperature.
- If the mill power draw can be increased (by further media addition), potential for increased mill throughput at the same product size target.
- Operation at lower power draw will result in reduced component wear rates in the IsaMill™, although the conical spacers are expected to become more of a wear item.

The conical spacers will occupy volumetric capacity within the mill and based on the work here will result in a lower drawn power. Depending on the typical media loading in the mill, it may not be possible to draw full power if no more media can physically fit into the mill.

If the percentage grinding efficiency improvement is the same as the percentage amount that the power draw decreases then the mill can continue operating at the reduced power without any effect on throughput. If the grinding efficiency improvement is greater than the percentage amount that the power draw decreases then the mill can operate with a reduced media load and the same throughput. If, however, the grinding efficiency improvement is less than the percentage amount that the power decreases, then an increased media volume will be required to allow the IsaMill™ to process material at the same throughput . This may be an issue if the spacers have taken up the remainder of the available operating media volume.

In the M1000 work discussed in this paper, the power decreased by approximately 22% for the given operating media volume. At an 18µm P_{80} target, the grinding efficiency improved by ~17%, meaning that if the IsaMill™ was at media capacity due to installing the conical spacers, the throughput would need to be reduced to maintain the same grind, as the media load could not be further increased.

Optimal Sites

Based on this work it appears that the sites which would benefit the most from installation of the new spacers are those with:

- Relatively fine grind targets
- Throughput constraints and a need to process more through the mill
- High specific energy applications running at close to the temperature limitations

Summary

The new conical spacer design was successfully tested from laboratory to pilot and then onto a full scale IsaMill™. At full scale, a reduction of approximately 13-17% in specific energy was observed over the range of throughput conditions tested (at standard operating mill speed) with an accompanying decrease of about 22% in drawn mill power. Based on the signature plots produced, the benefit was shown to increase to 21% at P_{80} of 15µm target. In fine grinding, where energy requirements increase exponentially as the target size decreases, this is a significant finding. The new design offers a number of potential benefits and options to existing and future IsaMill™ installations, the most significant of which is a substantial improvement in grinding efficiency.

Acknowledgements

The authors acknowledge the M1000 operating site for allowing the testwork to proceed, the assistance of Voltaire Villadolid and Katie Barns in conducting and organising some of the associated testwork and Glencore Technology for permission to publish this work.

References

Enderle, U., Woodall, P., Duffy, M., & Johnson, N. W. (1997). Stirred Mill Technology for McArthur River and Mount Isa Lead/Zinc Ores. *Proceedings of the XX IMPC Vol.2*, (pp. 71-78). Aachen.

Gao, M., Reemeyer, L., Obeng, D., & Holmes, R. (2007). Efficiency of the Detritor Mills at Zinifex Century Mine. *Proceedings Ninth Mill Operators' Conference.* Fremantle, WA: AusIMM, Fremantle.

Harbort, G., Murphy, A., Vargas, A., & Young Michael. (1999). The Introduction of the IsaMill™ for Ultrafine Grinding in the Mount Isa Lead/Zinc Concentrator. *Proceedings Extemin99.* Arequipa, Peru.

Johnson, N. W., Gao, M., Young, M. F., & Cronin, B. (1998). Application of the IsaMill™ (a Horizontal Stirred Mill) to the Lead/Zinc Concentrator (Mount Isa Mines ltd.) and the Mining Cycle. *Proceedings AusIMM Annual Conference.* Mount Isa: AusIMM.

Larson, M., Anderson, G., Barns, K., & Villadolid, V. (2012). IsaMill™ - 1:1 Direct Scaleup from Ultrafine to Coarse Grinding. *Proceedings Comminution 2012.* Capetown, South Africa.

Larson, M., Anderson, G., Morrison, R., & Young, M. F. (2011). Regrind Mills: Scaleup or Screwup. *Proceedings SME Annual Meeting 2011.* Denver: SME.

Larson, M., Young, M., & Morrison, R. (2008). Improving Grinding Efficiency with the IsaMill™. *Proceedings Comminution 2008.* Falmouth: ME.

Pease, J., Curry, D., Barns, K., Young, M. F., & Rule, C. (2006). Transforming Flowsheet Design with Inert Grinding - the IsaMill™. *Proceedings of the 38th Annual Canadian Mineral Processors' Conference* (pp. 231-249). Montreal: CIM.

MOUNT ISA MINES NECESSITY DRIVING INNOVATION

*V. Lawson, H. DeWaal, G. Heferen, N. Aslin, P. Voigt, and M. Hourn

Glencore Technology
Level 10, 160 Ann St
Brisbane, Australia, 4000
*(*Corresponding author: virginia.lawson@glencore.com.au)*

ABSTRACT

Mount Isa Mines (MIM) acquired a reputation for the successful application of R&D to develop break-through technologies for the mining industry starting in the 1978's through until the early 2000's. The ISAPROCESS™ tank-house technology has been licensed to copper refineries throughout the world, and a significant per cent of the world's copper is refined using this technology. Since development in the late 1980's more than 20 ISASMELT™ copper and lead smelting furnaces are now installed in countries around the world. Jameson Cell flotation technology developed jointly by Mount Isa Mines and Professor Graeme Jameson is widely used in the Australian coal mining industry and increasingly in the base-metal and gold industry. The IsaMill™'s developed at Mount Isa and McArthur River made it possible to develop the McArthur River and George Fisher orebodies and has been successfully implemented into base metal fine grinding applications around the world. The most recent commercialised innovation is the atmospheric leach Albion Process™ with its supersonic HyperSparge™ gas sparger, is being adopted as a solution to the increasing complexity of orebodies.

MIM's contribution to the industry was significant given the size and the remote location of its operations with Townsville Copper Refineries more than 1350 km and Mount Isa 1800 km from the nearest state capital of Brisbane. This paper will briefly discuss the development of each of these technologies and why MIM – now owned by Glencore - was so successful innovating and developing such technologies over a period of nearly 40 years.

KEYWORDS

Innovation, Mount Isa Mines, ISAPROCESS™, IsaKidd™, ISASMELT™, IsaMill™, Jameson cell, Albion Process™, HyperSparge™, ZipaTank™

INTRODUCTION

Mount Isa is located in the Gulf Country region of Queensland about 1800 kilometers North West of Brisbane (see Figure 1). It came into existence because of the world class mineral deposits found in the area. In 1923 the orebody containing lead, zinc and silver was discovered by the miner John Campbell Miles. Mount Isa Mines Limited (MIM) was founded in 1924 to develop the minerals discovered by Miles, but production did not begin until May 1931. It paid its first dividend in 1947 after 16 years of troubled production. In 1954 the 1100 copper orebody was discovered and with rapidly rising reserves during the 1950's and 1960's led to the construction of new concentrators to treat lead/zinc/silver ores in 1966 (#2 concentrator) and copper ore's in 1973 (#4 concentrator). The difficult nature of the Mount Isa lead-zinc orebodies has meant that the company had always needed to be at the forefront of mining technology. In the 1970's through to the 1990's, it became a world leader in developing new mining techniques and processing technologies as a response to declining metal prices and rising costs. Mount Isa has been smelting copper since 1953 and lead since the early 1930's. Copper Refining at Mount Isa's fully owned subsidiary of Copper Refineries Proprietary Limited (CRL) had commenced operations in 1959.

Figure 1 – Location of Mount Isa and Townsville relative to Brisbane the nearest Capital City

Technologies to come out of Mount Isa include the ISAPROCESS^{TM} copper refining technology, the ISASMELT^{TM}, The Jameson Cell, the IsaMill^{TM}, the Albion Process^{TM} and the Hypersparge^{TM}. Mount Isa Mines Ltd was acquired by Xstrata in 2003 and Xstrata was then merged with Glencore in 2015. The level of innovation achieved at Mount Isa Mines is unsurpassed and was the result of the difficult nature of the Mount Isa ore bodies and its response to declining metal prices and rising operational costs in the 1970's and 1980's. By the 1990's, Mount Isa had become a world leader in innovative mining techniques and state of the art processing technologies. The processing technologies are discussed below.

INNOVATIONS

Each of the innovations developed at Mount Isa Mines had a driver but the overarching desire was to make technology more efficient and cost effective. Each of these process developments will be discussed separately.

ISAPROCESS™

The development of the ISAPROCESS™ tank house technology had its beginning in the zinc industry. During the mid-1970s, MIM was considering building a zinc refinery in Townsville to treat the zinc concentrate produced by its Mount Isa operations. As a result, MIM staff visited the zinc smelters using the best-practice technology and found that modern electrolytic zinc smelters had adopted permanent cathode plate and mechanised stripping technology. MIM realised that the copper refineries performance was constrained by the conventional practice of copper starter sheets. The preparation of these copper starter sheets was labour intensive and the overall cycle was several weeks in duration.

MIM initiated a research program aimed at developing similar permanent cathode technology for copper refining. CRL, a subsidiary of MIM, had been operating in Townsville since 1959, using conventional starter-sheet technology and treating blister copper produced in the copper smelter at Mount Isa. Permanent cathode technology was developed and adapted over many years of in-plant experimental work and successfully introduced to the Townsville refinery in 1978. . The fundamental difference between the new ISAPROCESS™ and the conventional starter sheet technology is the use of a permanent reusable cathode blank instead of a non-reuseable copper starter sheet and the introduction of mechanised and automated electrode handling machines replacing labour-intensive manual operations. The vertical edges had plastic strips and the bottom cased in wax to prevent copper cathode from growing around the edges of the cathode plate during, stripping and allowing two separate copper sheets from each cathode plate. This technology led to major advances in the electrode handling systems and automation in copper tank houses. The improved geometry of the cathode plates and the significantly shorter cathode cycle times allowed for increased intensity and efficiency of the refining process. Introduction of permanent cathode technology resulted in higher capacity, better copper cathode quality with less defects, safer operation and a four-fold improvement in productivity. Considerable development work was required to modify the original stripping machines from their zinc cathode origins due to the heavier cathodes. The stripping capacity of the machines has increased from 250 plates per hour to 600 plates per hour in the latest designs. More recent developments include the elimination of wax masking from the cathode plate, robotic electrode handling machines, and the introduction Duplex Stainless Steel cathode plates giving greater durability and corrosion resistance. Through the use of ISAPROCESS™ user forums, to exchange ideas and developments in the technology and to share operational experiences, the technology has enjoyed continued improvement with higher productivity and improved quality at low cost.

Figure 2 – The IsaKidd process

In mid 1981 Falconbridge Limited commissioned a copper smelter near Timmins to treat concentrate from its Kidd Mine. The original copper cathode produced at Kidd suffered from the presence of higher concentrations of lead and selenium and could not meet customer specifications. It was determined that the use of copper starter sheets was preventing the Kidd refinery from meeting its cathode quality targets. Testwork began with the use of permanent stainless steel cathodes after preliminary tests showed a significant reduction in deleterious elements. The Kidd Process cathode used a solid copper header bar welded onto stainless steel resulting in a lower voltage drop than the ISAPROCESS^{TM}. Falconbridge began marketing the Kidd Process technology in 1992 providing competition between the two suppliers of permanent cathode technology. Between 1992 and 2006, 25 Kidd technology licenses were sold and 52 ISAPROCESS^{TM} licenses.

The development of the ISAPROCESS^{TM} and Kidd Process set the scene for a run of technology developments that continued until the mid 2000's. Xstrata took over MIM in 2003 and then Falconbridge in 2006. The Kidd Process technology consequently became part of the tank house package and together they have since been marketed as IsaKidd^{TM} representing the dual heritage of the technology. The current robotic stripping machine (Figure 2) is based on over 30 years of copper refining and winning technology. Today over 100 licensees are using IsaKidd^{TM} technology.

ISASMELT^{TM}

The sinter plant/blast furnace combination was the dominant technology for lead smelting throughout the 20th century. In the early 1970's companies using this technology came under sustained political and economic pressure as tighter environmental regulations were introduced, and energy costs increased, leading to higher capital and operating costs (Fewings 1988). It was in this environment that Mount Isa Mines sought a process that would improve the performance of the operations at their lead smelter in Mount Isa. After investigating the various processes under development, researchers turned their attention to the Sirosmelt lance. It had recently been developed on a laboratory scale at the Commonwealth Scientific and Industrial Research Organisation (CSIRO) in Melbourne. Following initial investigations Mount Isa Mines recognised the potential of the novel concept for smelting of lead concentrates and embarked on an extensive development program.

In 1978 a joint project was initiated between Mount Isa Mines and CSIRO to investigate the application of the Sirosmelt submerged-combustion technology to the smelting of Mount Isa lead concentrates. The ISASMELT™ process, as it became known, was developed to maturity for smelting copper, nickel, lead and zinc feeds by Mount Isa Mines through the 1980's and 1990's using incremental scale up. Commercialization only occurred once the process had been proven on laboratory, pilot and demonstration scale over many years. Approximately ten years were required for development of the lead and copper ISASMELT™ from crucible to demonstration scale (refer to Figure 4). During this decade the core know-how that was accumulated enabled the development team to reach the point where they were much better equipped to design and construct a full scale commercial plant – the final stage of the scale up process. Key aspects in this process were the selection of the scale up factors and the systematic design, development and re-engineering of several components of the technology. Figure 3 shows a comparison for the scale up stages for the lead and copper ISASMELT™ processes. Pilot scale was defined as unity for scale up comparison.

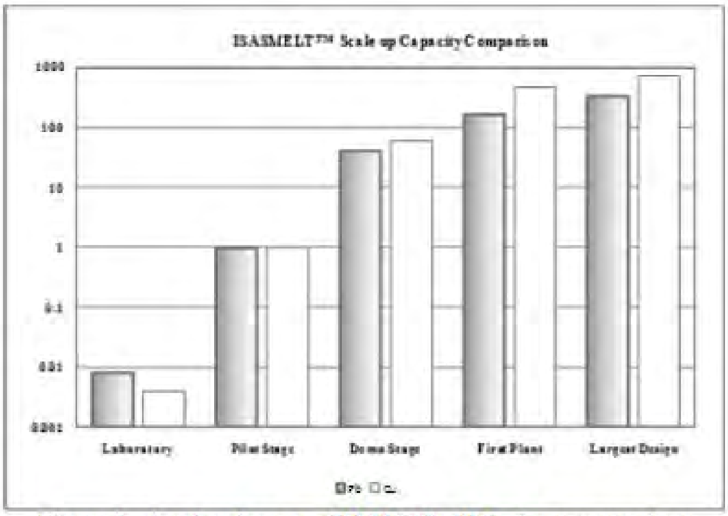


Figure 3 – Lead and copper ISASMELT™ Scale up comparison

During the scale up process, refer to Table 1, several aspects of the technology were developed to a high standard that allowed the ISASMELT™ technology to become a commercial success. As a result, ISASMELT™ technology now operates successfully at numerous plants around the world. The methodical approach to development of the technology has allowed owners to modernise their existing operations or create new businesses with significantly reduced technical risk.

An important parameter in the evolution of the ISASMELT™ technology has been the refractory campaign life. Figure 5 shows the history of the refractory campaigns at the commercial copper ISASMELT™ plant at Mount Isa since commissioning. At the time Mount Isa Mines management considered the installation of water cooling on the furnace refractories undesirable because of the potential for fatal incidents and increased operating costs. As a result the commercial scale furnaces were constructed with minimal water cooling. Although this led to shorter campaign lives initially, a development program was begun that focussed on optimising refractory materials selection and installation methodology. When coupled with process control strategies and continuous on-line monitoring of the bath temperature using systems developed over more than 10 years of operation, it allowed Mount Isa Mines to achieve campaign lives of more than 3 years without using any water cooling of the furnace refractories.

Figure 4 – Tapping matte from the copper ISASMELT at Kazzinc

Table 1 – Key Indicators of ISASMELT™ Plants from pilot to commercial scale

Topic	Unit	Pilot Scale		Demo Scale		First Full Scale		Current Design[1]	
		Pb	Cu	Pb	Cu	Pb[3]	Cu	Pb	Cu
Furnace ID	m	0.4	0.4	1.8	2.3	2.5	3.75	3.6	4.4
Lance Diameter	mm	38	38	150	250	250	350	250	500
Lance Control	-	Manual		Semi Automatic		Semi Automatic		Automatic	
Oxygen Enrichment	%	21	21	21	28	35	45	70	90
Nominal Feed Rate	tph	0.12	0.25	5	15	20	101	40	183
Offgas Treatment	-	Flue System / Baghouse		Gas cooler/ Baghouse		WHB		WHB[2]	

Notes:
ID: Internal Diameter; WHB: Waste Heat Boiler
[1] Refers to maximum throughput
[2] Some of the plants use a combination of radiation section and evaporative cooler for offgas treatment
[3] Refers to the smelting furnace from the two stage lead ISASMELT™ process

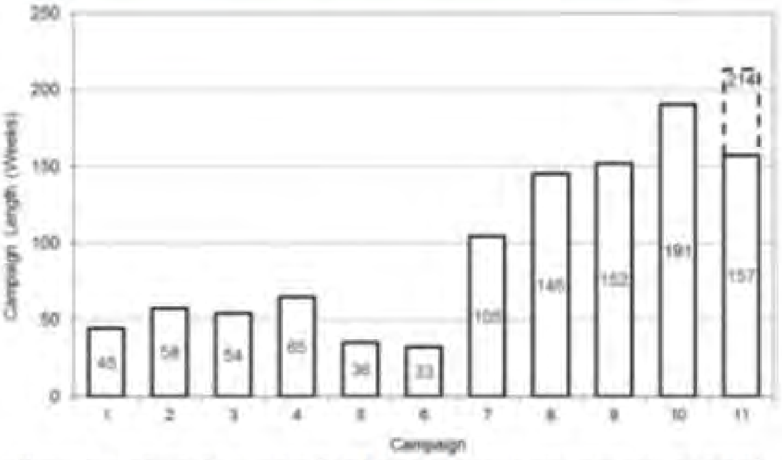


Figure 5 – Mount Isa copper ISASMELT™ plant campaigns (as of 2013)

Jameson Cell

The Jameson Cell (Figure 6) was jointly developed by Mount Isa Mines and Laureate Professor Graeme J Jameson (AO) of the University of Newcastle. Mt Isa had commenced operations with conventional flotation cells but was installing columns in cleaning duties in the mid 1980's. The columns had the benefit of froth washing that was likely to allow significant grade benefits in the very fine lead-zinc circuit. The first observations of the columns was that the collection process was slow necessitating long residence times and large volumes which remains a limitation of columns even today. In 1985 Professor Jameson was commissioned to undertake a project to improve the column sparger design.

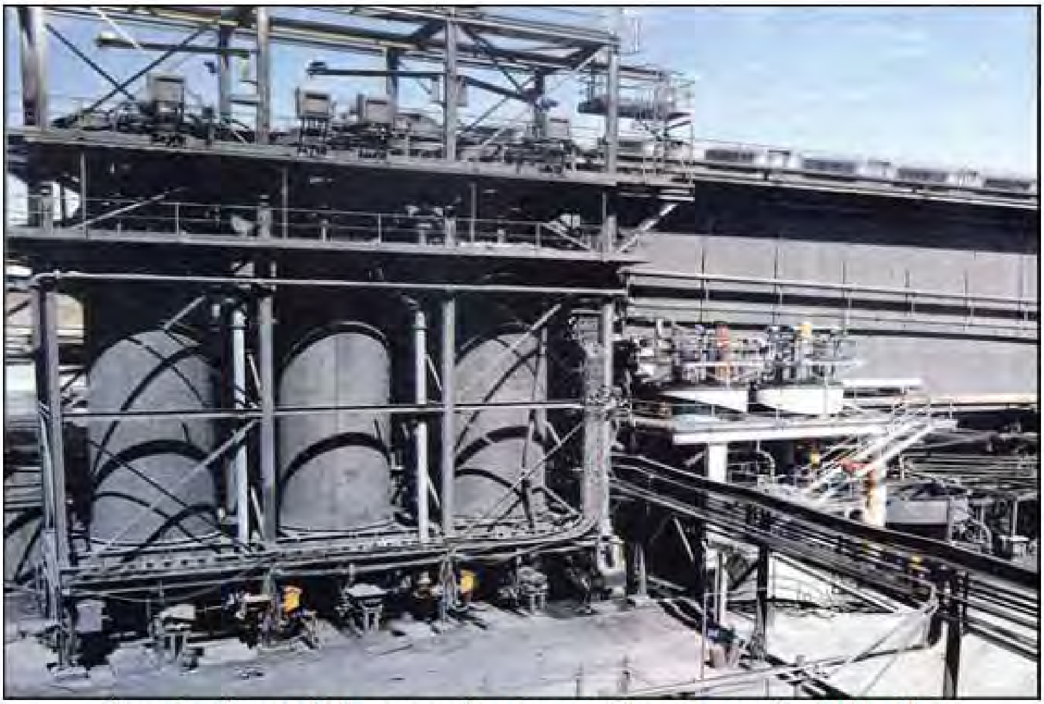


Figure 6 – Jameson Cells compared to columns of the same capacity at Mount Isa

Following initial work to provide an alternate method to bring together bubbles and particles, the downcomer was created. In the downcomer the air and the slurry are co-current with the air being entrained into the plunging jet under vacuum. Investigation showed that all of the bubble particle contact took place in the downcomer and thus the flotation tank could be much smaller. The first application at an industrial scale was in the lead zinc concentrator on the heavy media plant (HMP) lead slimes circuit. The initial improvement in performance were attributed to the very short residence time that allowed the minimisation of oxidation of galena fines. The cells were significantly smaller than the columns and there is no doubt the performance was superior as shown in Figure 3.

The testwork and trials in the early applications showed improved metallurgical performance when operated correctly. The challenge was operating them correctly. The technology hadn't been sufficiently developed to be successfully adopted into plant operations. The cell fell out of favour in base metals and in the 1990's was adopted into the Australia Coal industry and into niche SXEW applications where the main design challenges were resolved. The operability was improved by the introduction of a partial recycle to maintain constant flow and the maintainability of the cell was improved through various design modifications in operating plants. It was a period of continuous improvement. The result was a robust, low maintenance, easy to operate cell with the original features of excellent bubble particle contact.

The final obstacle was overcome when its adaption into the flowsheet was recognised to enable successful installations at the head of cleaner circuits and as low cost brownfield expansions. It is clear that the fast failures have had a significant effect on the success of the cell limiting its adoption into the industry. It is interesting that a significant proportion of sales are to return customers. Once you get over the hurdle of getting a Jameson Cell into your plant then seeing is believing. 2016 was the best year for Jameson cells into base metals and include the first sales back into South America where the cell had been abandoned after the difficulties of operations and maintenance of the Alumbrera installation. The metallurgical performance in Alumbrera was never the issue but the operators and maintainers hated the cells and they failed fast and hard.

The Jameson Cell celebrates its 30th birthday this year and has finally been adopted into mainstream base metals concentrators mainly as cleaner scalper at the head of the cleaner circuit. The cells generally recover up to 80% of the cleaner feed at high grades enabling much lower capital expenditure on the entire circuit. Process performance can be predicted from laboratory and pilot plant testing with demonstrated direct scale-up. It may have taken 30 years but the Jameson Cell is finally a success story. There are many lessons that can be learned from the implementation of innovation into industry from this case study.

IsaMill™

Unlike the developments of some of the other technologies at Mount Isa where efficiency was the main driver, the IsaMill was developed based on necessity. Figure 7 shows photomicrographs with the same scale of 40 micron demonstrating the increased complexity of Mount Isa ore over Broken Hill ore and the very difficult McArthur River ore. Although McArthur River was discovered in 1955 it was not able to be economically processed until the successful development of ultrafine grinding. McArthur River processing began in 1995 – 40 years after discovery when the IsaMill™ made it technically and economically feasible to grind all of the rougher concentrate to 7 micron to facilitate the rejection of non-sulphide gangue. Even at 7 micron galena liberation is not possible and a bulk zinc-lead concentrate is produced.

Figure 7 – Photomicrograph of a) Broken Hill ore b) Mount Isa ore c) McArthur River ore

Investigations into fine-grinding started at Mount Isa started in the 1970s using conventional grinding technology to increase mineral liberation by grinding to fine sizes. These technologies were not only found to have high power consumption but also proved to be detrimental to flotation performance as a result of pulp chemistry and iron contamination from steel media. These poor results were revisited during pilot plant and tower mill testwork in the 1980s which also showed an inability of tower mills to economically achieve the required sizes. When it became clear that the solution to efficient fine-grinding did not exist in the minerals industry, MIM looked for ideas to "crossover" from other industries that also ground fine particles – pigments, pharmaceuticals, foodstuffs (e.g. chocolate). While these mills operated at a much lower scale and treated high value products they demonstrated the principle that stirring fine media at high speed was highly efficient. The challenge was transferring this concept to continuous, high tonnage and lower-value streams in the minerals industry.

In 1991 the introduction of a Netzsch laboratory stirred mill to the Mount Isa site was a turning point in fine-grinding and ultrafine grinding. The ½ litre bench scale mill resembled a milk shake maker and used fine copper smelter slag as grinding media. Testwork on McArthur River ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the Mount Isa pilot plant. The testwork showed that high speed, inert, horizontal mills could efficiently grind to 7 microns at laboratory scale providing major improvements in metallurgical performance. To make ultrafine grinding applicable to full-scale production a program of development was undertaken between Mount Isa Mines Limited and NETZSCH-Feinmahltechnik GmbH.

After 7 years of development and testing of prototypes in the Mount Isa operations, the IsaMill™ evolved. It was large scale, continuous, and most importantly robust because it was developed by operators. The crucial breakthrough was the perfection of the internal product separator – this allowed the mill to use cheap natural media (sand, smelter slag, ore particles) and to operate in open circuit. These are significant advantages for operating cost and circuit simplicity. Scale-up was tested using trial installations at the Hilton and Mount Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill™ (1.1MW) was installed in the Mount Isa concentrator. Improvements to the technology were continually made by the operators, maintainers and engineers working with the technology.

In 1998 the rights for commercialisation of the IsaMill™ where transferred from Mount Isa Mines Limited to MIM Process Technologies (now Glencore Technology) and under an exclusive agreement with Netzsch. In December 1998, the IsaMill™ technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. The IsaMill™ is now a mainstream fine grinding machine with over 130 installations around the world.

The Albion Process™

In the 1990's, MIM were studying options for the development of the large Frieda River/Nena project in PNG through its subsidiary Highlands Pacific. The Nena ores were not amenable to smelting, due to the elevated arsenic content, and several hydrometallurgical options were examined. Out of this work, MIM developed the Albion Process™, named after the suburb in Brisbane where MIM's development laboratory was located. The Albion Process™ is a combination of ultrafine grinding using Glencore Technology's IsaMill™, followed by oxidative leaching at atmospheric pressure in a series of reactors designed to achieve high oxygen mass transfer efficiency. The HyperSparge™ was also developed to deliver oxygen to the reactors efficiently.

Various small scale continuous pilot plant campaigns were conducted in 1994 and 1995. A larger pilot plant (120kg zinc cathode/day) was constructed in 1997 to conduct testwork as part of a feasibility study on the zinc/gold resources of Pueblo Viejo in the Dominican Republic. Extensive piloting was also conducted on lower grade chalcopyrite concentrates for Cyprus Amax in 1998, and for Mount Isa Mines in 2000. Pre-feasibility and feasibility pilot testing was conducted on the zinc/lead bulk concentrates from McArthur River and Mount Isa in Australia between 2001 and 2005. During this time the Albion Process™ was successfully tested on over 70 different ores and concentrates. The process is designed to recover gold and base metals from refractory ores. The key to the process is the ultrafine grinding stage followed by a hot oxidative leach at atmospheric pressure.

In the period from 1994 until 2004, the Albion Process™ (see Figure 8) was seen as strategic to the MIM/Xstrata group, and was not marketed externally. In 2005, a decision was made to offer the technology to external clients under licence, and a marketing agent – Core Resources, was appointed to market the technology globally. Interest in the technology has been very strong in the subsequent period, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Certej Project. The technology moved into commercial production in 2010 with the commissioning of Glencore's Albion Process™ plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Glencore of a second plant in Germany (16,000 tpa zinc metal). The Las Lagunas refractory gold project commissioned in 2012, and the GPM Gold refractory gold project commissioned in 2013.

Figure 8 – The Albion Process oxidative leach plant in Armenia

The major scale up risk with any oxidative leaching technology is oxygen mass transfer. High agitator power demands are common to achieve the shear rates in the vessel required for effective mass transfer at a commercial scale. A different approach was taken in the design of the Albion Leach Reactor to lower the agitator power demand. Glencore developed the HyperSparge supersonic gas injection lance to provide gas injection velocities of the order of 500 $m.s^{-1}$ within the leaching vessel, compared to the 4 – 8 $m.s^{-1}$ achieved with a typical agitator. Supersonic oxygen injection is a far more efficient method of generating shear than conventional agitation, allowing the total power input into the vessel to be significantly reduced, and greatly reducing the scale up risk for the oxidative leach.

The Albion Process™ was enabled by the fine grinding of the IsaMill™ and the process was designed to deliver a lower cost processing option for treating refractory mineral resources. There are now six operating Albion Process™ plants and the process has now an extensive database of potential applications.

CONCLUSIONS

MIM developed a significant number of processing innovations that are technical and economic successes. The ability to innovate at MIM was enabled by very challenging orebodies and the need to process efficiently to remain economically viable. The success has been attributed to the development of these technologies on an operating site with the R&D group solving the technical issues on small scale. Each subsequent scale up was completed in the operating plants where the operators, maintainers, engineers and metallurgists were required to achieve production goals at each step of the scale up to ensure funding for the next step.

The number of innovations, at MIM, was disproportionate to the scale of operations and may have been enabled by the remoteness of the site and the researchers and operators working collaboratively to solve economic and technical problems. The research group were not capital city based but worked on the same site and were required to assist with installation, commissioning and operation of the various stages. This co-operation led to adoption into the plant and a fast feedback loop for improvements. The ultimate success of the innovations has been their widespread adoption into the mainstream industry where feedback from operating sites based on a user group model has enabled continuous improvement of each of the technologies.

ACKNOWLEDGEMENTS

The authors would like to acknowledge Glencore Copper and Glencore Technology for permission to publish and to all the research and production personnel who enabled and improved the

technologies in their plants. The success of these developments continues with the input from end users in the ongoing development.

REFERENCES

Armstrong, W. "The Isa Process and its contribution to electrolytic copper," paper presented at the Rautomead Conference, Scotland, August 1999.

Fewings J.H., Management of Innovation – The IsaSmelt Process. Presented at AMIRA "Innovation in Metal Production" Technical Meeting at Mount Isa, October 3-4, 1988.

Fountain C. Isasmelt and IsaMills – Models of successful R&D – AusIMM Young Leaders conference, 19-21 March 2002, Kalgoorlie, W.A.

Nihill D.N., Stewart C M. and Bowen P. the McArthur River Mine – The First Years of Operation. The AusIMM '98 Mining Cycle. Mount Isa 19-23 April 1998.

Pease, Joe (August 2005), "Complex leaching becomes much simpler" (PDF), Australian Mining, pp. 26–32, archived from the original (PDF) on 12 September 2009, retrieved 6 January 2010

Pease, Joe September 2016, "Crossing the innovation valley of death" Presented to JKMRC.

IsaMills™ at Kalgoorlie Consolidated Gold Mines – from the M3000 to the M10 000 and Replacement of the Roasters at Gidji Processing Plant

ABSTRACT

The IsaMill™ was originally developed to address liberation issues at Mount Isa Mines (now Glencore) operations, McArthur River and Mount Isa in the early 1990s. Following on from the industrial success at those operations, the IsaMill™ was commercialised. In the year 2000, Kalgoorlie Consolidated Gold Mines (KCGM) became the first external user of the technology with a 1.1 MW M3000 IsaMill™ installed to supplement the roaster capacity at its Gidji Processing Plant. The IsaMill™ was used to ultra-fine grind about 10 t/h of pyrite flotation concentrate to 10–12 μm prior to gold cyanide leaching. Based on the success of the Gidji installation, another was installed at the processing plant at KCGM's Fimiston operation in 2002 in a similar duty.

In 2014, a 3 MW M10 000 IsaMill™ was selected to replace the roasters at Gidji as part of the A$98 M emissions reduction project. The mill was commissioned in April 2015 at the design 30 t/h, allowing the immediate shutdown of the two roasters. Replacement of roasting with ultra-fine grinding completely eliminated air emissions at Gidji while maintaining overall plant production rates.

This paper details the history of the IsaMill™ at KCGM from the original M3000 to the latest M10 000. It covers the basis of the process selection, circuit design, operating and maintenance data.

INTRODUCTION

The IsaMill™ was originally developed by Mount Isa Mines (MIM, and now Glencore) [illegible] with Netzsch Feinmahltechnik to address the liberation issues of the McArthur River project and within MIM's lead/zinc [illegible] to [illegible] which had [illegible] sizes of [illegible] 80 per cent passing seven microns. The first 1.1 MW IsaMill™ using the [illegible] was commissioned at Mount Isa in 1994 and [illegible] 2.6 MW IsaMill™ became the enabling technology for the McArthur River project in 1995. The early development and implementation of the IsaMill™ is well described by a number of authors including Enderle *et al* (1997), Johnson *et al* (1998) and Harbort *et al* (1998). Modern industrial operation of the IsaMill™ has been described by numerous authors including Anderson and Burford (2006).

By 1998 there were five IsaMills™ in operation at McArthur River, two in the lead/zinc concentrator [illegible] with a further six mills on order as part of the George Fisher project. The IsaMill™ program had been an outstanding success, allowing both the development of the McArthur River project and significant improvements in the grade-recovery performance of the Mount Isa lead/zinc concentrator, including a ten per cent gain in recovery and two per cent increase in zinc grade after the commissioning of the six mills. This was achieved through improved flotation and concentration [illegible] due to the inert grinding environment (Pease *et al*, 2006).

The IsaMill™ technology was commercialised in [illegible] 1999 and made available to the general mining industry, [illegible]

Kalgoorlie Consolidated Gold Mines (KCGM) is a 50/50 joint venture between Barrick and Newmont. It currently produces 700 000 to 800 000 oz of gold annually. KCGM was formed in 1989 through the amalgamation of many individual leases and underground mines in an area known as the Golden Mile. Kalgoorlie's open pit mine, the Fimiston open pit, now popularly known as 'The Super Pit', was developed in place of the original underground operations. The ore is treated through the new Fimiston processing plant to produce a pyritic gold concentrate. The concentrate tailings were subject to cyanide leach at Fimiston while the concentrate was roasted to calcine and cyanide leached at the Gidji Processing Plant, 20 km north of Kalgoorlie-Boulder.

Due to a change in the mined areas in 1999, sulfur grades in the mined ore increased, resulting in a significant increase in the tonnage of flotation concentrate produced which exceeded Gidji Plant's roaster capacity (Ellis and Gao, 2002). Alternative treatment methods were investigated and evaluated, including KCGM becoming the first external user of the IsaMill™ and

---

1. [illegible] Technology Manager, Glencore Technology, Brisbane Qld 4000. Email: [illegible]@glencore.com.au

2. Senior Metallurgist, Gidji Roasting, Kalgoorlie Consolidated Gold Mines, Kalgoorlie WA 6430. Email: [illegible]@kalgold.com.au

the first gold application through the installation of an ultra-fine grinding and leach circuit at Gidji in 2001.

An additional ultra-fine grinding mill, IsaMill™ was installed at [illegible] in 2002. The IsaMill™ was successfully scaled up in [illegible] MW in 2003 (Curry, Clark and Rule, 2005) and to date 100 IsaMills™ ranging from 75 kW to 3 MW have been installed across the world in lead/zinc, gold, platinum, copper, nickel, magnetite and molybdenum grinding duties.

This paper discusses the history and performance of the IsaMill™ ultra-fine grinding from the first M3000 installation at Gidji roasting plant to the latest M10000 installation, which has allowed the two roasters at Gidji to be decommissioned.

THE GOLD PROCESS AT KALGOORLIE CONSOLIDATED GOLD MINES

The Gidji Roaster's flow sheet, shown in Figure 1, illustrates the process path for the Fimiston pyrite concentrates prior to leaching. The roasters were constructed in 1989 to process the sulfide concentrate from the Fimiston mill and were the first commercial application of a circulating fluid bed roaster in the gold industry. Mineral flotation concentrate at 25–30 per cent sulfur grade was trucked to the Gidji Plant where it was repulped to 60–65 per cent solids by weight and transferred into holding tanks. The slurried concentrate was pumped at a rate of 20–27 t/h into one of two Lurgi circulating fluidised bed roasters operating at around 650°C and oxidised through the following exothermic roasting reactions:

$$4FeS_2 + 11O_2 \rightarrow 2Fe_2O_3 + 8SO_2$$

$$3FeS_2 + 8O_2 \rightarrow Fe_3O_4 + 6SO_2$$

$$FeS_2 + 3O_2 \rightarrow FeSO_4 + SO_2$$

The roasting process oxidised the pyrite concentrate particles to iron oxides (calcine), breaking down the particles and liberating the entrapped gold to be accessed in the subsequent cyanide leaching process. The leached gold was processed through an eight-stage carbon-in-pulp circuit and the loaded carbon trucked back to the Fimiston Plant for elution and carbon regeneration. Gold recoveries generally achieved via the roasting cyanide leaching route were [illegible] 92–95 per cent.

As part of the roasting reaction, the sulfide sulfur was released as sulfur dioxide along with quantities of mercury and arsenic. The off-gas was processed through the electrostatic precipitator and then the gas released via the 100 m high stack.

The most frustrating constraint for the roasters was the strict requirements for air quality control (AQC) to meet sulfur dioxide environmental limits. This necessitated shutting down one of the roasters whenever the prevailing atmospheric conditions together with AQC modelling indicated exceeding atmospheric sulfur dioxide levels – these limits were originally as high as 1700 µg/m³, reduced down to 1400 in 1997, 700 in 2008 and just prior to closure was down to 400 µg/m³.

In a typical week each roaster might be down for 50+ hours or over 30 per cent of available time. The stopping and starting created both operational and production issues. If the AQC was prolonged there was often a chance of having to reheat with the auxiliary diesel burner to regain adequate temperature. The time required to restore temperature was as much as 136 hours from cold and as little as 24 hours from 550°C. This led to a considerable loss of production. Table 1 shows a typical period (quarter) in 2014 clearly demonstrating the level of downtime experienced by the roasters. Corresponding data from the IsaMill™ M3000 ultra-fine grinding (UFG) circuit (IsaMill™), which was not affected by AQC considerations, is included for comparison.

THE NEED FOR ULTRA-FINE GRINDING

During 1999, there was a significant shift in the areas being mined within the open pit operations such that the sulfur head grade of the ore increased by around 75 per cent from 0.8 per cent to 1.4 per cent (Ellis and Gao, 2003). As a result there was a significant increase in the amount of concentrate produced from the Fimiston concentrator, which exceeded the treatment capacity and availability of the roasters at Gidji. It was identified that additional roasting capacity was very expensive to employ, and the addition of alternative treatment routes were technically and economically reviewed, including

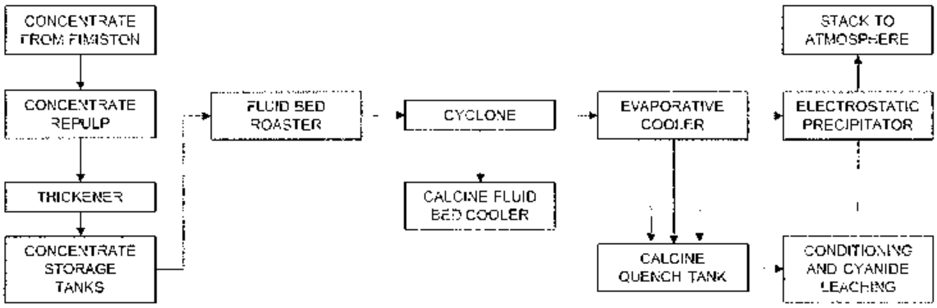


FIG 1 – Gidji Roaster flow sheet for treating Fimiston concentrates.

TABLE 1

Quarterly data from 2014 – Gidji Plant Roaster (R1, R2) and ultra-fine grinding circuit (IsaMill™) operability.

Month	Run hours R1	Run hours R2	Run hours ultra-fine grinding circuit	Utilisation R1 (%)	Utilisation R2 (%)	Utilisation of ultra-fine grinding circuit (%)
1	5[illegible]0	740	715	80.0	91.5	96.[illegible]
2	[illegible]	551	706	[illegible]	76.4	98.0
3	[illegible]	535	730	91.7	[illegible]	97.[illegible]
Average	[illegible]	609	717	[illegible]	[illegible]	97.[illegible]

addition of an acid plant, pressure oxidation, bacterial oxidation and UFG. Based on net present value calculations, UFG followed by direct cyanide leaching was identified as the best alternative to the roasting process, warranting further investigation. A large body of work had been completed on this option over preceding years demonstrating that gold recoveries of up to 92 per cent could be achieved after grinding to a P_{80} of 10 microns (Ellis, 2003). Further laboratory and then pilot test work was carried out by KCGM and is summarised by Ellis (2003).

TECHNOLOGY SELECTION

During the laboratory test work phase, KCGM concentrate samples had been sent to a number of stirred mill providers in order for them to provide estimates of the specific energy requirements to grind to a ten micron product. As reported by Ellis and Gao (2002), due to a wide and differing variety of operating parameters employed by the different manufacturers, it was not possible to make any conclusive comparisons other than that all samples tested indicated that the energy consumption was high to grind below 10 microns.

Following the pilot scale test work, there was a difficult decision for KCGM as to which fine grinding technology to select for the circuit design, due to the fact that stirred milling as a fine-fine grinding step was a relatively new concept in the minerals industry at that time. At the time, there were [illegible] operating sites [illegible]. As a result, the technology choice was narrowed down to the IsaMill™ and the Svedala Detritor (now the Metso SMD (Stirred Media Detritor)). To try and distinguish between the two mills, a head to head onsite pilot mill demonstration was conducted. After working through different media types, grind targets and other performance parameters, KCGM's conclusion was that there was no significant difference in the metallurgical performance of the two stirred mills (Ellis and Gao, 2003).

Due [illegible] for KCGM [illegible] proven operational and maintenance performance [illegible] the duty [illegible] size reduction and [illegible] process. As a result, with the IsaMill™ having a proven operational and maintenance record [illegible] at the time the [illegible] was only available at 355 kW, the IsaMill™ became the selected technology.

GIDJI INSTALLATION AND PERFORMANCE

The Gidji 1.1 MW M3000 IsaMill™ circuit was designed on the back of the laboratory, pilot and demonstration plant data to treat 12 t/h at an F_{80} of 120 microns to a product size of 10 microns. It was commissioned in early 2001. Commissioning performance of the Gidji IsaMill™ circuit has been described by Ellis and Gao (2002).

Commissioned circuit

Figure 2 shows the feed preparation and commissioned UFG flow sheet at Gidji. The new IsaMill™ UFG operated in parallel with the existing roaster circuit. The ground concentrate product from the IsaMill™ was thickened and processed through a three-stage cyanide leach circuit before joining with the leached calcine from the roaster circuit for combined processing through the eight-stage carbon adsorption circuit.

Due to the internal product separator arrangement (described by Anderson and Burford, 2006), most current IsaMills™ operate in open circuit configuration with feed generally being prepared by a preceding cyclone to avoid excessive operating density and to avoid bypass of material already at the target product size. The KCGM IsaMill™ was designed in closed circuit arrangement. Due to the high specific energy input required, open circuit arrangement would have created temperature issues within the IsaMill™, which is limited to a maximum 70°C to protect the internal rubber components. An additional benefit was the preference for high specific gravity (SG) pyrite (containing the gold) to report to the cyclone underflow at equivalent sizes where the lower SG gangue minerals report to the cyclone overflow (Ellis, 2003). This essentially resulted in the pyrite being preferentially ground to a finer size than the gangue minerals, leading to a more efficient use of the grinding energy as the target [illegible] minerals were not being ground finer than necessary.

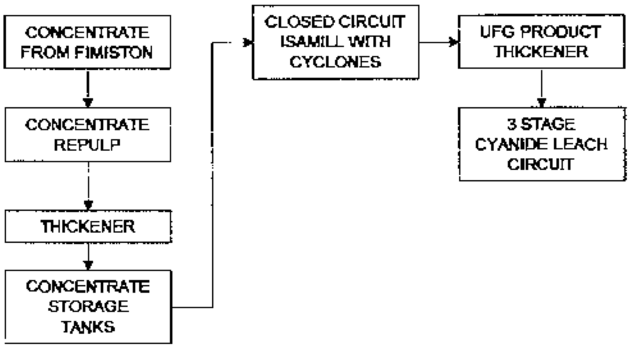


FIG 2 – Gidji flow sheet [illegible] IsaMill™ (Ellis, 2003).

Grinding media

Grinding media selection for optimised energy efficiency must consider both the slurry feed size distribution and the required product size target. Primarily, the selected media must be large enough to ensure adequate top size breakage and prevent mill bogging. Evidence of [illegible] in the IsaMill™ [illegible] due to build-up of the coarse [illegible] grinding media [illegible]. This was observed at KCGM [illegible] pilot operation and has also been observed in other mill [illegible] during the [illegible] grinding development work at McArthur River (Anderson, [illegible] and [illegible], 2012).

The coarse media required to deal with the top size was not the optimal size to grind efficiently to the fine target product size. Therefore, the decision to make the size reduction at KCGM in a single step ensured that a penalty in grinding efficiency would be incurred due to the need for a compromised media size selection. At the time of the original installation, sand media was the only identified viable economic option (Ellis, 2003) and a 6 mm top size sand media was employed.

In the years since the KCGM installation, there has been a rapid uptake of both IsaMill™ and alternative stirred mill technology – IsaMill™, for example, grew from an installed base of 17 MW at the time of the Gidji installation to over 200 MW by 2012. This growth resulted in ceramic grinding media becoming a much more cost-effective option, with many suppliers now vying for a share of the market, driving prices down. Over 90 per cent of all IsaMills™ installed now use ceramic media. Ceramic media provides significant advantages over sand due to its higher density, high hardness, smoother surface characteristics and lower consumption rates, leading to benefits in grinding efficiency and mill component wear (Curry and Clermont, 2005). Critically, it is manufactured and so can provide a more stable and consistent product compared to sand, which is subject to natural variations that can impact on quality, consumption and therefore on mill performance in terms of both efficiency and wear characteristics.

KCGM started to investigate ceramic media use in 2006 and eventually changed both mills to operate on ceramic media by 2011. The higher density of the ceramic at SG 3.7 allows a smaller size media to be used and still generate the same or larger stress intensity inside the IsaMill™. The ceramic media also has a much smoother surface profile compared to the sand. These two factors assist to reduce the wear rate of the internal mill components, which lead to lower downtime and lower spare parts costs. Test work reported from a two month trial at the Fimiston M3000 IsaMill™ (Wang, Glover and Clements, 2013) identified a 20 per cent reduction in specific energy, a 50 per cent increase in maintenance interval and a 25 per cent improvement in plant throughput for the same target grind size. Historical consumption rates of the sand at Gidji were around 17 kg/t of concentrate, whereas the current 3.5 mm Magotteaux Keramax ceramic media has a measured consumption rate of 1.10 kg/t, which is about 10 g/kWh. It is noted that the Fimiston UFG mill, also an M3000, uses a 3.5 mm Keramax ceramic media.

Wear performance

One of the main issues back in mid-2013 was the high outside edge wear on the discs at the feed end of the IsaMill™, which required mill shutdowns approximately every ten days for disc maintenance. The performance was significantly worse compared to the Mt Isa or McArthur River mills and was a direct outcome of the coarse and hard pyrite feed material combined with the coarse sand media in use (Ellis and Gao, 2002). Shell liner wear was also higher than the other sites. Through a concerted development effort between KCGM and Glencore Technology (GT), significant improvements were made in the wear performance during this operating period.

There have been many benefits gained from continuity in wear material development and media use since then. Currently, the M3000 IsaMill™ operates on a four to five month maintenance cycle, [illegible] the shell liners [illegible].

Smaller diameter disc configuration

In late 2014, the Gidji mill was configured with smaller diameter discs (SDD), which had been developed by GT to address wear issues at other sites experiencing higher than average wear rates as a result of processing coarse, hard feed material (Rule and deWaal, 2011). A total of two SDD were ultimately installed at the first two disc positions in the Gidji mill leading to further improvements in the mill operational stability. Longer term benefits from the SDD included measured reductions in shell liner wear rates of 50 per cent and a ten per cent increase in mill throughput for the same grind size target. Shutdown frequencies were increased from a typical 18 days to 38 days.

FIMISTON INSTALLATION

A second M3000 IsaMill™ in a similar duty was installed and commissioned at the Fimiston concentrator in 2012 to process additional stream concentrate as a result of higher concentrator throughput and increased sulfur grades. The IsaMill™ was selected for the UFG duty on the basis of the success of the Gidji installation. Based on learnings from the Gidji circuit, several changes were made to the Fimiston circuit including cyclones and pumping discussed by Ellis and Gao (2002).

EMISSIONS REDUCTION PROJECT

The emissions reduction project (ERP) commenced in 2013 with the aim of eliminating sulfur dioxide and mercury emissions from the Gidji Processing Plant. The ERP project included the installation of the new M10 000 IsaMill™ at Gidji with associated flotation equipment, additional leaching tanks, decommissioning of the roasters and installation of an additional carbon regeneration kiln and new mercury condensing and capture system at Fimiston for the carbon regeneration process.

SELECTION, DESIGN AND COMMISSIONING OF THE GIDJI M10 000 ISAMILL™

A number of signature plot tests were completed by ALS Ammtec in Perth under various conditions resulting in a 3 MW M10000 IsaMill™ being confirmed as the selected mill for the project to treat 36.5 t/h of fresh feed to a product size of P_{80} 13 microns. GT was contracted to supply its standard IsaMill™ package, which included the basic IsaMill™ supply (including drivetrain and associated lubrication/cooling systems) together with feed and discharge pump boxes and pumps, media addition bin and media charging system, instrumentation, associated piping, platforms and structure.

The M10 000 IsaMill™ design incorporates a two-stage cyclone circuit arrangement. Flotation feed from the existing concentrate holding tank is fed into the IsaMill™ feed hopper, from where it is pumped into the IsaMill™ with grinding media as required. Discharge from the IsaMill™ enters the primary cyclone feed hopper from where it is pumped to the primary cyclones (GL Warman CVX 100 mm) with the underflow stream gravitating to the secondary cyclone feed hopper. Underflow from the secondary cyclones (also GL Warman CVX 100 mm) is gravity fed to the IsaMill™ feed hopper to combine with the fresh feed.

Both cyclone overflows are pumped to a new 13 m diameter thickener. The thickener underflow is pumped to the existing leaching circuit with thickener overflow reporting to a dedicated process water tank for use in the new circuit. The overall Gidji flowsheet utilising the new IsaMill™ and no roasters is shown in Figure 3.

A late change to the project was the inclusion of a variable speed drive (VSD) and Toshiba motor in order to prevent potential harmonics associated with power fluctuations on the local embedded power supply. Typically the large-scale IsaMills™ operate with a fixed speed motor to reduce the capital costs – a wound rotor motor and liquid resistance starter (LRS) combination can be less than half the cost of a squirrel cage motor with VSD. However, inclusion of a VSD can provide the operator with additional control variables with which to optimise grinding and component wear performance. To date, the IsaMill™ has only been operated at its design speed but there does exist an option to investigate the impact of mill speed in the future.

Figure 4 shows an overview of the new M10 000 IsaMill™ installation. Installation of the IsaMill™ and surrounding equipment was largely completed by mid-March 2016 and initial commissioning of the IsaMill™ commenced from 9 March. IsaMill™ commissioning followed a basic standard plan and checklist which ensures all items are correctly commissioned in the correct order. Within the basic plan there is flexibility with the order, depending on the daily site conditions and schedule.

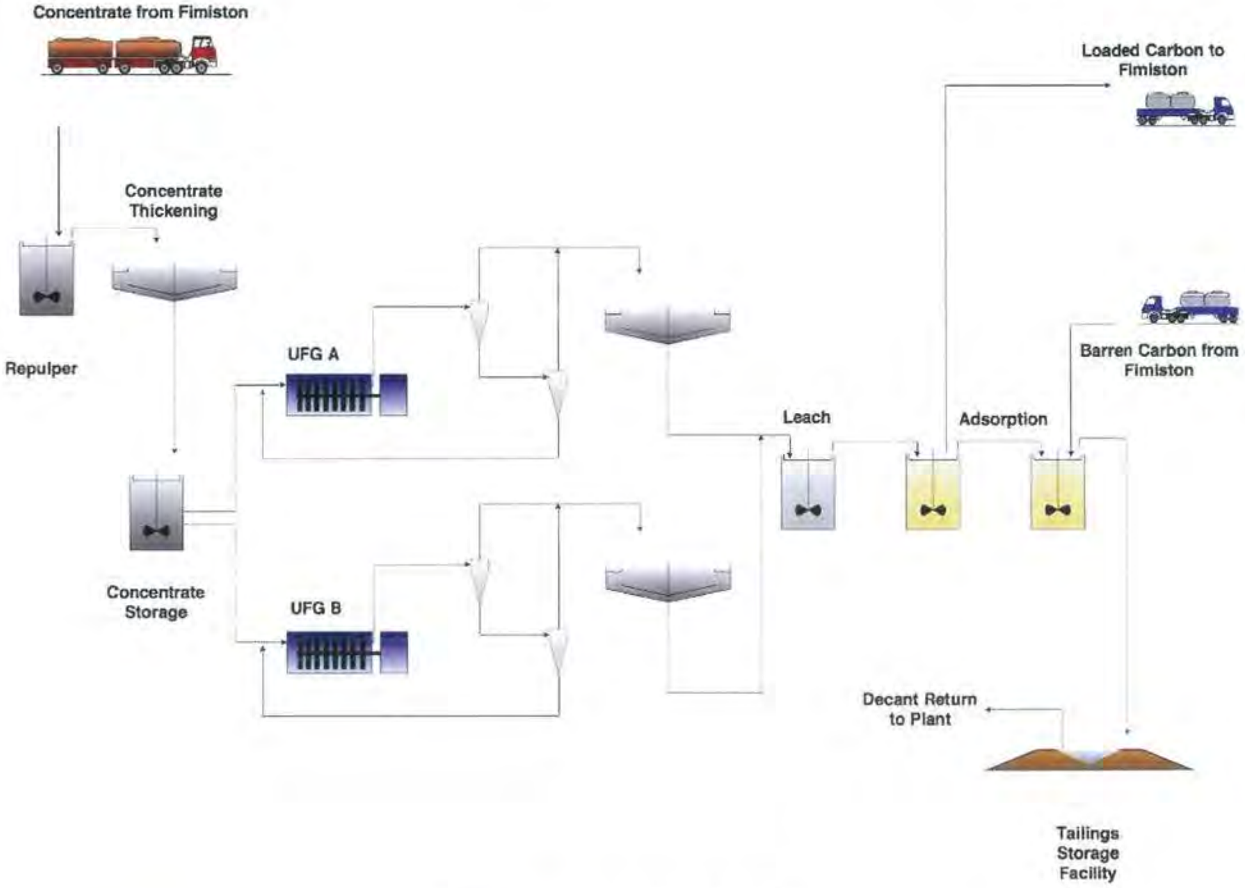


FIG 3 – Gidji – all ultra-fine grinding flow sheet.

FIG 4 – Kalgoorlie Consolidated Gold Mines' M10 000 IsaMill™ installation.

The initial priority was to prepare the necessary equipment to enable the uncoupled run of the IsaMill™ motor when it was available. This involved standard instrumentation/electrical and interlock checks through to the plant distributed control system (DCS) and flushing/commissioning of the mill motor bearing lubrication system. Availability of the motor for commissioning was delayed several days due to technical issues with the VSD. These were resolved and uncoupled motor testing was able to commence on 21 March. In parallel, commissioning of the IsaMill™ and surrounding equipment progressed. This included:

- standard instrumentation/electrical/interlock checks on each system through to the DCS
- operational testing of all valves
- flushing of all lines and tanks
- flushing and commissioning of the IsaMill™ bearing and gearbox lubrication systems
- flushing and commissioning of the IsaMill™ gland system
- water commissioning of the IsaMill™ media charging system (IsaCharger)
- water commissioning of IsaMill™ feed tank and pumps
- water commissioning of IsaMill™ discharge tank, pumps and cyclones
- control sequence logic (start-up, shutdown, media addition) simulation testing at the DCS.

The final step involved pumping water through the mill and actual testing of the start/stop sequences, without the motor coupled. Upon satisfactory completion of those tests, the motor was coupled to the gearbox and IsaMill™ for the final stage of water commissioning, which involved simulating slurry conditions, adding an initial charge of media into the mill and running final checks on the sequencing logic. In this testing the IsaMill™ was operated with both sets of cyclones in operation. Water testing was completed by 27 of March; however, due to other circuit issues, slurry remained unavailable to the IsaMill™ circuit and GT departed site, returning on 8 April to commence slurry commissioning.

The roaster suffered a significant failure on 5 April and the decision was made to not repair, which hastened the slurry commissioning in order to maintain site gold production. The IsaMill™ commenced treating slurry on 8 April and was quickly ramped up to the design target of 30 t/h at 2200 kW with measured thickener underflow circuit product distributions in the 11–13 micron range. As per standard commissioning practice, the IsaMill™ was shut down for an initial inspection on 12 April after 83 operating hours. The inspection highlighted no abnormal wear conditions in the mill, so the mill was closed and operation continued.

The IsaMill™ internal configuration was changed from three small diameter discs to two small diameter discs on the second inspection, which occurred on 13 April, to provide better distribution of the grinding media within the IsaMill™.

GIDJI M10 000 PERFORMANCE

Maintenance

Since commissioning, the M10 000 has operated at 95 per cent availability with shutdowns occurring approximately once every four to five weeks. The current data suggests about 11 discs and less than one shell liner change out will be required per annum. The most recent shell liner lasted 285 days.

For the M10 000, the most significant maintenance issue since commissioning has been the drive end bearing failure, which was traced to an issue related to an extended period between the mill installation and the commissioning process.

Grinding media

The same 3.5 mm Magotteaux Keramax media is used in the M10 000 as the M3000. Consumption is about approximately 1.25 kg/t of dry concentrate processed.

Current grinding circuit performance – Comparing the M3000 and M10 000 IsaMills™

Table 2 compares the operational parameters between the two IsaMills™ at Gidji. The M10 000 is currently operating at more than ten per cent above the original design throughput of 20.5 t/h and achieving the grind targets set by the plant metallurgists.

Leaching performance

When head grade variability is taken into account, average gold recovery of [illegible] is assessed as being currently around 0.5 per cent lower than the nominal recovery and four per cent lower than the previous recovery achieved via the roasting process. The roasting process produces very porous particles once the sulfides are oxidised – essentially hollowing the pyrite host. This high degree of porosity enhances the dissolution of gold by allowing access to a high percentage of the gold particle by the cyanide solution. The UFG route breaks the particle down to a fine size and induces stress into the mineral lattice allowing a high percentage of the gold to be accessed by the cyanide. Finer grinds promote further exposure; however, eventually this can introduce the formation of some surface coatings, which limit the extraction. A comprehensive study is underway to determine the full gold deportment and determine methods to improve the gold recovery by a minimum of two per cent to approach the same recoveries achieved under the roaster operation.

TABLE 2

M3000 and M10 000 operational parameters.

Parameter	Units	M3000	M10 000
Solids new feed	t/h	5.9	32.9
Recirculating load	wt%	197	183
New feed pulp density	wt%	52.5	50.0
Specific energy consumption	kWh/t	42	71.1
Media consumption	g/kWh	15	15
IsaMill™ circuit feed F_{80}	µm	125	125
IsaMill™ circuit product P_{80}	µm	13.1	12.0
Cyclone feed pulp density	wt%	30.6	32.6
Solids split to cyclone overflow	wt%	16.6	31.9
Water split to cyclone overflow	wt%	56.0	74.2
Cyclone underflow pulp density	wt%	51.7	62.5

Ultra-fine grinding versus roasting costs

Table 3 summarises the main differences of operating costs for the Gidji Processing Plant comparing the last 12 months of roaster operation with the M3000 circuit to the ten months up until March 2016 of the full grinding/leaching circuit (roaster decommissioned) comprising the M3000 and the M10 000. Operating cost has reduced by 0.5 per cent and maintenance cost by 23.4 per cent for an overall combined reduction of 11 per cent or nearly $13/t. On an annual basis this represents a A$3.5 M saving in operating and maintenance costs.

Emissions reduction

The reduction of sulfur dioxide (SO2) and dust emissions associated with the operation of the roasters at Gidji has been a long-term focus of KCGM and its joint venture owners Newmont and Barrick. Following the formation of KCGM in 1989, the establishment of the Gidji Processing Plant resulted in the decommissioning of the remaining three Kalgoorlie roasters. This resulted in a significant reduction in sulfur dioxide (SO2) levels in the City of Kalgoorlie-Boulder. With the decommissioning of the roasters at Gidji Processing Plant, KCGM has eliminated 170 000 t of sulfur dioxide from entering the atmosphere annually. Gidji was previously listed in the National Pollutant Inventory (NPI) top two highest mercury and sulfur dioxide emitters. Replacement of the roasters with the UFG circuit means these air emissions have now been totally eliminated from the site.

GIDJI M10 000 ACOUSTIC MONITORING

Typically, an IsaMill™ operator can determine the charge and diagnose the mill's performance based on listening to the sound from the IsaMill™ shell with their ear pressed directly onto the shell. However, this is a difficult skill to learn and is subject to individual differences and interpretations. In conjunction with the CSIRO, GT has developed an acoustic analyser for the IsaMill™ to aid quantitative measurement and optimisation of the media charge position within the mill. Stress waves, generated by the impact of grinding media with the internal mill shell liner, are detected by broadband accelerometers and processed into an acoustic emission mean signal power to allow comparison at various points along the length of the mill shell including the product separator compartment. The unit itself has been well described by Jackson *et al* (2014). Two of these units were originally installed as part of the acoustic development program. One of the acoustically equipped IsaMills™ was idled during 2015, so the opportunity was taken to shift the analyser to the KCGM M10 000 site for the commissioning process. It has proved very useful in both assisting the

TABLE 3

Ultra-fine grinding and roasting cost comparison.

	Operating expenditure	Maintenance expenditure	Combined costs
Change	-0.5%	23.4%	-11.0%

operators to understand the condition and operation of the mill and in diagnosing several issues.

Media in the product separator

The product separator is located at the discharge end of the IsaMill™ and essentially allows ground slurry to exit the mill while retaining the grinding media in the mill through its centrifugal pumping action (Anderson and Burford, 2006). Under ideal operating conditions, the product separator area is free of media; however, certain conditions can cause media to enter the product separator. This can reduce the performance of the product separator, eventually leading to more media in the area, loss of media from the mill, wear of the rotor and unstable power draw.

After a mill shutdown on 5 May 2015, the mill was charged to a higher power draw set point than intended due to an issue with the control system around the media charger. This immediately resulted in an unstable power draw situation with a corresponding distinct change in the acoustic reading in the product separator area. This reading indicated the presence of media in the product separator, which could be confirmed by direct listening to the mill shell and suggested that the mill had been overcharged with media. Due to the closed circuit configuration, any media lost from the mill was ultimately returned back via the cyclone underflow so the actual media load could not be reduced without draining media from the mill back into the media hopper.

Higher flow rates can push media into the product separator area as the pumping action of the rotor is overcome. Reducing the flow rate can change the hydraulic balance back in favour of the product separator, allowing media to clear the area and compress more towards the feed end of the mill. Reductions were made to the flow rate in this case with minimal effect suggesting that the media was in the rotor because the mill was full of media rather than as a direct result of any specific process condition. Another observation was that the mill density was operating around 46–48 per cent solids by weight. The density/viscosity can have a significant impact on the overall mill power draw and grinding efficiency.

The mill was shut down for inspection on 26 May, the only changes being an end-to-end rotation of the shell liner. Some additional wear was observed on the product separator related to the presence of media. Prior to the shutdown the media bin was run as empty as possible so that the media volume dumped from the mill could be inspected and monitored during the recharge procedure.

It was recommended to restart the mill at a 50 per cent solids target as this should allow a higher power draw for a given volume of grinding media in the mill. Upon restart, the only process change was the increase in density and the mill was able to reach the target power draw set point of 2430 kW with no significant acoustic measurement detected in the rotor area. The before and after trends are shown in Figure 5. From observation of the charge level in the media bin, it was estimated that once the mill had reached the power draw set point, there were at least 4 t of media still remaining in the bin that had previously been in the mill (an M10 000 operating at full capacity would normally hold around 20 t of media). This confirmed that the mill had been overcharged with media. By remaining at the 50 per cent solids target, the mill was able to be operated successfully without any noise in the rotor area. Figure 5 also demonstrates the response of the acoustics to the variations in the mill power draw.

Shell liner wear

A disposable rubber shell liner supported in a 6 mm steel backer is mounted between the two outer shell halves and the end flanges. During an internal maintenance inspection,

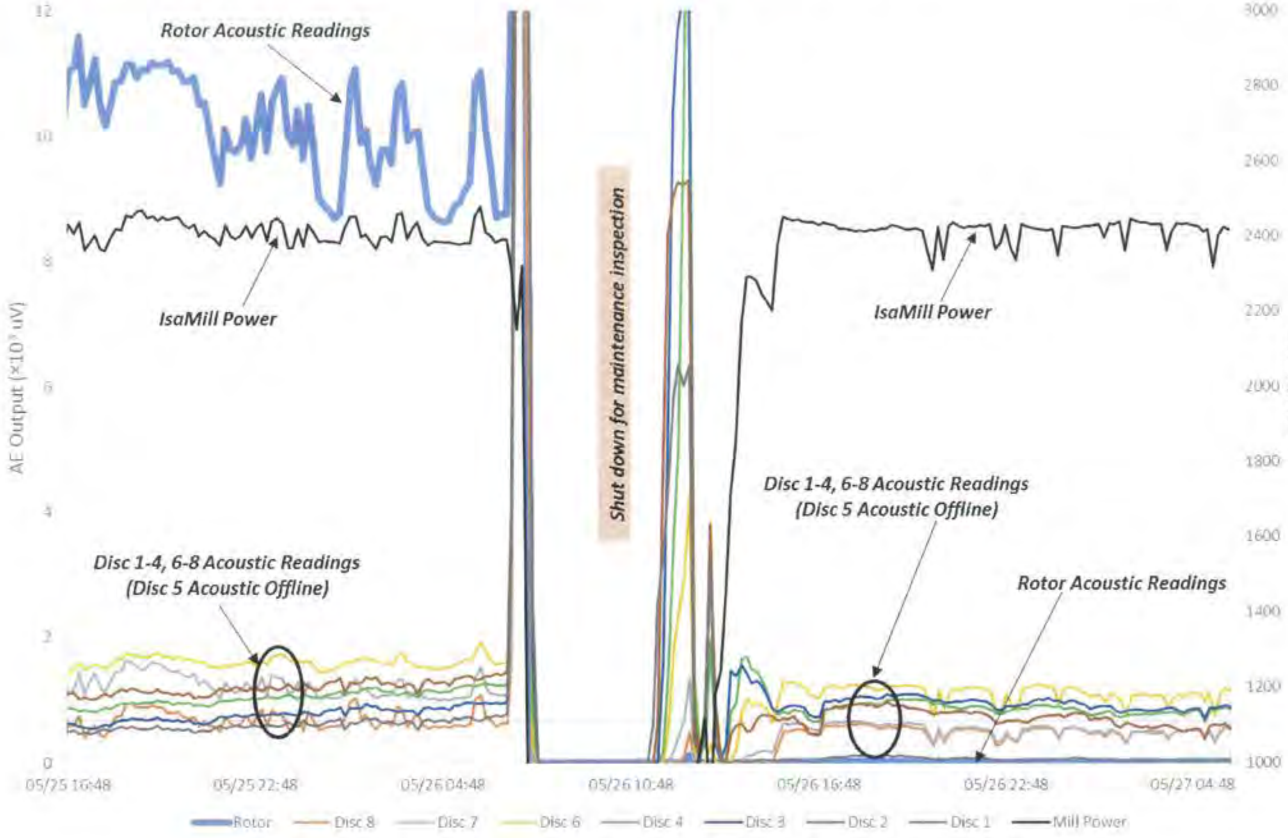


FIG 5 – IsaMill™ acoustic readings before and after shutdown.

the shell liner is rotatable about its own axis and end-to-end to allow the overall life of the liner to be maximised. The liner wears at different rates along its length and diameter dependant on the slurry/media characteristics and process conditions within the mill.

Due to the over-charging incident in May (referred to previously) there was an abnormal wear pattern on the shell liner. On 26 May shutdown directly after the over-charging period, the shell liner was rotated end-to-end and 180 degrees about its axis. Unfortunately this placed a more worn area of the liner into a section of the mill that was subject to higher wear under the normal operating conditions. As the run hours progressed during June, it was apparent that the acoustic reading at the disc four area steadily increased, in relation to the other acoustic channels, up until the afternoon of 9 June when it started to increase rapidly as shown in Figure 6. It is believed that the rapid increase corresponded with the last of the rubber wearing away and the media then starting to wear into firstly the shell liner steel backer and then into the actual mill shell, culminating in the leakage of slurry from the shell.

The acoustic system was not set up with any alarms at the time; however, on reviewing the acoustic data after the event, a series of alarms were implemented to warn of such an issue in the future. These alarms are now linked to the mobile phones of key plant personnel. This event demonstrated the ability of the acoustic system to pick up abnormal conditions within the mill and act as an early warning system, allowing potential issues to be addressed before they become problems.

SERIES GRINDING

The KCGM grind is a very arduous duty with a significant size reduction in a single step. Typical size reductions for the IsaMill™ are in the order of five to six times, whereas KCGM is around ten times. Previous test work by GT has demonstrated significant energy efficiency improvements through the use of series grinding, particularly where reduction ratios are greater than eight times and the product is ultra-fine.

Series grinding initially utilises a larger media to reduce the top size of the incoming feed down to a point where a smaller, more efficient media can complete the energy intensive grind to the final product target. At ultra-fine product targets, energy efficiency is dependent on media size. However, the media size must still be adequate to break the coarsest particles in the feed to prevent build-up and potential bogging of the mill.

The 3.5 mm ceramic media, employed in each of the parallel IsaMill™ circuits at Gidji, was a compromise between being large enough to break the coarse particles in the feed and having enough fine media present to enable production of sufficient sub 12 micron material. A 4-5 mm media was considered more ideal for breaking the top size of the incoming feed material but would simply not be efficient to grind all the way to 12 microns. A 2 mm media is much more suited to efficiently grinding to the 12 micron target; however, it would not provide adequate breakage of the coarse end of the incoming feed.

Considering that two IsaMills™ (the new M10000 and the existing M3000) would be on the Gidji site, investigations were conducted into possible energy efficiency improvements through series grinding. Based on the power available in each of the mills it was established that the M3000 would be used to do the initial grind of the coarse feed using larger media to produce a suitable transfer size for the more energy intensive finer grind in the M10000 using the smaller media. This represented about a 25/75 power split between the two mills.

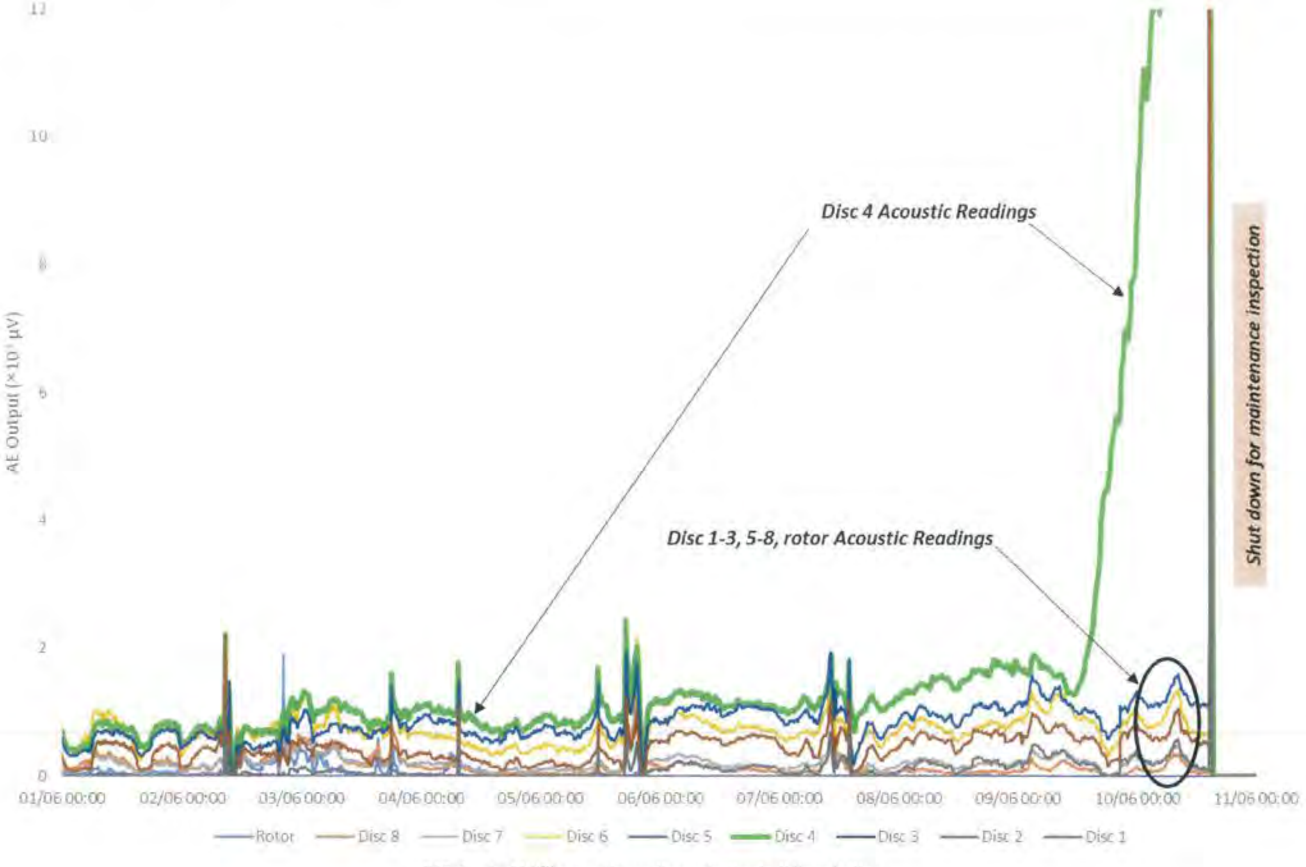


FIG 6 – IsaMill™ acoustic reading prior to shell liner holing.

Samples of KCGM regrind circuit feed were sent to ALS in Perth where they were prepared for standard IsaMill™ signature plot test work – ensuring that the samples for each test were identical. Analysis of the feed indicated a P_{80} of 117 microns. A single-stage test was conducted using the standard graded 3.5 mm media and resulted in a specific energy of 60 kWh/t to grind to a P_{80} of 12 microns. A two-stage grinding test was then completed whereby a graded 4–5 mm media was used as the first stage consuming 13.8 kWh/t to produce a P_{80} of 42 microns. The first-stage product became the feed to the second-stage grind using a graded 2 mm media, which consumed a further 35.7 kWh/t to reach the P_{80} 12 micron final grind target. The two stages combined consumed a total specific energy of 49.5 kWh/t to the P_{80} 12 micron target. This represents a 17 per cent reduction in specific energy compared to the single-stage grind using the 3.5 mm media and a potential substantial energy saving for KCGM. The P_{98} was maintained at 24.3 microns for both the single-stage and two-stage grinds where the P_{80} of 12 microns was achieved. Figure 7 shows the signature plots of both the P_{80} and P_{98} data for the single-stage and series grinds. The equations are shown for the P_{80} curves and the decrease in exponent (line slope) indicates the change in efficiency. The difference in total energy requirement at the 12 micron P_{80} target, and in the corresponding P_{98} curves, can be clearly seen and highlights the improved breakage rates of the two-stage grind.

To maintain the existing combined plant tonnage, overall throughput would be 45 t/h. The M3000 IsaMill™ would operate with 4–5 mm media and be configured with SDD to increase the disc-shell gap. Power draw would be limited to around 750 kW or about 16 kWh/t. The M10 000 would operate with 2 mm media in conventional configuration and be able to input up to 60 kWh/t. If the predicted efficiency gains are realised, then there is also opportunity for overall plant tonnage to be increased.

In addition to improved efficiency, maintenance benefits may also arise. Observations by GT over many years have established that coarser feed distributions result in higher IsaMill™ internal component wear rates. The coarsest feed particles would be broken down more rapidly in the M3000 by the use of the 4–5 mm media (compared to the 3.5 mm media) in the M3000. The M10 000 will not see any coarse particles due to it receiving the product from the M3000 and will only use fine 2 mm media. Both of these configuration changes will likely result in reduced wear compared to the current configurations using 3.5 mm media. Operationally, the 4–5 mm media will have a wider operating window to handle coarse material compared to the 3.5 mm media. This should result in the two-stage grind being better positioned to handle variations in feed size and hardness into the plant.

Ultimately the new M10 000 circuit was installed in parallel to the existing M3000 circuit in order to simplify the commissioning process and develop operational and maintenance understanding of the new IsaMill™ at the plant. Plans are currently in place to investigate the options to transition to the series circuit.

CONCLUSION

Gold was traditionally produced at KCGM through roasting of the refractory concentrate followed by cyanide leaching. Since 2001, KCGM has operated the M3000 IsaMill™ technology, initially to provide additional gold output alongside the roaster at Gidji followed by a second unit at Fimiston. In 2015, with the installation and successful commissioning of an M10 000 IsaMill™, the roasters were shut down with all gold production via the UFG route. This enabled atmospheric contamination by sulfur dioxide (170 000 t/a) to be eliminated. As a result of the operation no longer being constrained by AQC, March 2016 saw the highest monthly production tonnage at Gidji since 2002.

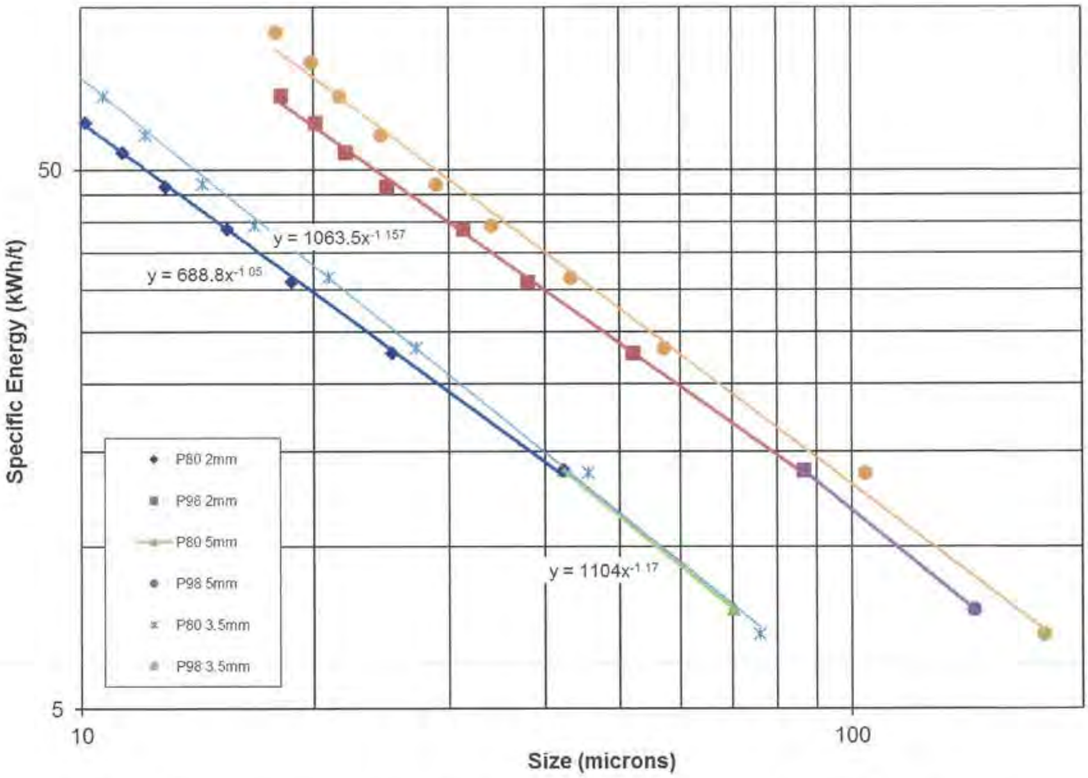


FIG 7 – Kalgoorlie Consolidated Gold Mines' signature plots for single stage (3.5 mm) and two stage (5 mm and 2 mm).

Gidji operating and maintenance costs were reduced by 11 per cent or A$4.8 M/a. Projects are currently underway to investigate increasing gold recovery.

A new acoustic monitoring system was added to the [illegible] IsaMill, which has proven invaluable in assisting the plant operators to better diagnose issues within the mill. There now exists opportunities at Gidji to further improve the efficiency of the UFG circuit, by modifying both mills to have a series grinding configuration.

ACKNOWLEDGEMENTS

The authors wish to thank the KCGM Fimiston Processing and Gidji Maintenance staff for their ongoing support. The authors further acknowledge and appreciate the management of Kalgoorlie Consolidated Gold Mines and Glencore Technology for permission to publish this information. The authors thank and acknowledge KCGM's [illegible] owners Barrick Australia Pacific and Newmont Australia Pty Ltd for their support of the Emissions Reduction Project.

REFERENCES

Anderson, G S and Burford, B D, 2006. IsaMill – the crossover from ultrafine to coarse grinding, in *Proceedings MetPlant 2006*, pp 18-32 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Anderson, G S, [illegible] and [illegible], S J, 2015. IsaMill [illegible] in the primary grinding circuit, in *Proceedings SAG Conference 2015*, Vancouver.

Blake, D, Clarke, C and Clermont, B, 2012. Performance assessment of direct and indirect ceramic grinding media in the IsaMill fine grinding application, *AusIMM Bulletin*, [illegible].

Curry, D C, Clark, L W and Rule, C, 2005. Collaborative technology development – design and operation of the world's largest stirred mill, in *Proceedings Randol Perth Conference*.

Curry, D C and Clermont, B, 2005. Improving the efficiency of fine grinding – developments in ceramic media technology, in *Proceedings Randol Perth Conference*.

Ellis, S, 2003. Ultrafine grinding – a practical alternative to oxidative treatment of refractory gold ores, in *Proceedings Eighth Mill Operators' Conference*, pp 11–17 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Ellis, S and Gao, M, 2002. The development of ultrafine grinding at KCGM, in *Proceedings 2002 SME Annual Meeting* (Society for Mining, Metallurgy and Exploration: Littleton).

Enderle, U, Woodall, P, Duffy, M and Johnson, N W, 1997. Stirred mill technology for regrinding McArthur River and Mount Isa zinc/lead ores, in *Proceedings XX IMPC*, Aachen, pp [illegible].

Harbort, G, Murphy, A, [illegible] and [illegible], [illegible]. The intersection of the IsaMill for ultrafine grinding in the [illegible], in *Proceedings* [illegible].

Jackson, C A, [illegible], 2014. [illegible] of M10000 IsaMill [illegible] performance by [illegible] acoustic emissions, in *Proceedings 12th AusIMM Mill Operators' Conference*, pp 241–254 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Johnson, N W, Gao, M, Young, M F and Cronin, B, 1998. Application of the IsaMill (a horizontal stirred mill) to the lead-zinc concentrator (Mount Isa Mines Ltd) and the mining cycle, in *Proceedings 1998 AusIMM Annual Conference*, pp 291–295 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Pease, J, Curry, D, Barns, K, Young, M and Rule, C, 2006. Transforming flow sheet design with inert grinding – the IsaMill, in *Proceedings 38th Annual Canadian Mineral Processors Conference 2006*, pp 231–249 (Canadian Institute of Mining, Metallurgy and Petroleum: Montreal).

Rule, C and de Waal, H, 2011. IsaMill design improvements and operational performance at Anglo Platinum, in *Proceedings MetPlant 2011*, pp 176–195 (The Australasian Institute of Mining and Metallurgy: Melbourne).

MetPlant 2015

Metallurgical Plant Design & Operating Strategies – World's Best Practice

7–8 September 2015, Perth, Western Australia

Image courtesy of Outotec. Atlas Mining's state-of-the art copper concentrator plant upgrade at its Carmen Copper Mining Complex in Toledo City, Cebu, Philippines.

Principal Sponsor

Outotec

Proceedings Sponsor

MAGOTTEAUX

SHAPING A WORLD OF PERFORMANCE

Proceedings

AusIMM
THE MINERALS INSTITUTE

Commissioning and Ramp-up of the Albion Process at the GPM Gold Project*

P Voigt[1],M Hourn [2], D Mallah[3] and D Turner[4]

1. Manager - Hydrometallurgy, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, paul.voigt@glencore.com.au
2. General Manager - Technology, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, mike.hourn@glencore.com.au
3. Metallurgist - Hydrometallurgy, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, daniel.mallah@glencore.com.au
4. General Manager Technology, Core Resources, 44 Corunna St Albion QLD 4010

* This paper was originally published in the proceedings of ALTA 2015.

ABSTRACT

The GPM Gold Project is located in Armenia, and consists of an open cut mine at Zod, near the Azerbaijan border, and a CIL processing plant at Ararat near the Turkish border. Mining at Zod commenced in 1976, and focused on near surface oxide ores, which overlay deeper refractory sulphides. Historical mining has now almost depleted the oxide ores, and the sulphide content of ore delivered to the processing plant at Ararat is increasing. Gold and silver recoveries through the Ararat plant were declining steadily.

GeoProMining, the owners of the project, have expanded the Ararat facility to deal with the increasing sulphide content in the ore. In 2014, GeoProMining refurbished an existing concentrator on site to recover a sulphide concentrate from the ore, and have constructed an Albion Process^{TM} plant to oxidize the refractory concentrate. Glencore Technology (GT) provided the Albion Process^{TM} plant as a technology package.

In July 2014, the progress of the GPM Gold Project Albion Process^{TM} Plant was reported (Hourn, Voigt and Turner, 2014). At the time of writing, the construction of the plant was nearly complete. This paper presents an update of project progress, covering commissioning and ramp up of the GPM Gold Project.

The commissioning phase occurred over June to August 2014, which included an M3,000 IsaMill^{TM} fine grinding plant, a 6 tph limestone milling plant, a 60 tpd vacuum pressure swing adsorption oxygen plant, a 10 m residue thickener and a 12 tph Albion Process^{TM} oxidative leach plant. Since commissioning was completed, ramp up occurred over the following three months with downstream gold recoveries from cyanide leaching reaching over 98%.

INTRODUCTION

The GPM Gold Project

The GPM Gold Project is owned by GeoProMining Gold LLC and is located in Armenia. The project consists of an open cut mine at Zod, near the Azerbaijan border, and a processing plant at Ararat near the Turkish border. The Ararat plant has a milling and flotation facility built during the soviet era, with a capacity of 1 million tonnes annually, and a CIL plant built in 1997, with a capacity of 1.5 million tonnes annually. The gold bearing ore, mined at the Zod Mine, is transported to the process plant at Ararat via a state owned rail link.

The Zod deposit originally consisted of weathered oxide ores overlying deeper sulphides. Arsenopyrite and pyrite are the major sulphide minerals. Historical mining has depleted the oxide ores, and the processing plant at Ararat currently treats significant quantities of sulphide ore with increasing amounts of gold locked within refractory sulphides.

The mineral reserves for the Zod mine, at August 2011, were estimated to be 14.2 million tonnes at a gold grade of 4.3 g/t. The mineral resources were estimated to contain 28 million tonnes of ore in the indicated category, at a gold grade of 4.2 g/t and 16 million tonnes of ore in the inferred category, at a gold grade of 4.2 g/t.

GeoProMining are expanding the Ararat plant by re-commissioning an existing flotation concentrator to recover a sulphide concentrate from the ore, and constructing an Albion Process (Hourn & Turner, 2012) plant to oxidize the refractory concentrates. The Albion Process plant will convert the sulphides to oxides, breaking down the sulphide matrix and liberating gold and silver for recovery. Tailings from the concentrator and the Albion Process plant will be combined and transferred to an existing CIL plant to recover the gold and silver as bullion.

Refurbishment of the concentrator and construction of the Albion Process plant commenced in 2013, with commissioning completed in August 2014. The Albion Process plant will has a design capacity to process up to 110,000 tonnes per annum of refractory concentrate from the concentrator.

Deposit Geology

The Zod deposit is located in the Vardenis District of Western Armenia within a setting of volcanogenic and volcanogenic-carbonate sequences, with gabbro-peridotite intrusions that have metamorphosed to serpentinite (Konstantinov & Grushin, 1970).

Gold mineralization is associated with carbonate alteration of ultramafic rocks and is commonly hosted within hydrothermal alteration zones, represented by talc carbonate and quartz-carbonate assemblages. The ore is moderately hard with a medium level abrasion index.

Gold occurs as native free gold, finely dispersed gold in arsenical sulphides, gold tellurides and secondary native gold remaining after oxidation of sulphides and tellurides. Silver occurs in its native form in quartz, chalcopyrite and pyrite, and as silver tellurides.

The deposit has an average sulfur grade of 1.4 % w/w, with an average gold and silver grade of 4.54 g/t and 4.65 g/t, respectively. The arsenic grade across the deposit is 0.3% w/w. The majority of the sulphides occur as relatively coarse mineral grains. The dominant gangue minerals are quartz, talc and chlorite, with minor magnesite, dolomite and calcite.

Development Testwork

Development testwork for the GPM Gold Project began in 2009 initially with batch testwork, and culminated in a continuous flotation and Albion Process pilot plant run over the months May and June 2010. Approximately 4,600 kg of sulphide ore samples were collected from across the Zod ore body to support the testwork program. The samples were classified by ore type, spatial location and sample type and blended into 163 composites. The composites were then grouped into the four major orebodies identified in the primary sulphide resource – orebodies 1, 4, 16 and 23.

Diagnostic leaching and ore characterization testwork (Rohner & Andreatidis, 2010) confirmed that an average gold recovery of only 48% w/w was possible from the blended Life of Mine ore adopting conventional carbon in leach (CIL) leaching methods, and that the majority of the refractory gold was present in arsenical minerals, such as arsenopyrite and arsenical pyrite. Laser ablation work showed that the majority of the pyrite had levels of arsenic in the lattice, averaging 0.9% w/w.

Comminution testwork focused on generating comminution modeling parameters to determine the capacity of the existing crushing and grinding circuit at the Ararat plant. The ore displayed an average Bond Crushing Index of 10 kWh/t, an abrasion index of 0.085 and an unconfined compressive strength of 59 kN. The Bond Ball Mill work Index was 16.5 kWh/t and the Bond Rod Mill work index was 15.8 kWh/t. Modeling work by SMMC (Morrell, 2010) confirmed the milling circuit at the Ararat plant would be capable of processing between 0.9 and 1 Mt/a of ore from the Zod deposit, with minor refurbishment.

Batch and locked cycle flotation campaigns were completed on the testwork samples and a flow sheet consisting of a bulk roughing and single cleaning stage was developed and taken forward into a continuous pilot run. The continuous pilot plant testwork proved a sulphide recovery of 93% could be achieved from the Zod ores, at a mass recovery of 9 – 10%. Gold recovery to the sulphide concentrate was 87%, at a silver recovery of 91%. The sulphide grade of the concentrate was in the range 16 – 18%.

The analysis of the blended pilot plant cleaner concentrate is presented in Table 1.

Table 1 – Cleaner concentrate XRD data.

Mineral	**Chemical Formula**	**Conc (%w/w)**
Unidentified		20.03
Arsenopyrite	$FeAsS$	6.78
Chalcopyrite	$CuFeS_2$	2.43
Clinochlore	$(Fe,Mg)_3Fe_3AlSi_3O_{10}(OH)_8$	1.58
Dolomite	$CaMg(CO_3)_2$	1.30
Galena	PbS	0.40
Magnesite	$MgCO_3$	2.80
Plagioclase (Albite)	$NaAlSi_3O_8$	4.50
Pyrite	FeS_2	36.63
Pyrrhotite	FeS	5.60
Quartz	SiO_2	3.90
Talc	$Mg_3Si_4O_{10}(OH)_2$	17.58

The flotation tailings contained 13% of the gold, and CIL testwork indicated a gold recovery of 60% could be achieved from the flotation tailings at modest reagent demand. The CIL plant at Ararat has a capacity well in excess of the 1 million tonnes per annum treatment rate for the project, and so co-treatment of both the oxidized flotation concentrate and the flotation tailings was incorporated in the design.

IsaMill signature plot testwork on the pilot plant composite concentrate sample returned a specific grinding energy of 59 kWh/t to grind the concentrate to the target 80 % passing size of 10 µm. The current plant is operating at a coarser grind of around 12µm with a specific energy of 45 kWh/t. To date the coarser grind has not impacted precious metals recovery.

Extensive testwork was carried out to determine the best oxidative leach pH for the finely ground concentrates. The testwork examined oxidative leaching under mildly acidic conditions for selective oxidation of the arsenopyrite minerals, and leaching at a more neutral pH. All tests were carried out under atmospheric pressure, with oxygen gas as the oxidant.

Leaching at near neutral pH was ultimately chosen for the oxidative leach. Leaching at near neutral pH allowed lower cost materials of construction to be used in the leaching circuit, and resulted in a final residue with more stable arsenic phases when tested in accordance with the USEPA TCLP protocol. Cyanide and lime demands were lowest for the residue generated under near neutral pH, and the gold and silver recoveries were higher.

The two major oxidative leach reactions observed under the near neutral oxidative leaching conditions were:

Pyrite: $FeS_2 + 15/4O_2 + 9/2H_2O + 2CaCO_3 = FeO.OH + 2CaSO_4.2H_2O + 2CO_2$

Arsenopyrite: $FeAsS + 7/2O_2 + 2H_2O + CaCO_3 = FeAsO_4 + CaSO_4.2H_2O + CO_2$

Confirmatory bench scale oxidative leaching testwork was then carried out on flotation concentrates from the four main orebodies at the Zod deposit. Economic modeling work that compared the capital and operating costs for the Albion Process plant at varying levels of sulphide oxidation was carried out using the batch leach test results. The modeling work showed that a sulphide oxidation of 70% returned the highest Net Present Value for the project, at a hurdle rate of 10%. The gold recovery at this level of oxidation was 93%.

Continuous pilot plant oxidative leaching testwork was then carried out on a blended concentrate. The continuous pilot testwork confirmed a sulphide oxidation of 70% was required to achieve an average gold recovery from the blended feed of 93%. The average silver recovery was 80%. A design oxidation target of 75% was taken forward into detailed design of the oxidative leach circuit. The oxygen demand for the concentrate to achieve the design oxidation of 75% was 336 kg/tonne, and the limestone demand was 326 kg/tonne. Mass and heat balance modeling indicated an average operating temperature in the oxidative leach circuit of 96^{o}C.

PROCESS PLANT DESCRIPTION

Plant General

The Ararat Process Plant experiences a hot, dry summers and cold winters, with an absolute maximum temperature of +42°C; and a minimum of -30°C. The region is classified as semi-desert with an average rainfall of 238 mm. The maximum 10 days depth of snow mantle is 35 cm, at a design snow load pressure of 70 kg/m^2. The area is seismically active, and the plant is designed to survive a magnitude 7.2 earthquake. A flow sheet for the GPM Gold Project is shown in Figure 1.

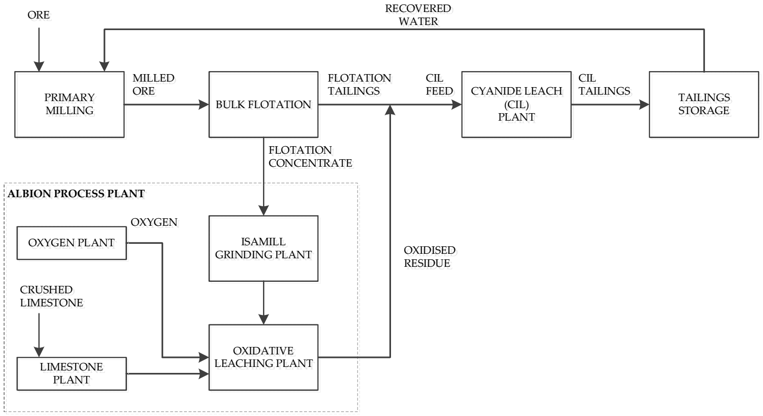


Figure 1 – GPM Gold Project overall flow sheet.

Comminution and Flotation

Ore is mined at the Zod Mine by open cut methods and delivered to a run of mine stockpile for blending ahead of crushing. Crushed ore is trucked to a stockpile at a rail siding and loaded into rail cars for transport to the Ararat Plant. Ore is recovered at the plant by tippler, into ore storage bins. Ore is recovered from the bins by apron feeder and conveyed to the comminution circuit.

The comminution circuit consists of two parallel milling lines. Each line has a primary 1600 kW Autogenous Grinding mill, operating in closed circuit with spiral classifiers, followed by two secondary 630 kW Ball Mills, each operating in closed circuit with cyclones. The Ball Mills are configured in parallel. Cyclone overflow from the Ball Mills are directed to the flotation feed tank.

The flotation circuit had not been operated for over 10 years and refurbishment works were part of the overall project. The existing flotation plant equipment consisted of two banks with five cells in each, and two banks with four cells in each. Each cell is fitted with dual 45 kW agitators, with a cell volume of approximately 32 m^3. The detailed design, procurement and installation was conducted by GPM and their nominated engineering company separate from Glencore Technology.

Slurry is fed to an agitated 25 m^3 flotation feed tank which overflows to the first cell for conditioning with frother. Conditioned slurry then gravitates to Pre-flotation cells, which will be used for pre-flotation of talc and carbonaceous slimes. The pre-flotation concentrate is gravitated to the tailings pumpbox and then to the 15 m diameter Flotation Tailings Thickener.

Tailings slurry from the pre-flotation stage gravitates to a Rougher Conditioner cell, which is used for conditioning of the slurry with copper sulfate and collector prior to rougher flotation. Conditioned slurry flows to the Rougher/Scavenger cells to produce a rougher concentrate for cleaning. The tailings slurry from Rougher/Scavenger flotation is transferred to the Flotation Tailings Thickener, with the thickened underflow slurry pumped to the CIL circuit.

Rougher concentrate will be transferred to the Cleaner Flotation Bank. The two cleaner banks operate in series with both cleaner concentrates combined as the final concentrate. The final Cleaner Concentrate is pumped to the 10 m diameter Flotation Concentrate Thickener, with the thickened underflow concentrate pumped to the IsaMill circuit for fine grinding. Cleaner tails are recycled to the scavenger circuit.

Albion Process

The Albion Process is a combination of ultrafine grinding and oxidative leaching at atmospheric pressure. The Albion Process™ technology was developed in 1994 by Glencore Technology and is patented worldwide. There are five Albion Process plants currently in operation.

The first stage of the Albion Process is fine grinding of the concentrate. Most sulphide minerals cannot be efficiently leached under normal atmospheric pressure conditions. The process of ultrafine grinding results in a high degree of strain being introduced into the sulphide mineral lattice. As a result, the number of grain boundary fractures and lattice defects in the mineral increases by several orders of magnitude, relative to un-ground minerals. The introduction of strain lowers the activation energy for the oxidation of the sulphides, and enables leaching under atmospheric conditions. The rate of leaching is also enhanced, due to the increased mineral surface area.

Fine grinding also prevents passivation of the leaching mineral by products of the leach reaction. Passivation occurs when leach products, such as iron oxides and/or elemental sulfur, precipitate on the surface of the leaching mineral. These precipitates passivate the mineral by preventing the access of oxidants to the mineral surface.

After the concentrate has been finely ground, the slurry is then leached in agitated vessels with oxygen to oxidize the sulphide minerals. Leaching is carried out under atmospheric pressure, and autothermally. Excess heat generated from the oxidation process is removed through humidification of the vessel off gases.

The average nominal throughput for the Albion Process Plant is 94,007 t/a of cleaner concentrate, with a design factor of 15 % applied to the average rate to achieve a design rate of 108,108 t/a. The average gold and silver throughput is 127,000 and 131,000 ounces per annum, respectively.

The feed rate to the IsaMill Fine Grinding circuit is 12.1 t/h, with a design feed rate of 13.9 t/h of concentrate and a final 80% passing size of 11 µm. Since commissioning, the IsaMill energy demand to achieve this grind size is typically 800 kW. The available drawn power for an M3000 IsaMill is 1,120 kW and this mill was chosen for the ultrafine grinding circuit. The completed IsaMill circuit at Ararat is shown in Figure 2. The current grind size is typically 80% passing size of 11 – 12 µm with a specific energy of 45 kWh/t which results is above target downstream gold recovery.

Finely ground slurry is then pumped to an agitated ground concentrate storage tank. The oxidative leach circuit consists of nine 240 m^3 Albion Leach Reactors, each with a live height of 9.4 meters and a diameter of 5.4 meters. Each reactor is agitated by a 160 kW dual impeller agitator, with oxygen delivered by a bank six HyperSparge oxygen injection lances in each reactor. The HyperSparge units are shown in Figure 3.

The slurry pH is maintained at 5.0 – 5.5 in each reactor by limestone slurry dosing.

Figure 2 – IsaMill grinding circuit.

Figure 3 – HyperSparge units.

The design rate of sulphide oxidation within the oxidative leach is 1800 kg/h. Under the near neutral pH conditions employed in the oxidative leach, sulfate is the reaction product of sulphide oxidation, with a design oxygen requirement of 3750 kg/h. The Albion Leach Reactors have all been designed to achieve an oxygen transfer rate of 4700 kg/h. The design oxygen capture efficiency in the leach train was 80 %. Site survey data collected to date suggests that the oxygen capture efficiency currently being achieved exceeds 90%.

The oxygen mass transfer rate for the oxidation of the sulphide minerals is defined by the following equation (Shuler and Kargi, 2002):

$$\text{Oxygen Transfer Rate} = K_L a\,(C_{sat} - C) \qquad (1)$$

where:

K_L = liquid film mass transfer coefficient for oxygen into the slurry, in units of $m.s^{-1}$

a = the specific gas surface area, in units of $m^2.m^{-3} = m^{-1}$

C_{sat} = the solubility of oxygen in the slurry at saturation, in units of $g.m^{-3}$

C = the steady state oxygen level in the slurry, in units of $g.m^{-3}$

The "K_L" and "a" terms are typically combined in the form of a mass transfer coefficient for the system. The design K_La for the Albion Leach Reactors is 0.12 s^{-1}. Oxygen gas has poor solubility in water, and so mechanical devices such as agitators and spargers are required to assist the mass transfer. In the Albion Leach Reactor, oxygen gas is sparged into the vessel using the HyperSparge supersonic gas injection lances. The HyperSparge oxygen injection system achieves very high oxygen mass transfer rates at the interface between the supersonic gas jet and the impinging slurry, reducing the amount of power required from the agitation system.

The agitator drawn power required to achieve the design mass transfer coefficient was determined using an empirical correlation of the form (Nielsen and Villadsen, 1994):

$$K_L a = A * U_s^{\alpha} * (P_g/(p_{SL}\ V))^{\beta} \qquad (2)$$

where:

A = a constant specific to the ionic strength of the leach solution

U_s = the gas superficial velocity in the reactor, in units of $m.s^{-1}$

P_g = the agitator drawn power under gassed conditions, in units of Watts

p_{SL} = the density of the slurry, in units of $kg.m^{-3}$

V = the volume of the slurry, in units of m^{-3}

α, β = dimensionless empirical constants

The A, α and β parameters used for sizing the agitator were determined based on over 900 laboratory and pilot mass transfer tests. This correlation has been used successfully in the scale up of all operating Albion Process plants to date. A drawn power requirement of 120 kW per Albion Leach Reactor was determined using the correlation.

The residence time for the oxidative leaching circuit was designed based on the specific rate constant for pyrite leaching measured in the batch and continuous leaching testwork. Pyrite oxidation under near neutral pH conditions is first order (Singer and Stumm, 1970), allowing a simple scale up. The residence time scale up was based on the method of Henein and Beigler (Henein & Beigler, 1988). A design residence time of 40 hours was calculated for the oxidative leach circuit.

Each Albion Leach Reactor was fabricated from lean duplex alloy steel having a diameter of 5460 mm and a live height in the range 9100 – 8100 mm. The Albion Leach Reactors were supplied in modular sections for rapid assembly on site. Each Reactor was constructed from 15 panels, each with a height of approximately 2.0 m and an arc length of 5.9 m. These panels were all fabricated off site and imported to the plant site in shipping containers. Baffles, slurry risers, leach tank lids, agitator support platforms and off gas stacks were all provided as part of the modular Glencore Technology equipment supply. Assembly of the oxidative leach train was rapid, with all nine leach reactors and two slurry storage tanks complete within 8 weeks. The final two tanks were erected in approximately three days each. The completed oxidative leach train is shown in Figure 4.

Overflow slurry from the oxidative leaching circuit will gravitate via a slurry sampler to a 10 m diameter thickener and be thickened to 45 %w/w solids prior to transfer to the CIL circuit. Thickener overflow is returned to the leach circuit to compensate for evaporative losses in a density control loop.

Figure 4 – Oxidative leach circuit.

A limestone plant with a capacity of 6 t/h was installed to generate limestone slurry for neutralizing duty. Limestone for the oxidative leach will be milled to an 80 % passing size of 75 microns in a 132 kW overflow ball mill operating in closed circuit with cyclones. Cyclone overflow will report to a 150 m^3 agitated distribution tank and be circulated through the oxidative leach train by a ring main. Individual dosing lines will add limestone slurry to each Albion Leach Reactor. The limestone distribution tank was a 150 m^3 ZipaTank zip join tank – the first of its kind the in the world. The tank was erected in approximately 5 days and was internally sealed with specially selected paint. The joins sealed on the first filling. The limestone distribution tank is shown in Figure 5.

Figure 5 – Limestone storage tank.

Two 60 t/d VPSA oxygen plants will operate in parallel to provide oxygen to the Albion Process Plant. Oxygen will be delivered from each plant at a maximum flowrate of 1,745 Nm^3/h, at a purity of 93% v/v.

The thickened oxidative leach residue and thickened flotation tailings will report to a 100 m^3 mixing tank and be blended prior to feed to the CIL plant. The CIL Plant will process 137.5 t/h of feed comprised of oxidized residue and flotation tailings. All six existing CIL tanks will be utilized, providing a total residence time in the CIL circuit of 41 hours. The CIL Plant is expected to consume

5.3 kg/t of sodium cyanide and 10 kg/t of lime. Carbon levels in the CIL Plant will be 10 – 15 kg/m^3, with a design carbon loading of 2,500 g/t. Carbon movements will total 7.5 t/d, and the existing dual AARL elution circuits will be used for carbon processing.

CIL Plant tailing gravitates to a cyanide destruction plant prior to being pumped to tailings. The tailings will be deposited within the existing tailings impoundment, approximately 6 km from the Ararat plant site.

Plant control is achieved through a Distributed Control System (DCS) located in a centralized control room between the concentrator and Albion Process plants. Training for field and control room operators was provided by GT and sub-contractors as part of the commissioning process. The control room is shown in Figure 6.

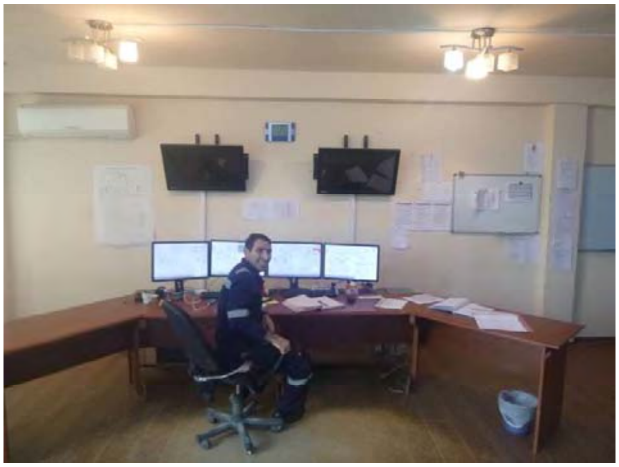


Figure 6 – Central control room for concentrator and Albion Process^{TM} plant.

PROJECT STATUS AND PLANT PERFORMANCE

The Albion Process plant was provided to the GPM Gold Project as a Lump Sum technology package by GT. The package included all detailed design, mechanical equipment, electrical, instrumentation and control equipment, structural steel, flooring, handrails, piping and valves. The scope of supply includes the fine grinding plant, oxidative leaching and thickening plant and the supporting limestone, oxygen, flocculent and caustic reagent plants.

Mechanical design was completed in December 2012, with the majority of mechanical equipment and fabricated components delivered to site by May, 2013. Site civil works were completed in March 2013. Construction was completed in April 2014.

The pre-commissioning phase was conducted during April/May 2014 with wet commissioning commencing in May/June 2014 and completed in August 2014. The commissioning was managed and coordinated by Glencore Technology. The commissioning team comprised three permanent Glencore Technology personnel (manager/process engineer, mechanical engineer and instrumentation engineer) supported by equipment specialists brought to site during crucial commissioning events. The GPM site team provided all other support.

The main setback during commissioning was the failure of an oxygen plant blower which had a lead time of 12 months to replace. Dual oxygen plants were supplied to the project, each with the capacity to oxidise 70% of the design sulphide feed, and so the blower failure has not impacted on plant throughput to date. The blower will be repaired and the second oxygen plant will be in service by May 2015.

The main setback for ramp-up has been the lack of feed quantity and quality from the refurbished concentrator. A project is in place to improve concentrator performance with expected results by October 2015.

As of March 2015 the refurbished concentrator was running at around 60% capacity with recent assistance from GT during March increasing throughput by 30%. The Albion Process plant performance has not been impacted by the slow ramp up of the concentrator, with the plant regularly achieving 95% gold recovery with around 50% sulfur oxidation.

A survey of the nine Albion Leach Reactors was collected to determine tank-by-tank sulfur oxidation and resulting gold extraction. The sulfur oxidation was determined using a Leco sulfur analyzer. Gold extraction was determined by subjecting each collected sample to a bottle roll test at the GPM laboratory and cross checked with an agitated CIL test at hrltesting laboratory in Brisbane. The sulfur oxidation against gold recovery from the plant and compared to pilot plant results is shown in Figure 7.

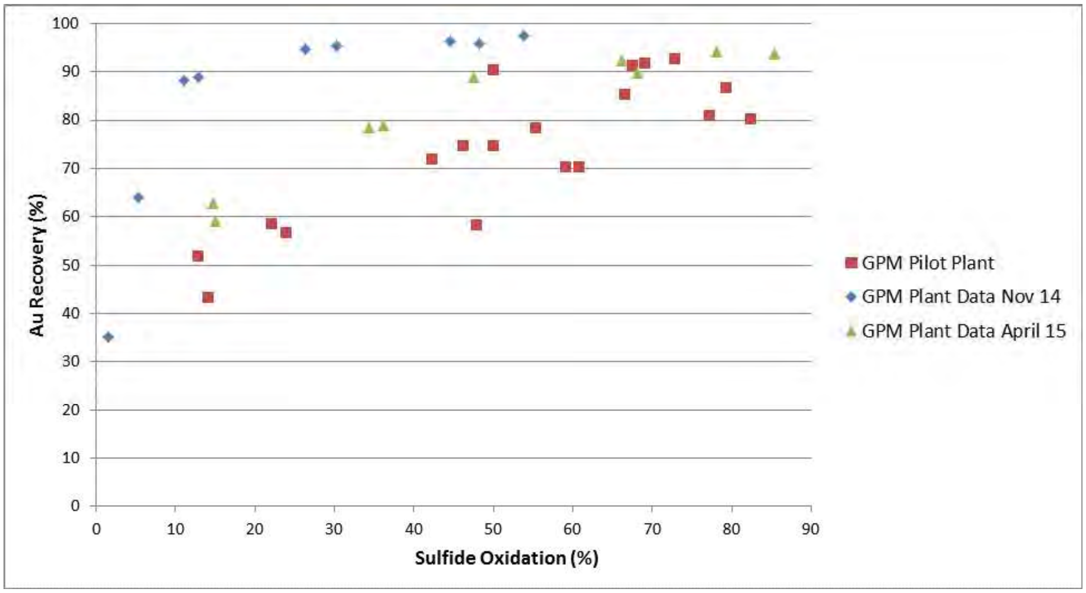


Figure 7 – Sulfide oxidation against gold recovery comparing pilot plant data and actual operating data.

Testwork done on individual components of the pilot plant feed are consistent with this level of oxidation and corresponding gold recovery.

A profile of sulfur oxidation and gold recovery was collected down the nine leach reactors. The data are presented in Figure 8.

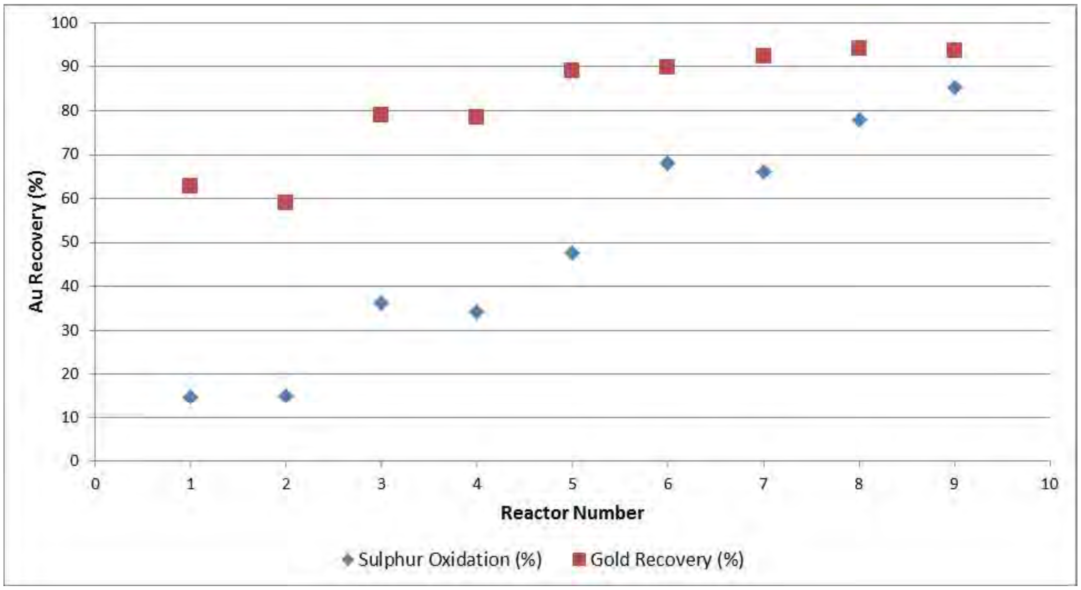


Figure 8 – Profile of sulfide oxidation and gold recovery down the leaching train. Plant is on reduced rates.

Figure 8 shows the profile of the sulphide oxidation and gold recovery down the leach train. Although one oxygen plant is operating, there is sufficient oxygen available to increase the throughput rate if feed was available from the concentrator.

The current oxygen consumption is operating at around 215 kg / t concentrate which is below the design value of 336 kg / t owing to higher than design oxygen utilization and lower oxidation levels for this feed material compared to the design case.

Current limestone consumption is very low owing to higher entrainment of acid consuming gangue with the flotation concentrate. Steady feed composition at full production rates will allow limestone consumption to be better analysed.

Cyanide consumption in the CIL plant is within the range expected during pilot testwork. The plant is currently operating at 1.8 to 2.2 kg sodium cyanide per tonne of feed to the CIL plant which is the combination of leach residue and flotation feed.

CONCLUSIONS

The GeoProMining Albion Process^{TM} plant was commissioned successfully over a 14 week period. The plant is achieving greater than 95% gold recovery in the cyanide leaching plant, consuming 1.8 – 2.2 kg cyanide / t combined leach residue and flotation tails.

The plant is running on reduced rates due to concentrate feed availability, as mine development has been slower than planned. GT continues to work with GPM to improve the concentrator performance. At the end of April 2015, GPM achieved a 30% increase in concentrator throughput with the assistance of GT. The second oxygen plant commissioning is scheduled for June 2015, and full plant throughput should be achieved by July 2015. An update will be provided after the plant is at full capacity with more comprehensive performance data including oxygen and limestone consumption as well as sulfur oxidation, oxygen utilization and resulting gold recovery.

ACKNOWLEDGEMENTS

GT wish to thank GeoProMining Gold LLC for the permission to publish this paper. This paper was first published in proceedings of ALTA Conference in 2015, Perth, Australia.

REFERENCES

Henein H, & Biegler, L, 1988. Optimization Of Reactor Volumes For Topochemical Reactions. *Trans. Inst. Min. Metall.*, 97:215-223.

Hourn, M, and Turner, D, 2012. Commercialisation of the Albion Process, in *Proceedings ALTA Gold Conference 2012*, pp 231-248 (ALTA Metallurgical Services: Melbourne)

Hourn, M, Voigt, P, and Turner, D, 2014. Development of the Albion Process plant to treat refractory concentrates from the GPM Gold Project, in *Proceedings Hydroprocess Conference 2014*, (Gecamin; Santiago)

Konstantinov, M and Grushin, V, 1970. Geologic Position of the Zod-Agduzdag Gold-ore Node in Transcaucasia. *International Geology Review*, 12(12):1447-1453.

Morrell, S, 2010. Estimates of the Throughput of the Ararat Comminution Circuit. Personal communication. June.

Nielsen, J and Villadsen, J, 1994. *Bioreaction Engineering Principles*, pp 27-32 (Plenum Press: New York).

Rohner, R and Andreatidis, J, 2010. Bankable Feasibility Report, Geoprominmg Gold LLC, Zod Upgrade Project Volume 3, Metallurgy And Processing. Personal communication. June.

Shuler, M and Kargi, F, 2002. *Bioprocess Engineering: Basic Concepts*, 2nd ed. pp171 (Prentice-Hall: New Jersey).

Singer, P and Stumm, W, 1970. Acidic Mine Drainage: The Rate Determining Step. Science, 167(3921):1121-1123.

Premium coal fuels with advanced coal beneficiation

Louis J Wibberley[1] and Dave Osborne[2]
[1]Leader DICE Development Program
CSIRO Energy Technology, Mayfield West 2304, Australia
Email: louis.wibberley@csiro.au
[2] Coal Technology Consultant and Managing Director,
Somerset International Australia, Brisbane, 4000, Australia
Email: dosborne@somersetpty.com

Abstract
Research and pilot plant trials over the last 5 years have shown that a range of premium quality coal fuels, with ash contents as low as 1%, can be produced economically from a wide range of coals (including coal tailings). This capability can improve both the economic and environmental performance of coal through increased grade recovery, recovery of saleable coal from existing tailings emplacements, and new higher value products such as a fuel oil replacement in boilers, enhanced coking blends, and for higher efficiency power generation using the direct injection carbon engine (DICE). While a conventional coal preparation plant can be used to produce suitable ultra-low ash coal for these fuels, changes to both the philosophy of coal preparation and operating strategy are needed for achieving the best results. Further improvements are possible via the application of the latest milling and dewatering technologies. These fuels also require a rethink of the supply logistics. The paper discusses findings from both laboratory and pilot scale trials in Australia, in the context of new export products for DICE and boiler fuels.

Key words: ultra-low ash coal, coal grain analysis, coal slurry fuel, advanced beneficiation, DICE, high efficiency engine

INTRODUCTION

Producing very low ash coal products has for many years been a challenge for coal treatment specialists and researchers. Much of the early work in the 1970s and 1980s used chemical cleaning processes involving leaching with acidic or caustic solutions, most of which proved uneconomic when scaled up to commercial application. The most successful of these include the AMAX 2-stage leach process developed with US-DOE funds in the mid-1980s[1] and the Australian UCC process[2].

The use of physical cleaning to ultra-low ash levels, using more conventional coal processing technologies, has proven more challenging. Old school thinking typically regards ash contents below 2-3%, as both technically and economically unviable. This is due to several factors/misnomers:

- The so-called "inherent ash" content of coal is usually regarded as the lowest ash possible
- Lower ash contents require finer grinding to increase liberation which is costly and produces a product for which there are few markets
- Flotation of ultra-fine coal can be problematic, and requires high reagent consumption

---

[1] Berggren, M. H., Chemical coal cleaning and cost refinement for coal-water slurry manufacture. Report DOE/MC/20700-T1, February 1985.

[2] http://www.yancoal.com.au/page/key-assets/technology/UCC

- Fine coal products are inevitably high moisture (> 35%) which means costly dewatering and/or drying to produce saleable products
- Ultra low ash coal products are uneconomic to produce

All of these factors are incorrect, or at best very misleading, as recent research/pilot plant tests have shown.

Fine coal cleaning is generally regarded as being the treatment of coal sized below 0.5mm (500µm) but the challenges have tended to descend further down the size distribution to 0.1mm (100µm) and lower. Most recently, with the widespread use of micro-flotation using the Jameson and Microcell technologies, the elusive size fraction is now <0.05mm (50µm) and this is probably where the current research and development focus lies. Such a size range usually means that coal particles are extensively liberated from the mineral content and are recoverable, providing that the concentrate can be recovered and dewatered to a commercially acceptable level.

Not surprisingly, much of the research and development that has occurred in ultra-fine coal cleaning has taken place for higher value products, often metallurgical, and almost all techniques for both cleaning and dewatering have undergone some form of improvement or development during the past decade in this context. These improvements have been in all areas - optimised design, improved materials of construction and improved wear resistance[3], process control and monitoring, integrated automation and sampling/analysis.

This paper describes recent work carried out at laboratory and pilot scale in Australia, that has successfully produced coal concentrates with <2% ash content from a wide range of coals and tailings, using a combination of fine grinding and froth flotation. The potential to achieve even lower ash levels is also discussed. The main objective of this work has been to produce a stable, coal-slurry fuel (i.e. micronised refined coal or MRC) for use in DICE, an application which demands the lowest possible ash content in the slurry fuel (preferrably <1%) and very fine particle size distribution (typically d_{80} of 20-40µm). This is much finer than the conventional coal water slurry fuel (CWSF) for boilers or gasifiers.

CHEMICALLY CLEANED COAL

Ultra-clean coal can be produced by chemically cleaning coal, and a 2- and 3-stage caustic leach process has been successfully demonstrated in Australia by UCC Energy Pty Ltd. A significant feature of the process is that the coal needs only to be coarsely milled (typically to -2mm) to provide sufficient contacting: all other processes for producing ultra-low ash coals require a higher degree of milling. The main process consumables are sulphuric acid, and lime; caustic soda is regenerated in the process. The waste materials are environmentally benign (gypsum and calcium alumina silicates) and the final ultra-clean coal product has 0.5-0.7% ash (with less than 0.2% possible for some coals). UCC Energy Pty Ltd (a wholly owned subsidiary Yancoal Australia Pty Ltd) owns and has operated a pilot plant at Cessnock in the Hunter Valley, New South Wales. The nominal 350kg/h facility (see Figure 1) was commissioned in 2002[4]. The main focus has been in the production of low ash coal fuels for gas turbines and DICE, as well as for electrode carbon and other high value applications – several using briquettes.

---

[3] Osborne, D. G.; 2012. Milestones in Fine Coal Development, *Challenges in Fine Coal Processing, Dewatering and Disposal*, editors M.S. Klima, B. J. Arnold and P.J. Bethell, SME, TN816.C375 2012, pp3-32

[4] http://www.uccenergy.com.au/the-process/

Fig 1 Ultra-Clean Coal pilot-plant, Cessnock, NSW (courtesy UCC Energy Pty Ltd)

Fuel from this plant has been used for longer duration tests in an adapted diesel engine (*ie* DICE) at the CSIRO, and scale up of the process is now under consideration by Yancoal and partners.

PHYSICAL CLEANING

Although chemical cleaning has been proven to produce very low ash coal products for the highest value applications (e.g. electrode carbon and MRC for DICE – especially smaller engines), the process is more costly than physical cleaning, is less tolerant of coal rank, and requires a coarser lower ash feed coal for optimum results. Physical cleaning is therefore preferred for the production of coals with ash contents <2% from a variety of coals and tailings, and is the focus of the present paper.

Liberation – removing the "inherent ash" constraint

Efficient analysis of the occurrence of mineral matter within the feed coal is key to economically producing premium coal products by physical cleaning. CSIRO in Australia has developed an optical reflected light microscopy system for assessing coal petrography samples. This system collects and creates mosaic images so that quantitative information can be obtained on individual coal/mineral grains. Size and compositional information (the amount of vitrinite, inertinite, liptinite and mineral type) is obtained for each particle, and this information can then also be used to estimate the density and ash value of each particle (both macerals and minerals) and that of the sample.

Although coal grain analysis (CGA) is generally conducted on a small representative sub-sample of -1 mm material, the technique has also been successfully used to characterise particles up to 4 mm in size (a size which perhaps report to a middlings circuit), and for samples which have been micronised to give particle size of less than 20 µm (e.g. for producing ultra-low ash concentrate by flotation).

For the ultra-fine material, CGA is particularly useful to assess the degree of liberation and therefore assist in optimising the beneficiation process to produce the lowest possible ash products (for uses such as DICE, or even direct carbon fuel cells[5]).

---

[5] O'Brien G., Firth B., Adair B. The Application of the Coal Grain Analysis Method to Coal Liberation Studies, 2011, *International Journal of Coal Preparation and Utilization 31*: pp 96–111

For advanced beneficiation, CGA enables liberation to be assessed and the expected yield at different target ash levels to be estimated. As the CGA information also provides detail on the intrinsic and entrained minerals in a sample, it benchmarks the lowest ash value which could be obtained if all entrained minerals are removed from the sample at that top-size.

An example of its application in the production of MRC for DICE is given in Figure 2 below. The characterised CGA images shown illustrate the extent of liberation for raw tailings (-250µm) and the corresponding micronised concentrate with a d_{80} of 15µm. The results show that milling has enhanced significantly the particle liberation, and has produced essentially single component grains. This indicates that it is physically possible to reduce the intrinsic mineral being carried over to the product. This implies that an appropriate circuit of grinding and flotation operations should be capable of producing a very low ash product concentrate. In this case sufficient liberation has occurred to enable a product coal with less than 1% ash to be obtained, provided entrainment within the froth can be eliminated. Although it is anticipated that the fineness of the entrained mineral particles will make separation more difficult, there are a number of process optimisations available to achieve this (e.g. use of reagents which enhance the hydrophobicity of the coal and hence allow more aggressive application of wash water to be used, optimisation of the pull between flotation stages). Both have been shown to be effective in practice at the pilot scale.

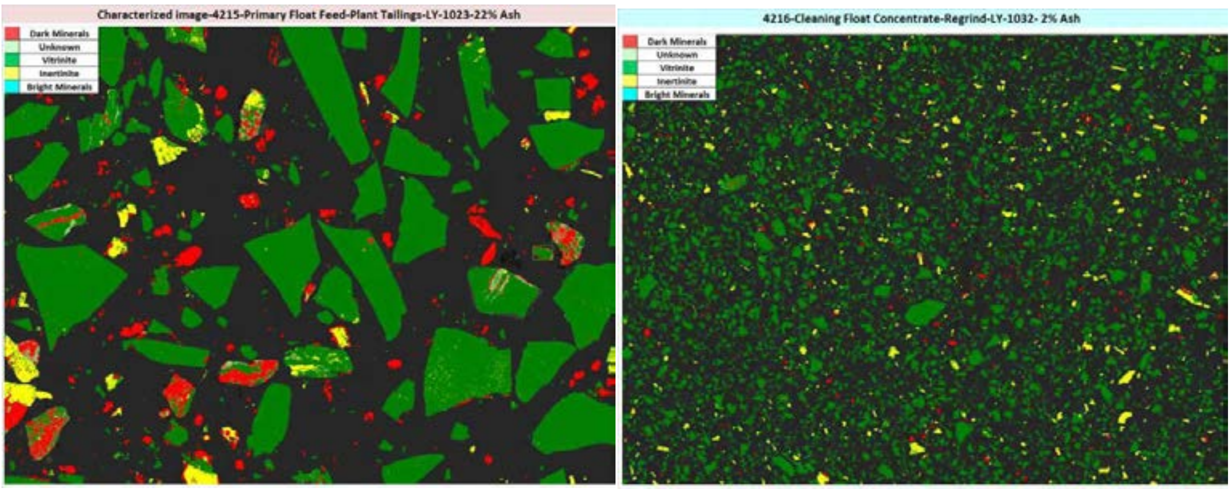


Fig 2 Characterised images for raw coal tailings feed and final concentrate (courtesy QCAT-CSIRO)

The results obtained from several CGA studies with a pilot plant, and other minerals tracking studies by the CSIRO with 9 Australian bituminous coals/tailings, have provided a valuable insight into the separation and recovery performance of the current beneficiation work, by tracking the selective recovery of individual grain types. In particular, this has enabled optimising the milling and flotation steps for processing raw tailings samples.

Overall, CGA has proved a valuable method for applying research in a commercial environment.

Milling/flotation process approach

In order to achieve lower ash levels, a liberation step needs to be introduced by fine grinding. Typically fine grinding for coal water slurry fuels is by ball mills; -for example, ball mills are commonly used in China where over 30Mt of slurry fuel is currently produced annually with ash levels below 4%. However, to enable even lower ash contents and maximum liberation (and with a reduced energy consumption), the present work has used bead mills (IsaMills) for both laboratory and pilot scale tests.

Using this mill, a 3 tonne/day pilot-plant, see Figure 3, owned and operated by Glencore Technology (formerly Xstrata) has been used for proving the feasibility of producing ultra-low ash coal from freshly generated raw coal tailings. The plant is sited at Bulga mine, a large thermal coal operation in the Hunter Valley. The pilot plant comprises two Jameson flotation cells, a small IsaMill and a membrane filter press, which can be configured to simulate a number of circuit designs.

Fig. 3 The 3 t/day coal tailings treatment pilot-plant (Courtesy Glencore)

A range of different coal feed types have been tested with raw coal ash contents ranging from ~ 30% feed ash (ad) from easily treated coal seams, to much higher ash coal seams containing difficult-to-separate colloidal clays. The plant is usually configured in a rougher-scavenger circuit where the tailings from the rougher (primary cell) feeds to the scavenger (secondary cell).

The cleaning efficiency of this circuit is clearly shown by the difference in colour of the two product streams in Figure 4 below; the black concentrate is high in vitrinite and the white tailings stream is predominantly kaolinite clay.

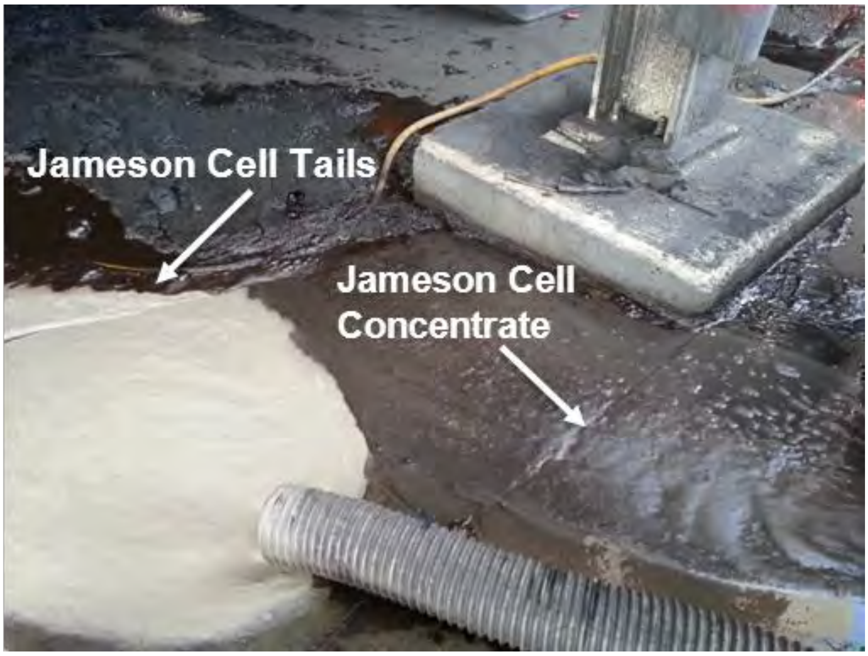


Fig. 4 Concentrate and tailings streams from pilot plant test-work (courtesy Glencore)

A secondary effect of cleaning is a marked improvement in dewatering: Fine mineral matter and clay in a coal concentrate significantly reduces filtration rate and increases product moisture. By removing much of this material, a significant reduction in product moisture can be achieved – essential for metallurgical and briquette products.

Cost effective and efficient dewatering is thus essential to producing premium coal products, for treating both the concentrate and barren tailings streams. Dewatering options available are listed in Table 1. Recent advances in "by zero" fines dewatering has enabled total moistures of flotation product to be reduced to a target whereby it has become commercially viable to include flotation concentrates into the final thermal coal product at a greater number of mine sites. Tailings dewatering still has some challenges, but high-G solid-bowl centrifuges offer promise for on-line dewatering.

Table 1 Equipment Used for Dewatering Coal (adapted from Table 13-1, *The coal handbook*[6])

Equipment	Footprint size	Throughput (t/h dry solids)	Product moisture (% w/w)	Feed preparation	Application
High frequency screen	0.6-2.4 x 3m	10-100 t/h	15-25	cyclone underflow	fine coal
Screen scroll centrifuge	0.5-1.5m dia.	45-100 t/h	11-18	cyclone underflow	fine coal
Horizontal vacuum belt filter	75-150m^2	50-130 t/h	20-30	flocculation	ultrafine coal
Screen bowl centrifuge	1.1m dia. x 3.3 m long	20-60 t/h	16-27	thickening	ultrafine coal
Solid bowl centrifuge	1.1 m dia. x 3.3 m long	15-20 t/h	15-20	thickening	ultrafine coal slimes

---

[6] Bickert, G. 2013. Solid-separation Technologies for Coal, Chapter 13, The coal handbook; Volume 1; Woodhead Publishing Series in Energy, pp 422 – 444, edited by D. G. Osborne, 2013

Equipment	Footprint size	Throughput (t/h dry solids)	Product moisture (% w/w)	Feed preparation	Application
Disc filter	120-200 m^2	50-150 t/h	20-32	thickening /flocculation	ultrafine coal
Hyperbaric disc filter	70-200 m^2	30-150 t/h	17-25	thickening /flocculation	ultrafine coal
Paste thickening	25 m dia x 6-12 m high	100 t/h	45-55	flocculation	barren tailings
Solid bowl centrifuge	1.1 m dia	20-60 t/h	30-45	thickening	barren tailings
Belt press filter	3-3.5 m wide	10-20 t/h	25-45	thickening /flocculation	barren tailings
Filter press	200-800 m^2	15-30 t/h	14-32	thickening	ultrafine coal & barren tailings

The economic argument for pursuing fines recovery in this application is now very compelling. Technology advances, particularly in dewatering, allow ultrafine product to be included in final product streams without penalising product quality (or introducing handling difficulties) as shown by the simplistic revenue scenario below.

Based on the following assumptions:

- Raw plant feed containing 10% passing 0.1 mm
- Thermal coal operation of 16 Mt/y ROM (1.6 Mt/y of raw feed currently sent to waste)
- Assuming a nominal 50% yield which equates to 0.8 Mt/y of potential saleable product
- At a benchmark price of US$85/product tonne (equivalent to US$69M revenue loss per annum, excluding freight, port, tonnage adjustments, etc.)

Including conservative capital and operating costs produces a reasonably attractive investment opportunity as given in Table 2 below.

Table 2 Economic evaluation of brownfields flotation installation

Capex $M	Operating $/t feed	Coal rate t/h	Direct costs $/t	Tax rate %	Discount rate %	NPV $M	IRR %	Payback years
50	15	230	30	30%	10%	78	51	2.5

Future direction – reducing product ash and moisture

The challenge of cost-effectively recovering a saleable fines component from tailings has been with us for many years, and periodically an apparent solution emerges. Glencore Technology has been operating the pilot plant described above for over 4 years, and this plant incorporates the combination of ultra-fine grinding by an Isamill and Jameson Cell flotation technology[7]. The inclusion of a fine grinding stage enables slurries to be prepared whereby mineral components are almost completely liberated from the carbonaceous material, thereby facilitating recovery of a highly concentrated ultrafine, low ash coal product, with ash levels well below traditional inherent ash limits – even from tailings.

This combination has already been proven capable of achieving very good combustible recoveries (material dependent, but normally over 90%) for coal derived from the raw tailings stream. The milling step produces a feed with a d_{80} of 15-30µm, enabling enhanced flotation recovery by a combination of increased liberation and the formation of fresh surfaces on the ultra-fine coal particles. This also reduces reagent consumption to levels significantly below traditional fine coal flotation.

Further enhancement by the addition of improved ultrafine dewatering of the flotation concentrate using a membrane filter press, or equivalent, and appropriate mixing system has resulted in the preparation of ultra-low ash, highly stable slurries with solids concentrations over 60% (w/w). This product should be very suitable for DICE, a potential large new market for a range of power generation markets:

- To replace fuel oil and natural gas,
- To provide highly efficient, highly flexible and modular coal-based power to backup increased renewables (giving an extremely low CO_2 and secure electricity system), and
- For incremental, low CO_2, coal-based generation capacity[8].

In the shorter term, slurries with a slightly coarser particle size distribution (typically with a d_{80} of 75µm) can be prepared in a similar way to create conventional coal-water slurry fuels, at over 70% solids, for direct firing into boilers as a replacement for heavy fuel oil (HFO) and for gasifiers.

Figure 5 below shows a nominal flowsheet based around an Isamill for micronising, and the Jameson Cell for separation. Adding the milling step can be optional for boiler grade product, but is essential for producing MRC for DICE. Various dewatering options, as described earlier, need to be carefully evaluated to achieve the desired solid-liquid outcome for each coal source/product combination. Alternatively, a briquetted product can be included in the normal product stream, thus avoiding slurry or fines related problems in product handling and transportation.

The integrated plant design thus has the functionality of the dual product offering, i.e. coal briquettes that can be added to the conventional product and/or coal-water slurry fuel for heavy fuel oil replacement for boilers, or for more innovative, value added applications such as DICE.

To facilitate logistics, and to enable early adoption of the technology, the concept of coal water slurry supply chains is being promoted to industry, whereby slurry fuels employ existing heavy fuel oil infrastructure to transport and store the fuel at the customer's facility. The use of these systems was been extensively demonstrated in Japan in the 1990s, and more recently in China.

---

[7] Mercuri, F., Osborne, D.G. and Young, M; 2014. The Future of Thermal Coal Flotation. Australian Coal Preparation Society conference 2014, Gold Coast, Qld.

[8] Wibberley, L.J. DICE – The best option for coal? Publication on the DICEnet web-site www.dice-net.org http://www.dice-net.org/images/pdf/Media/Papers

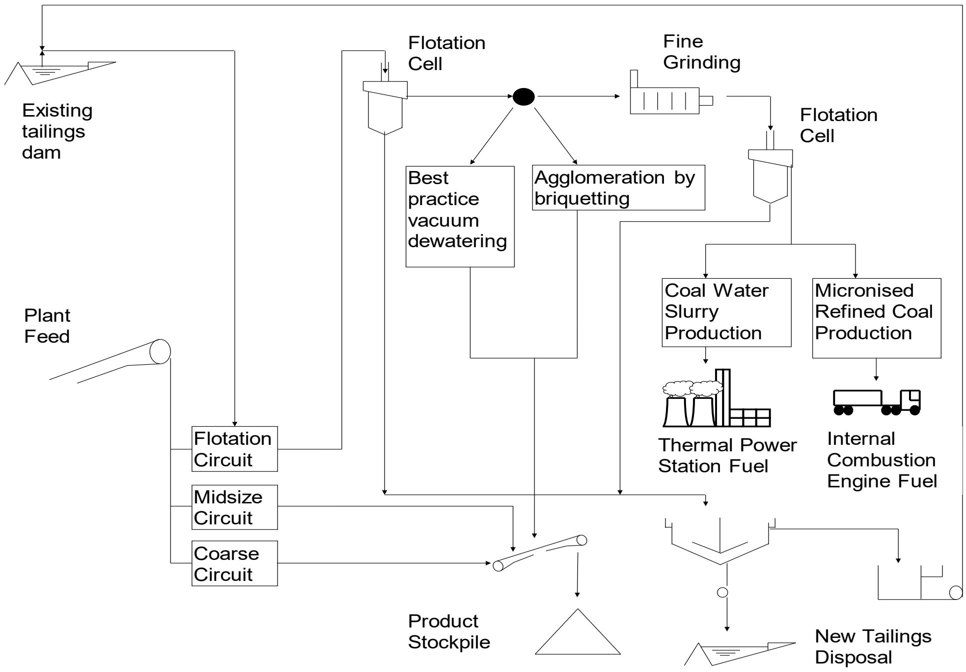


Fig. 5 Alternative fines treatment for thermal coals

Conclusions

For more than two decades the appeal of so-called "deep cleaning" of coal via liberation and subsequent beneficiation has been recognised in terms of the significant downstream improvements that would result - maximised resource recovery, minimised transport and handling costs, numerous end-user process improvements, reduced maintenance and wear, lower environmental impacts and sustainable improvements.

While the capability of ultrafine particle separation has matured via progressive improvements in froth flotation, the capability of dewatering the concentrate has continued to be the major barrier until recently. The emergence of larger capacity membrane filter presses, hyperbaric dewatering via decanter centrifuges or large disc filters now offer commercial solutions.

Binderless briquetting has also progressed to machines with capacities of up to ~40t/h for fine coal applications, and providing an alternative product for conventional transportation infrastructure.

The manufacture of stable coal-water slurries with over 65% solids content, and MRC slurries with over 60% solids content, have also reached commercial adoption; such products can also be produced from tailings.

With these technical barriers now overcome, the scene is set for these new applications to progress to become the new generation of clean coal technologies for a wide range of applications, with key economic drivers being a much higher fuel cycle efficiency (i.e. lower carbon intensity), highest grade recovery, and lowest solid waste disposal.

References

1. Osborne, D. G.; 2012. Milestones in Fine Coal Development, *Challenges in Fine Coal Processing, Dewatering and Disposal*, editors M.S. Klima, B. J. Arnold and P.J. Bethell, SME, TN816.C375 2012, pp3-32.

2. O'Brien G., Firth B., Adair B. The Application of the Coal Grain Analysis Method to Coal Liberation Studies, 2011, *International Journal of Coal Preparation and Utilization 31*: pp 96–111.
3. O'Brien G., Firth B., Warren, K and Hapugoda, P; 2015. Principles of Coal Selection for Advanced Beneficiation. 2015 Clearwater Clean Coal Conference; *40th International Technical Conference on Clean Coal & Fuel Systems*
4. Mercuri, R., Osborne, D.G. and Young, M; 2014. The Future of Thermal Coal Flotation. *15th Australian Coal Preparation Society Conference*, Gold Coast. QLD.
5. Bickert, G; 2013. Solid-separation Technologies for Coal, Chapter 13, The coal handbook; Volume 1; *Woodhead Publishing Series in Energy*, pp 422 – 444
6. Wibberley, L.J. 2014: DICE – The best option for coal? Publication on the DICE-net web-site http://www.dice-net.org/images/pdf/Media/Papers

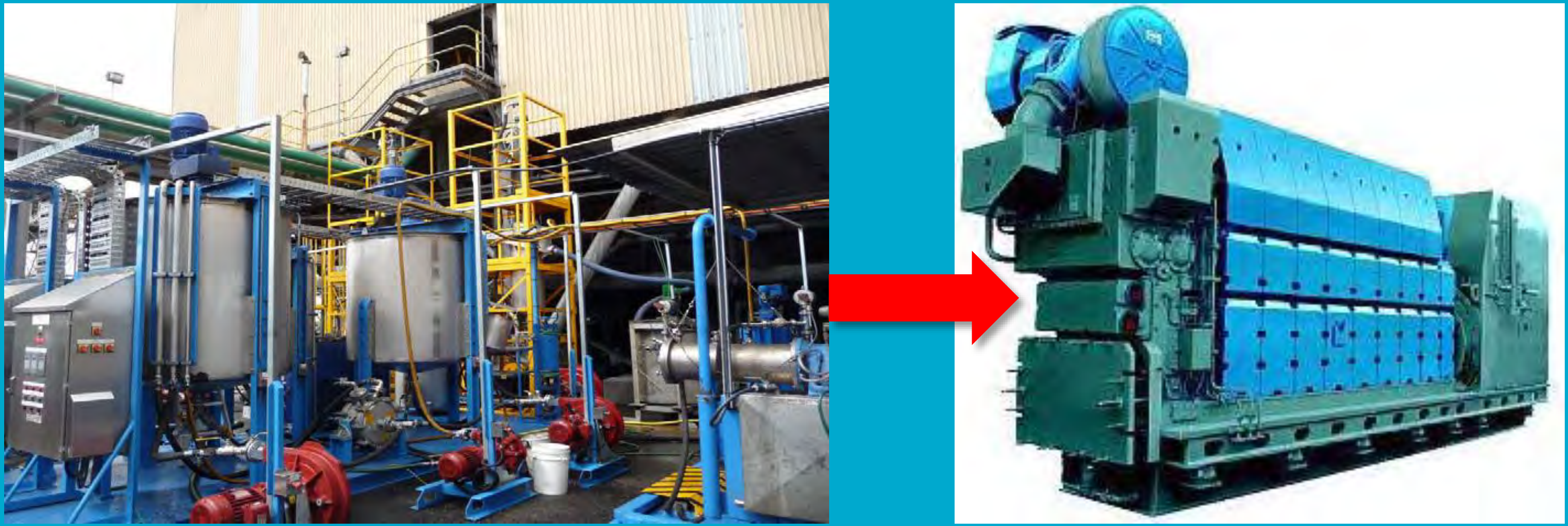
Acknowledgements

The authors acknowledge the contributions provided by Glencore Technology, Glencore Coal, Yancoal and the CSIRO towards the preparation of this paper and for permission to include figures and data from other recently published papers and articles on this subject.

Dr Osborne thanks Somerset Coal International for encouragement and support in attending and participating in the 2015 Clearwater Clean Coal conference.

Premium coal fuels with advanced coal beneficiation

Louis Wibberley - CSIRO
Dave Osborne – Somerset International



Premium coal fuels with advanced coal beneficiation

Contents

Introduction

- Greater resource recoveries are being sought by mine operators to maximize investment returns.
- Current industry trend - “by zero” recovery of fine coal to maximize resource yield and minimize environmental footprint.
- Fine coal size fraction faces the greatest barriers towards qualification as a product component.
- However, advances in liberation, flotation and dewatering create new opportunities for thermal coal operations. So…………

What are the historical barriers and how are they being overcome?

Why have advanced processing options now become viable?

Ultra-fine Coal Beneficiation

Two distinct approaches:

1. Chemical cleaning – coal structure is changed via chemical decomposition – potential is <0.2% ash residue.
2. Physical cleaning – coal structure not changed, but comminution may be applied for liberation – potential is <1% ash residue.

“Old school” thinking typically regards ash contents below 2-3%, as both technically and economically unviable because:

- “Inherent ash "of coal is usually regarded as the lowest achievable ash content
- Lower ash requires milling to finer particle size to increase liberation
- Flotation of ultra-fine coal can be problematic often requiring higher reagent dosages
- Fine coal concentrates are inevitably high in moisture (> 35%) which means costly dewatering and/or drying to produce saleable products.

Chemically Cleaned Coal

- Caustic leach process has been successfully demonstrated in Australia.
- Similar to the well-proven Bayer alumina process and also the AMAX 2-stage leach process developed with US-DOE funds in the mid-1980s.
- Ultra-low ash residue <0.2%
- Uses include slurry fuel or briquettes
- Costly option difficult to justify in current climate.

Ultra-clean Coal pilot-plant, Cessnock, NSW (Courtesy UCC Energy Pty Ltd.)

Liberation - Inherent ash constraint

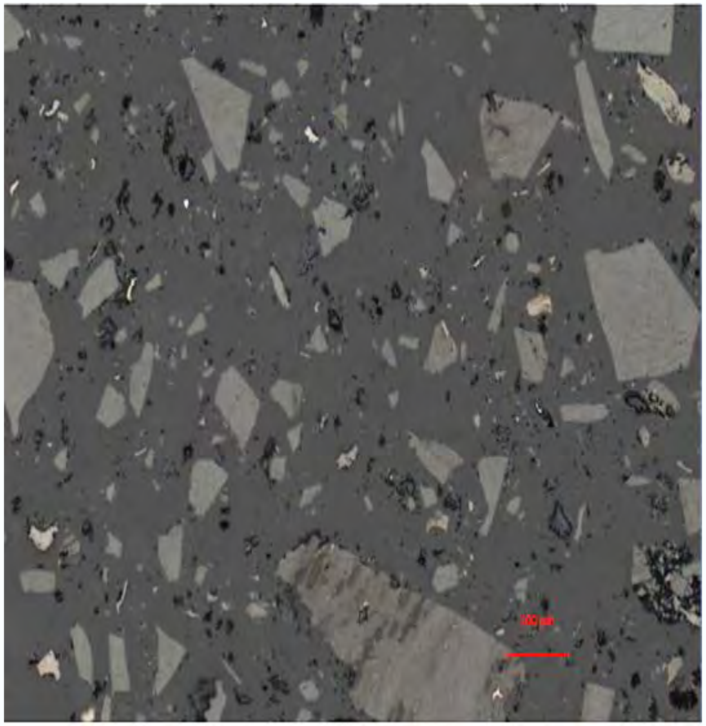
CSIRO has developed an optical reflected light microscopy system for assessing coal petrography samples.

System collects and creates mosaic images so that quantitative information can be obtained on individual coal grains, i.e. **Coal Grain Analysis** (CGA).

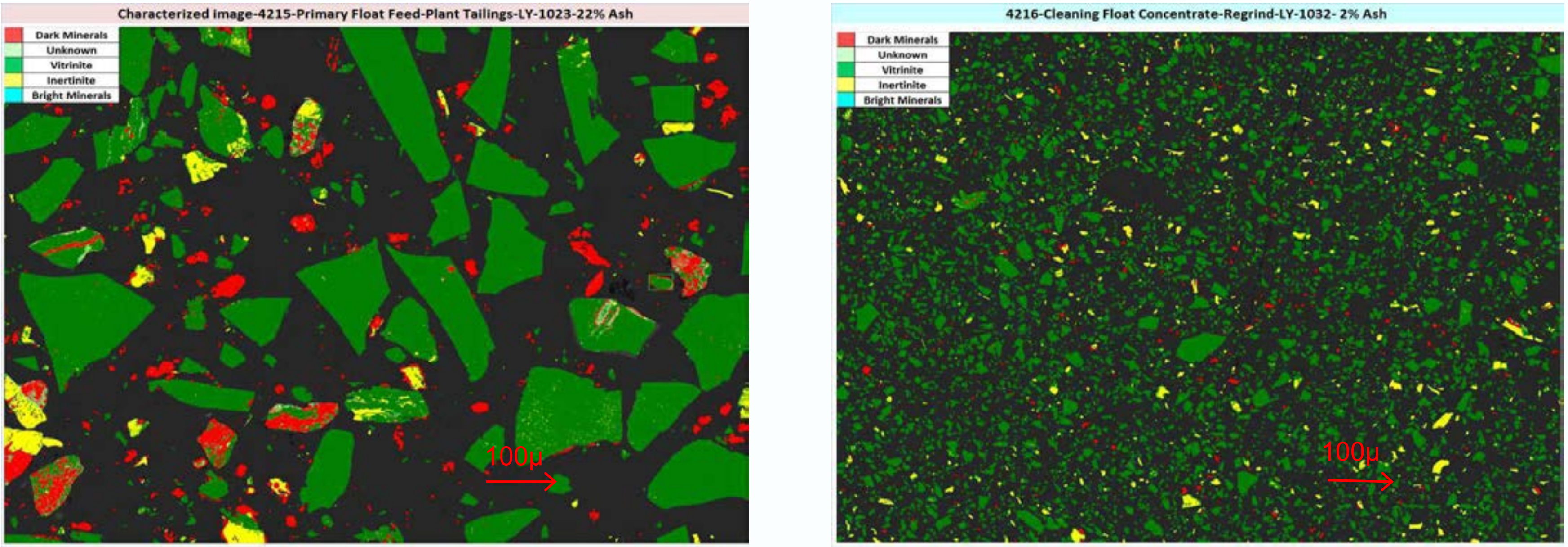
CGA generally requires only small representative sub-samples of <1mm material.

Size and compositional information, i.e., **macerals** vitrinite, inertinite, liptinite and **minerals** can be determined for each particle.

Information can then also be used to estimate % mass, density & "ash" value of each particle.



Liberation – CGA Images Confirm Status



Characterised images for Raw Coal Tailings Feed compared with Final Concentrate. (Courtesy QCAT-CSIRO)

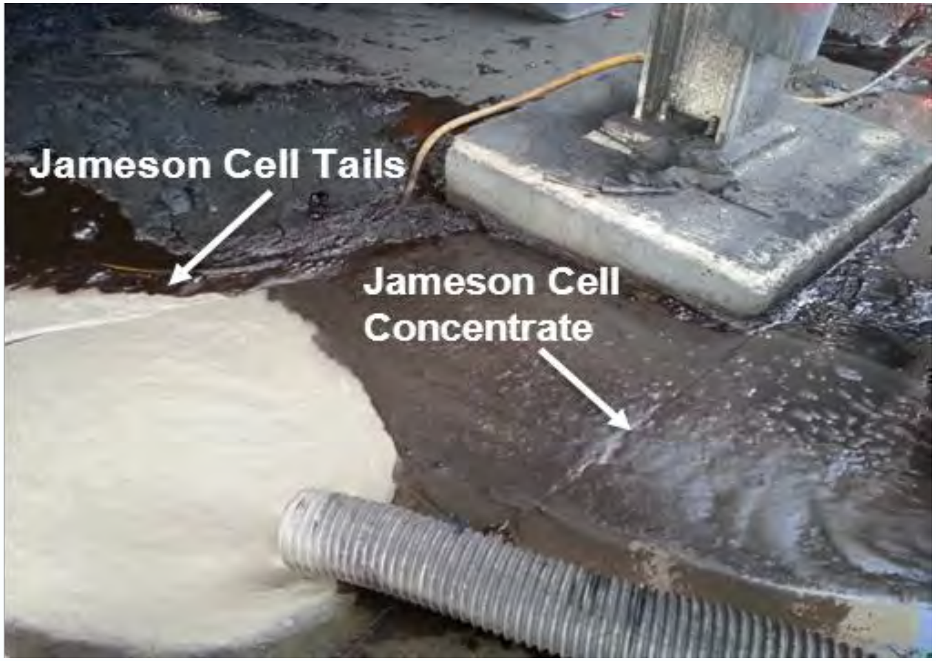
Flotation/Milling process approach

A 1tonne/h pilot-plant, owned and operated by Glencore Technology (formerly Xstrata) comprising

- two Jameson flotation cells,
- a small IsaMill, and a
- membrane filter press, etc.,
- Located at a large thermal coal operation in the Hunter Valley.

Currently testing freshly generated raw coal tailings to produce coal water slurry fuels

Milling and Sub-50μ Coal Flotation



NSW Coal Water Slurry Fuel (CWSF) Pilot Plant

- Successfully produces stable Coal Water Slurry Fuel (CWSF) from coal tailings
- CWSF can then be further refined to produce very low ash (<1% ash) Micronized Refined Coal (MRC)
- MRC produced from 2011 - 2015 for diesel engine tests
- Process information obtained also used for design of CWSF modules including a package plant and fuel handling rig.

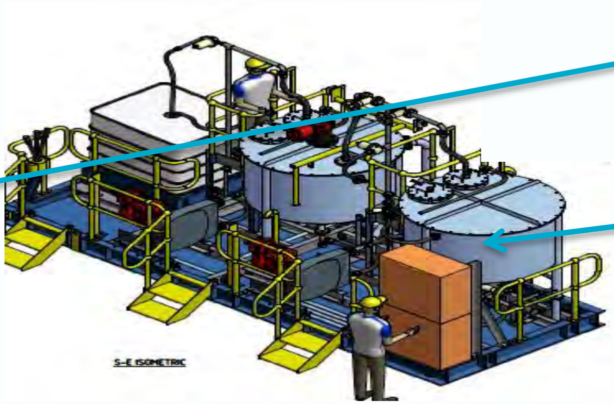
Pilot Plant Fuel Production

Jameson cells

Flotation rig

Isa Mill

Fuel Preparation



Fuel Delivery

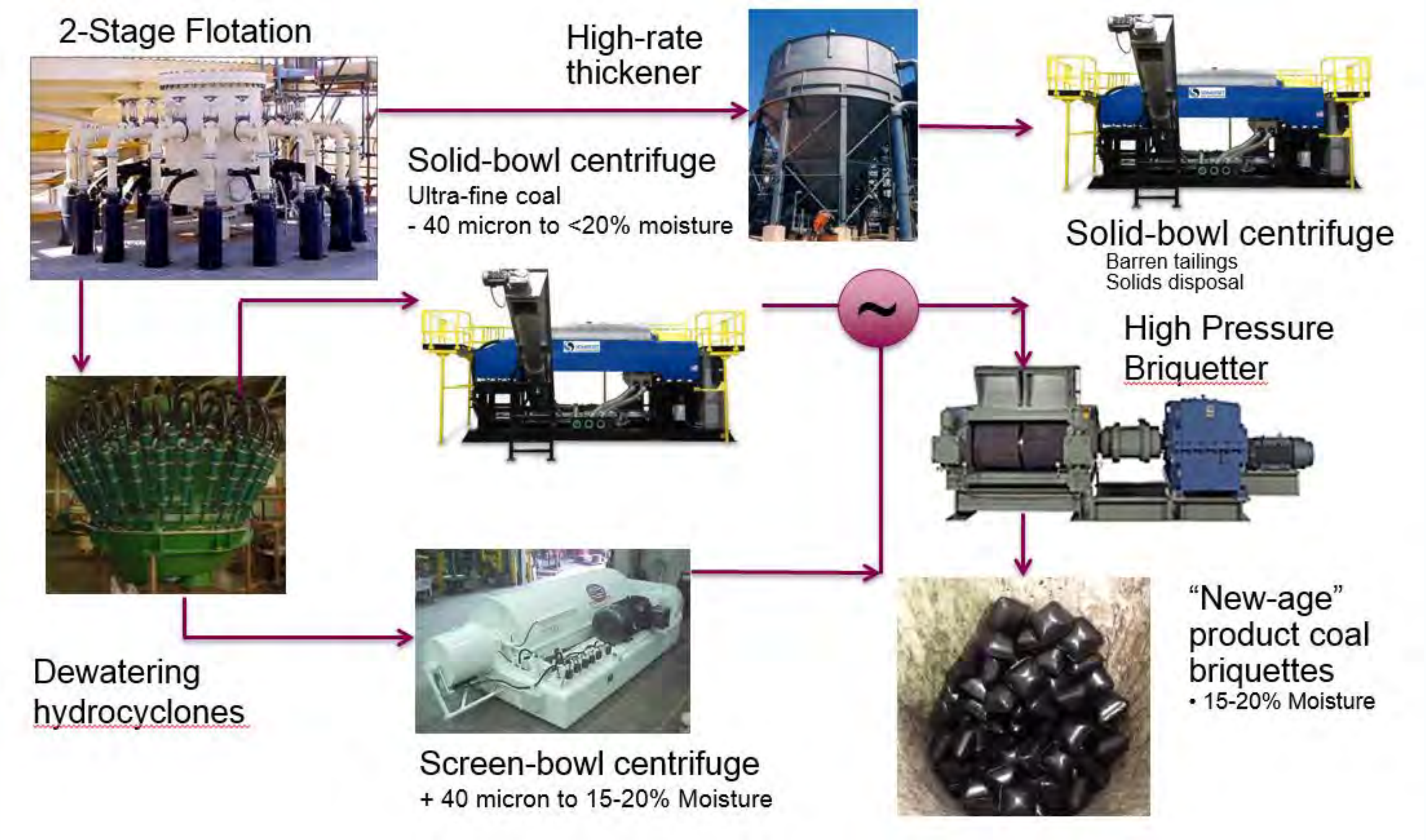
Dewatering Technology

Equipment	Throughput (dry solids)	Product Moisture (% w/w)
Horizontal belt filter	50-130 t/h	20-30
Screen bowl centrifuge	20-60 t/h	16-27
Centribaric centrifuge	15-20 t/h	15-20
Vacuum disc filter	50-150 t/h	20-32
Hyperbaric disc filter	30-150 t/h	17-25
Solid bowl centrifuge	10-18 t/h	18-25
Membrane Filter press	15-30 t/h	14-32

Ultrafines

Key objective is to achieve a cake moisture of 20% or lower for ultrafines

Emerging Coal Fines Treatment Circuit



2-Stage Flotation

High-rate thickener

Solid-bowl centrifuge

Ultra-fine coal

- 40 micron to <20% moisture

Solid-bowl centrifuge

Barren tailings

Solids disposal

~

High Pressure Briquetter

Dewatering hydrocyclones

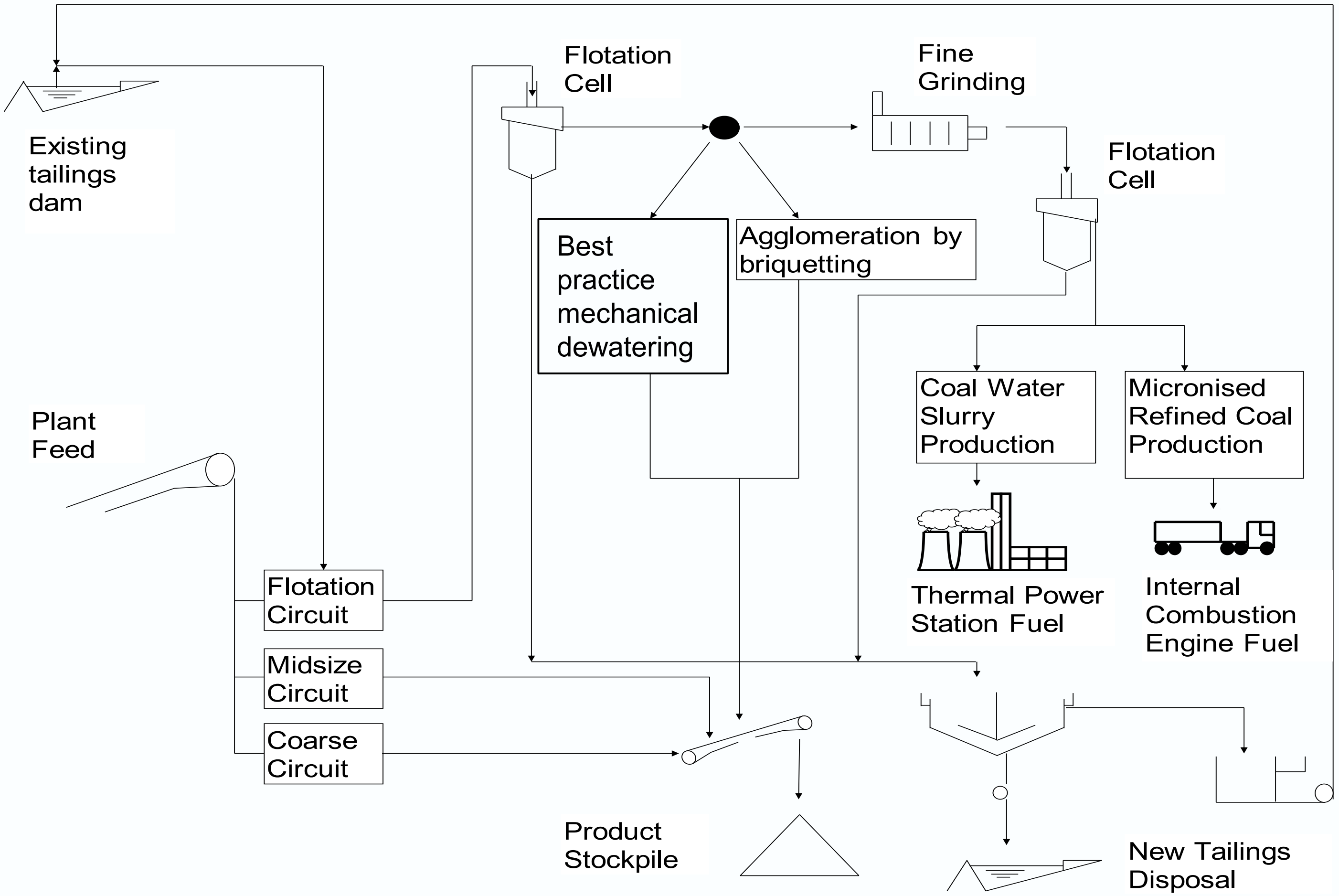
Screen-bowl centrifuge

+ 40 micron to 15-20% Moisture

“New-age” product coal briquettes

- 15-20% Moisture

Wider Fines Treatment Options



Existing tailings dam

Flotation Cell

Fine Grinding

Flotation Cell

Best practice mechanical dewatering

Agglomeration by briquetting

Coal Water Slurry Production

Micronised Refined Coal Production

Plant Feed

Thermal Power Station Fuel

Internal Combustion Engine Fuel

Flotation Circuit

Midsize Circuit

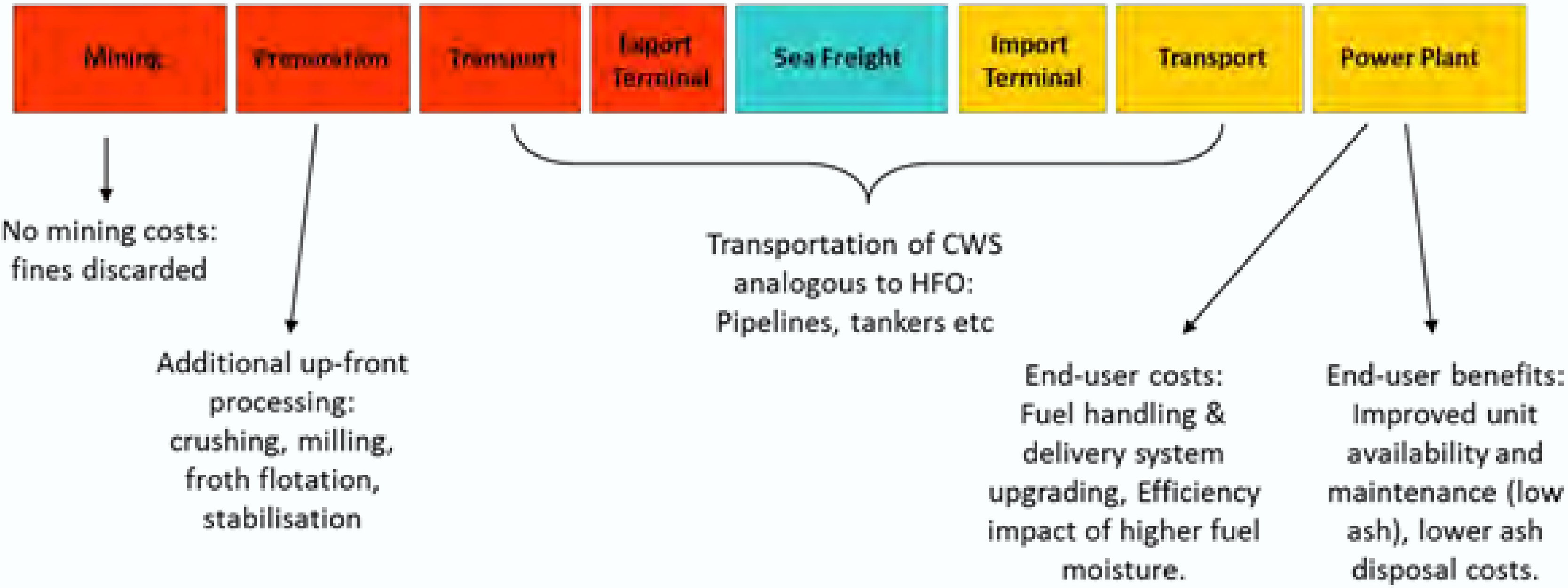
Coarse Circuit

Product Stockpile

New Tailings Disposal

Coal Supply Chain

- Current Coal Supply Chain (CSC) has been hampered by an inability to dewater and efficiently transport fine coal.
- Innovative approach - recover and use all the ultrafines via <u>coal-water slurry</u> thereby recovering potential "lost coal" creating higher yield and lower cost/tonne.



Mining

Preparation

Transport

Export Terminal

Sea Freight

Import Terminal

Transport

Power Plant

No mining costs: fines discarded

Additional up-front processing: crushing, milling, froth flotation, stabilisation

Transportation of CWS analogous to HFO: Pipelines, tankers etc

End-user costs: Fuel handling & delivery system upgrading, Efficiency impact of higher fuel moisture.

End-user benefits: Improved unit availability and maintenance (low ash), lower ash disposal costs.

Innovative Coal Supply Chain

New Low-cost Fuel

- Coal-water slurry fuel (CWSF) at ~70% solids prepared from coarser (bi-modal) particle size distribution (p80 of 0.075 mm)
- Use for direct firing to boilers as a potential replacement for heavy fuel oil (HFO), or partial replacement for Pulverized Fuel (PF)
- Transport as slurry fuel - avoids sticky, wet or dusting coal problems
- Lower tailings disposal cost - paste-thickening, further dewatering for co-disposal with coarse plant discards and mining waste.

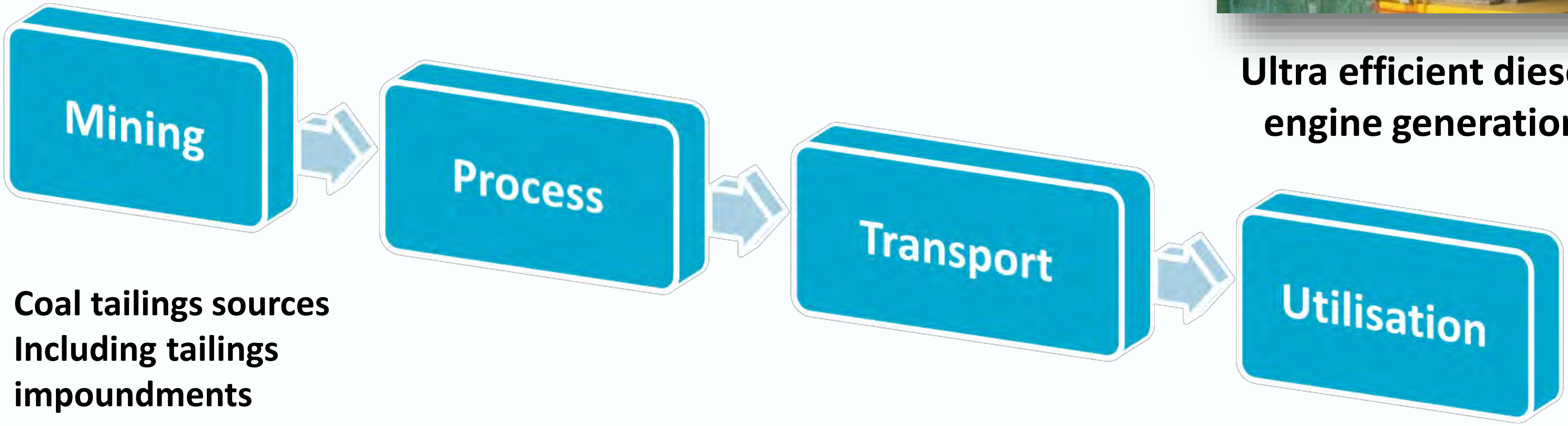
User Benefits

- No further grinding needed, significantly lowering cost
- Major O&M savings and lower ash disposal cost
- Reduced thermal efficiency offset by cost reductions from recovering lost coal from tailings.
- Potential to replace > 30% of the pulverised coal capacity.
- **Value Proposition:** a 1.0 to 1.5 c/kWh saving once the boiler has been converted for CWSF.

Micronized Coal Water Slurry

Optimization of the fuel cycle (DICE)

Ultra efficient diesel engine generation



Mining

Process

Transport

Utilisation

- Coal tailings sources
- Including tailings impoundments

- Increased grade recovery
- Recovery of ultra fines
- Minimal dewatering

- Road/rail/ship – cake or slurry
- Pipeline coal water fuels
- Higher solids paste for longer distance

- Mine-mouth or centralized
- Distributed generation
- Support of renewables

Conclusions

- "Deep cleaning" via liberation and subsequent beneficiation has offered significant potential downstream improvements, i.e.,
  - maximised resource recovery,
  - minimised transport and handling costs,
  - numerous end-user process improvements,
  - reduced maintenance and wear,
  - lower environmental impacts and
  - other sustainable improvements.
- Ultrafine coal beneficiation has matured via progressive froth flotation improvements
- Dewatering the concentrate was a barrier, but emergence of membrane filter presses, hyperbaric disc filters or high-g decanter centrifuges now offers commercial solutions.
- Briquetting and agglomeration has progressed to machine capacities of up to ~40 ton/h for fine coal applications to improve product handling.
- Manufacture of stable coal-water slurries with > 65% solids and stable micronized slurries with > 60% solids have reached commercial adoption.
- **Scene is now set** for new generation clean coal technologies with minimal wasted energy, lowest ash disposal costs and reduced SO_x, NO_x and CO_2 emission costs.

Acknowledgements

The authors acknowledge the contributions provided by

- Glencore Technology,
- Glencore Coal,
- Yancoal, and the
- CSIRO

towards the preparation of this paper and for permission to include figures and data from other recently published papers and articles on this subject.

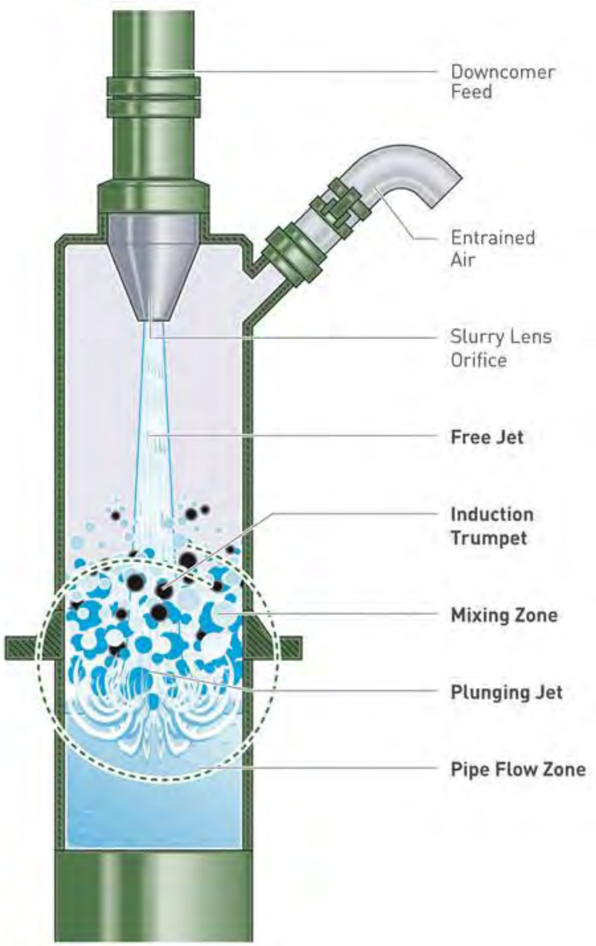
Dave Osborne thanks Somerset Coal International for encouragement and support in attending and participating in the 2015 Clearwater Clean Coal conference.

Thank you for your time

Questions?

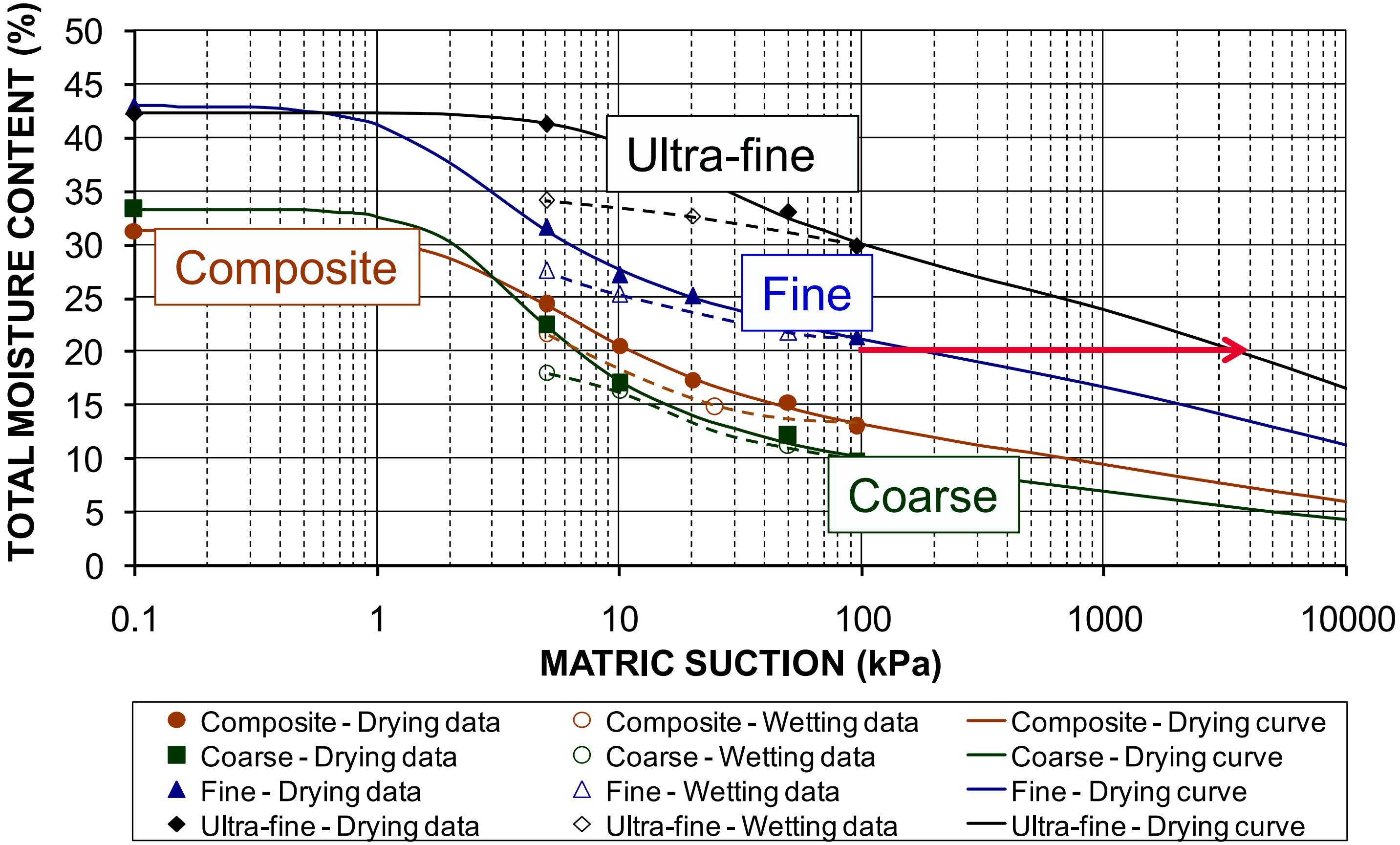
Louis Wibberley Phone: +61 7 3327 4457 Email: Louis.wibberley@csiro.au
Dave Osborne Phone: +61 7 3010 9443 Email: dosborne@somersetpty.com

The Jameson Cell - Downcomer



Product Coal Moisture Relationships

- Soil Water Characteristic Curve (SWCC), related to pore size distribution, in turn related to Particle Size Distribution[1].
- Matric Suction is related to Applied Pressure



[1] Source: Prof David Williams; Univ Queensland - *D.Williams@uq.edu.au*

Coal Water Slurry Fuel

Comparison of Coal Supply Chain Costs for Electricity Generation

Conventional Coal

FOB 80 $/t

CIF 90 $/t

Total Costs = 144 $/t

Total Costs = 5.5 c/kWh

Mining	Preparation	Transport	Export Terminal	Sea Freight	Import Terminal	Transport	Power Plant
60 $/t	5 $/t	10 $/t	5 $/t	10 $/t	5 $/t	5 $/t	44 $/t
2.0 c/kWh	0.2 c/kWh	0.5 c/kWh	0.2 c/kWh	0.5 c/kWh	0.2 c/kWh	0.2 c/kWh	1.6 c/kWh

Coal Water Slurry

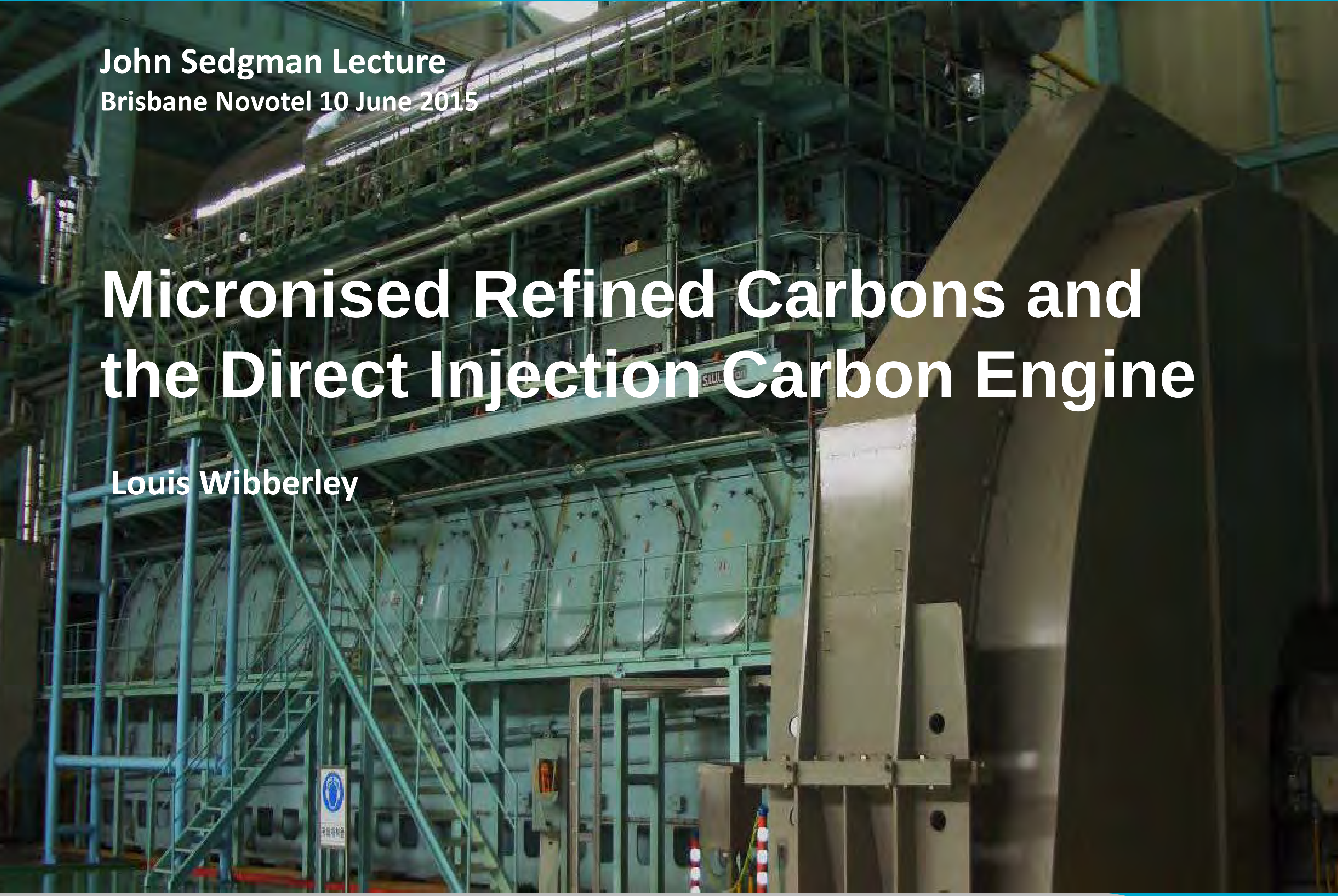
FOB 36 $/t

CIF 47 $/t

Total Costs = 105 $/t

Total Costs = 4.0 c/kWh

Mining	Preparation	Transport	Export Terminal	Sea Freight	Import Terminal	Transport	Power Plant
	20 $/t	11 $/t	5 $/t	11 $/t	5 $/t	6 $/t	47 $/t
	0.8 c/kWh	0.5 c/kWh	0.2 c/kWh	0.5 c/kWh	0.2 c/kWh	0.2 c/kWh	1.7 c/kWh



John Sedgman Lecture

Brisbane Novotel 10 June 2015

Micronised Refined Carbons and the Direct Injection Carbon Engine

Louis Wibberley

ENERGY FLAGSHIP

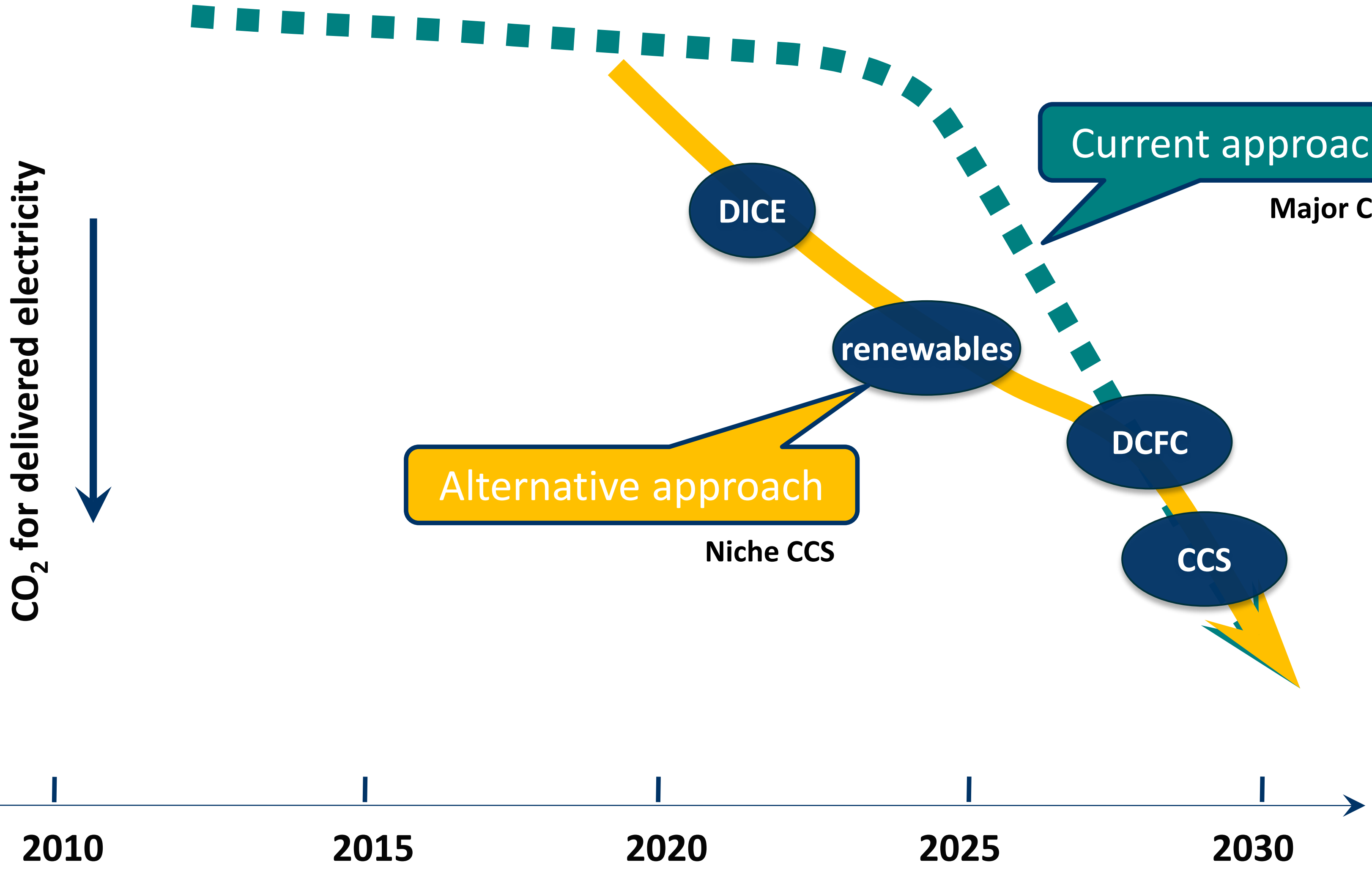
Key messages

1. Combining the low cost and availability of coal with the superior thermal efficiency, flexibility and lower capital cost of the diesel engine provides a <u>step change technology</u> for coal
   – this requires <u>ultra low ash coal</u>

2. Ultra low ash (and other premium coals) can be economically produced from a wide range of coal sources, including tailings - using <u>conventional equipment</u>
   … but this requires a <u>change in philosophy</u>

Why? … provides an alternative LE pathway

Philosophy: **higher efficiency + underpinning renewables + niche CCS**



MRC-DICE fuel cycle

Carbons

Premium water-based slurry fuel

micronised refined carbons (MRC)

Ultra efficient diesel engine generation

direct injection carbon engine (DICE)

MRC – the most efficient way of converting carbons into liquid fuels

MRC from Victorian brown coal

Micronised refined carbon (MRC) has been produced from a range of sources

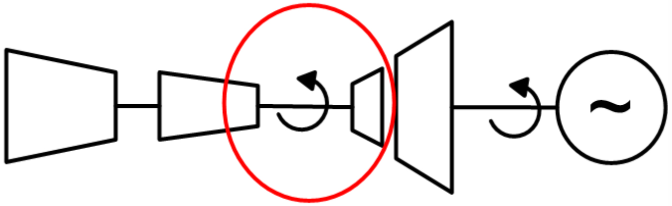
- desanded and hydrothermally treated low rank coals
- deashed black coals (including tailings)
- chars and algal matter (blended)

Fuel choice determines carbon footprint

Process has very high energy conversion efficiency >97% (LCA basis)

Diesel engine – efficient, flexible and fuel tolerant
(but some adaptation required for coal)

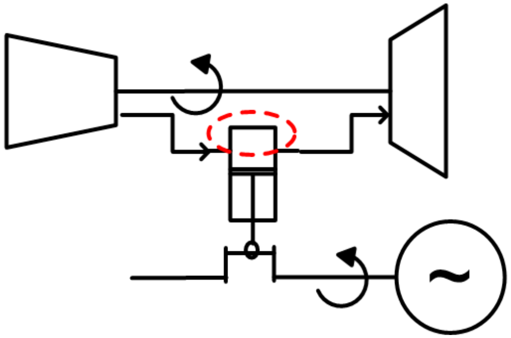
Gas turbine



Continuous combustion - hot section is exposed continuously to 1450-1500°C gas at 2-4 MPa
Exotic alloys, hot strength, oxidation
Fouling issues for impure fuels

Turbocharged diesel engine

Higher "Carnot" efficiency
Cyclic hot space allows dirtier fuel without fouling or the need for exotic alloys
Large expansion ratio = smaller waste heat recovery



Batch combustion - hot section hot for <10% of the time, cyclic @ 1-5 Hz (larger engines)
Higher T & P possible and without fouling, >1500°C, 15-25 MPa

Cycle comparisons – too much water?

	External cycle				Power cycle				
Cycle-fuel $P/T_1/T_2$[1]	Fuel (dry t)	Water (t)	Air (t)	η (%HHV)	Fuel (dry t)	Water (t)	Air (t)	η (%HHV)	η sent out (%HHV)
Steam-black 250/650/650	1	0.1	11	88%		12		48%	42%
Steam-brown 250/650/650	1	2.2	10	73%		11		48%	35%
Steam-CWM 250/650/650	1	0.7	11	84%		11.5		48%	40%
Diesel-HFO 200/1500					1	0-1	15	54%	52%
DICE-MRC 200/1450					**1**	**1**	**15**	**51%**	**49%**

[1] *bar/°C/°C*

DICE offers game-changing attributes in 5-6 years

1. Match and compete with natural gas; rapid start/stop and load following capability
   - excellent match to electricity grid with high intermittent renewables
2. Step reduction in carbon emissions for electricity generation without CCS
   - 20-35% reduction in carbon emissions versus current black coal
   - 35-50% reduction in carbon emissions versus current brown coal (in Victoria)
3. Cost competitive with new conventional coal

… not possible with other coal technology

4. Small capital investment steps
   - can achieve large power plant size incrementally using 20-100 MW units
   - shorter construction time
5. No cooling water
6. Can be used for various biomasses
7. Capture ready/capture efficient
   - 30-40% lower cost of CO_2 abatement over conventional coal
8. Short path to commercialisation
   - adaptation of current large engines, short cycle time to implement changes, relatively low development cost

ability for commercial-scale demonstration at a small scale (say 30 MW)

What about USC? … development opportunities restricted and costly

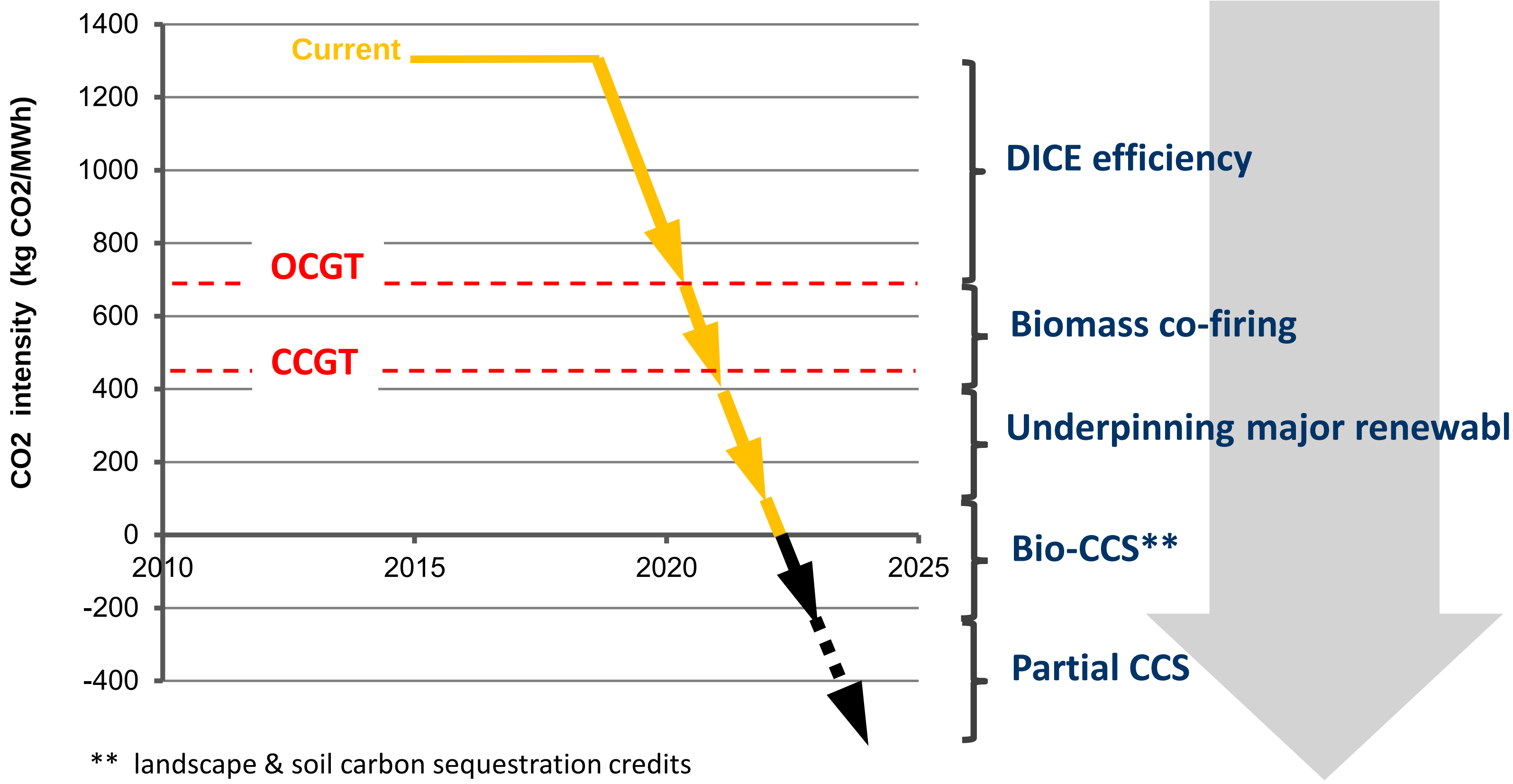
- Thermodynamic efficiency of pf generation is severely limited by the availability of materials that can operate at these conditions for practical service lifetimes
  - EU, USA, Japan, India and China all have extensive material research programs aiming for steam temperatures of 700°C (advanced ultra-supercritical)
  - development cost of billions of $ and long lead times (creep testing)
  - anticipated that a commercial unit could be brought on-line in 2031 (IEACCC/229)
  - high capital cost of advanced ultra-supercritical is of particular concern (high pressure steam pipes currently 80% of the boiler cost)
- While the combustion conditions in the diesel engine are more extreme, the diesel cycle is a batch process
  - high temperature conditions are present for less than 10% of the time, which avoids the need for major exotic alloys

A pathway to net negative CO_2 emissions?

(DICE efficiency first, then high penetration renewables with bio-CCS and lastly partial CCS)

Example: Victorian generation

Carbon management sequence



- DICE efficiency
- Biomass co-firing
- Underpinning major renewables
- Bio-CCS**
- Partial CCS

** landscape & soil carbon sequestration credits

If successful DICE could address many aspects of the coal dilemma

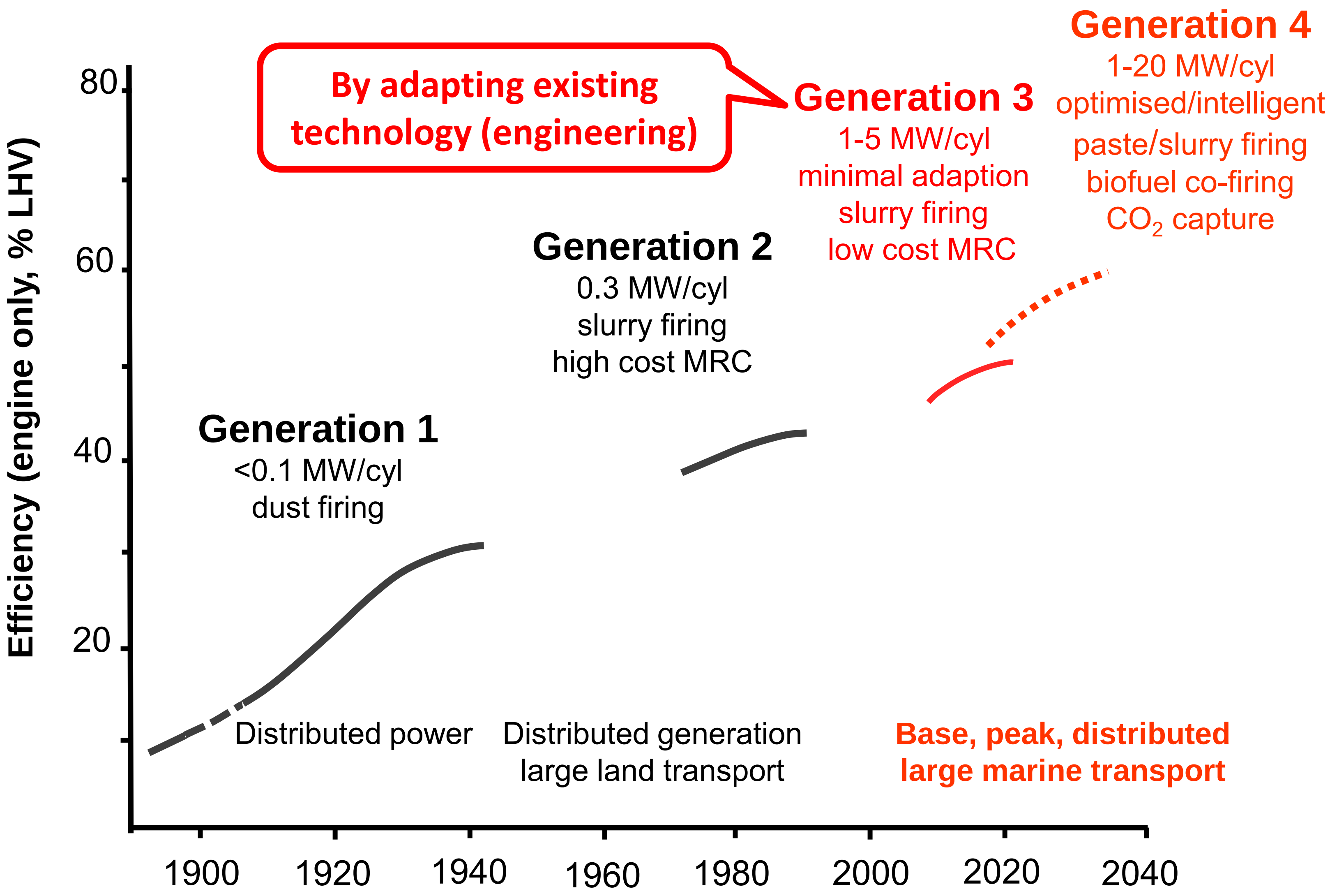
Including:

- that large centralised plants are needed for efficiency
- the nexus between water/CO_2 and cost (dry cooling)
- technology development by a fragment of generation industry
- inefficient (even if cheap) is no longer acceptable
- very poor image of low rank coals
- poor project economics from long development times
- the higher flexibility needed for current and future electricity markets

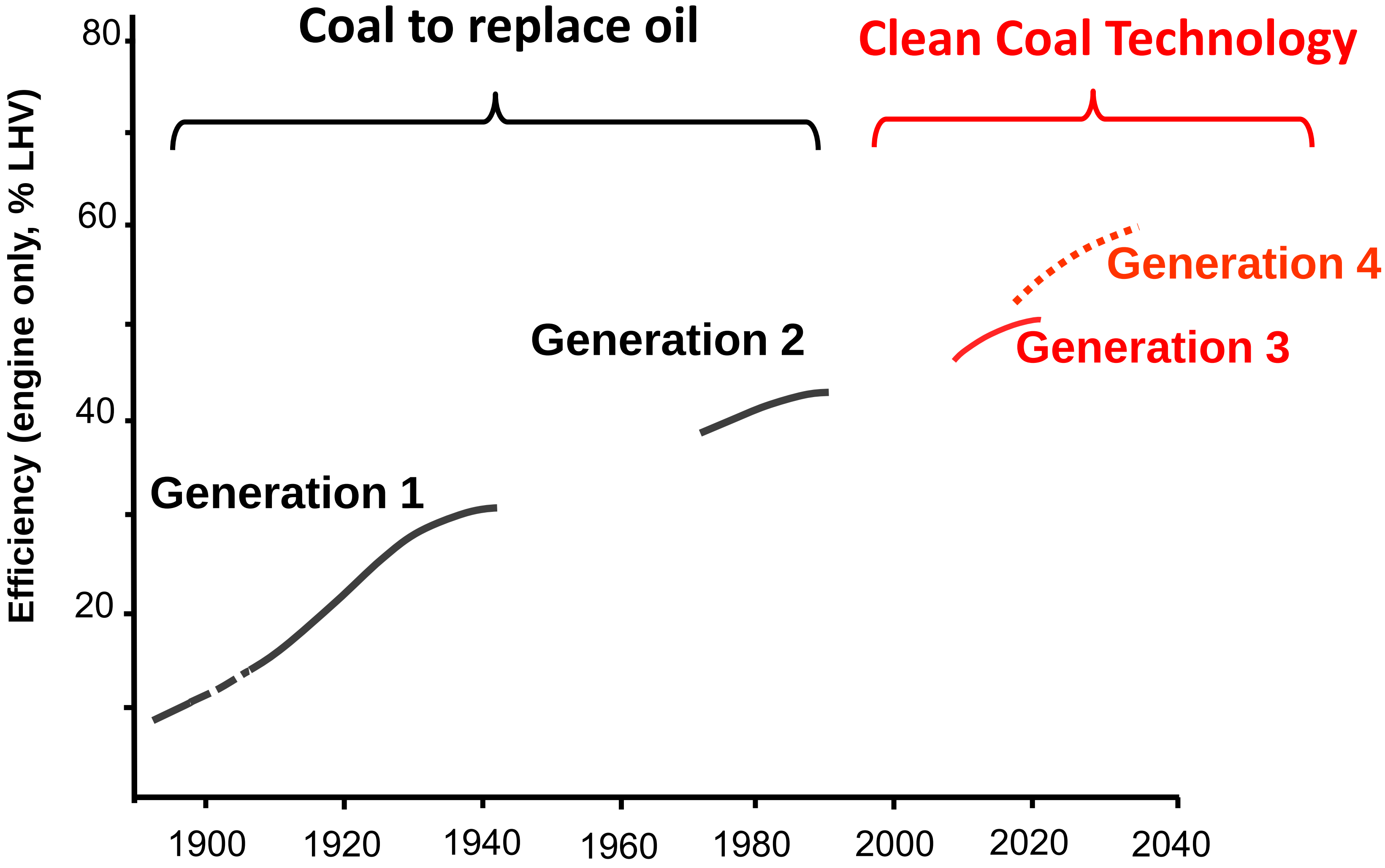
Could DICE become the benchmark coal generation technology?

Not new … commercialisation in G3?

Considerable upside



… with a changed philosophy



The fuel side … towards commercial production of MRC

Coal water fuel in China – stepping stone to MRC?

Coal water mixtures for boilers

~40 Mtpa in China
5-8% ash, d90 ~150µm
typically 70% coal
2000 mPa.s @ 100/s

Micronise and float

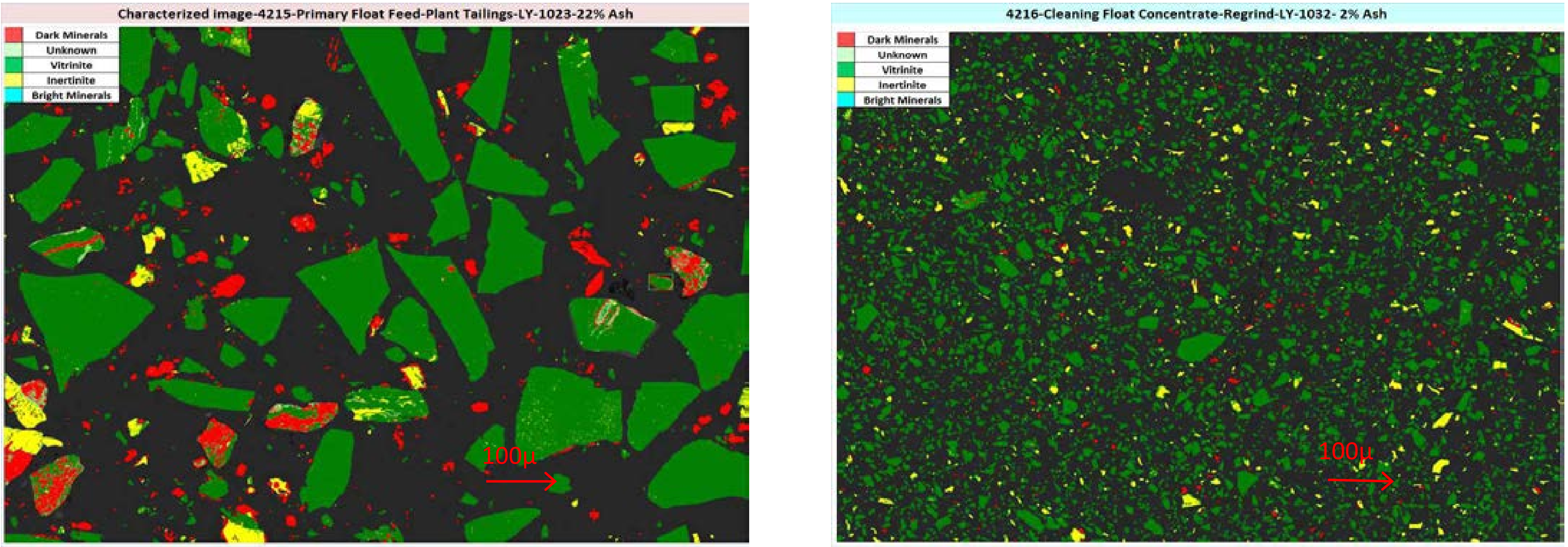
MRC for DICE

nominal specifications
1-2% ash, d90 ~40µm
typically 55% coal
<300 mPa.s @ >200,000/s

MRC is based on ultra-fine coal beneficiation

- Physical cleaning – coal structure not changed, but comminution is needed for optimal liberation/improved flotation response
  - potential product <1% ash, including from tailings
- “Old school” thinking typically regards ash contents below 2-3%, as both technically and economically unviable because
  - “inherent ash“ of coal is usually regarded as the lowest achievable ash content
  - lower ash requires milling to impractical ultra-fine sizes for liberation
  - … but flotation of ultra-fine coal is problematic requiring higher reagent dosages
  - fine coal concentrates are inevitably high in moisture (> 35%) which means costly dewatering and/or drying to produce saleable products
- All of these factors are incorrect, or at best very misleading, as recent research/pilot plant tests have shown

Excellent liberation by micronising



Feed

After liberation by milling

Characterised images for raw coal tailings feed compared with final concentrate (QCAT-CSIRO)

Micronising and sub-50µm coal flotation – excellent separation and recovery

Courtesy Glencore Technology

Pilot scale production of MRC for MAN Diesel & Turbo

Micronised Refined Coal

- coal water slurry (54% coal; 46% water)

Batch No.

MRC 200-175

Hazard

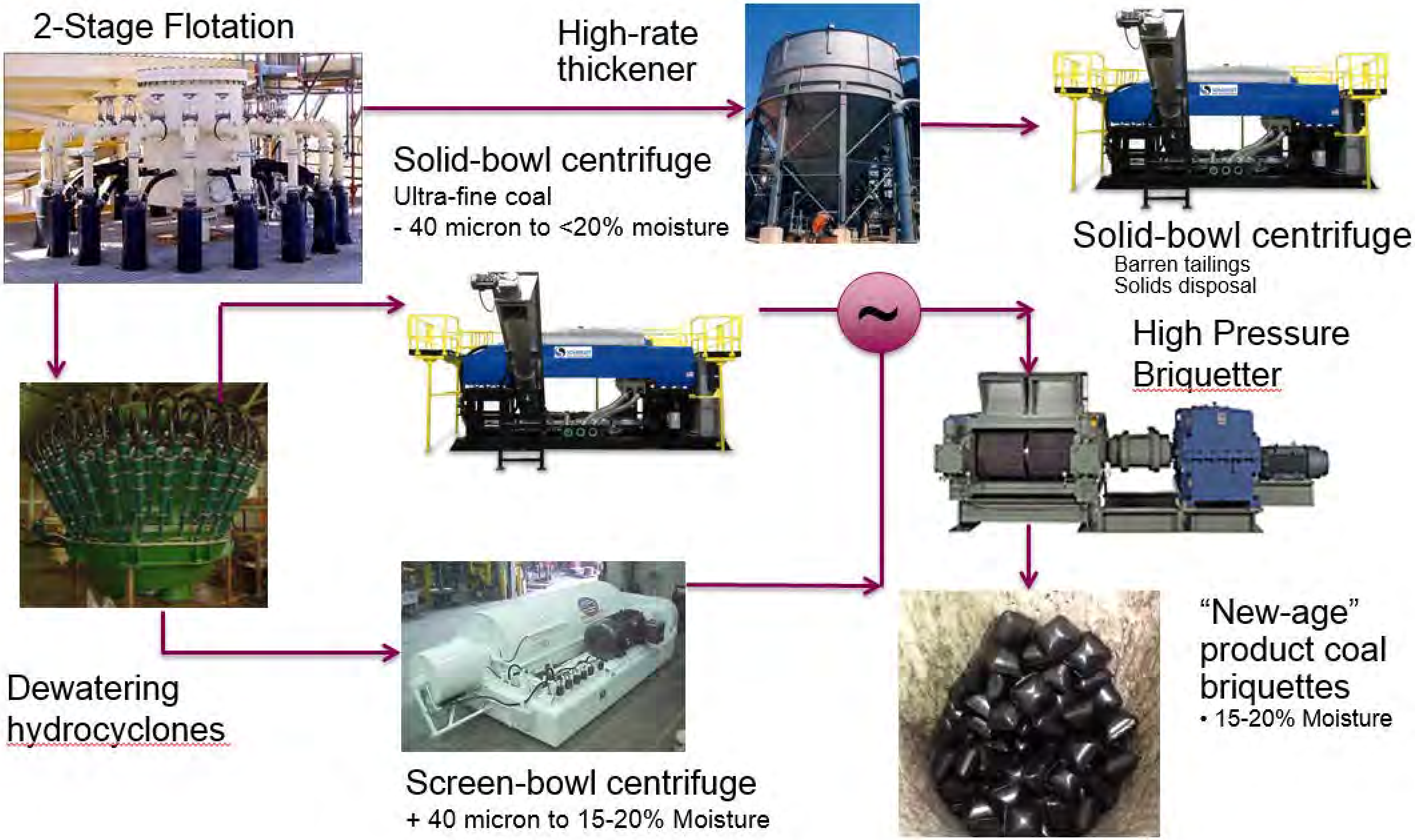
NIL

Target Organs

Mild irritation to eyes due to fine coal particle in slurry; stains skin surface (pigment effect, temporary).

1. MRC cake production at Bulga Pilot Plant *(Glencore)*
2. Formulation & rheology trim *(CSIRO)*
3. Certification *(ALS)*

MRC … strategic part of a bigger picture of “Premium Coal Products”



2-Stage Flotation

High-rate thickener

Solid-bowl centrifuge
Ultra-fine coal
- 40 micron to <20% moisture

Solid-bowl centrifuge
Barren tailings
Solids disposal

High Pressure Briquetter

Dewatering hydrocyclones

Screen-bowl centrifuge
+ 40 micron to 15-20% Moisture

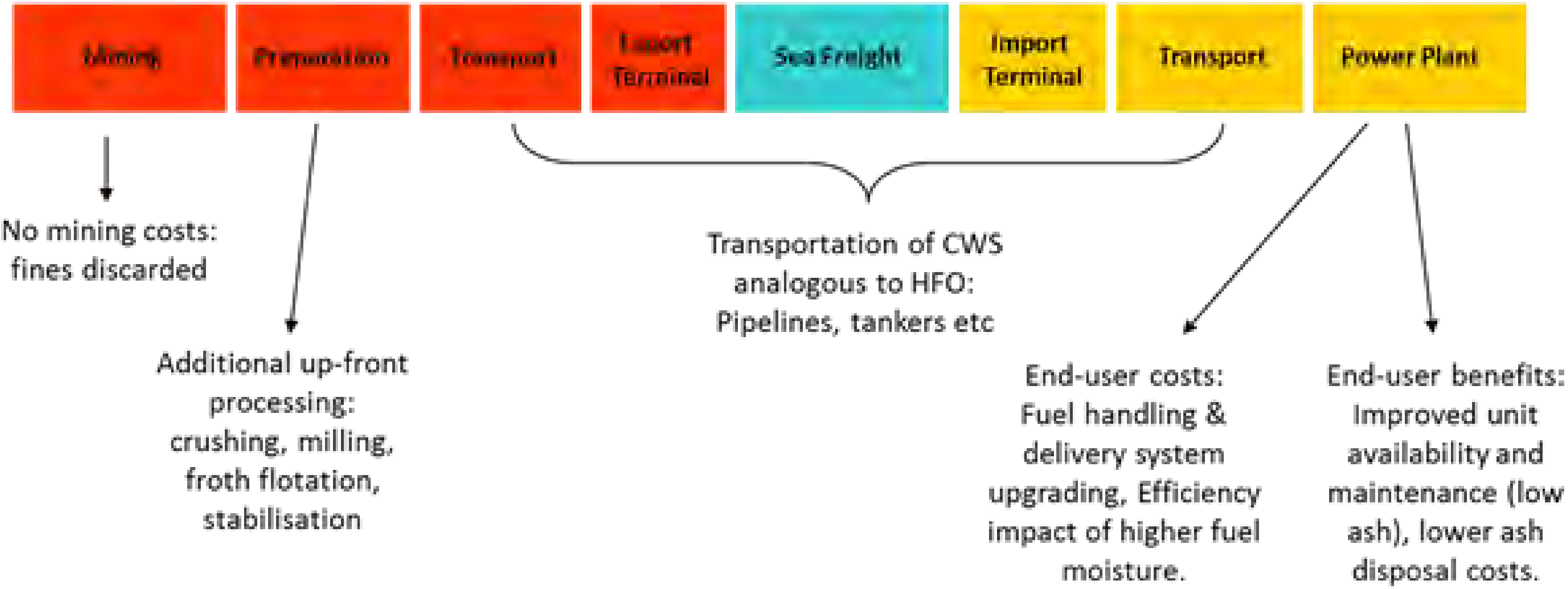
“New-age” product coal briquettes
- 15-20% Moisture

Context

Coal products	Conventional	Steaming
		Coking
“Premium” (<3% ash)		Coal water fuel
		Electrode carbon
		Micronised refined carbons for DICE (MRC)
Electricity generation	Conventional	Pf
	Alternative	IGCC
	Emerging	**Direct injection carbon engine (DICE)**
	Novel	Direct carbon fuel cell (DCFC)

… and an alternative Coal Supply Chain

- Current Coal Supply Chain hampered by an inability to dewater and efficiently transport fine coal
- Innovative approach - recover and use all the ultra-fines as coal water slurry thereby recovering potential “lost coal” creating higher yield and lower cost/tonne



No mining costs: fines discarded

Additional up-front processing: crushing, milling, froth flotation, stabilisation

Transportation of CWS analogous to HFO: Pipelines, tankers etc

End-user costs: Fuel handling & delivery system upgrading, Efficiency impact of higher fuel moisture.

End-user benefits: Improved unit availability and maintenance (low ash), lower ash disposal costs.

The engine side … towards commercial DICE

Recent developments

www.dice-net.org

CSIRO DICE program since 2008

- **de-risking based R&D** program (Yancoal, Exergen, Newcrest/JGC, BCIA, Ignite Energy Resources, Xstrata)

MAN Diesel & Turbo have taken a lead position in DICE development

MAN Diesel

Umbrella organisation established to facilitate DICE development internationally

- 17 participants includes MAN, RWE (Germany), JGC (Japan), Sinarmas (Indonesia), Exergen, Ignite Energy Resources, BCIA, Energy Aust, AGL, Newcrest, Yancoal, Worley Parsons, GHD, ACALET and CSIRO

Recent interest from groups in Korea and China

Stage-gated development

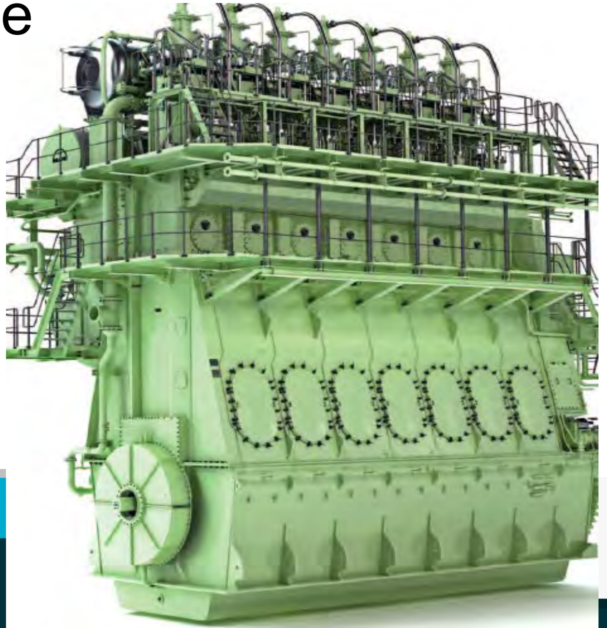
2014-16 Small scale demonstration, initial demonstration/validation DICE,1MW single cylinder (brown and black coals)

Joint funding with BCIA, ANLEC, CSIRO and industry

2016-17* Development/design of components for prototype engine

2017-19* Full scale demonstration MRC production with a 12-30 MW prototype engine for 8000h campaign

2020*- First commercial DICE power plant [$1.4-2 M/MW] possible given appropriate funding support



** based on MAN D&T estimates of 3-5 years for engine dev*

DICE deployment strategy

DICE favoured when natural gas price >$6-7/GJ forecast to occur by 2020 for most countries

- Australia; $5-6/GJ; forecast >$8.50/GJ in 2020
- China (import); $13.70/GJ; forecast >$10/GJ in 2020
- Europe; $10.80-12.20/GJ; forecast >$8.80/GJ in 2020
- UK; $10-14.60/GJ; forecast >$10/GJ in 2020
- Japan/Korea; $14.20-16/GJ; forecast >$13/GJ in 2020
- Limited incentives in USA (low cost gas,1100 lb CO_2/MWh regulations)

DICE suitable for new coal capacity, and to replace old capacity nearing the end of its economic life (or as it becomes socially unacceptable)

Final comments

1. DICE could provide coals with a innovative step technology to increase its cost competitiveness and environmental acceptance
2. Barriers to commercialisation are <u>mostly engineering</u>
    - adaptation of commercial process & engine technologies
3. Rapid development possible - can be demonstrated at commercial scale at a relatively small cost
    - <u>short lead time</u> between technology development & implement
4. Logistical barriers to commercialisation of the fuel cycle needs broad intra-industry support
    - <u>as part of premium coal products for maximum benefit</u>

Energy Technology
Louis Wibberley
Leader DICE Development Program

t +61 2 4960 6050
e louis.wibberley@csiro.au
w www.csiro.au/lorem

MetPlant 2015

Metallurgical Plant Design & Operating Strategies – World's Best Practice

7–8 September 2015, Perth, Western Australia

Image courtesy of Outotec. Atlas Mining's state-of-the art copper concentrator plant upgrade at its Carmen Copper Mining Complex in Toledo City, Cebu, Philippines.

Principal Sponsor

Outotec

Proceedings Sponsor

MAGOTTEAUX

Proceedings

AusIMM
THE MINERALS INSTITUTE

The Arrium IsaMill from Design through Commissioning and Optimisation

M Larson[1], G Anderson[2], M Mativenga[3] and C Stanton[4]

1. Senior Metallurgical Engineer, Glencore Technology, 160 Ann Street Level 10, Brisbane QLD michael.larson@glencore.com.au
2. IsaMill Technology Manager, Glencore Technology, 160 Ann Street Level 10, Brisbane QLD, greg.anderson@glencore.com.au
3. Principle Development Manager, Arrium Mining, PO Box 21, Whyalla, SA. Email: martin.mativenga@arrium.com
4. Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: chris.stanton@arrium.com

ABSTRACT

Arrium's magnetite concentrator has recently completed a modification program to improve the process that has been in operation in Whyalla, South Australia for the previous seven years. The magnetite concentrate is pumped via pipeline to the coast and serves as pellet plant feed, eventually being turned into steel in the Whyalla blast furnace. With the addition of a new 3MW IsaMill at the concentrator, the plant has seen increased production from a reduced feed grade, enabling Arrium to treat ores previously stockpiled as waste. This paper will focus on the design work and commissioning of the fine grinding circuit along with the year of optimisation that followed, taking the IsaMill up to 350+ tph of throughput.

INTRODUCTION

The original Arrium magnetite flow sheet that began operation in 2007 consisted of two stages of grinding, utilizing two HPGRs with a total installed power of 1.8 MW and a 7.5 MW ball mill. The complete flow sheet is shown in Figure 1.

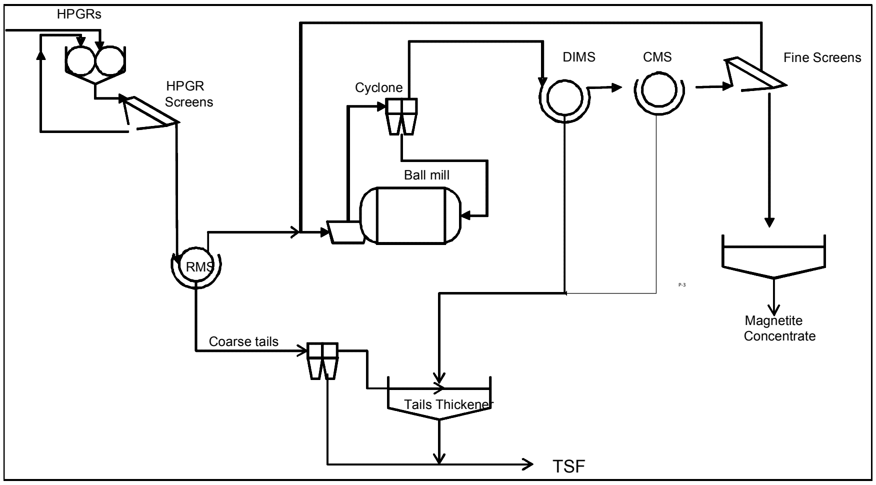


Figure 1 – Original Arrium magnetite flow sheet.

In 2009 Arrium began a review of the plant with the aim of debottlenecking and optimisation. Much of this early work is covered in the AUSIMM Iron Ore 2015 paper *Unlocking Plant Capability through Targeted Debottlenecking of Arrium's Magnetite Concentrator* (Mativenga, et al 2015). The main outcome of this work was the conclusion that if the plant was upgraded to improve liberation a large portion of stockpiled waste could be converted to plant feed while also increasing mill throughput. Arrium then began investigating different options to add grinding power to the plant through both conventional and stirred milling options.

BALL MILL VS ISAMILL COMPARISON

In 2011 Arrium Mining ran their first IsaMill lab test at ALS AMMTEC in Perth WA. This was a straight comparison of the existing ball mill circuit to the IsaMill. With an F_{80} of +350 microns it is coarser than most operating IsaMills. With 5-6 mm media the IsaMill was able to break the coarsest feed and still efficiently reach the final product size in significantly less energy than the ball mill. The ball mill reached a P_{80} of 32 microns in 24 kWh/t, not including any cyclone pump energy. The M4 IsaMill with the same feed from the ball mill survey reached a 32 micron P_{80} in 17 kWh/t as shown in Figure 2 (Steele, 2011). This results in a 29% improvement over the ball energy to 32 microns.

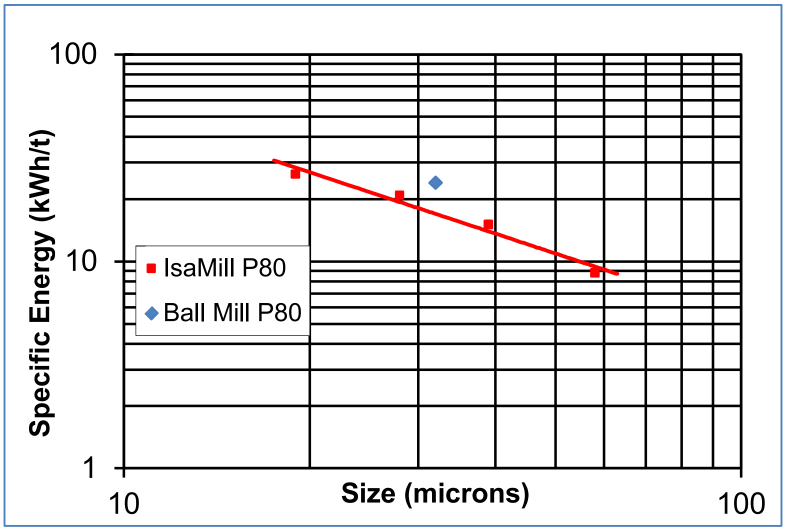


Figure 2 – Arrium ball mill versus M4 IsaMill.

The basic principle of the IsaMill is that it fluidizes a charge of small ceramic grinding media. Media ranging in size from 1.5 to 6.5 mm and coarse particles from the mill feed are centrifuged to the outside of each disc where backflow compression from the product separator keeps them in the mill until the particles are ground fine enough by attrition and compressive forces to pass down the middle of the mill and discharge. Due to this process external classification is not necessary.

As long as the IsaMill can break down the coarsest particles as quickly as they enter the mill it can compete with a ball mill. The M4 IsaMill testwork showed this was possible when grinding a 300 micron Arrium ore feed to P_{80} values below 100 microns.

However, the Arrium flowsheet has a functional ball mill and the goal of this exercise was not to replace it, but to modify the flowsheet to increase plant capacity and improve grinding energy efficiency.

Through previous Australian magnetite testwork it was known that running the IsaMill with the same feed (i.e. in parallel) as the ball mill is not the most efficient flow sheet. From pilot and lab work presented at MetPlant 2011 in the paper *Optimising Western Australian Magnetite Circuit Design* it is shown that 3 stages of grinding efficiently utilizes each grinding mill to the best of its abilities (David *et al*, 2011). As shown in Figure 3 the ball mill is perfectly suited for a coarser grind which then reduces the feed size to the IsaMill. From the graph it can be seen the IsaMill is only more efficient at grinding to product sizes below 100 microns in this particular case. With smaller reduction ratios both mills can be optimised with the correct media sizes. As an added benefit the separation step between the two milling stages reduces the amount of gangue material needing to be ground down to 32 microns.

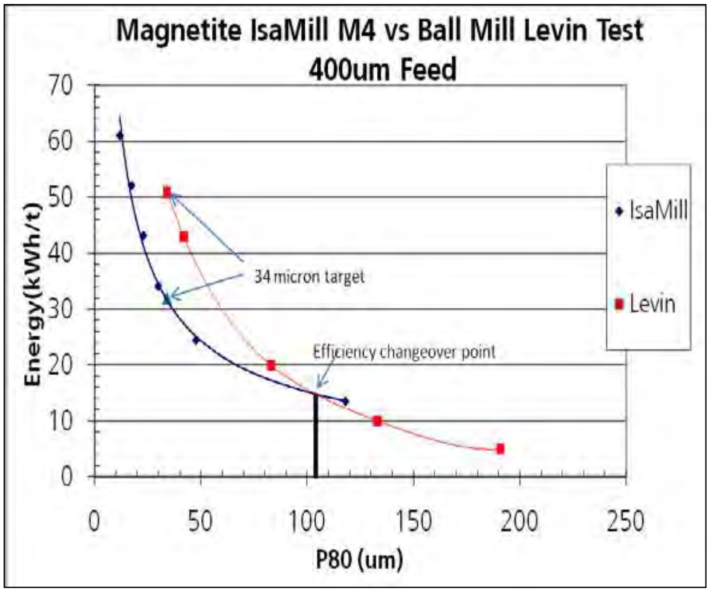


Figure 3 – Previous magnetite (not Arrium) IsaMill vs Levin ball mill lab tests.

Large reduction ratios and lack of intermediate gangue rejection work against the two stage flow sheet shown in Figure 3. The ball mill, though not ideal for grinding fine, is well suited for an intermediate grind, which can then be followed by an extra step of magnetic separation. This produces a reduced mass of feed which is ideally sized for the IsaMill. This idea was incorporated into the next phase of the Arrium M4 IsaMill work at ALS AMMTEC completed in 2012.

ISAMILL VARIABILITY SIGNATURE PLOTS

Test work was next undertaken at ALS AMMTEC with Arrium personnel present to understand how different sections of the ore body grind in the M4 IsaMill. Figure 4 shows three of these tests which display a considerable variation in the energy required to reach 32 microns, with values of 14.2, 12.3 and 18 kWh/t respectively (Ladhams, 2012). These three samples were each taken from different areas of the ore body. It should be noted these three samples were a harder, higher waste material than the first material tested in Figure 2. The test feed was ground down to an F_{80} of 90 microns with no intermediate magnetic separation prior to being run through the IsaMill. Removing liberated gangue material would have the effect of reducing this variability somewhat. It does show the importance of understanding the ore body and not relying on just one sample to design a regrind circuit.

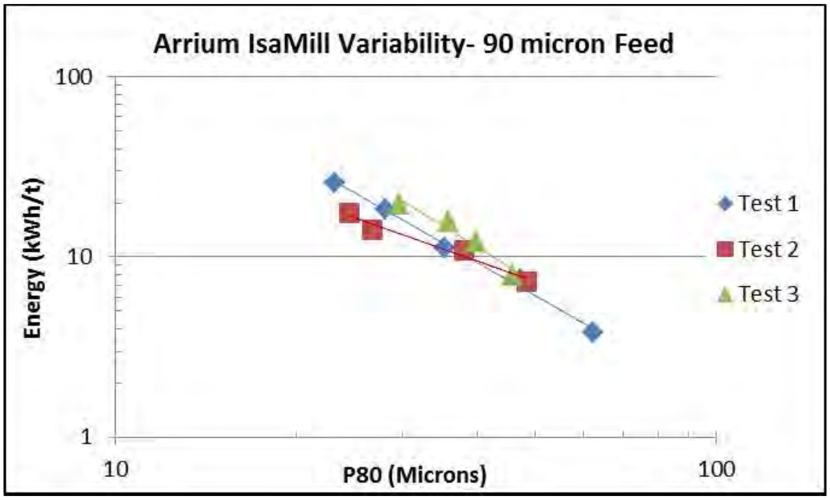


Figure 4 – Arrium variability signature plots.

FINAL MILL FLOW SHEET

After numerous signature plots and some pilot testing it was determined that one 3 MW M10000 IsaMill would be suitable to grind a maximum design 300 tph of 60 micron feed to a desired 32 µm P_{80} and Davis Tube silica grade of 2-3%. The mill would be in a tertiary grinding step rather than run in parallel with the ball mill. The design concept is to allow the HPGRs to produce a coarser product at higher tonnage, optimise the ball mill to treat this coarse material and then provide a suitably fine feed for IsaMilling. The IsaMill product would be fed to an installation of Derrick Stack Sizer screens with 63 micron panels to ensure steady silica grade. The role of the Derrick Stack Sizers is primarily to eliminate any remaining oversize material both for silica grade control and pipeline protection. This flowsheet is shown in Figure 5. There are additional 1 mm screens ahead of the IsaMill serving to remove trash from the feed rather than providing actual classification.

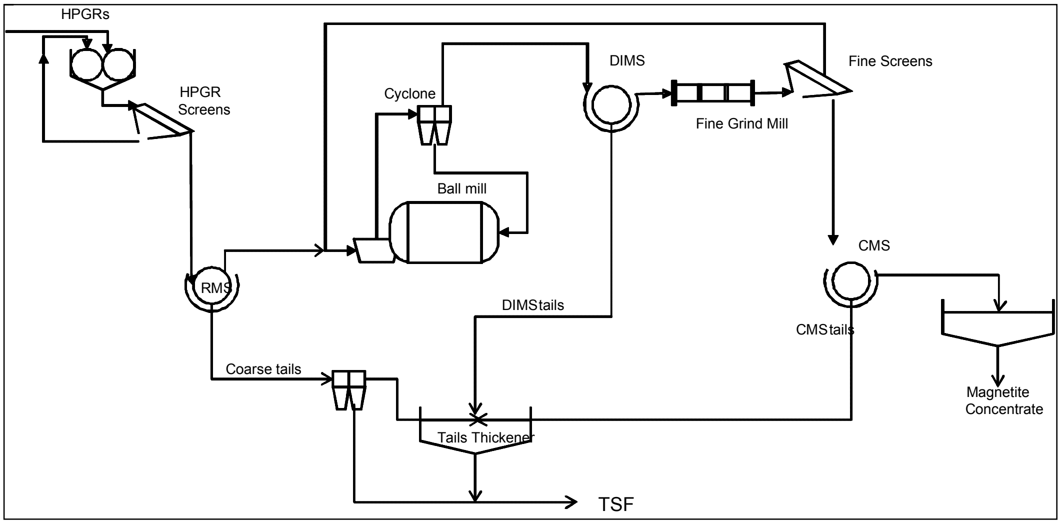


Figure 5 – Arrium modified flow sheet.

ISAMILL START UP

The M10000 IsaMill at Arrium is shown in Figure 6. Feed enters the mill at the non-drive end and discharges from the drive end already classified. Grinding is achieved by discs on an internal rotating shaft, and the discs are designed to keep a ceramic grinding media load of about 70% of the mill volume agitated.

Figure 6 – Arrium M10000 IsaMill installation.

The IsaMill was commissioned in November of 2013. Performance Acceptance Testing (PAT) was conducted on a dedicated ore stockpile shortly after start up. Comparison test work had been run on an M4 IsaMill at ALS AMMTEC from samples of this same stockpile. Minor differences from these feed samples included the media size used and lower solids SG due to there being no DIMs upgrading applied in the case of the M4 feed. The M4 sample was also slightly coarser, resulting in more residence time and a sharper size distribution once it reaches the final P_{80}. A comparison of the average results from both the lab and full scale is given in Table 1 (Bandarian, 2014).

Table 1 – Average PAT plant sizing data versus average AMMTEC signature plot test work data.

	Specific energy input (kWh/t)	Feed F80 (µm)	Discharge % passing 32 microns
Average Plant Data (PAT)	8.0	41.5	86.1%
Average AMMTEC Data	8.6	41.9	85.6%

Considering the differences in operating conditions and feed, the agreement between the two data sets is acceptable. For all of the M10000 product size distributions measured during the PAT period the average percent passing 32 microns was 86.1%. The average percent passing 63 microns was 98.1%. With an extrapolated P_{80} of 27 or 28 microns that gives a P_{98}/P_{80} ratio of about 2.33. The ratio for the three AMMTEC PAT comparative tests averaged around 2.3. The percent passing 75 microns averaged 99.1% in full scale surveys. ALS AMMTEC test C at pass 3 had a P_{80} of 26 microns and 99.5% passing 75 microns. Test B was a P_{80} of 28 microns and 99.6% passing 75 microns. The slight discrepancy between full scale and the M4 is likely due to the larger media and longer residence time in the M4. Consequently, the M4 produced a sharper size distribution once it reduced the feed to the ~50 micron P_{80} size that was feeding the M10000. It should be noted that test A is not included in these comparisons simply because none of the passes were at sizes directly comparable to individual survey results.

Average comparisons of the full feed and product size distributions between the M4 and M10000 for this test are shown in Figure 7 (Bandarian, 2014).

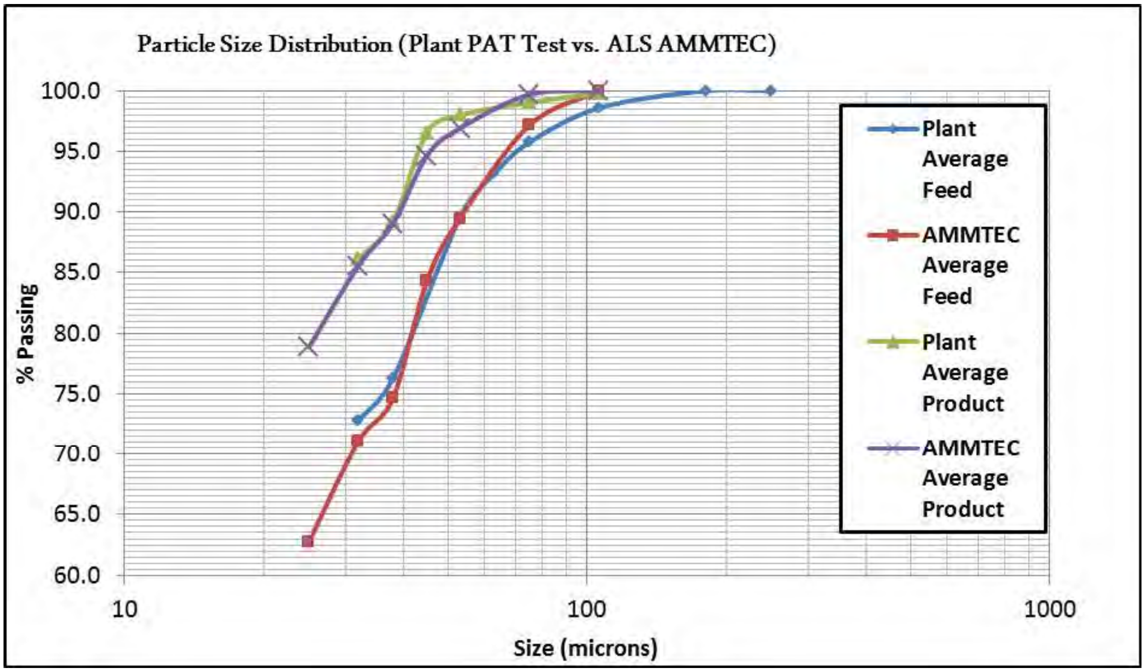


Figure 7 – M4 and M10000 complete size distribution comparisons.

MILL CONFIGURATION OPTIMIZATION

As is standard commissioning and operating practice, temperature checks are regularly taken along the IsaMill shell. These are tracked to monitor for compression and build-up of heat in the mill. Through the first two months of commissioning it was clear that there was more compression at the shell than is recommended for optimum mill liner life. Different mill configurations were implemented to improve the fluidization inside the mill and reduce shell liner and disc wear. Through small incremental changes the ideal mill setup was found. This spread out the temperature profile in the mill and also acoustically gave a more fluid sound inside the mill. Full mill operation with minimal internal wear was realized in early February 2014, about 3 months into commissioning and well ahead of the June deadline to reach full run time and throughput.

The main indicator of this success was the condition of the mill when opened for inspection, shown in Figure 8 (Ziki, 2014). After the optimisation phase wear of the discs and shell liner was found to be more consistent across all surfaces leading to an increased maintenance interval with better predictability of spares use.

Figure 8 – IsaMill maintenance opening February 2014.

Currently the IsaMill is inspected every 6 weeks in the line with the rest of the plant. Having the IsaMill maintenance line up with the rest of the plant was critical to ensuring a minimization of disturbances to the pellet plant feed, both in terms of grade and throughput.

ISAMILL SCALE-UP AND OPTIMISATION

Once it was confirmed that the mill was achieving the design grind duty and required maintenance interval, work began to fully optimize the operation. An M4 IsaMill was once again used, this time at the University of Queensland. A short survey of the IsaMill was performed to collect sufficient feed sample for multiple M4 tests. A graded charge of media from the M10000 IsaMill was shipped to Brisbane to perform test work. Product was sampled to determine the discharge sizing by sieve screens. The average results of the full scale survey are shown in Table 2.

Table 2 – IsaMill feed and product sizes, February 2014 survey for M4 scaleup (Villadolid, 2014).

Size µm	**%Passing Feed**	**% Passing Discharge**
106	98.3	99.9
75	94.3	99.5
53	87.3	97.6
38	61.3	84.9
25	52.7	74.6

The media size of 4-5 mm was initially decided upon by IsaMill rules of thumb as it was thought to give the best product size distribution for the 32 micron P_{80} of product and ~50-60 micron F_{80} design feed at a higher than average expected feed rate. At the University of Queensland a GT engineer tested this media against a laboratory 3.5 mm charge and a 5-6 mm charge that had been used at Ernest Henry Mine in their much coarser M10000 magnetite tails operation (Villadolid, 2014).

The M4 at UQ accurately predicted the required Arrium M10000 energy with 4-5 mm media, although the signature plot line shown in Figure 9 needed to be extrapolated due to the fine first pass produced in the M4 (Villadolid, 2014).

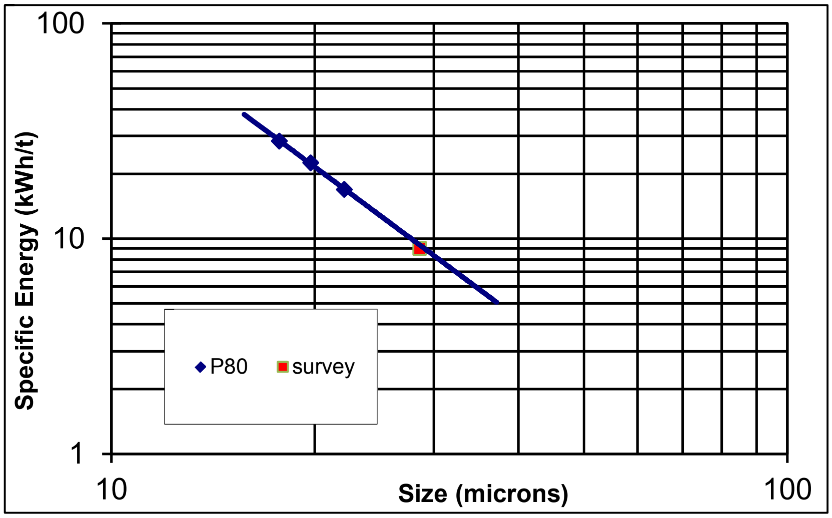


Figure 9 – IsaMill M4-M10000 scale-up.

When comparing the different media size options it would appear that the current 4-5 mm is correct for the full scale application. To a P_{80} of 32 microns the 4-5 mm media required 7.2 kWh/t, the 5-6 mm 11.2 kWh/t and the 3.5 mm 7.6 kWh/t. Due to the superior product size distribution inherently produced by the larger 4-5 mm media compared to any smaller media sizes, and its ability to cope with potentially coarser feed, it was recommended that Arrium continue to use the current size of ceramic media.

Satisfied that the media size was appropriate, a program to push the IsaMill well past its design maximums was implemented in April/May of 2014 just over two months after commissioning was officially ended. Mill feed density was slowly increased while monitoring the mill for proper fluidization and compression. It was found the mill slurry density could be increased from 1.64 to 1.68 kg/l in this period, increasing the solids feed rate by about 20 t/h at a constant flow rate. Next the feed flow rate was increased while monitoring the mill media load to be sure rotor pumping capacity was not exceeded. The flow was gradually increased until it reached 400 m^3/hr. The end result was an increase in tonnage from about 300 tph to 350 tph. Through this period the P_{80} increased slightly along with the feed size but the product top size and Davis Tube silica were barely changed, increasing by less than 0.5% silica.

One realization during this trial was that the Derrick Screen oversize typically contained less than 4 t/h of material that was actually coarser than 63 microns (the oversize mass was much more than 4 t/h but the majority of it was misreporting fines). In essence, the Derrick screens were underutilised because the IsaMill does not leave much +63 µm in the product. It was possible to use the excess capacity in the Derrick Screens and final grade to process additional material equivalent to IsaMill feed, while still processing the full flow if IsaMill product. The Derrick screens control the top size reporting to the cleaner magnetic separators and all oversize is returned to the ball mill cyclone feed rather than the IsaMill feed. In this way total tonnage for the plant could be increased further without exceeding the silica grade limit and without coarsening the product P_{80} significantly. Typically the liberation from the IsaMill even when flow constrained was above what was necessary. With a bypass line installed to the Derrick screen the plant could increase total new feed to the HPGR and ball mill. The coarsest fraction containing the most silica that bypassed the IsaMill would be removed by the Derrick screens, providing the necessary control over silica grade.

CURRENT OPERATING CONDITIONS

From May of 2014 until March 2015 the IsaMill was treating an average of 350 tph while the design maximum tonnage was 300 tph. Mill power draw has run consistently at 2,600 kW. Design disc consumption was predicted to be 26 per year. It is currently under 14 discs per year and continues to drop as the IsaMill Information System (IMIS) program provided by GT receives more information to better predict and utilize the full life of a disc. Though accelerated wear was experienced at the start of operations, current indications predict a liner life of 8-12 months, which is about on budget. Arrium has taken on a methodical approach to moving the shell liner every second mill opening to spread wear to all available rubber areas. Availability for the IsaMill has been excellent, with almost all downtime related to issues outside of the IsaMill circuit. Mill stability has been steady due to consistent upstream operation. The yearlong average for the Cenotec grinding media wear has been 6 g/kWh, versus a budget of 10 g/kWh. The media has worn smooth and round as shown in Figure 10, also contributing to improved disc and shell liner wear.

Figure 10 – Arrium IsaMill graded media charge.

Future work

Beginning in March of 2015 the IsaMill internal configuration was changed again to allow more flow through the mill. This brought the IsaMill to 485 m3/hr of flow and about 425-435 t/h of solids. As of the writing of this paper analysis continues into the sustainability of this production increase. Publications on progress will be made available in the future.

CONCLUSIONS

This project has seen the magnetite operation undergo an overhaul, allowing it to produce magnetite concentrate using a lower grade feed material at an increase in production capacity of greater than 400,000 tonnes per year. (Arrium news webpage, 2014) This throughput has continued to be expanded as the maximum capacity of the IsaMill has been pushed well beyond the original design. After a challenging start, handover of the plant was completed on budget and ahead of schedule. Through methodical implementation procedures, step by step improvements in throughput and mill wear have been realized allowing the plant to process tonnage well beyond anything for the original design.

ACKNOWLEDGEMENTS

The authors would like to thank Arrium Mining and Glencore Technology for permission to publish this work. We would also like to thank ALS AMMTEC, AMEC and GRES for the lab testing, design work and engineering for this project.

REFERENCES

Arrium news website, Accessed 4/1/2014, http://www.arrium.com/about%20us/latest%20news/fy14/major%20project%20complete%20%2024%20feb%202014

Bandarian P, 2014. Arrium Commissioning Report. Internal and to client.

David D, Larson M, Li M, 2011. *Optimising Western Australia Magnetite Circuit Design*, Proceedings of 2011 AUSIMM MetPlant Conference.

Ladhams I, 2012. February 10, 2012 ALS AMMTEC IsaMill Signature Plot Test Report.

Mativenga M, Stanton C, Jenkins C, Gerber J 2015, in press. *Unlocking Plant Capability through Targeted Debottlenecking of Arrium's Magnetite Concentrator*, Proceeding of 2015 AUSIMM Iron Ore conference.

Steele L, 2011. March 10, 2011 ALS AMMTEC IsaMill Signature Plot Test Report.

Villadolid V, 2014. Arrium M4 Scaleup Report (internal).

Ziki T, 2014. Arrium IsaMill Maintenance Report February 7, 2014.

Iron Ore 2015 Conference

Paper Number: 111
Unlocking Plant Capability through Targeted Debottlenecking of Arrium's Magnetite Concentrator

C. Stanton[1], M. Mativenga[2], J Gerber[3], C Jenkins[4]

1.
Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: chris.stanton@arrium.com

2. Process Development Manager, Arrium Mining, PO Box 21, Whyalla, SA. Email: martin.mativenga@arrium.com

3.
Principle Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: johan.gerber@arrium.com

4.
Senior Process Engineer, Arrium Mining, PO Box 21, Whyalla, SA. Email: Clinton.jenkins@arrium.com

ABSTRACT

In 2014, Arrium completed an optimisation program on its magnetite concentrator, enabling a reduction in the overall cost base of producing concentrate through an increased ability to treat a wider range of ore types, previously considered waste. The development was the culmination of four years of work, conducted in house by Arrium's mining development group. The overall intent of the plant design focused on improving the concentrator's ability to treat a wider range of magnetite ore types. Metallurgical analysis of these feed ore groups outlined a need for more grinding power if production quality was to be maintained and throughput increased. As such, a tertiary grinding circuit was designed which, alongside the rectification of key operational bottlenecks, would allow the plant to continue to produce concentrate at the required specification and rate. The technology chosen for the tertiary grinding application was the XT IsaMill^{TM}, which is known for efficient fine grinding when compared to conventional ball milling. The transfer sizes between the existing comminution circuits were coarsened, allowed for the treatment of different ores at the required rates. The optimisation project was successfully completed in February 2014 after a three month commissioning phase. The initial metallurgical testing and engineering of new equipment proved highly successful with all equipment meeting or exceeding design parameters.

INTRODUCTION

Arrium operates a magnetite concentrator at its Iron Duke mine, located 67km south west of the township of Whyalla, South Australia. The concentrator was commissioned in 2007 as part of Project Magnet, a successful initiative to convert the Whyalla pellet plant from hematite feed to magnetite. Operation of the concentrator in subsequent years highlighted several areas of improvement in the original plant design. A project was commenced in 2009 to investigate, design, and construct solutions to optimize the plant. In addition, stockpiles of high silica material previously disregarded presented a business opportunity if made treatable by the upgraded flowsheet. The Magnetite Optimization Project (MOP) represented a body of work consisting of four years' work in plant design, working with an array of highly experienced subject matter experts, to develop, test and construct an upgraded flowsheet for the processing of Arrium's magnetite reserve. This paper will discuss the identification of problem areas, development of key process solutions and finally, the implementation measures that occurred.

ORIGINAL FLOWSHEET

The original Project Magnet flowsheet consisted of two stages of grinding with low intensity magnetic separation following each comminution circuit. Crushed ore (-32mm) was fed to two parallel 1.8MW High Pressure Grinding Rollers (HPGR) in closed circuit with 2.2mm banana deck screens. Undersize from the screens was slurried and pumped to the Rougher Magnetic Separators (RMS). Rougher concentrate was then screened at 0.5mm, with all oversize going back to the HPGR, in an attempt to liberate and reject as much non-magnetic material as possible. Minus 0.5mm was ground down to an 80% passing size (P_{80}) of 38µm within the 7.5MW ball mill. The ball mill was closed out with hydro cyclones, with the overflow gravity discharging to a bank of De-slime Intermediate Magnetic Separators (DIMS). The DIMS concentrate was fed to the Cleaner Magnetic Sseparator (CMS) before passing through a protection screen prior to the slurry storage tanks.

INITIAL CIRCUIT ANALYSIS

In 2009, Arrium began an optimization program on the concentrator plant to improve production rate and reduce cost. An initial review was conducted on plant performance in conjunction with Process Technology and Innovation (Baguley, Jankovic and Valery, 2010). It comprised of metallurgical surveys, assay mass balancing, and historical data analysis. This review provided the following main conclusions:

- The HPGR circuit was power limited at harder ore types; restricting fresh feed due to the recycling load rising beyond conveyor limitations.
- At 0.5mm, the HPGR close out screen size was causing excessive front end recycling load, limiting throughput.
- The ball mill was shown to be overgrinding the heavier magnetics, reducing efficiency.

The existing cyclone feed pumps were operating at the limit of their duty due to the preferred cyclone configuration at the time.

CHANGES TO THE FEED: BUSINESS OPPORTUNITY

During the initial years of processing magnetite, a significant amount of high silica magnetite was stockpiled. This material was not treated through the concentrator due to final product quality concerns that existed at the time.

The Geometallurgy group developed a modified Davis Tube Recovery (DTR) process, with the results able to be translated into expected plant performance utilising a regression model. Laboratory test work conducted on these high silica magnetite stockpiles in 2011-2012 indicated that the material achieved good metallurgical performance when ground to the target 80% passing size of 38µm (P_{80}) of the concentrate. The test work also indicated that due to the increased quartz content, the material was harder than the traditionally dominant carbonate magnetite usually treated by the plant. This material also had a lower mass recovery when compared to the average magnetite feed blend. These two factors together would require an increased capacity in the concentrator grinding circuit to maintain the required final product production rates.

Full scale plant trials were conducted, progressively increasing the ratio of high silica feed to standard feed, ranging from 10% through to 30%. Initial results on the trial date blend showed encouraging results in the concentrate quality that was produced. During subsequent plant trials feeding up to 100% high silica material, it was shown that while the concentrator could achieve the target grind and an acceptable concentrate quality, production was significantly reduced due to lower mass recovery. It was also found that the final product quality was sensitive towards the grind achieved, with any variation from the target grind causing the product quality to fall outside the required levels.

This added impetus to the notion of conducting a major debottlenecking exercise as the most applicable means of increasing production.

EARLY BOTTLENECK IDENTIFICATION

Prior to commencing a design study with an engineering company, Arrium conducted an internal review of the plant bottlenecks identified previously by PTI and other subject matter experts. In order to increase the feed rate beyond the design these key bottlenecks would need to be identified first to save time later on.

HPGR Circuit

At the time of review, the feed conveyors around the HPGR units were running at near maximum capacity. Each HPGR was tasked with reducing 1000tph of -32mm feed material down to below 0.5mm, taking into account the fine recycle from the RMS screens. The plant operators could not physically increase the circuit raw feed rate without exceeding the capacity of the HPGR feed conveyor belts. Prior optimisation work to increase the recycling load was recommended by PTI as a means of increasing the utilisation of installed power (Baguley, Jankovic and Valery, 2010), which should also increase the proportion of fine material being generated by the circuit (Dundar et al, 2009). However, a site review by the HPGR vendor, Koeppern, noted that the amount of fines in the HPGR feed was already well above the original design specification (Gardula, 2010). It was also noted that there was up to 40% -2mm in the HPGR feed, as opposed to the 15% catered for in the origin design calculations. The inability to process more tonnage because of the belt limitations and recycling load was a major conclusion to this stage of the review.

Ball Mill Circuit

From metallurgical surveys and reviews, it was established that the recycling load around the ball mill was industry average at 250%, but that the cyclone feed pumps were operating at their limits. This data led to the conclusion that the mill could be optimized for higher throughput or coarser grind size, and that attention should be instead paid to cyclone pump capacity.

An in house review of the cyclone pump capacities confirmed that both cyclone feed pumps were not adequate for the volumetric duty required to achieve the 38µm grind. The actual static head on the system at the time was 17 metres, as opposed to the original design of 13.8m (Phillips, Westbrook, 2010). This was further compounded by a back pressure of 166kPa, as opposed to the original design of 100kPa. Because of these disparities, the overall static head increase was 9.2m over the design, and the pump could not be expected to reach its design duty rate (Phillips, Westbrook, 2010). From the review, the calculated sustainable volumetric flow rate was 712m^3/hr, far less than the original mass balance rate of 1114.5m^3/hr.

For the plant operators, this translated simply into a lack of capacity to control product quality if the feed ore was harder than normal, or an inability to increase feed rate when the opposite was true.

OPTIMISATION APPLIED TO INITIAL SCOPING DESIGN

While optimization of the existing concentrator concluded with PTI at the end of 2011, work had already began within Arrium creating multiple base flowsheets upon which a feasibility study could be developed. These flowsheets would target the bottlenecks identified and include, where possible, proposed remedies based on the best available information at the time. Major changes put forward for engineering design were:

Removal of the RMS screens

This would coarsen the feed 80% passing size (F_{80}) to the ball mill from approximately 300µm to 1400µm. Fresh feed tonnes into the HPGR circuit would also be increased to make up the recycling load.

To check the ball mill capacity, a simulation of the ball mill performance at the optimised conditions was carried out, and only through significant variance in the design parameters did the model predicted power requirement exceed the installed capacity. An internal report outlined that with the original design 80% particle passing size (F_{80}) of 416µm and product 80% passing size of (P_{80}) of 38µm, the calculated installed grinding power ranges from 5800kW to 7250kW. In essence, there was more than enough installed power to achieve the original grinding requirements, yet as the resultant mill grinding efficiency was lower than anticipated the mill represented a bottleneck on the plant. A paper written by Partyka outlined the limitations of ball milling when fine grinding, noting that below 20 to 30µm, they become inefficient (Partyka, Yan, 2007). Of particular interest, the paper also showed a worldwide trend against using ball mills in a regrind application with similar parameters to Arrium's operation (Table 1).

This hypothesis was later re-iterated with ongoing design work with Amec. The ongoing feasibility study also pointed out the inefficiency within the mill could be attributed to its very fine grinding duty and wide reduction ratio (Lilford, Nofal, 2012). As such, the tonnage and grind reduction ratio required at Arrium's concentrator was not conducive for any increase in throughput, and the design focused on coarsening the mill and increasing grinding efficiency.

Installation of a new grinding mill

To cater for the throughput increase and coarsened product from the ball mill, an additional grinding circuit was required. For the size range being considered (-100µm), stirred milling technology was deemed appropriate for consideration. The two possible mill configurations considered suitable for the duty were the Metso VertiMill and XT IsaMill^{TM}. The early steps involved in the data collection for the decision making were:

1. Visit to Arrium by Metso and Xstrata engineers,
2. Bench-scale testwork on each technology – samples were sent to independent laboratories for grinding testwork with both technologies. This included multiple tests with the XT M4 lab scale IsaMill,
3. Communication and visits to reference sites – a number of reference sites were visited, including Ernest Henry Mine in Queensland and McArthur River Mine in the Northern Territory.
4. Advice sought from independent subject matter experts,
5. Literature review of laboratory mill technology, focusing on the grinding performance and successful scale-up.
6. Construction lead time of equipment, and what auxiliary capital is required (for example, cyclones).

Criteria from the data collection, such as capital and operational expenditure, lead times, and risk were weighed in a selection workshop. From this, the XT IsaMill^{TM} was chosen as the preferred technology to move into feasibility design. For the duty, a single 3MW M10,000 IsaMill^{TM} was required. By the start of 2011, Arrium commenced a feasibility study into the new plant design, with Amec selected as engineering partner.

FEASIBILITY – CONFIRMING ASSUMPTIONS

With the scoping study concluded, the feasibility phase of design aimed to provide a revised flowsheet and capital estimate to +/-25% accuracy.

Flowsheet Simulation Model

In order to mitigate the risk of the flowsheet changes and new equipment, a JKSimMet model was developed simulating the proposed changes on the plant. In order to fulfil this task an extensive metallurgical survey of the concentrator was conducted, including obtaining a bulk sample of the RMS feed. The bulk sample would be processed by the selected lab on bench scale equipment as per the proposed flowsheet steps, including the IsaMill stage. Bond work indices were also conducted on the nominated size range to verify the mill power required. As a final measure for risk mitigation, three separate plant surveys were conducted on different dates to ensure some variance in feed mineralogy. Three bulk samples were in turn submitted for analysis to the laboratory to run through the testwork program independently.

To ensure adequate model comparison against the actual industrial sized concentrator, the metallurgical surveys sampled every major unit of operations within the plant. All samples generated from the lab scale and full scale surveys were subject to identical analytical tests to ensure comparative data. As the M4 IsaMill was situated at ALS Perth, this laboratory was selected to conduct the majority of the analytical work, with the aid of vendors for specific equipment tests.

Results Analysis

The flowsheet simulation program successfully verified many of the assumptions carried through the feasibility study. The model and the final flowsheet design were in turn updated ensuring the rigour of third party review was continued. It was concluded from the testwork program that the results substantiated the flow sheet selection, successfully mitigating a significant portion of the process risks. In turn, an independent third party review conducted by PTI (Baguley, Jankovic, 2012) outlined no major issues or problems with the flowsheet.

A high level comparison of the grinding transfer point changes between the original plant and that at the end of feasibility is shown below in figure 1.

FINAL DESIGN AND CONSTRUCTION

Leighton Contractors (LCPL) and GR Engineering Services (GRES) were selected as final design and construction engineering partners. A final review of flowsheet and testwork results were conducted by GRES, with a few key changes made. These included modifications to the final screening plant, including reducing the complexity from two stages to one, and installing Derrick StackSizer^{TM} screens for greater efficiency. Another notable change to the design was the recycle of the oversize from the screening plant back into the process, as opposed to sending it to tail. Construction commenced in 2013, with LCPL working in conjunction with Arrium's Major Capital Development Group. All brownfield construction was completed within the monthly 24 hour production shutdowns leading into November, upon which a 10 day production halt enabled the completion of all tie-ins and pre-commissioning activities.

THE FINAL FLOWSHEET

The magnetite optimization project targeted multiple areas of the existing plant to install additional capacity, and while this paper focused primarily on the grinding circuits, there were numerous other upgrades that were included and not discussed. Figure 2 is a depiction of the improved flowsheet as constructed and commissioned in 2013.

COMMISSIONING

Transitioning the project from design to construction, commissioning, and finally to plant handover was managed through the use of risk management and coordination tools. During the risk management process, all risks associated with the project were classified, ranked and prioritized with timed action plans and performance monitoring processes put in place (figure 3).

The transition and coordination steps involved bringing all stakeholders together to review construction progress and map out when the different phases of commissioning would commence. The overall goal was to ensure the plant was handed over to the operational owner as soon as possible.

A Management Operating System (MOS) was put in place to ensure all stakeholders were able to attend and give input into the required progress review meetings. The project progressed into process commissioning by mid November 2013, with performance acceptance tests successfully concluded in mid-December, 2013. By the beginning of January, the plant was formally handed over to operations and the ramp up continued.

During the first two months from commissioning, process optimization was undertaken to bring the plant to design production rates. The work included optimisation of the IsaMill^{TM} wear components to maximise throughput and grind. Plant surveys and sampling campaigns were conducted to validate the mass balance and design assumptions. Classification (ball mill cyclones and screening) optimisation was also performed to achieve the required concentrate grade. The optimisation work under taken resulted in the process operating at above 10% of the design throughput prior to concluding the commissioning phase.

The performance monitoring system put in place to transition from construction to project handover was done successfully, with seamless transition between the different phases of the project delivery model. This success also led into an above target ramp up process after commissioning, shown in figure 4.

RESULTS

Measurement of plant capability post commissioning was demonstrated by the newfound ability to process lower feed magnetic content ores while maintaining production rate. The following parameters have been assessed since commissioning:

- The removal of the RMS screens resulted in a 98% reduction in HPGR recycling load, resulted in an increase of up to 25% in fresh feed rates, depending on the feed magnetic content,
- The de-constraining of the ball mill circuit and installation of the IsaMill has allowed for the processing of the coarser HPGR product, while maintaining design throughput rate and maintaining final grind of 38µm.
- Average concentrate pumping rate has been increased by 6% on the previous financial year, however 15% additional throughput has been demonstrated on higher mass recovery material.
- Finally, the plant has been successfully treating feed blended with 30% silica material successfully since commissioning. This was not achievable prior to the project completion.

CONCLUSIONS

The targeted de-bottlenecking of the magnetite concentrator has achieved its overall goals. The desired feed changes have been incorporated into the production budget and the plant is more than able to maintain grade and throughput.

ACKNOWLEDGEMENTS

The authors would like to acknowledge firstly Arrium Mining Pty Ltd for the approval to publish this paper. Secondly, to all the subject matter experts over the life of the project who have lent their expertise to allow us to achieve our goals.

REFERENCES

Baguley P, Jankovic A, Valery W, 2010, Preliminary Data Analyses of the Whyalla Magnetite Concentrator (unpublished), Metso Process Technology and Innovation.

Baguley P, Jankovic A, , 2012, Review of proposed MEP flowsheet and mass balance (unpublished), Metso Process Technology and Innovation.

Dundar H, Aydogan N A, Benzer H, Lynch A J, 2009, Effect of Recycling Load on the High Pressure Grinding Rolls' Performance – A Case Study with the Industrial Unit During Cement Clinker Grinding, Proceedings 10th Mill Operators' Conference, p161-166, The Australian Institute of Mining and Metallurgy

Lilford B, Nofal P, 2012, Magnetite Expansion Project FEL3 Study Report (unpublished), Amec Australia Pty Ltd

Partyka T, Yan D, 2007. Fine Grinding in a Horizontal Ball Mill, in WA School of Mines, Curtin University

Phillips G, Westbrook K, 2011. SMR Concentrator Optimisation Study (unpublished), OneSteel Manufacturing Pty Ltd.

FIGURE CAPTIONS

FIG 1 Grinding transfer point comparison between old and new plants

FIG 2 – Revised flowsheet, as constructed

FIG 3 – Risk Management Process

FIG 4 – Ramp Up Curve

TABLE CAPTIONS

TABLE 1: Examples of fine grinding ball mills (Partyka, Yan, 2007).

FIGURES

FIG 1 – Grinding transfer point comparison between old and new plants

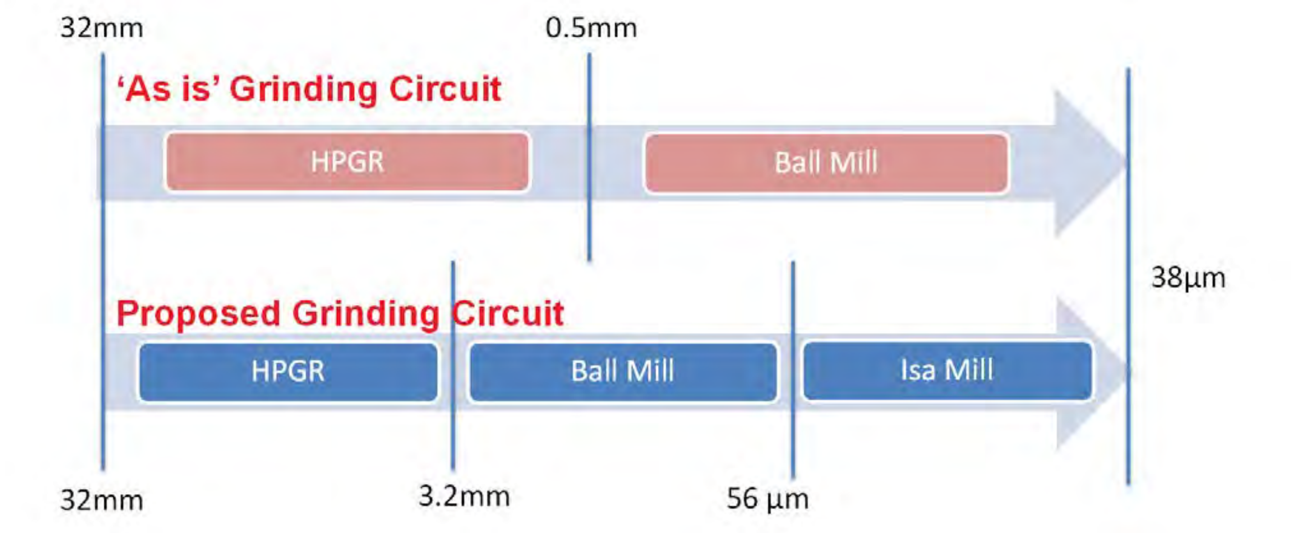


FIG 2 – Revised flowsheet, as constructed.

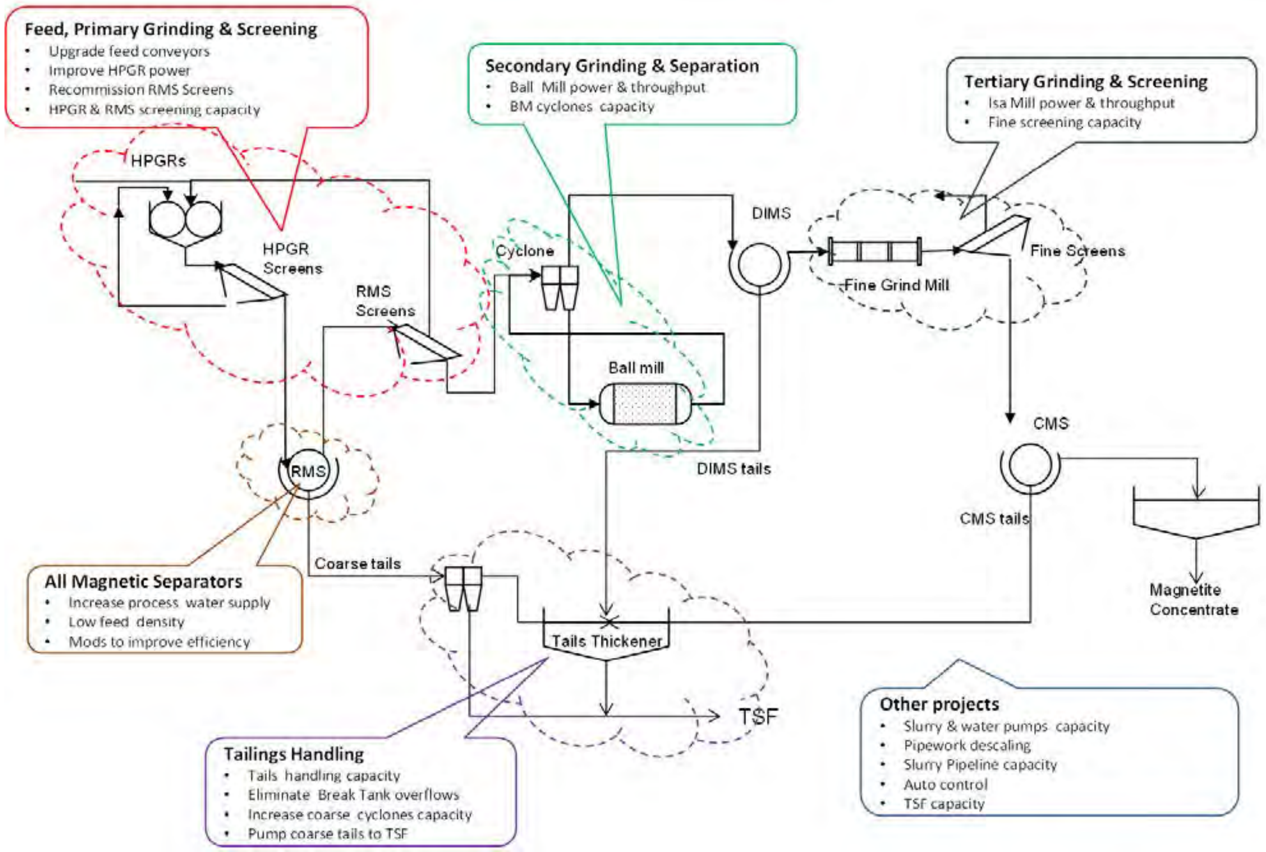
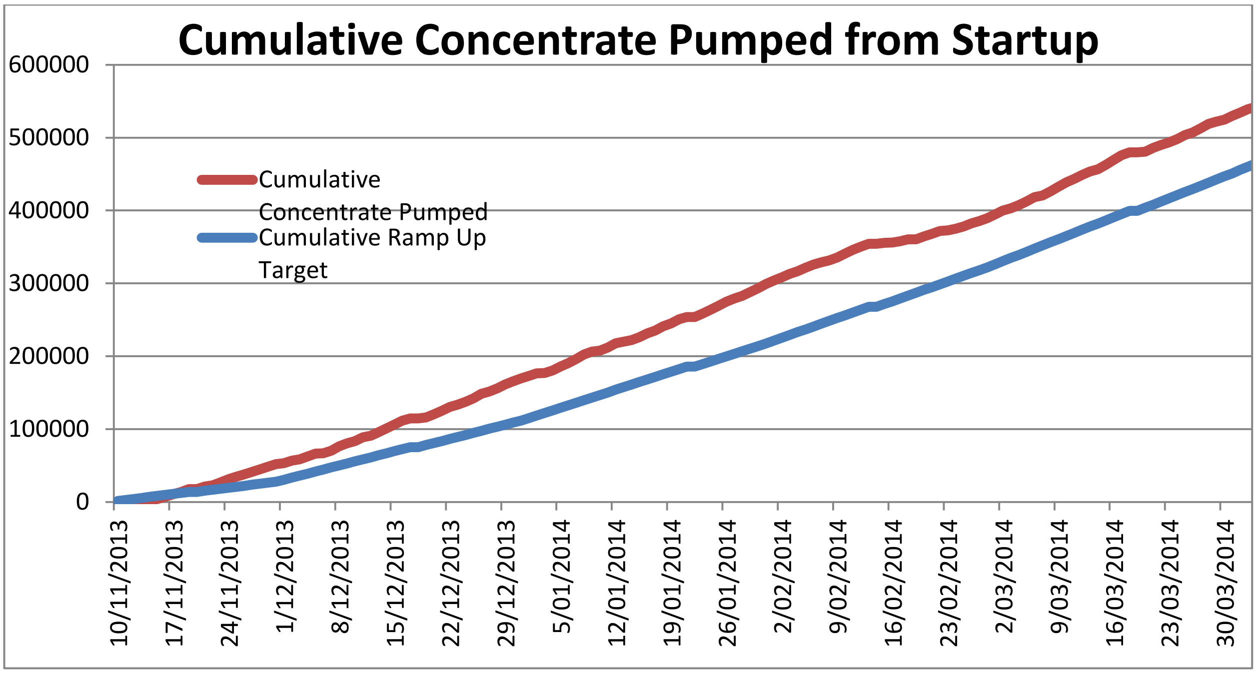


FIG 3 – Risk Management Process.

FIG 4 – Ramp Up Curve.



TABLES

TABLE 1: Examples of fine grinding ball mills (Partyka, Yan, 2007).

	Particle Size		**Mill Dimensions**	
Site	**Feed Size μm**	**Product Size μm**	**Dia (m)**	**Length (m)**
Pajingo Gold	200	38	3.66	4.18
Germano Iron Ore	120	32	5.18	10.36
Savage River	140	43	3.9	8.8
Macraes		20	3	8.2
Pena Colorada	125	38	5	10.67
Beaconsfield		20	1.83	2.44
Tritton Copper	45	30	2	3.4
Brunswick Mining	30	25	3.2	4
Porgera Gold	106	30	3.05	4.27
Arrium	**500**	**38**	**6.1**	**11**

Using Geometallurgy during Process Optimisation Activities at the Southern Middleback Ranges Magnetite Concentrator

J Gerber[1], C Stanton[2] and D Olwagen[3]

ABSTRACT

From December 2013 to February 2014 the Southern Middleback Ranges magnetite concentrator went through a successful optimisation process that included the installation of extra grinding power to treat a wider range of ore types. After the initial improvements, sampling surveys were conducted to identify the mineral deportment through the concentrator. Samples collected during the survey were analysed using X-ray fluorescence, quantitative X-ray diffraction and QEMSCAN™ to cross-validate the analysis. Certain streams were also put through a laboratory-scale Davis Tube Recovery washing process to produce clean concentrates for analysis. The QEMSCAN™ analysis identified various metallurgical characteristics (mineral association, particle sizes, grain sizes, particle densities, etc) and these were evaluated for each separation and comminution stage to both understand the processes involved and identify opportunities for improvement. Key findings from the investigation indicated the importance of understanding the elemental deportment of target elements throughout the various minerals. The standard practice of tracking iron deportment through assay provided misleading results on losses to tail through the plant. The investigation provided insight into both wanted and unwanted minerals and the efficiency of each separation stage. The efficiencies of separation are critical due to their impact on downstream capacity, and the investigation provided insight into where opportunities exist to extract more benefit. The findings of the geometallurgical analysis have been implemented on the concentrator, and improvements in the rejection of non-magnetic minerals have been seen.

INTRODUCTION

The Southern Middleback Ranges (SMR) magnetite concentrator was commissioned in 2007 to produce feed for the OneSteel blast furnace in Whyalla, South Australia. The concentrator went through an optimisation project in December 2013, which included the installation of a M10000 IsaMill™ and Derrick Stack Sizer™ screens to give it the ability to treat ore with higher silica content.

Magnetite ore is mined from the Iron Magnet pit and then crushed and stockpiled for feed to the concentrator. The stockpiled feed is fed to the high-pressure grinding rolls circuit that produces a coarse grind feed for the first magnetic cobbing stage. The magnetic concentrate then reports to the ball mill for another stage of grinding. The ball mill product goes through a classification stage before reporting to the second stage of magnetic separation. The magnetic concentrate from the second stage reports to the newly installed IsaMill™. The finely ground product is passed through the newly installed Derrick Stack Sizer™ screen system before reporting to the final magnetic cleaning stage. This stage produces the final concentrate that is thickened and pumped down a 63 km pipeline to the filter and pellet plant located in Whyalla.

During the optimisation process, it was identified that the concentrator had increased levels of unwanted elements (predominantly silica) in the final concentrate. While metallurgical characterisation of the ore indicated that marginal increases in the concentration of silica could be expected, the levels seen in the concentrator were significantly higher.

Initial investigations started with a snapshot survey of the major processing streams during stable operation. The analysis of the snapshot surveys included X-ray fluorescence (XRF), quantitative X-ray diffraction (QXRD), QEMSCAN™ and Davis Tube Recovery (DTR) washing.

The findings from the snapshot survey indicated high levels of entrainment of non-magnetic minerals in the magnetic concentrates. Discussions with subject-matter experts confirmed that the efficiencies around the magnetic separation stages were lower than expected. The snapshot

---

1. MAusIMM, Principal Process Engineer, Arrium Mining, PO Box 21, Whyalla SA 5600. Email: johan.gerber@arrium.com

2. GAusIMM, Process Engineer, Arrium Mining, PO Box 21, Whyalla SA 5600. Email: chris.stanton@arrium.com

3. Metallurgy Team Leader, Arrium Mining, PO Box 21, Whyalla SA 5600. Email: dian.olwagen@arrium.com

survey also found that a significant quantity of non-magnetic iron was being rejected at the first stage magnetic cobbing.

The surveys indicated that the IsaMill™ improved the liberation of the silicates from the magnetite. This led to a more than 30 per cent reduction in the magnesium oxide (MgO) levels in the final concentrate.

These findings led to more detailed focus points around:

- first stage magnetic cobbing, with an emphasis on both the magnetic concentrate and non-magnetic rejects
- the IsaMill™ feed and discharge streams, with an emphasis on silicates liberation.

MINERALOGY

Mineralogical analysis helps to identify and understand the mineral deportment through a minerals processing plant. Understanding which minerals are reporting to the product stream and the reject stream can assist with optimising the efficiency of separation. It also provides understanding of quality constraints on the product stream and the losses of valuable metals to the reject/tail stream.

The current analysis method employed by Arrium Mining's Geometallurgy department uses a combination of measuring techniques to validate results across analysis methods. The analysis techniques that are currently employed are:

- XRF
- QXRD
- QEMSCAN™
- optical microscopy.

It was found that the combination of XRF, QXRD and QEMSCAN™ works well for samples that have a top particle size of 1–3 mm. Samples that contain particles larger than 3 mm are analysed with XRF, QXRD and optical microscopy. As discussed by Donskoi *et al* (2011), the QEMSCAN™ has trouble differentiating between minerals with similar chemical composition (such as hematite and magnetite). The use of other techniques improves the mineralogists' ability to differentiate these minerals when setting up the species identification profile.

Arrium has a good working relationship with the mineralogists at the Bureau Veritas laboratory in Adelaide. This has led to Arrium using the QEMSCAN™ iExplorer viewing software on-site to comprehensively investigate the analysis results. The use of the software has assisted greatly in reviewing metallurgical characteristics such as:

- grade-density profiles
- grade-particle size profiles
- particle shape analysis
- elemental deportment.

The biggest benefit associated with using the iExplorer software on-site is the ability to modify graphs as the investigation progresses and new avenues of investigation are pursued.

MAGNETIC SEPARATORS

Process improvement

Discussions held with Wennen (April 2014, personal communication) indicated that the weight per cent solids of the feed slurry to the magnetic separators was too high. Plant trials were completed and the feed slurry density was decreased in multiple stages to identify any improvements in the entrainment levels of non-magnetic minerals. The lowest density achievable was limited by the capacity of the water supply system. The samples taken during the trial were washed using a DTR method. The amount of non-magnetic material measured by this test was assumed to be the entrained material that should have been rejected. The results were reported as mass per cent magnetic recovery. Figure 1 shows the magnetic recoveries for the various trials. A lower magnetic recovery indicates a higher level of entrainment. The test work confirmed that at lower slurry densities, the entrainment decreases. The concept of lower feed density was embraced by the production personnel, who immediately reduced the density to the lowest possible level with the current infrastructure. A review is currently underway to optimise the water distribution system to allow more flexibility in controlling the slurry densities to the magnetic separation stages.

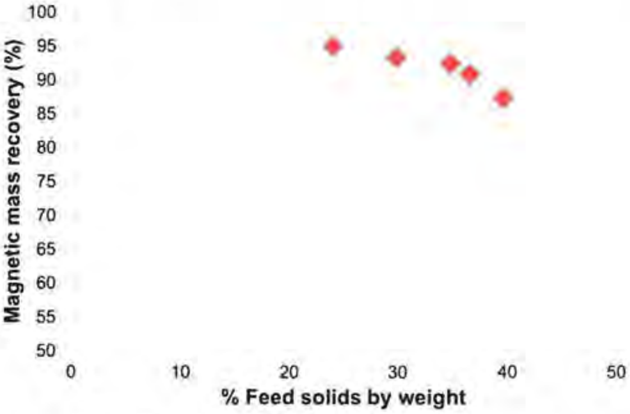


FIG 1 – The magnetic mass recovery for various per cent feed solids by weight for the first magnetic cobbing stage.

Mineral deportment

The samples taken during the trials were sent for mineralogical analysis to understand which minerals were being entrained, recovered or rejected at the magnetic separation stages. The feed to the various magnetic separation stages showed that the main mineral groups contributing to iron content were:

- iron oxides (magnetite, hematite) – 78 per cent
- iron silicates (chlorite, minnesotaite) – 14 per cent
- carbonates (siderite, ankerite) – seven per cent
- iron sulfides (pyrite, pyrrhotite) – one per cent.

The magnetic separation recovers any particle that contains magnetic minerals as long as the magnetic force is larger than any of the other forces being applied to the particle (drag force, gravity, particle interaction forces etc). Depending on the size and weight of a particle, only a small amount of magnetic material is needed for the particle to be recovered (Figure 2). From the minerals listed, it was found that only magnetite and pyrrhotite responded positively in the presence of the magnetic field strength used in the concentrator. This limited the iron recovered to those magnetic minerals plus any of the other minerals being entrained during the separation stage.

The magnetite-hematite association is such that a large portion of the hematite is recovered in the concentrate. The QXRD analysis (Table 1) showed hematite and magnetite levels as high as nine per cent and 61 per cent respectively. Other main minerals identified were quartz, talc (in the form of minnesotaite), dolomite/ankerite, siderite and chlorite.

Analysis of the tails indicated that the hematite lost was not associated with the magnetite but rather with iron silicates or as liberated hematite particles. The pyrrhotite was present

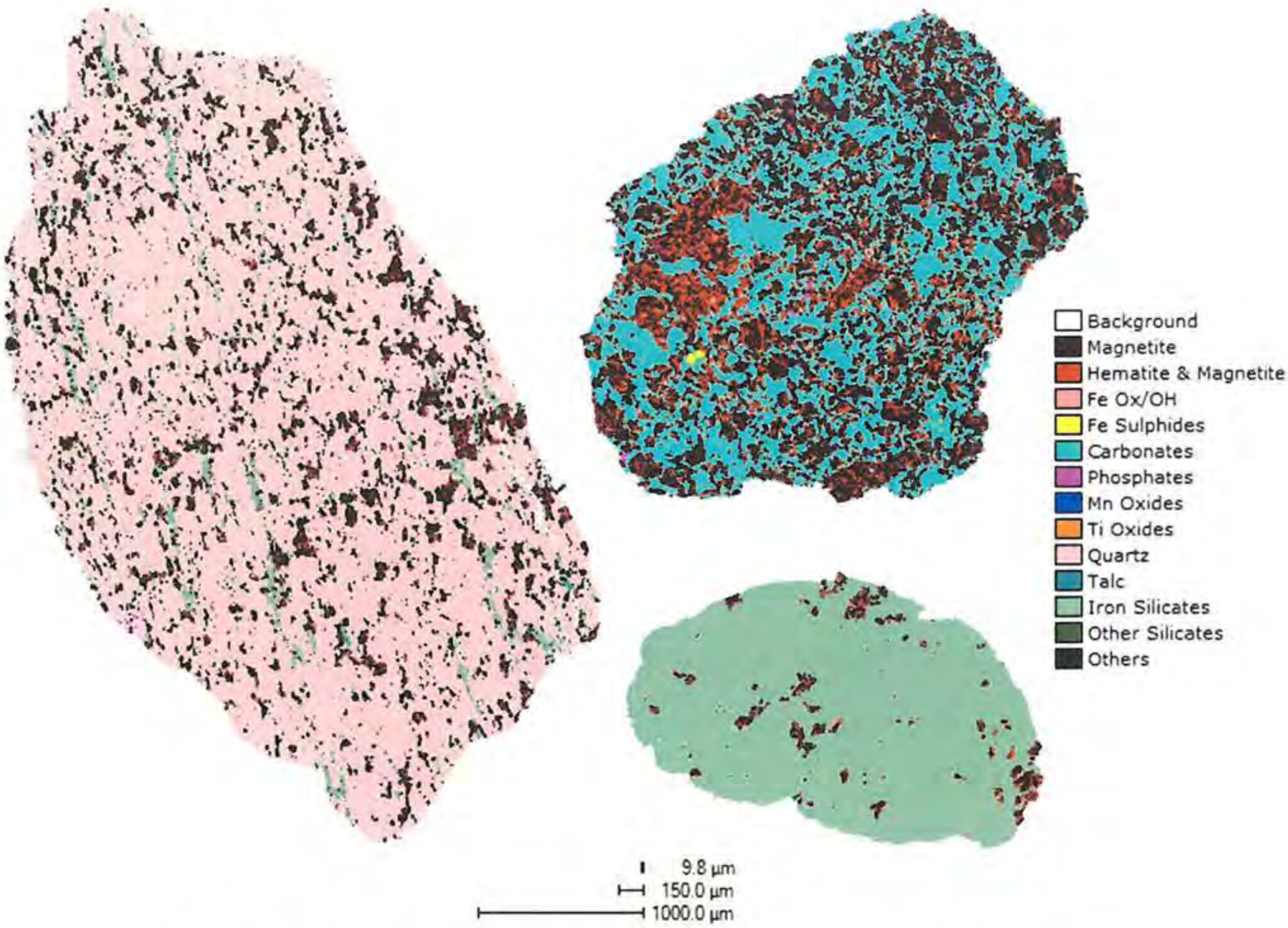


FIG 2 – Three magnetite concentrate particles from the first magnetic cobbing stage, each containing the three dominant non-magnetic minerals (quartz, carbonates and iron silicates) together with magnetite.

TABLE 1

Quantitative X-ray diffraction results from the surveys for the concentrate from the first magnetic cobbing stage.

Mineral	Composition	Survey 1	Survey 2	Survey 3	Survey 4	Survey 5
Quartz	SiO_2	10	16	11	11	12
Magnetite	Fe_3O_4	61	53	56	60	59
Hematite	Fe_2O_3	9	9	7	7	7
Chlorite	$(X_5Al)(AlSi_3)O_{10}(OH)_8$	5	5	4	3	2
Talc (minnesotaite)	$(Fe^{2+}, Mg)_3Si_4O_{10}(OH)_2$	4	3	10	8	11
Dolomite/ankerite	$Ca(Fe^{2+}, Mg, Mn)(CO_3)_2$	5	8	7	7	6
Siderite	$FeCO_3$	5	4	5	4	4
Sepiolite	$Mg_4Si_6O_{15}(OH)_2(H_2O)_6$	2	2	-	-	-
Total		100	100	100	100	100

Chlorite where X = Mg, Fe, Ni and Mn.

in very small quantities and, from the QEMSCAN™ particle view data, was typically rejected due to the particle sizes it was found in. The particle density and particle size for hematite present in the rejects from the magnetic cobbing stage is shown in Figure 3. Sixty-four per cent of the particles in the cobbing rejects were larger than 45 µm. These particles were rejected due to low magnetic content, but still contained significant amounts of iron (12 per cent by mass) at grades of around 60 per cent. Recovery of these particles with a gravity-based concentration stage is possible.

ISAMILL™

Process improvement

The original Project Magnet concentrator flow sheet consisted of two stages of grinding to produce 1.8 Mt/a of slurry concentrate at a grind of 38 µm (P80). This grind was necessary to ensure that the concentrate quality was within the blast furnace requirements; however, to meet ongoing throughput and grind requirements and increase the range of feed materials that can be processed, the optimisation study concluded that the plant required a third grinding stage to ensure ongoing compliance with concentrate specification. For the upgraded concentrator flow sheet, the installed M10000 IsaMill™ sits downstream of the existing 7.5 MW ball mill. With the installation of the IsaMill™, the throughput capability of the concentrator was increased by 20–30 per cent while maintaining the final grind and quality specification.

Mineral deportment

The IsaMill™ was installed as an energy-efficient tertiary grind stage to improve the liberation of magnetite from

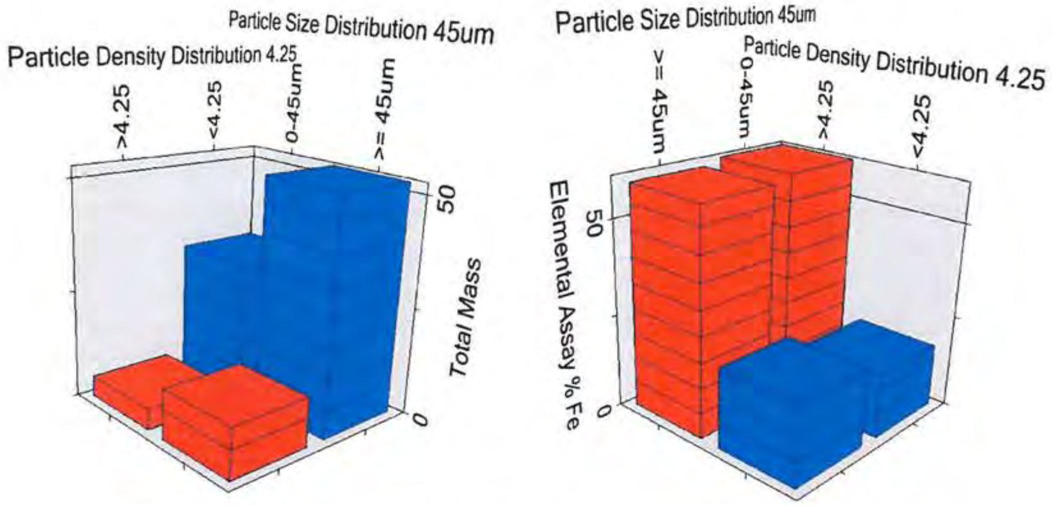


FIG 3 – Three-dimensional figures showing the relationships between particle density, particle size and both the mass distribution and iron assays for each grouping of the magnetic cobbing rejects.

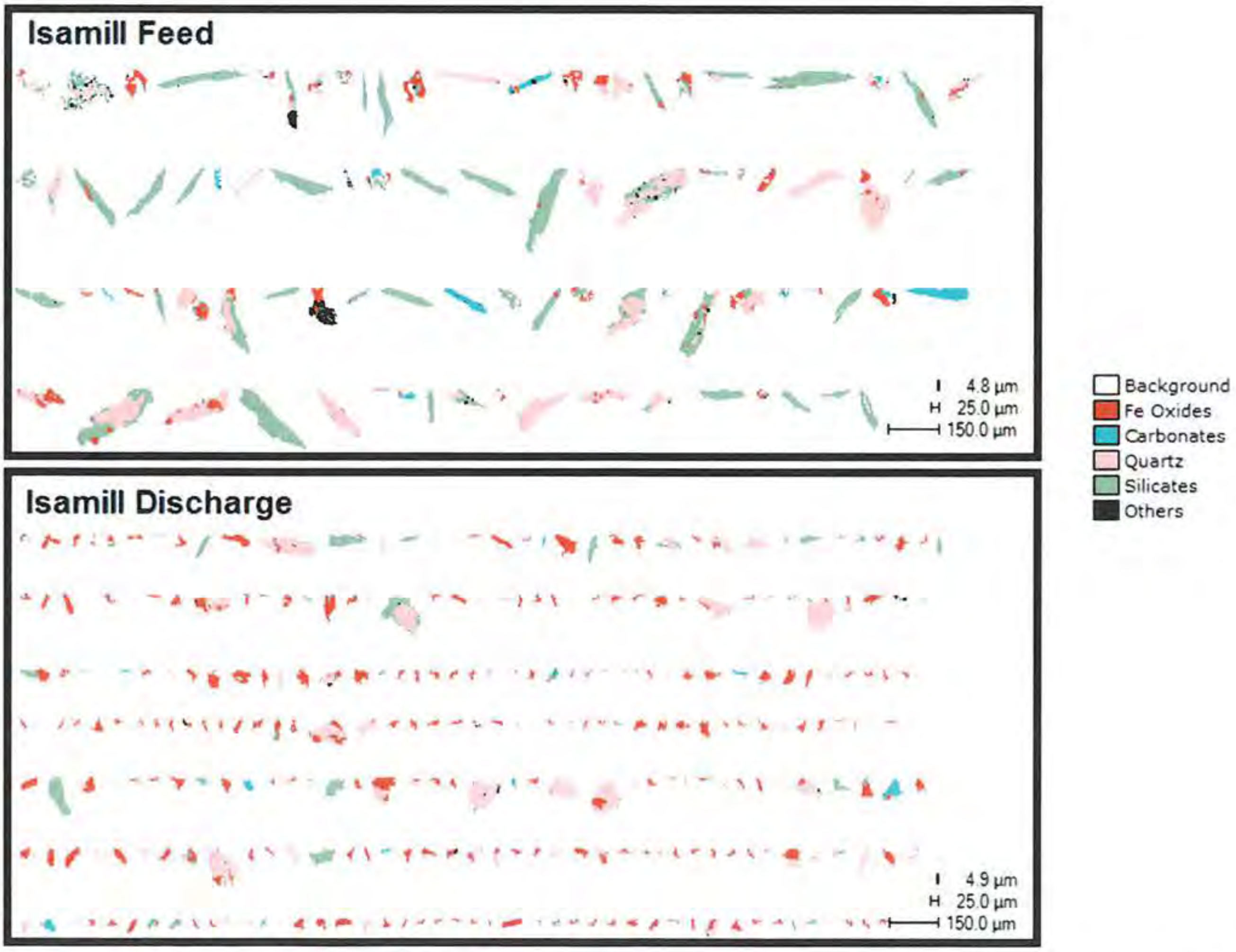


FIG 4 – Particle view report for the IsaMill™ feed and discharge streams arranged by the shape factor of the particles using the QEMSCAN™ iDiscover software package.

the other non-magnetic minerals, predominantly quartz, ankerite and talc. The samples analysed to identify the improved liberation from the tertiary grinding stage were the IsaMill™ feed and discharge. A DTR wash was completed on both samples as well as mineralogical analysis with the QEMSCAN™. The DTR wash in the feed showed a magnetic recovery of 84 per cent and the discharge was 78 per cent. The drop of 6 per cent in magnetic recovery indicates that more non-magnetic material was liberated from the magnetic particles in the IsaMill™ discharge.

The grade of the concentrate between the feed and discharge showed an improvement from 7.2 per cent to 4.9 per cent for silica and 0.9 per cent to 0.07 per cent for MgO. The improvement in MgO liberation is a significant improvement when compared to historical performance. Figure 4 shows how the elongated shapes of the IsaMill™ feed particles have been ground smaller and the amount of elongated particles present in the material reduced. The majority of the elongated particles are dominated by the 'silicates' mineral grouping

The improvement has led to adjustments in the fluxing philosophy at the pellet plant, which is delivering cost benefits by using cheaper dolomite, rather than limestone, to achieve the required flux levels.

CONCLUSIONS

Mineralogical investigations have proven quite useful for Arrium Mining. The use of analysis tools is expanding towards not just fault finding, but also to optimisation exercises and opportunity investigations. It is important to use multiple techniques to ensure that measurements and interpretations are representative. Mineral deportment analysis on the concentrator has identified potential iron recovery gains. These gains are currently being investigated for feasibility.

ACKNOWLEDGEMENTS

The authors would like to thank Mike Bannear for allowing this paper to be submitted and the groups involved during the optimisation process. A special thank you to Wade Hodgson and Barry Whittington from Bureau Veritas for taking the time to train one of the authors in using iExplorer.

REFERENCES

Donskoi, E, Manuel, J R, Austin, P, Poliakov, A, Peterson, M J and Hapugoda, S, 2011. Comparative study of iron ore characterisation by optical image analysis and QEMSCAN™, in *Proceedings Iron Ore 2011*, pp 213–222 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Материалы 42-ой Ежегодной конференции Канадского общества переработки минерального сырья - 2010

ДОКЛАД 14

“Нужное оборудование в нужном месте”: увеличение производительности по никелю на комбинате Cosmos компании Xstrata Nickel Australasia

Дэн Керри[1], главный управляющий комбината

Майкл Купер[1], управляющий комбината

Джош Рубенстайн[2], старший инженер-технолог

Том Шоулдис[3], президент

Майкл Янг[2], главный металлург

---

[1] Xstrata Nickel Australasia, Level 3, 24 Outram Street, West Perth, WA 6006
Тел. +61 (8) 9213 1588
dcurry@xstratanickel.com.au
mcooper@xstratanickel.com.au

[2] Xstrata Technology, Level 5, 509 Richards Street, Vancouver, BC V6B 2Z6
Тел. (604) 699 6402
jrubenstein@xstratatech.com.au
myoung@xstratatech.com.au

[3] G & T Metallurgical Services Ltd., 2957 Bowers Place, Kamloops, BC V1S 1W5
Тел. (250) 828 6157
tshouldice@gtmet.com

Ключевые слова: IsaMill™, флотомашина Джеймсон, Wemco® SmartCell™, никель, технологическая минералогия

42-ое Ежегодное собрание
Канадского общества
переработки минерального
сырья

19-21 января 2010 года,
Оттава, Онтарио, Канада

АВТОРЕФЕРАТ

Комбинат Cosmos компании Xstrata Nickel Austrasia (шт. Западная Австралия) в настоящее время осуществляет проект устранения узких мест в технологической схеме с целью увеличения производства никеля при сохранении металлургических показателей и низкой удельной себестоимости продукции. Результаты минералогического анализа показали, что узкие места в технологической схеме можно устранить, изменив отдельные участки технологического цикла, а также увеличив мощность производства с использованием новых технологий, таких как IsaMill™, флотомашина Джеймсон, Wemco® SmartCell™ и фильтр Larox.

Флотомашина Wemco® SmartCell™, идеально подходящая для флотации грубого минерального сырья, и флотомашина Джеймсон, признанная оптимальной для быстро флотируемого и мелкого минерального сырья, могут быть установлены на основной технологической линии цикла флотации в качестве флотомашин скальпирования основной флотации. Установка двух дополнительных флотомашин позволит получить достаточную мощность флотации с учетом ожидаемого повышения содержания никеля в питании и тоннажа.

Установка мельницы IsaMill™ для доизмельчения позволит сделать более грубым первичный продукт мельницы полусамоизмельчения, тем самым повысив производительность. Среди преимуществ, предлагаемых мельницей IsaMill™, – возможность работы в открытом цикле (внутренняя классификация), инертная (керамическая) среда измельчения, облегчающая последующую флотацию (более чистые новые поверхности), компактная конструкция (удобство установки на действующем производстве) и энергоэффективность. Одновременно с увеличением мощности основной флотации, схема установки остальных существующих флотомашин была изменена с учетом новой мельницы доизмельчения и требований цикла перечистки. Существующий фильтр-пресс Larox 24м^2, используемый для обезвоживания концентрата, был заменен фильтром 32 м^2 с учетом увеличенного выхода концентрата.

ВВЕДЕНИЕ

Горно-обогатительный комбинат Cosmos расположен в 680 км к северо-востоку от Перта, шт. Западная Австралия. Разработка месторождения была начата в 1999 году компанией Jubilee Mines NL, предыдущим владельцем комбината Cosmos, а первый никелевый концентрат был получен в апреле 2000 года. Комбинат перерабатывает высокосортную никелевую сульфидную руду в коматиитовой вмещающей породе; сульфидные минералы преимущественно представлены пентландитом и пирротитом, в меньшей степени пиритом и халькопиритом.

Компании Jubilee принадлежали одни из самых перспективных и относительно малоизученных никелевых месторождений в мире, что обеспечивало отличный потенциал расширения производства. Это хорошо сочеталось со стратегией развития компании Xstrata Nickel, и в октябре 2007 года Xstrata сделала предложение о покупке Jubilee Mines NL. В феврале 2008 года Xstrata приняла на себя функции управления активами Jubilee и учредила компанию Xstrata Nickel Australasia (XNA), новое производственное подразделение Xstrata Nickel.

Стратегия интенсивного роста XNA потребовала внесения изменений в перерабатывающее производство с учетом увеличения производства никеля из руды, добываемой на руднике Просперо комбината Cosmos (рис. 1). Увеличение производства никеля было связано с увеличением не только тоннажа, но и содержания никеля в руде. Обогатительная фабрика комбината Cosmos был спроектирован более 10 лет назад и был рассчитан примерно на 3 года отработки одноименного месторождения. Учитывая возраст и проектный срок службы фабрики, а также его компактную планировку, нельзя было не ожидать определенных трудностей, связанных с его модернизацией для увеличения мощности. В ноябре 2008 года обогатительная фабрика Cosmos имела массовую производительность 25 т/ч и производительность по никелю в питании 1,3 т/ч. Чтобы получить запланированный на 2009 год выход никеля, необходимо было увеличить производительность по никелю в питании 2,8 т/ч (+215%) и массовую производительность мельницы полусамоизмельчения 45 т/ч (+180%).

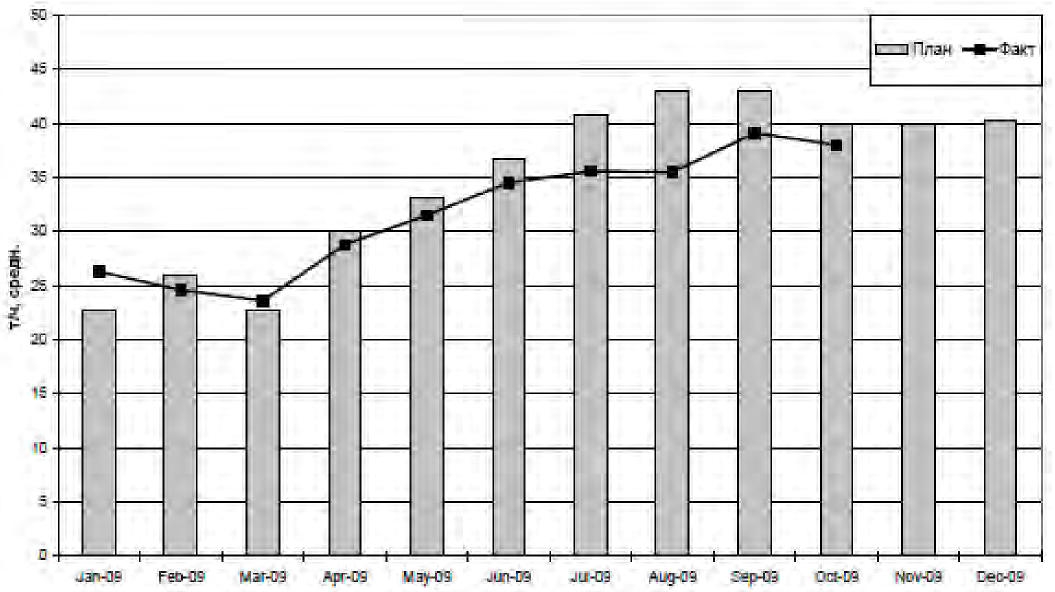


Рисунок 1. Производительность мельницы полусамоизмельчения на комбинате Cosmos (фактическая в 2009 году на указанную дату против плановой)

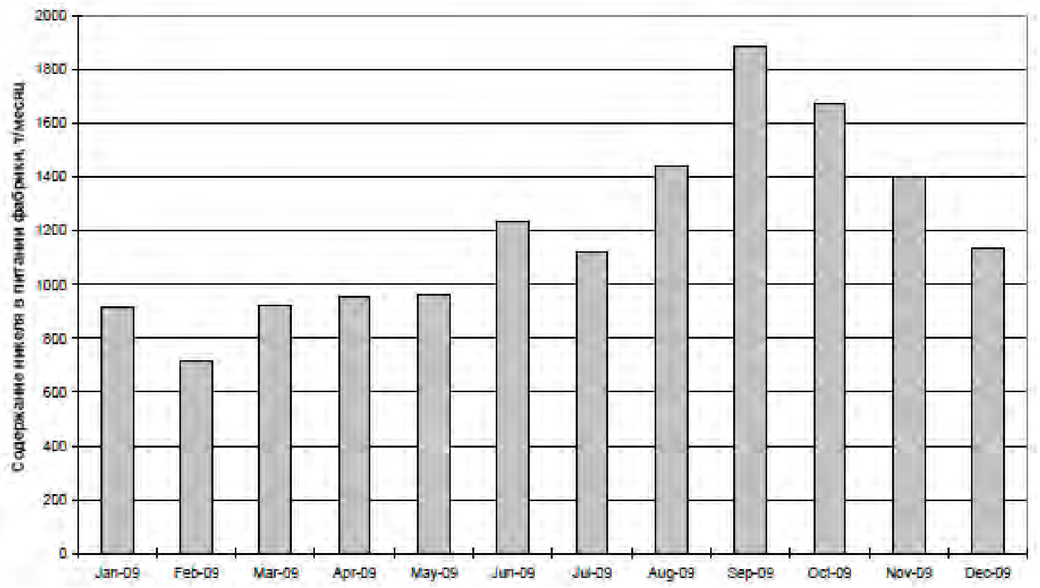


Рисунок 2. Производительность обогатительной фабрики Cosmos по никелю в питании, т/месяц согласно плану на 2009 год

Были выполнены минералогические и металлургические исследования, результаты которых показали, что выйти на высокие плановые показатели можно при условии установки и реконфигурации нескольких единиц ключевого оборудования. На основании полученных результатов и с использованием принципа применения оптимального оборудования была начата программа устранения узких мест в технологической схеме. Хотя она еще не завершена, программа уже доказала свою эффективность и будет выполнена согласно плану (за 12 месяцев) и с минимальными (по меркам отрасли) затратами.

Решения, используемые при устранении узких мест в технологической схеме, были и остаются основаны на минералогическом подходе. Вносимые в технологию изменения основаны на данных анализа извлечения по фракциям крупности, высвобождения и минералогических подсчетах. Для оценки эффективности вносимых изменений ключевое значение имели определения извлечения целевых минералов по фракциям в основных технологических потоках. Принцип “нужного оборудования в нужном месте” подразумевал реконфигурацию части оборудования для получения измененной технологической схемы и применение новых технологий, таких как IsaMill™, флотомашина Джеймсон и Wemco® SmartCell™, для получения требуемых металлургических показателей.

Ниже представлено краткое описание исходной технологической схемы и соответствующей технологической минералогии; далее следует описание изменений, внесенных при устранении узких мест, и их влияние на технологические показатели обогатительной фабрики по фракциям крупности.

ИСХОДНАЯ ТЕХНОЛОГИЧЕСКАЯ СХЕМА И ЕЕ ЭФФЕКТИВНОСТЬ (РЕЗУЛЬТАТЫ ИССЛЕДОВАНИЯ, ВЫПОЛНЕННОГО В ИЮЛЕ 2008 ГОДА)

В этой части будет рассмотрена схема фабрики, характеристики питания флотации, показатели флотации, состав конечного концентрата и показатели уноса с хвостами до программы устранения узких мест, причем последние два аспекта имеют особое значение. Иными словами, что присутствует в концентрате из того, чего в нем не должно быть, и наоборот, - что присутствует в хвостах из того, чего в них не должно быть. На рис. 3 показана исходная технологическая схема обогатительной фабрики Cosmos.

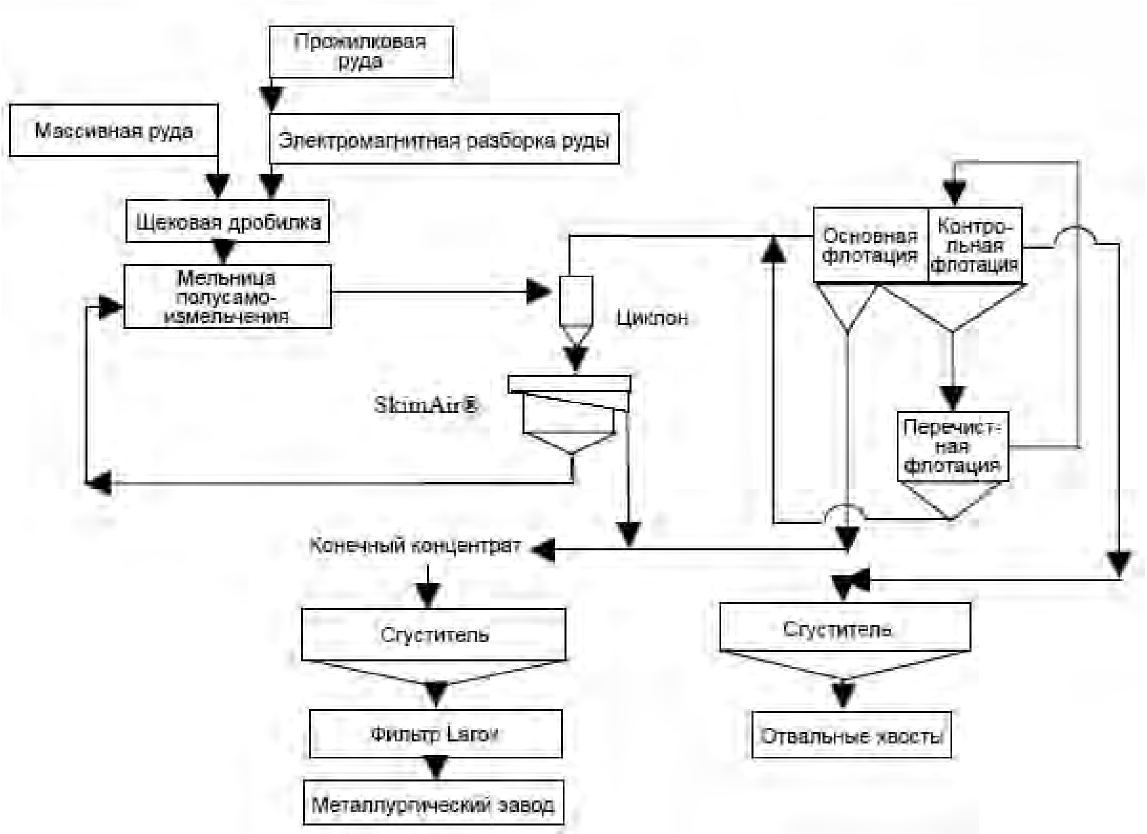


Рисунок 3. Исходная технологическая схема

Цикл измельчения состоял из одной щековой дробилки, продукт которой поступал непосредственно в одностадийную мельницу полусамоизмельчения закрытого цикла с одной парой циклонов Cavex 400 мм. Схема классификации не отличалась гибкостью и эффективностью, в результате чего значительное количество высвобожденного пентландита попадало в нижний продукт циклона. Поэтому флотомашина SkimAir® демонстрировала высокие показатели эффективности, однако это указывало на неудовлетворительную эффективность цикла измельчения, поскольку эти высвобожденные частицы должны были поступать в верхний продукт циклона. SkimAir® извлекала большую часть никеля (61%) при содержании в концентрате 20%. Остальная часть цикла флотации производила концентрат с содержанием 16%, обеспечивая извлечение еще 30% никеля. Общее извлечение никеля в цикле достигало 92% при содержании в концентрате 19% (таблица 1).

Цикл флотации состоял из двух стадий основной флотации и трех стадий контрольной флотации в основном технологическом потоке питания флотации (таблица 2 и 3). Концентрат первой основной флотации поступал в конечный концентрат вместе с продуктом SkimAir®. Примечательно, что концентрат и хвосты перечистной флотации находились в закрытом цикле с основной/контрольной флотацией. В результате во всех потоках флотации присутствовала смесь различных флотореагентов, которые были вынуждены одновременно выполнять функции активации, сбора и депрессии. Это делало цикл проще в управлении, однако большие рециркулирующие нагрузки, сложности, возникающие при дозировании реагентов, и непрерывное разубоживание концентрата перечистной флотации питанием основной флотации указывали на возможность несложного и значимого усовершенствования цикла.

Верхний продукт циклона содержал высвобожденный на 86% пентландит и высвобожденные на 97% несульфидные жильные минералы, что также указывало на переизмельчение. Столь высокие показатели высвобождения более подходили питанию перечистной флотации, чем циклу основной флотации, и позволяли сделать вывод, что можно увеличить производительность мельницы полусамоизмельчения и получать более грубый верхний продукт циклона без ущерба для извлечения в цикле основной флотации.

Таблица 1. Общий металлургический баланс, июль 2008 года

Продукт	Масса, %	Содержание, %					Распределение, %				
		Ni	Cu	Fe	S	As	Ni	Cu	Fe	S	As
Питание, пересчет	100	4,61	0,17	12,2	8,69	0,08	100	100	100	100	100
Концентрат SkimAir®	13,8	20,6	0,68	34,5	30,6	0,24	61	54	39	49	43
Концентрат 1-й основной флотации	8,5	16,4	0,77	27,7	26,1	0,13	30	37	19	25	15
Конечный концентрат	22,2	19,0	0,70	31,9	28,9	0,21	92	91	58	74	58
Отвальные хвосты	77,8	0,5	0,02	6,5	2,9	0,04	8	9	42	26	42

Таблица 2. Питание флотации, июль 2008 года

Минерал	**Символ**	**Содержание, %**
Пентландит	Pe	13,5
Халькопирит	Cp	0,5
Пирротит	Po	9,7
Пирит	Py	1,0
Несульфидные жильные минералы	Gn	75,3

Таблица 3. Состав минералов в питании флотации, июль 2008 года

Состояние минерала	**Питание флотации, 104 мкм, K_{80}**		
	Pe	**Po**	**Gn**
Высвобожденный	85,9	77,4	97,0
Бинарные частицы с Pe	—	6,0	1,1
Бинарные частицы с Po	3,2	—	0,6
Бинарные частицы с Gn	6,9	9,4	—
Бинарные частицы с проч.	0,7	0,6	0,3
Мультифазные частицы	3,3	6,6	1,0

На рис. 4 показано извлечение в зависимости от крупности в цикле SkimAir®. Извлечение из крупной фракции имеет отличные показатели при незначительном вовлечении жильной породы (особенно учитывая относительно высокую плотность пульпы, необходимую в этом цикле для управления водным балансом в мельнице полусамоизмельчения). Извлечение из мелких фракций (<50 мкм) хуже, однако извлечение происходит позже в цикле основной флотации, как показано на рис. 5. Примерно 30% высвобожденного пирротита в верхнем продукте циклона извлекается в концентрат первой основной флотации (максимальное извлечение наблюдается при диаметре частиц от 10 до 50 мкм). Вовлечение несульфидных жильных минералов в концентрат первой основной флотации выше, чем в концентрат SkimAir®.

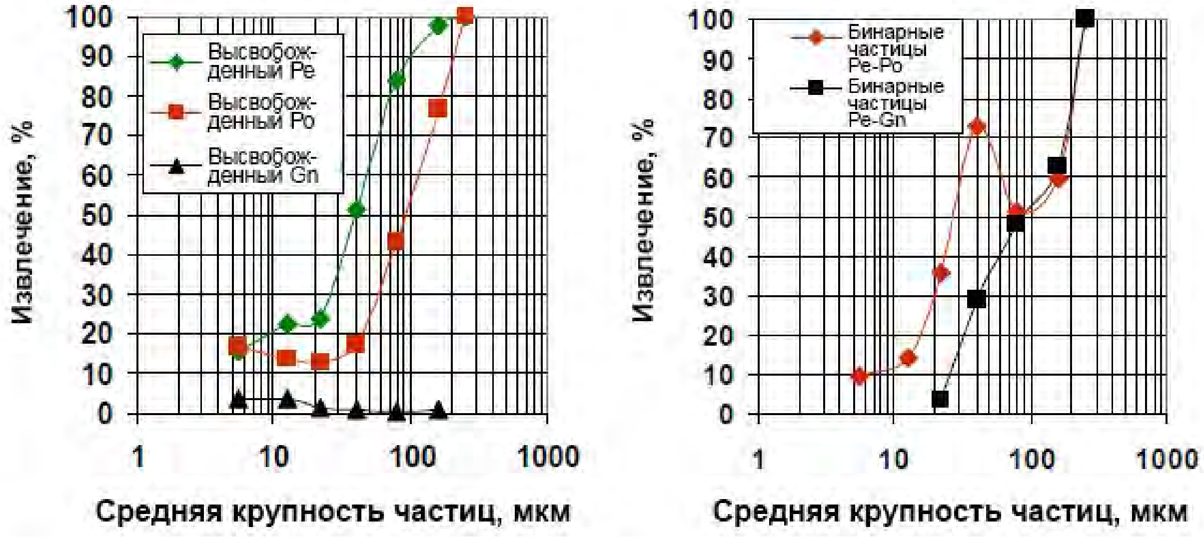


Рисунок 4. Извлечение в концентрат SkimAir® в зависимости от крупности и класса

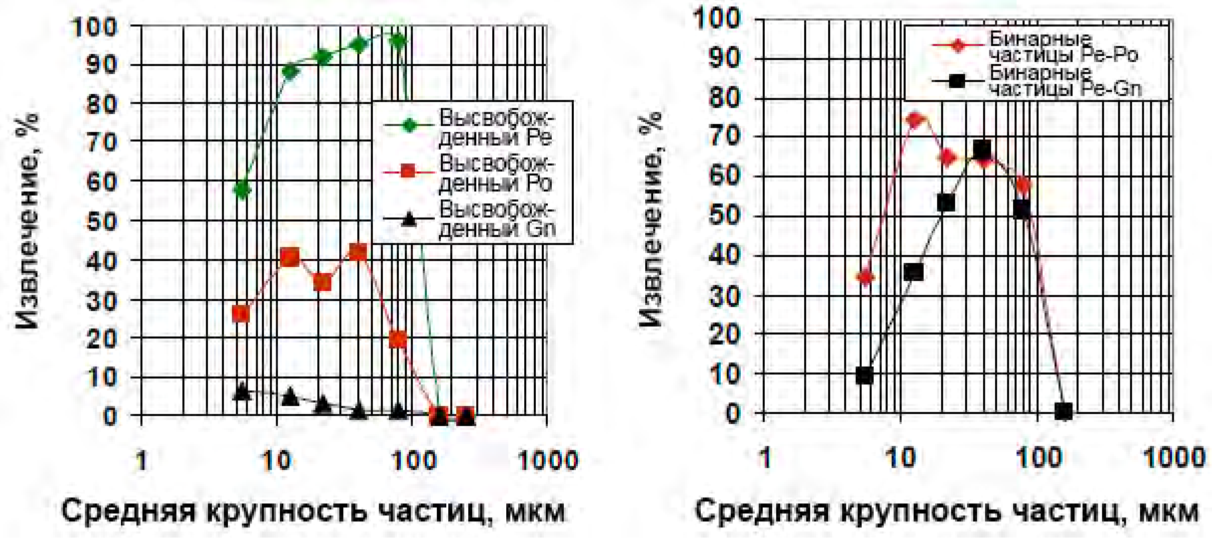


Рисунок 5. Извлечение в концентрат первой основной флотации в зависимости от крупности и класса

Как показано на рис. 6, основными разубоживающими минералами в концентрате SkimAir® являются пирротит и несульфидные жильные минералы (21% и 12% от массы концентрата, соответственно). Концентрат первой основной флотации содержал примерно в два раза больше несульфидных жильных минералов и примерно столько же пирротита, сколько содержал концентрат SkimAir® (рис. 7). Основные разубоживающие минералы преимущественно присутствуют в виде высвобожденных зерен. Теоретическая кривая зависимости содержания от извлечения на рис. 8 показывает, что при условии удаления большей части высвобожденных разубоживающих минералов можно получить концентрат более высокого сорта.

В то же время, увеличение крупности измельчения (приводящее к меньшему высвобождению) может не повлиять на содержание в конечном концентрате. Если применить традиционный цикл перечистки и/или промывку пены для удаления высвобожденных вовлеченных несульфидных жильных минералов, существует возможность получения концентрата с более высоким содержанием.

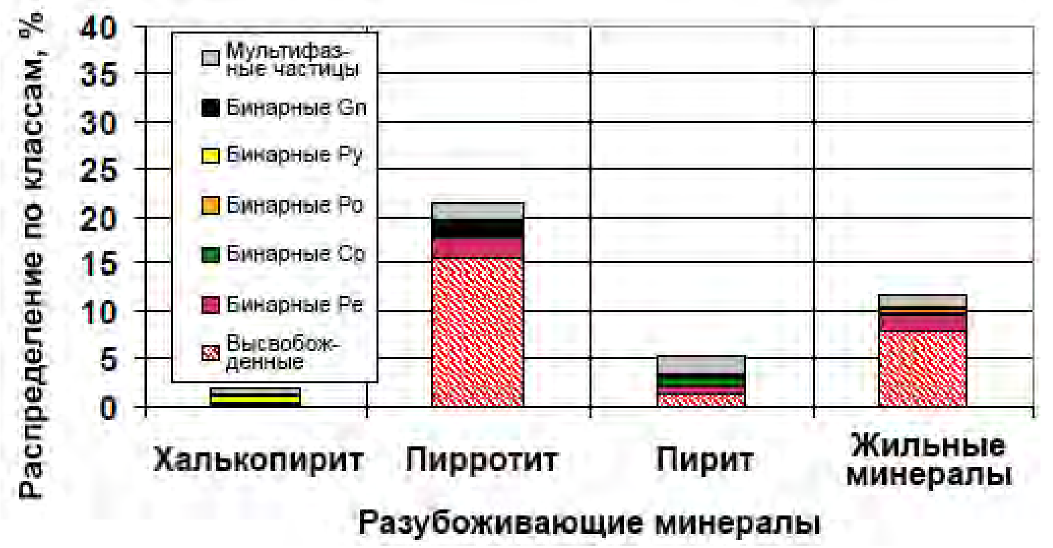


Рисунок 6. Распределение разубоживающих минералов в концентрате SkimAir®

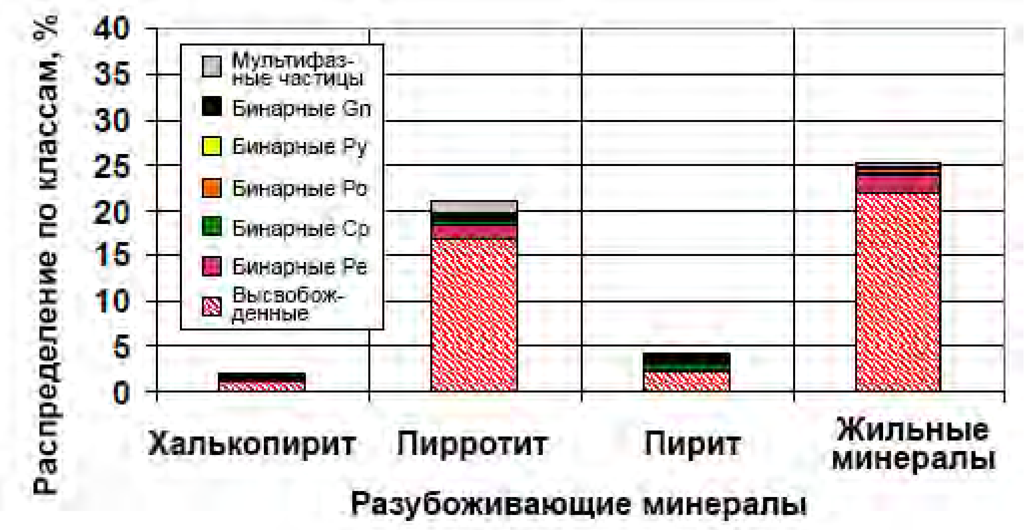


Рисунок 7. Распределение разубоживающих минералов по массе в концентрате основной флотации

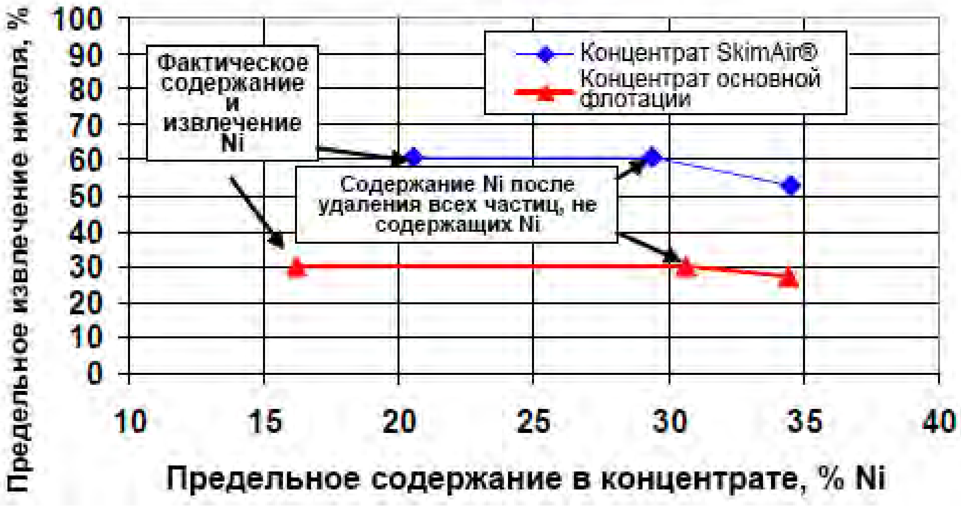


Рисунок 8. Предельные минералогические характеристики содержания и извлечения

8% никеля в руде теряется с отвальными хвостами. 60% потерь приходится на унос высвобожденных зерен пентландита (рис. 9), значительная доля которых имеет крупность менее 11 мкм. Это закономерно, учитывая отсутствие полноценного цикла перечистки и сложный химический состав пульпы.

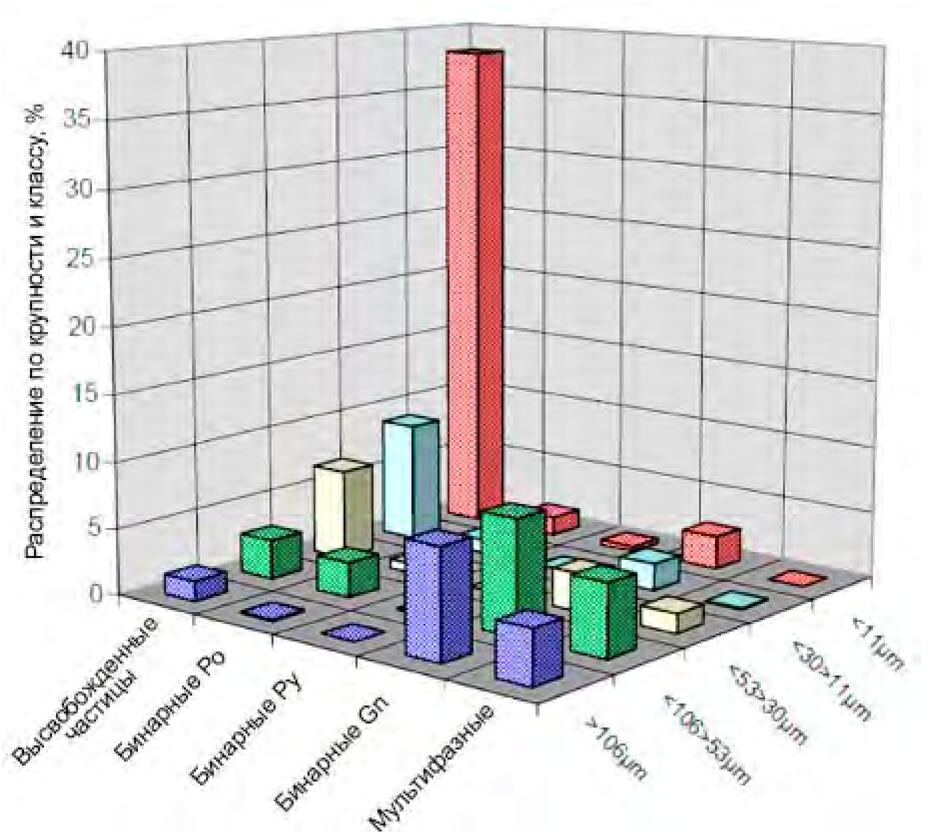


Рисунок 9. Распределение уноса никеля с хвостами по крупности и классу, % от содержания пентландита в хвостах

ТЕХНОЛОГИЧЕСКАЯ СХЕМА ПОСЛЕ УСТРАНЕНИЯ УЗКИХ МЕСТ И ЕЕ ЭФФЕКТИВНОСТЬ (РЕЗУЛЬТАТЫ ИССЛЕДОВАНИЯ, ВЫПОЛНЕННОГО В АПРЕЛЕ 2009 ГОДА)

Производительность мельницы полусамоизмельчения была увеличена с учетом большего тоннажа добычи с получением более крупного питания флотации. Увеличение производительности мельницы полусамоизмельчения показано на рис. 10. Среди существенных доработок цикла полусамоизмельчения можно назвать переход со смеси шаров 80/100/125 мм на шары 125 мм, увеличение угла лифтера корпуса мельницы полусамоизмельчения с 15 до 25 градусов с учетом повышенной скорости мельницы, увеличение размера отверстий и живого сечения сит на стороне разгрузки. Из-за невозможности добиться сбалансированного режима работы двух циклонов Cavex 400 мм при больших колебаниях расхода они были заменены четырьмя циклонами Cavex 250 мм. Результаты минералогического анализа показали, что уровень высвобождения пентландита и несульфидных жильных минералов в питании флотации может быть снижен при большей крупности измельчения без ущерба для показателей извлечения основной и контрольной флотации. Это позволило бы извлекать большую долю композитных частиц. Чтобы добиться постоянства содержания в конечном концентрате, композитные частицы необходимо доизмельчать до перечистной флотации. Мельница IsaMill™ была установлена в цикле для доизмельчения.

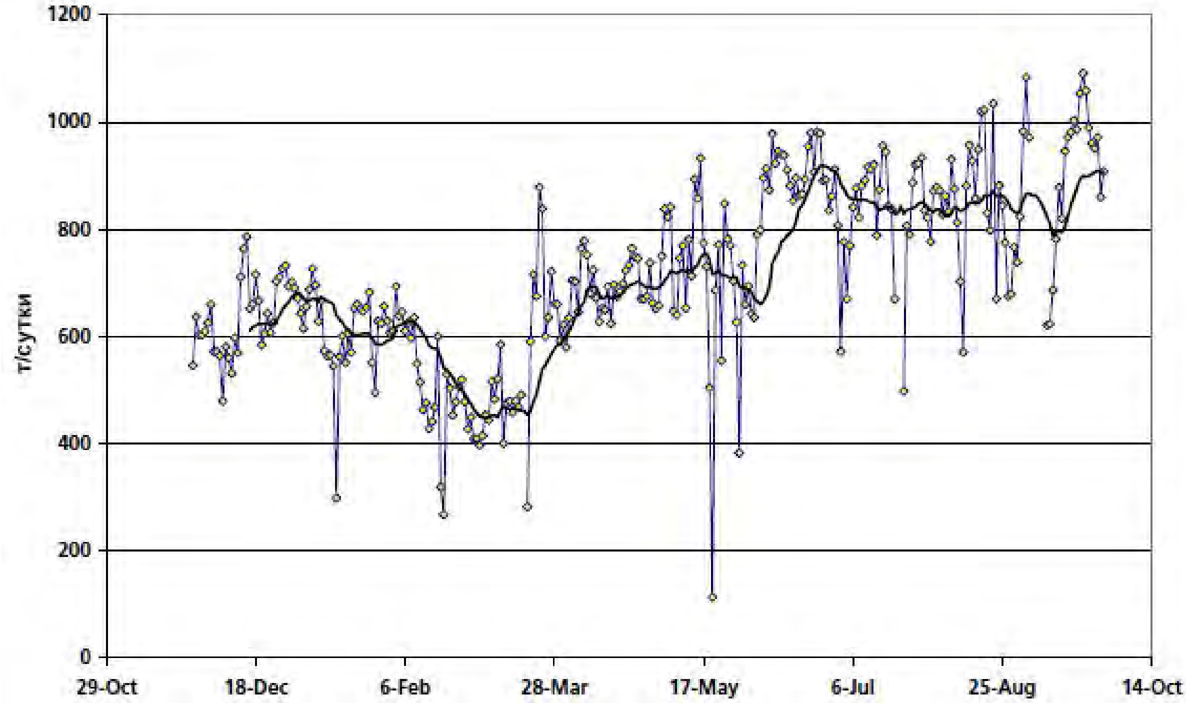


Рисунок 10. Производительность мельницы полусамоизмельчения

Для доизмельчения была выбрана мельница IsaMill™ M500 (показана на рис. 11), поскольку благодаря ее компактности она идеально подходит для проектов расширения действующих производств. С точки зрения минералогических и металлургических характеристик фабрики Cosmos, большую ценность представляла возможность повысить благодаря применению IsaMill™ содержание в концентрате, использовать инертную среду измельчения (чистые поверхности частиц), получить узкий гранулометрический диапазон частиц (в открытом цикле), эффективно измельчать материал при низкой плотности пульпы. Чистые свежие поверхности минеральных частиц, которые могут быть получены при использовании инертных сред, позволяют оптимизировать разделение минералов, снизить расход реагентов и ускорить кинетику флотации (Côté and Adante, 2009; Finch, Rao, and Nesset, 2007; Huang, Grano, and Skinner, 2006). Пример плохо высвобожденного пентландитового композита, требующего доизмельчения, показан на рис. 12.

Рисунок 11. Мельница IsaMill™ с двигателем 200 кВт на обогатительной фабрике Cosmos

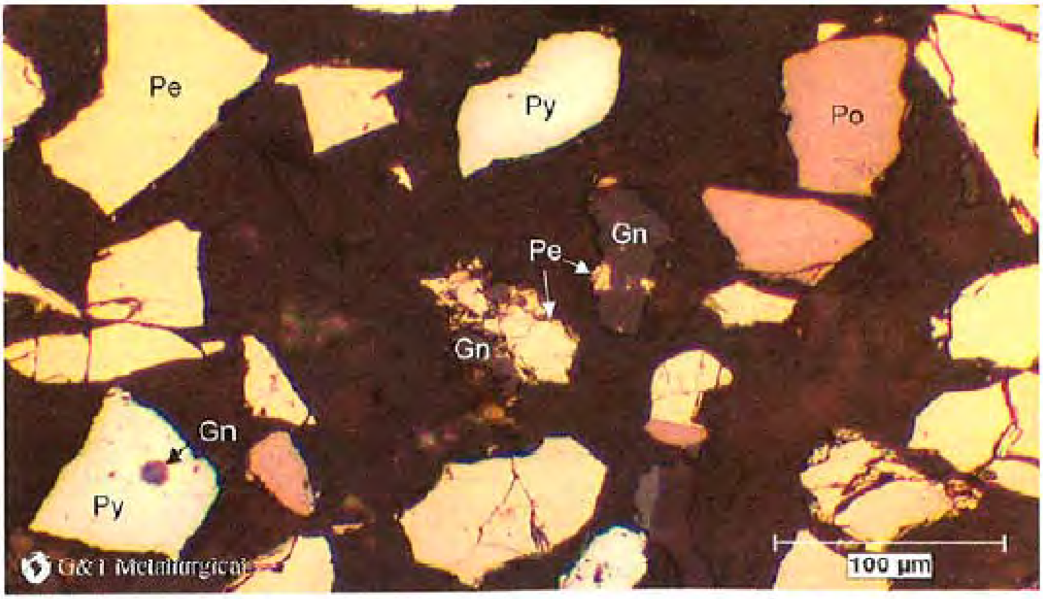


Рисунок 12. Микрофотография плохо высвобожденных частиц пентландита

Лабораторные пробы выбранных концентратов были доизмельчены и подвергнуты флотации, чтобы проверить предположение. Было установлено, что при более мелком питании флотации кривая зависимости содержания никеля от извлечения смещалась вверх (рис. 13). IsaMill™ обеспечивает намного более тщательное удаление MgO и, при создании благоприятных химических условий в цикле перечистной флотации, -оптимальное удаление As.

Характерная гранулометрическая характеристика продукта IsaMill™ M500 показана на рис. 14. F_{80} и P_{80} составляют 40 и 20 мкм, соответственно. Удельный расход энергии на такое измельчение составляет примерно 20 кВт-ч/т. Примечательно, что гранулометрическая характеристика продукта круче гранулометрической характеристики питания. Мельница IsaMill™ направляет энергию измельчения на получение крупных частиц, а не на увеличение доли мелких. На это указывает минимальное изменение крупностей классов P_{10}, P_{20} и P_{30}. Для доизмельчения применяется керамическая среда крупностью 2 мм Keramax® MT1™ производителя Magotteaux.

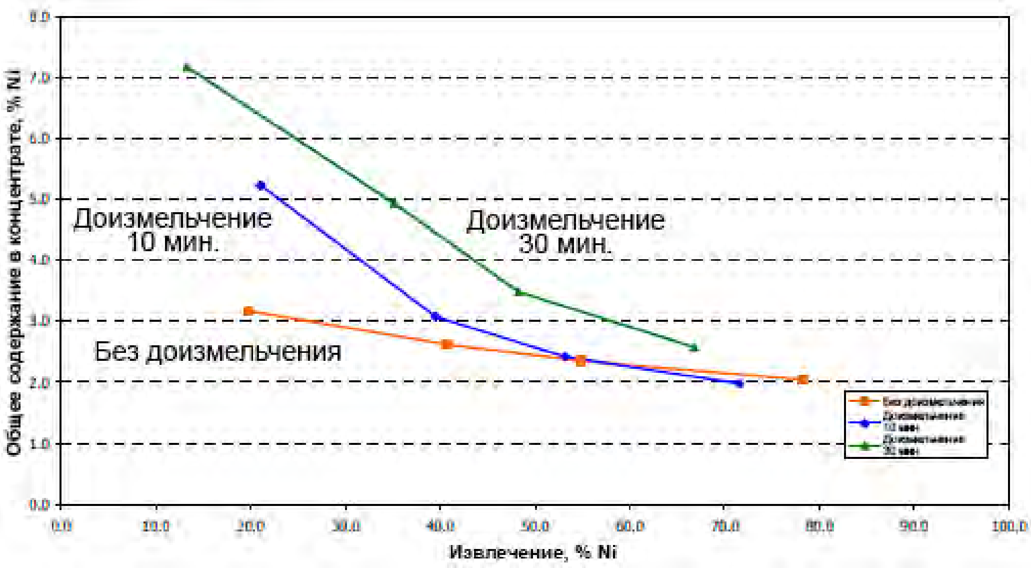


Рисунок 13. Пример положительного влияния доизмельчения на характеристику зависимости содержания от извлечения

Рисунок 14. Характерный гранулометрический состав питания и продукта мельницы IsaMill™ M500

В связи с более высоким исходным содержанием и мощностью производства потребовалась установка дополнительных мощностей флотации. В первую очередь была установлена флотомашина Джеймсон Z1600 для переработки верхнего продукта циклона мельницы полусамоизмельчения в качестве флотомашины скальпирования основной флотации, а цикл флотации был сделан полностью открытым (концентраты перечистной флотации поступают в конечные концентраты, а хвосты перечистной флотации – в отвальные хвосты). После огрубления верхнего продукта циклона мельницы полусамоизмельчения уменьшились содержание высвобожденного пентландита в нижнем продукте циклона и извлечение в SkimAir® (рис. 15). После этого флотомашина SkimAir® стала использоваться в качестве флотомашины основной флотации, получающей питание из верхнего продукта циклона.

Рисунок 15. Извлечение в SkimAir® в зависимости от крупности верхнего продукта циклона мельницы полусамоизмельчения

Во вторую очередь была установлена флотомашина Wemco® SmartCell™ емкостью 40 м^3 для увеличения мощности основной флотации. Сперва она была установлена после флотомашины Джеймсон с целью получения достаточной мощности основной флотации, чтобы остальные первоначально установленные флотомашины могли быть включены в соответствующие циклы перечистной флотации. После пуска флотомашины Wemco® SmartCell™ было установлено, что вспениватель, попадающий из предшествующих технологических стадий, вызывает чрезмерное пенообразование, а также проблемы при перекачивании концентрата. Существующая трубная обвязка позволила подключить флотомашину Wemco® SmartCell™ перед флотомашиной Джеймсон. Такая схема подключения используется по настоящее время.

Флотомашина Джеймсон была выбрана для фабрики Cosmos благодаря ее компактности и высокой производительности. Флотомашина Джеймсон, установленная на фабрике Cosmos, показана на рис. 16. Подобно мельнице IsaMill™, компактная конструкция облегчает установку при переоборудовании действующих предприятий.

Рисунок 16. Флотомашина Джеймсон Z1600 на обогатительной фабрике Cosmos

Флотомашина Джеймсон была запущена в конце марта 2009 года, что совпало с увеличением содержания никеля в концентрате на 2 единицы при прежнем извлечении – 90% (таблица 4). Как следует из рис. 17, содержание в концентрате флотомашины Джеймсон стабильно выше содержания в конечном концентрате. Кроме того, для уменьшения вовлечения несульфидных жильных минералов применяется промывка пены, как показано на рис. 18. Следует отметить, что флотомашина Джеймсон была пущена на верхнем продукте циклона, за несколько недель до реконфигурации флотомашины SkimAir®.

Таблица 4. Металлургические балансы за различные месяцы

Месяц	Конечный продукт	Масса, %	Содержание, %					Распределение, %				
			Ni	Cu	Fe	S	As	Ni	Cu	Fe	S	As
Июль 08	Концентрат	22,2	19,0	0.71	31,9	28,9	0,2	92	91	58	74	58
	Хвосты	77,8	0,5	0,02	6,5	2,9	0,04	8	9	42	26	42
Август 08	Концентрат	18,7	18,7	0,74	29,9	29,6	0,28	91	92	56	76	77
	Хвосты	81,3	0,4	0,01	5,4	2,2	0,02	9	8	44	24	23
Декабрь 08	Концентрат	19,9	19,6	0,65	25,7	24,5	0,25	90	87	50	67	46
	Хвосты	80,1	0,5	0,02	6,4	3,0	0,07	10	13	50	33	54
Февраль 09	Концентрат	22,8	19,9	0,68	24,8	23,5	0,19	90	81	44	60	45
	Хвосты	77,2	0,7	0,05	9,3	4,6	0,07	10	19	56	40	55
Апрель 09	Концентрат	22,5	21,8	0,75	30,6	27,6	0,11	90	88	36	50	44
	Хвосты	77,5	0,7	0,03	15,5	8,0	0,04	10	12	34	50	56

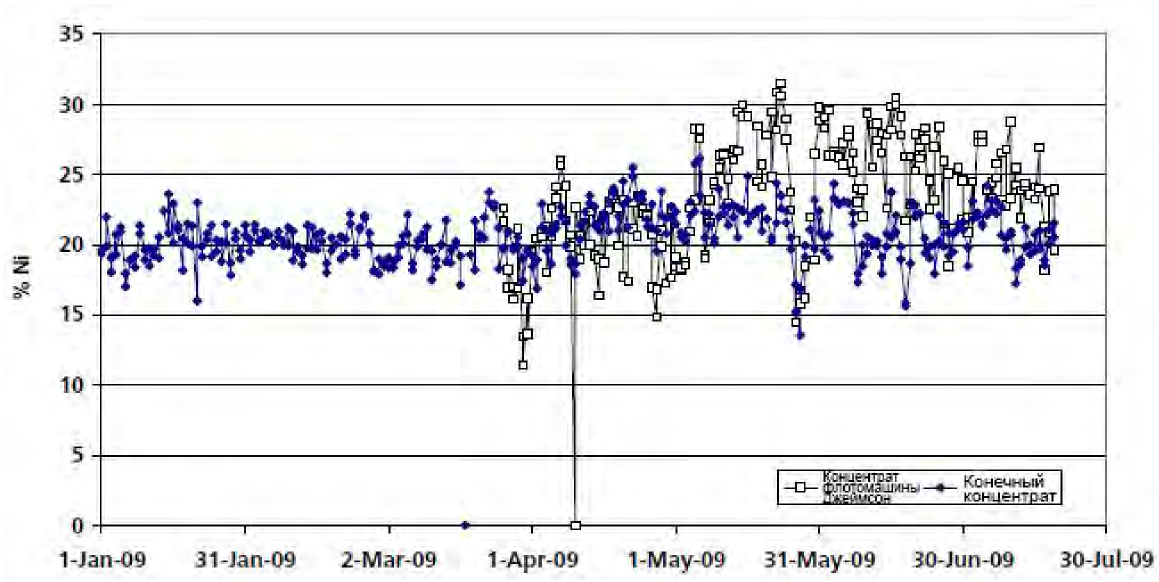


Рисунок 17. Содержание в концентрате флотомашины Джеймсон и в конечном концентрате

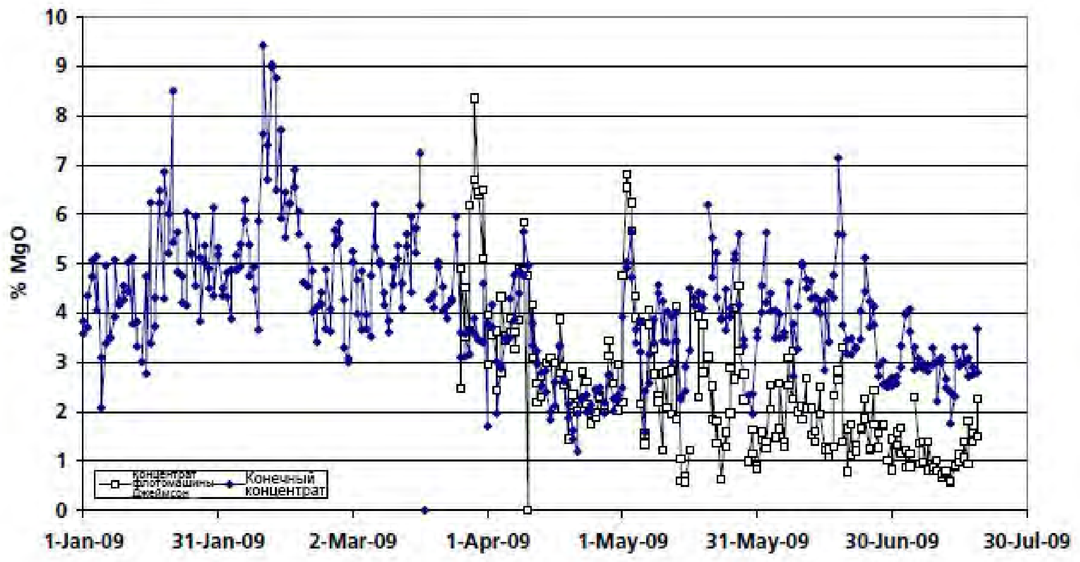


Рисунок 18. Содержание MgO в концентрате флотомашины Джеймсон и в конечном концентрате

В последнюю очередь флотомашины, первоначально работавшие в цикле основной/контрольной флотации, были подключены к циклу перечистной флотации. Организация открытого цикла флотации позволила более четко определить последовательные стадии флотации. Схема трех основных компонентов цикла показана на рис. 19. В данной схеме флотомашины основной и контрольной флотации обеспечивают максимальное извлечение никеля. На стадии перечистной флотации происходит депрессия и удаление несульфидных жильных минералов при помощи доизмельчения, перечистной флотации с промывкой/разубоживанием и применения депрессантов. При необходимости может применяться вторая перечистная флотация для избирательного удаления мышьяка посредством контроля pH и добавления цианида.

Данная схема имеет три преимущества. Во-первых, благодаря последовательному введению, реагенты добавляются точно на нужной стадии, а не до нее через рециркулирующий поток. Так, прошлые исследования показали, что добавление цианида и сульфата меди замедляет кинетику пентландита. Во-вторых, благодаря разделению функций циклом проще управлять. В-третьих, большее время пребывания позволяет увеличить производительность по руде и никелю.

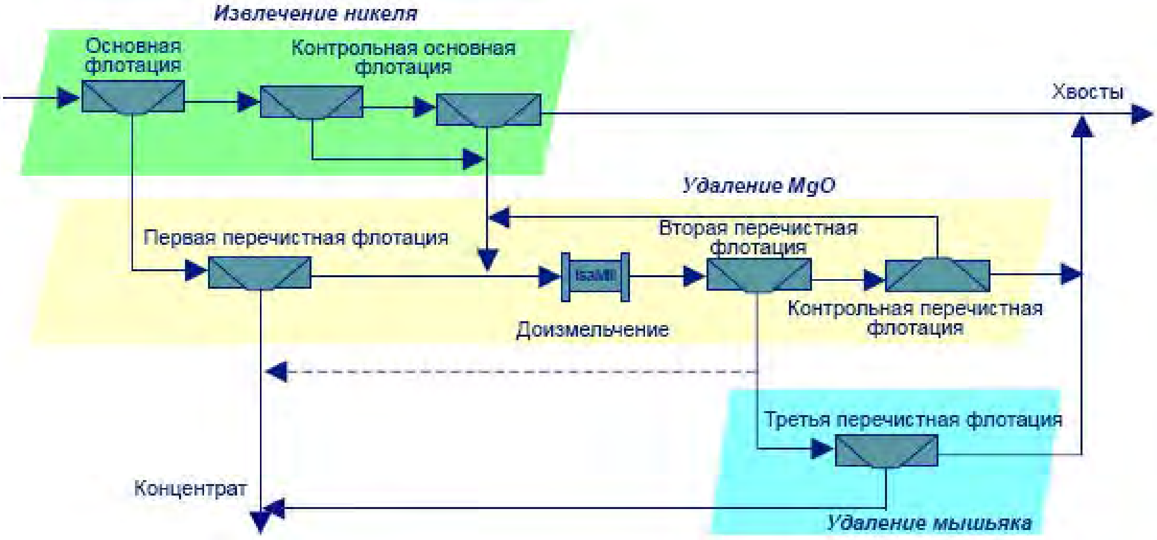


Рисунок 19. Последовательные стадии флотации

Анализ извлечения в конечный концентрат в зависимости от крупности после устранения узких мест указывает на хорошие показатели извлечения высвобожденного пентландита в цикле флотации (рис. 20), хотя по-прежнему наблюдается существенный унос материала мельче 10 мкм. На момент исследования, проведенного в апреле 2009 года, доработка цикла перечистной флотации не была завершена, поэтому время увеличенное пребывания сопровождалось высокой плотностью пульпы. Перечистка с разубоживанием в доработанном цикле позволит извлекать больше ценной мелкой фракции и уменьшить вовлечение высвобожденных несульфидных жильных минералов. Показатели извлечения невысвобожденного пентландита также улучшились (рис. 21). Новый цикл флотации также отличается более эффективным удалением высвобожденного пирротита. Как показано на рис. 22, высвобожденный пирротит составил менее 9% конечного концентрата – по сравнению с 15% в июле 2008 года в цикле до доработки.

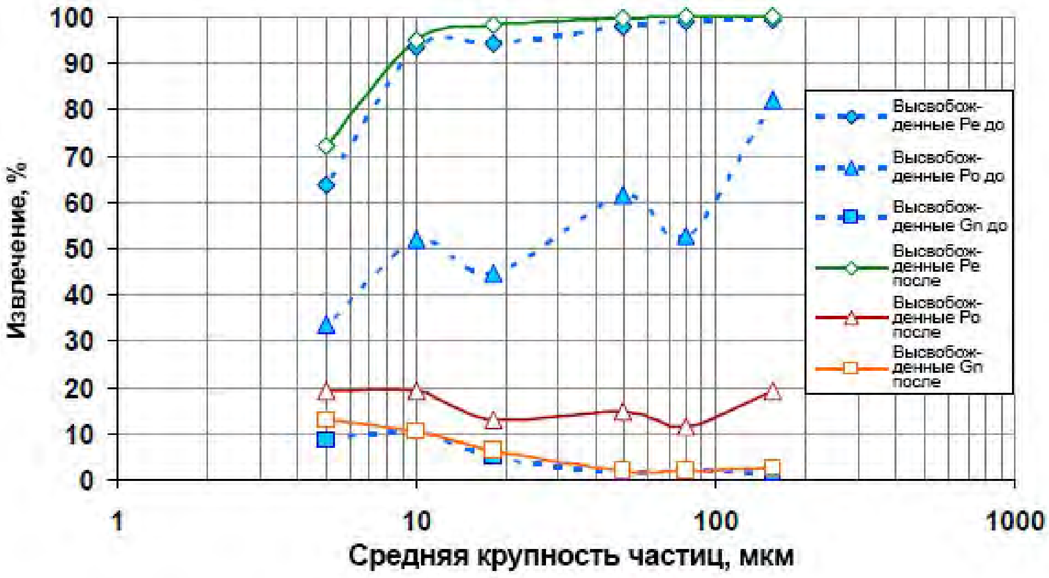


Рисунок 20. Извлечение в конечный концентрат в зависимости от крупности и класса

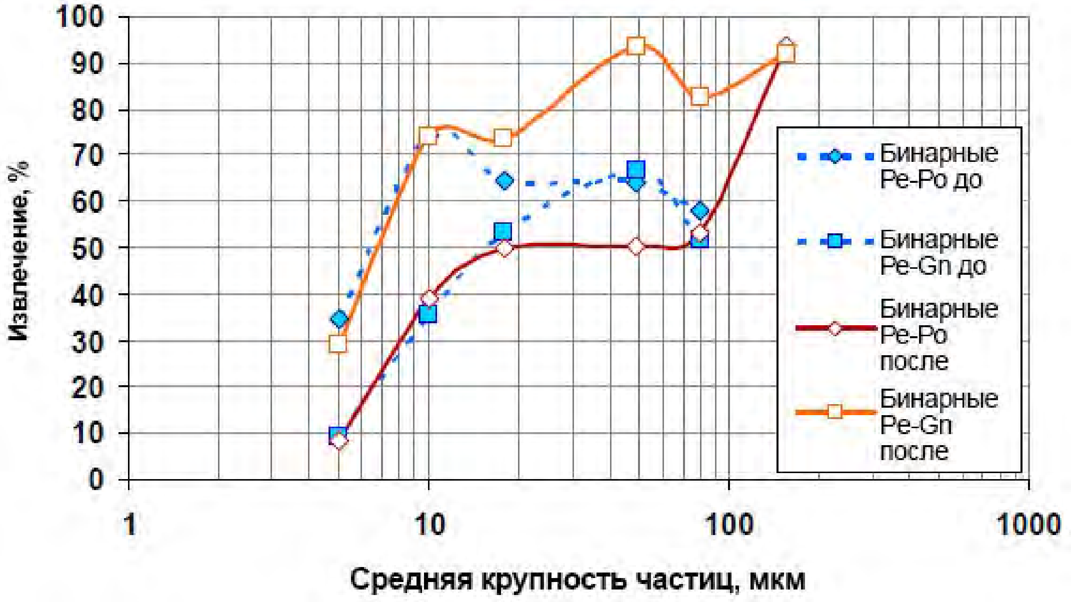


Рисунок 21. Извлечение бинарных частиц пентландита в конечный концентрат

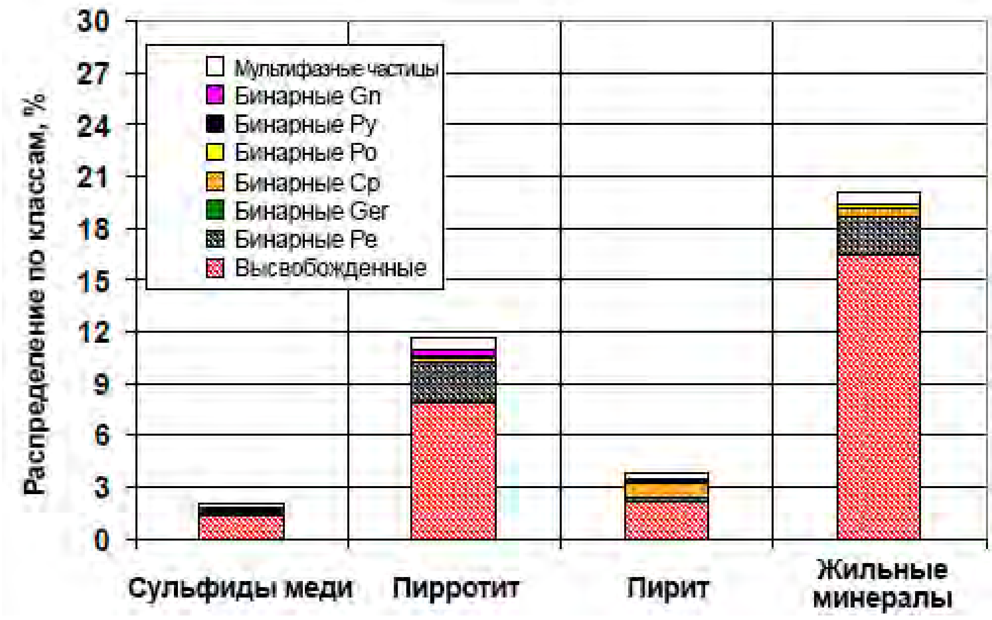


Рисунок 22. Состав разубоживающих минералов в конечном концентрате, апрель 2009 года

Исходя из теоретической характеристики зависимости между содержанием и извлечением, представленной на рис. 23, можно заключить, что, несмотря на некоторое улучшение относительно исходных показателей остаются возможности повышения содержания в концентрате без ущерба для извлечения никеля.

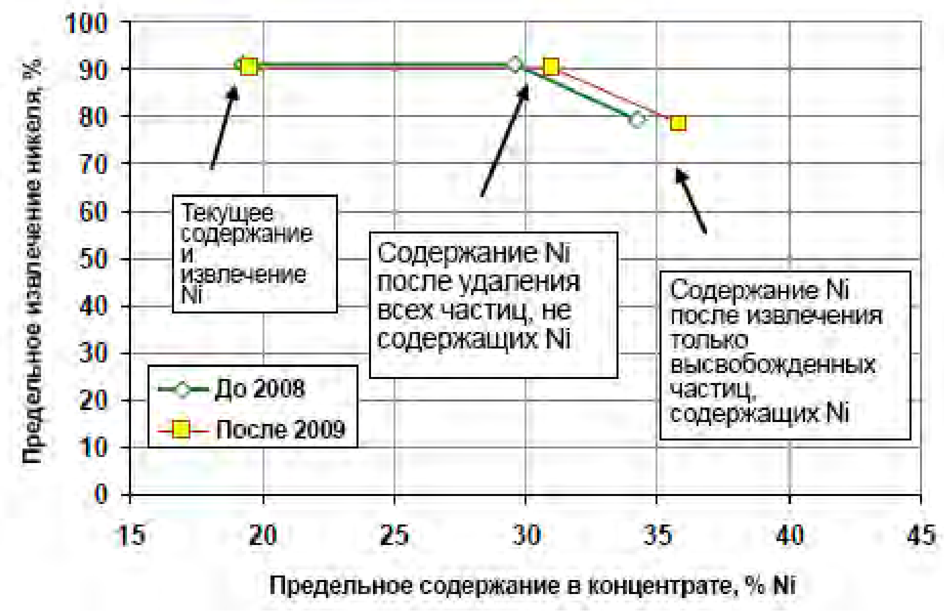


Рисунок 23. Предельные минералогические характеристики содержания и извлечения, до и после

Как показано на рис. 24 и в таблице 5, унос никеля в отвальные хвосты в апреле остается сопоставимым с результатами предыдущего исследования (рис. 9).

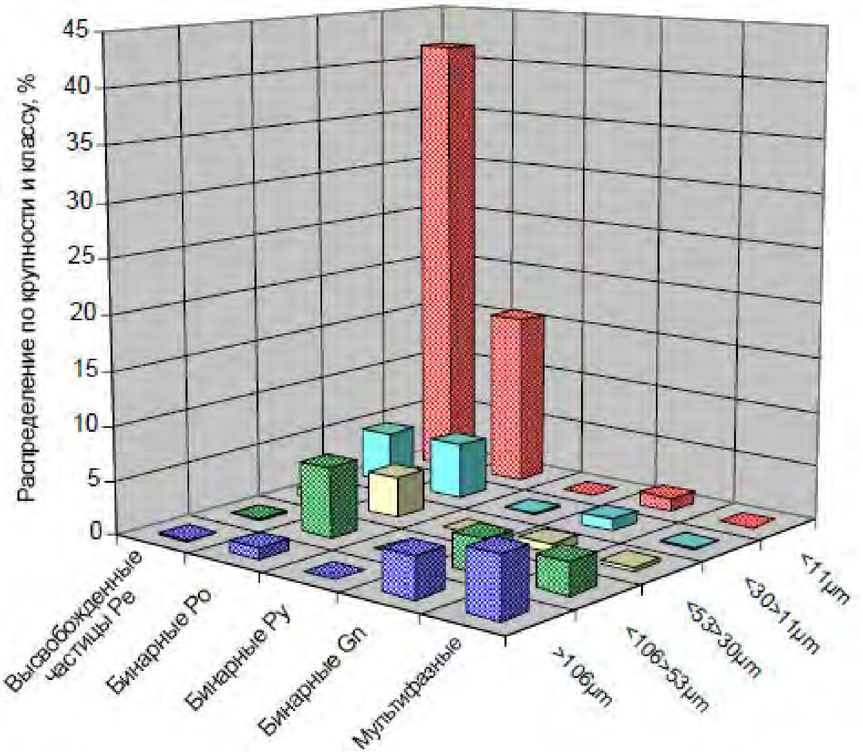


Рисунок 24. Распределение уноса никеля с хвостами по крупности и классу, апрель 2009 года

Таблица 5. Общее распределение уноса никеля в отвальные хвосты

Класс минералов	**Распределение по крупности и классу**					**Унос по классам**
	>106 мкм	**<106>53 мкм**	**<53>30 мкм**	**<30>11 мкм**	**<11 мкм**	
	Среднее за 2008 год					
Высвобожденные	0,1	0,2	0,4	1,0	3,1	4,8
Бинарные Po	0,0	0,2	0,2	0,1	0,2	0,7
Бинарные Py	0,0	0,0	0,0	0,0	0,0	0,0
Бинарные Gn	0,7	0,8	0,2	0,3	0,3	2,3
Мультифазные	0,4	0,4	0,1	0,1	0,0	1,0
Унос по крупности	1,2	1,6	0,9	1,5	3,6	8,8
	Апрель 2009 года					
Высвобожденные	0,0	0,0	0,1	0,5	4,0	4,5
Бинарные Po	0,1	0,6	0,4	0,5	1,5	3,1
Бинарные Py	0,0	0,0	0,0	0,0	0,0	0,0
Бинарные Gn	0,4	0,3	0,1	0,1	0,1	1,0
Мультифазные	0,5	0,3	0,0	0,0	0,0	0,9
Унос по крупности	1,0	1,2	0,5	1,1	5,6	9,4

ОКОНЧАТЕЛЬНАЯ ТЕХНОЛОГИЧЕСКАЯ СХЕМА

Предполагаемая окончательная технологическая схема показана на рис. 25. Доработка позволит завершить программу увеличения производительности по никелю и снижения производственных затрат.

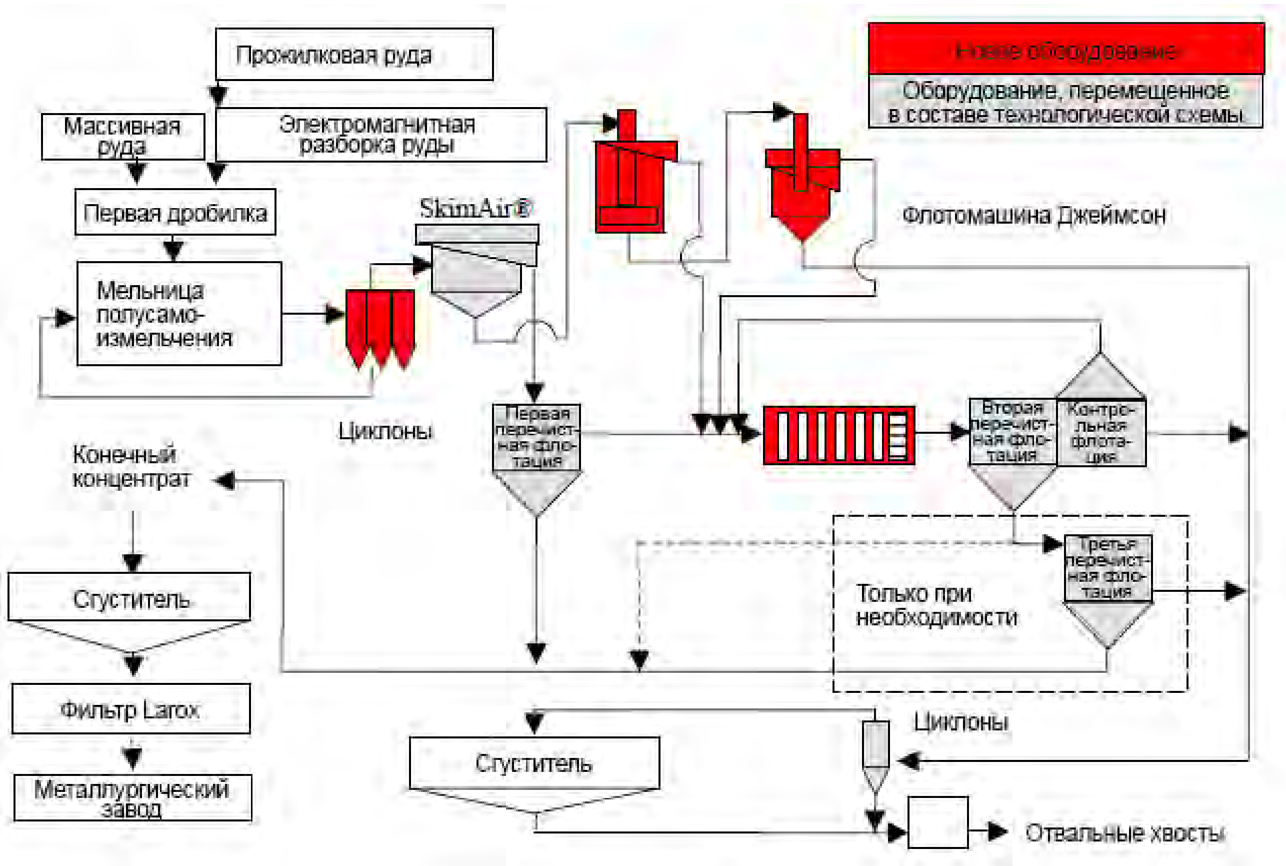


Рисунок 25. Новая технологическая схема

ЗАКЛЮЧЕНИЕ

Производственные показатели обогатительной фабрики Cosmos, в целом, соответствуют плановым показателям, принятым в программе устранения узких технологических мест. Общие показатели содержания в концентрате и извлечения по фабрике сохранены несмотря на то, что на переработку поступает более труднообогатимая руда (исходное содержание никеля в питании уменьшилось, отношение никеля к мышьяку в питании повысилось, отношение массивной руды к вкрапленной понизилось – все эти факторы негативно влияют на извлечение никеля), а производительность фабрики увеличена.

С минералогической точки зрения, рисунки 27 – 29 указывают на увеличившееся извлечение пентландита во всех фракциях. Извлечение несульфидных жильных минералов осталось. в целом, прежним, однако этот показатель станет основным объектом внимания при планируемой доработке стадии перечистной флотации в новой технологической схеме.

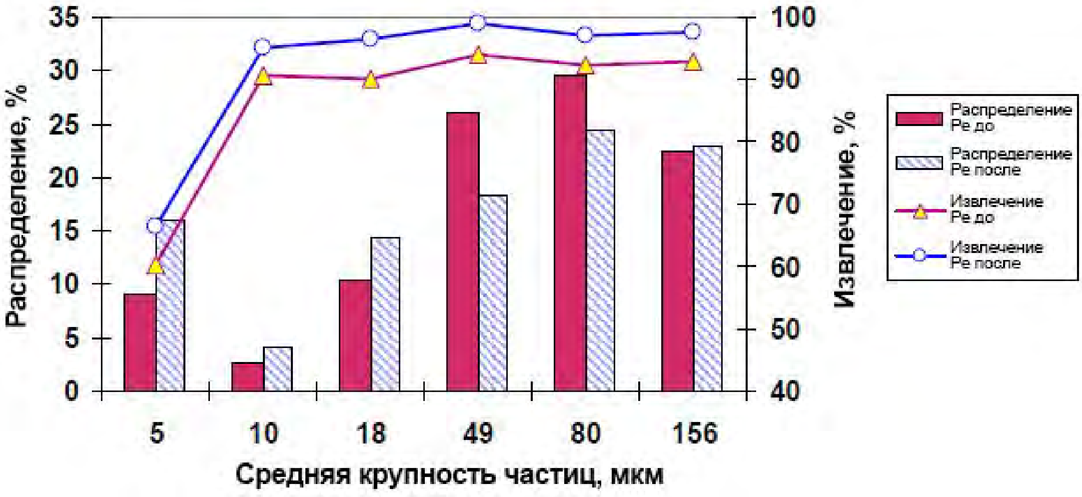


Рисунок 26. Сравнение показателей извлечения пентландита

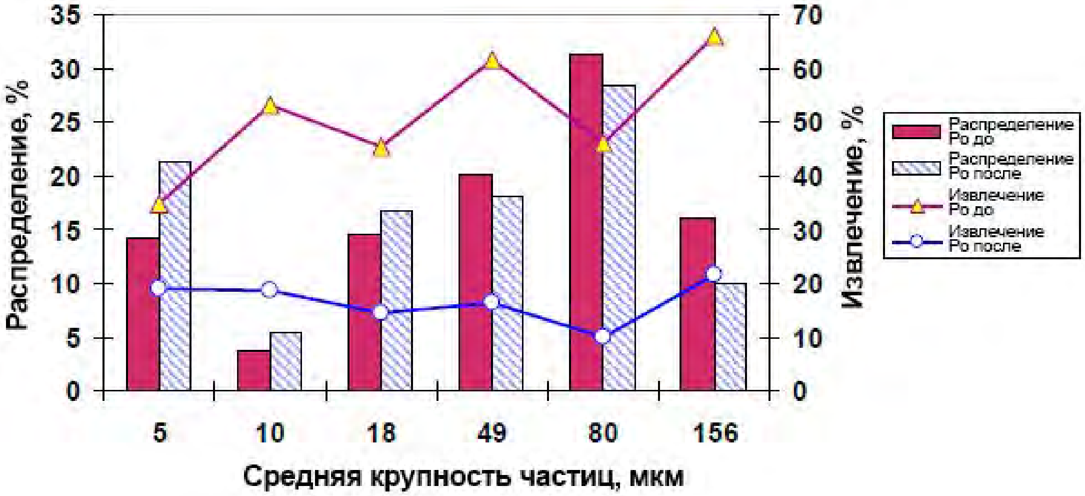


Рисунок 27. Сравнение показателей извлечения пирротита

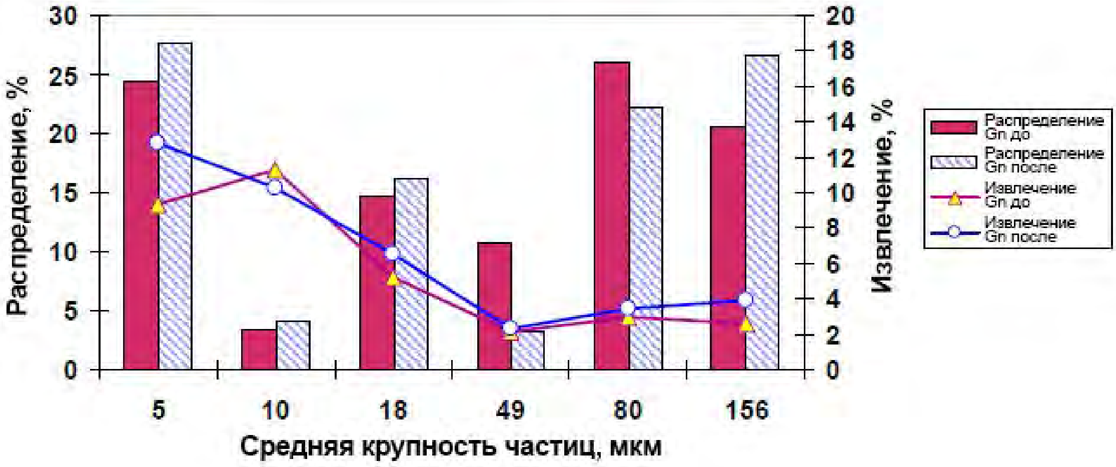


Рисунок 28. Сравнение показателей извлечения несульфидных жильных минералов

Большая часть принятых к настоящему времени мер по устранению технологических мест была направлена на доработку существующего оборудования и установку основного дополнительного оборудования, такого как мельница IsaMill™ M500, флотомашина Джеймсон Z1600 и флотомашина Wemco® SmartCell™, преимущественно с целью увеличения производительности обогатительной фабрики. Как показано на рисунках 27 – 29, эта задача была выполнена без ущерба для производственных показателей фабрики.

Предстоящая доработка цикла перечистной флотации позволит в полной мере реализовать принцип “нужного оборудования в нужном месте” для повышения металлургической эффективности. Аналитические средства, такие как модальный анализ и расчеты извлечения по фракциям крупности, были и остаются незаменимым инструментом определения наилучшего применения имеющихся технологических средств для получения оптимальных показателей разделения минералов. Новый цикл позволит обогатительной фабрике Cosmos сохранить в будущем способность производить никелевый концентрат с минимальной себестоимостью.

БЛАГОДАРНОСТИ

Авторы выражают признательность компании Xstrata Nickel Australasia за разрешение на публикацию настоящей работы.

СПИСОК ИСПОЛЬЗОВАННОЙ ЛИТЕРАТУРЫ

Shouldice, T. and Lafreniere, T. (2008) G&T Metallurgical Report of the July 2008 Survey of the Cosmos Concentrator (Internal)

Côté, I. and Adante, J. (2009) Design, Start-Up and Optimization of Pb & Zn Flotation Circuits Utilizing Ultra-Fine Grinding Technology, 41st Annual Meeting of the Canadian Mineral Processors, Ottawa, Canada, January 20 to 22.

Huang, G., Grano, S. and Skinner, W. (2006) Galvanic Interaction between Grinding Media and Arsenopyrite and its Effect on Flotation: Part II. Effect of Grinding on Flotation, Int. J. Miner. Process. 78, pp. 198–213.

Finch, J. Rao, S., and Nesset, J. (2007) Iron Control in Mineral Processing, 39th Annual Meeting of the Canadian Mineral Processors, Ottawa, Canada, January 23 to 25

Shouldice, T. and Ma, W. (2009) G&T Metallurgical Report of the April 2009 Survey of the Cosmos Concentrator (Internal)

Comminution 14

*Larson M, **Morrison R, **Xie W, *Young M

Development of the Larson/Morrison IsaMill JKSimMet Model

*Glencore-XT
**JKMRC, SMI, the University of Queensland

Abstract

The IsaMill is a high intensity horizontal stirred mill utilizing small 2-6 mm ceramic grinding media for attrition grinding. Grinding duties range from feeds of up to 300 microns being ground to 40 microns and UFG grinds as fine as 5 microns. Being completely dissimilar to normal ball mill breakage it was desired to produce a model for this process in JKSimMet. The result of this study was the discovery that IsaMill breakage can be reliably predicted on a basis of energy versus the squared value of the percent passing given sizes. This relationship can be used to analyse circuit efficiency for varying feed sizes. By using this new relationship the product size distributions for new feed sizes and energies can be reliably predicted with simple math, something not previously possible with only a signature plot. This paper will provide details on the model development, validation and implementation.

Introduction

The IsaMill is a horizontal high speed attrition mill developed by Mt Isa Mines and Netzsch for the treatment of fine grained minerals at McArthur River Mine and Mt Isa where ultrafine grinds of sub 7 microns are necessary. (Johnson et al, 1998) As the development of the mill and ancillary items has progressed the mill has expanded its range of duties into relatively coarser concentrate and mainstream grinding duties with F_{80}'s of up to 300 microns. As acceptance of the mill has increased so too has the need to better understand the grinding action taking place. In 2006 Glencore-XT funded the JKMRC to develop a simulation model of the mill breakage. While the mill scaleup based on the proven signature plot is widely accepted, the signature plot has one major drawback in its comparison of P_{80} versus energy. The plot is only valid for the feed size and conditions tested. If a coarser feed size became part of the design criteria no adjustment was available as the signature plot line does not intersect zero energy at the feed size and cannot be reliably adjusted. A method is required to accurately predict the results of not only changing feed sizes but also different media size and rheological constraints.

IsaMill Grinding Variables

There are many variables within the IsaMill or any stirred mill; however, only a handful of these actually impact the grinding efficiency. Residence time and energy intensity may change but the actual energy input per unit of feed (kWh/t) to a given grind target is a quite robust number. Variation in factors such as disc shape, spacing, and size; mill speed, feed slurry density within a range, separator configuration, flowrate and media type as long as relatively round and an SG of 3.5-4 do not have much impact on the energy per tonne required for the same feed size distribution to achieve a target product size. This is well supported by experience as the original scaleup as performed by MIM went from lab units of 1.5 to 4 litre capacity to pilot and industrial scale units of 100 to 3000 litres per mill.

If the different mill configuration variables made a difference in grinding efficiency the team at MIM could be considered some of the luckiest engineers in the world in that each step scaled along the signature plot. Further, past work by Ming Wei Gao at CSIRO has shown the lab four litre IsaMill scaled from energy versus P_{80} to the full size 355 kW Century Zinc SMD's (Gao et al, 2007). This is despite the fact that the vertical SMD uses pegs to stir the media and slurry and the horizontal M4 and all IsaMills use discs at a higher speed. Similarly the lab scale IsaMills use round pegs with a large open area for a separator while full scale mills use square or rhomboidal fingers with far less relative open area. In all cases the mills are just stirring grinding media and slurry. With regards to mill speed the M4 has a tip speed of about 8.5 m/s, the M20 about 12-14 m/s and full scale mills about 20 m/s, yet they all scale from one another. The relative mill configurations also vary from size to size with regards to disc spacing and size compared to the shell volume. This suggests that the energy being absorbed by shearing the mixture of media and slurry may be the controlling factor. The sensitivity to changes in the shape of the feed size distribution also suggests that coarser fed particles may be contributing to grinding of the finer ones. Over complicating any of this serves no real purpose other than to distract from issues which actually do make a difference.

These can be summarized as media shape, media size, ore type, slurry viscosity (above an optimum point but not below that point) and feed size distribution.

Media shape is generally not considered as all but three IsaMill sites use modern ceramics which in most cases are generally round i.e. close to spherical. Full scale plant test work and pilot test work has shown that the round ceramic is generally 15-20% more efficient than competent sand or oblong shaped ceramics. Different ceramic manufacturers may claim energy efficiency advantages over competitors but the reality is that this media has become much more of a commodity where purchase price and wear rate (g/kWh) is a distinguishing factor, not grinding efficiency.

Media size has only been examined in limited detail but it is known that different size media will produce different energy versus grind slopes for each size fraction. Optimum media size will depend on the incoming ore feed size, desired product size, ore hardness and a myriad of other factors to lesser extents.

Slurry viscosity and its effect on stirred milling efficiency at fine sizes is poorly understood at this time. In general Glencore-XT will try to operate mills at about 20% solids by volume or 20-24 centipoise on the mill discharge. Above these values energy efficiency is lost. Below these values the efficiency will flatline before decreasing once the slurry is dilute enough that energy is spent contacting media against media with no ore particles present to be ground. Ideally in the future there will be a reliable model to predict viscosity based on the surface area present in a given volume calculated from the solid SG, slurry SG and laser sizer product size distribution.

Signature Plot and Scaleup

The ability to predict energy for varying feed size distributions is the main goal of the work described in this paper. The IsaMill signature plot with energy (kWh/t) versus P_{80} on a log-log plot is widely accepted as scaling to full scale IsaMills. This has been published over the past two decades showing accurate 1:1 scaleup from M4 and M20 test mills to the M1000, M3000 and M10000 IsaMill. Glencore-XT maintains relationships with a dozen independent laboratories around the world. The robustness and repeatability of the signature plot test was shown in IsaMill 1:1 Direct Scaleup from Ultrafine to Coarse Grinding presented at Comminution 12. However, it has one large limitation in that it does not cross zero energy at the feed size. Thus it can only be used for the feed size tested. If the same P_{80} is to be fed to the mill and the rest of the size distribution is different - for example far fewer fines or a sharper size distribution - then the signature plot will not be valid. A new method is necessary to provide the flexibility required to be useful for a JKSimMet model which must to able to handle reasonable variation in feed rate and feed size distribution in a reasonable manner. The proven robustness of the signature plot though means that any model will still use the signature plot. Given the fact that the entire size distribution, not just the P_{80}, can be determined from the signature plot test it can be assumed that any new model method based on the same information will be successful.

Test Mill Coarse Material Retention

Based on previous work it cannot be stressed enough that proper test conditions and data analysis are vital to the successful development and implementation of any IsaMill model. Without careful planning and observation results can deceive the user into thinking the mill performed better than reality. The combination of too small media and not enough sample volume can easily result in coarse material being retained in the lab mill. In these cases the too small media will only grind the fines and not have enough energy to break the coarsest material. The coarse material will be held in the mill by the centripetal forces as there is not enough material passing through the mill for those particles to fully build up where they would be discharged. This is usually recognized by either an unusually high power draw or a low density reading on the discharge.

Flaws in Other Models

It is well accepted that the Bond equation becomes much less reliable at product sizes finer than about 70 microns. As ore is ground finer, the key Bond assumption that the next increment of input energy will produce a similar increment of fine material (i.e. g/rev of the test mill) becomes invalid and the required energy for each increment of fines generated increases at finer product sizes. This means that the relationship between grind size and energy is not linear at fine sizes.

As shown in Figure 1, grinding from 50 to 33 µm requires significantly more power than the BWI suggests, with a much higher slope (-1.0345). Grinding finer still needs ever more energy. All three points on the graph come from the same test but the finer points look to be less energy efficient in terms of a Bond comparison.

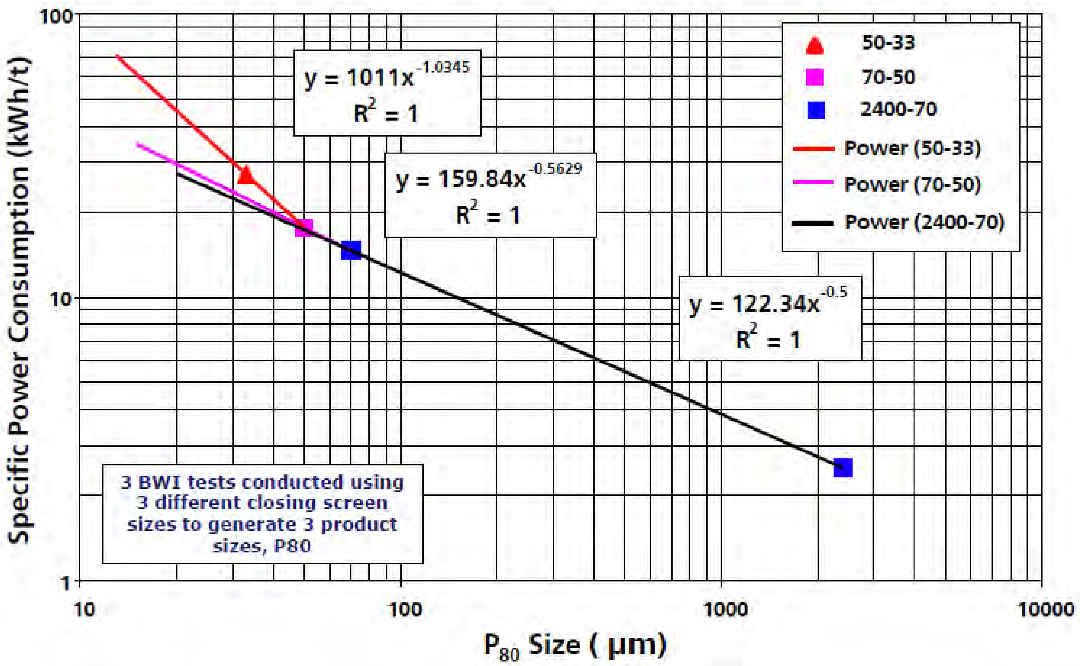


Figure 1. Flaws in using the Bond equation to compare fine grinding energy efficiency

Further, most Bond Work Index data is developed by testing run of mine ore. Most IsaMill feed will be a stream with quite different characteristics. The Bond test uses steel balls, while the IsaMill uses ceramic grinding media. In almost all cases the ceramic media will be harder than the ore it is grinding and of roughly similar stiffness. The steel balls used in a Bond test can vary widely in hardness compared with the ore being ground and will always be much more elastic. Stiffer media transfer energy more efficiently than elastic media.

The commonly used appearance function will also not be considered in this case as it assumes every particle breaks with the same progeny. The breakage rates are modelled but not changed for different energy, feed sizes or media size.

Other researchers (Mannheim, 2011) have attempted to utilize the existing signature plot line to predict results at different feed sizes. This is shown in Figure 2.

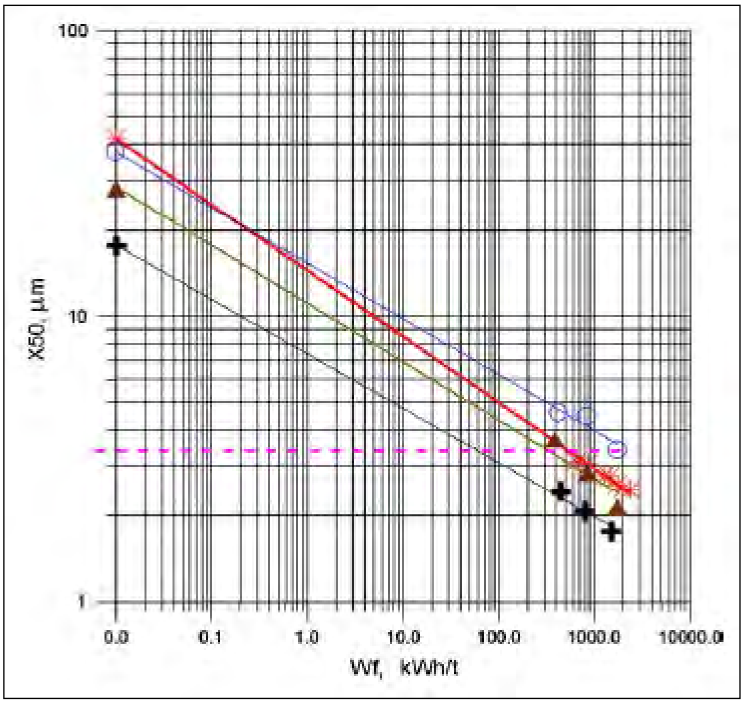
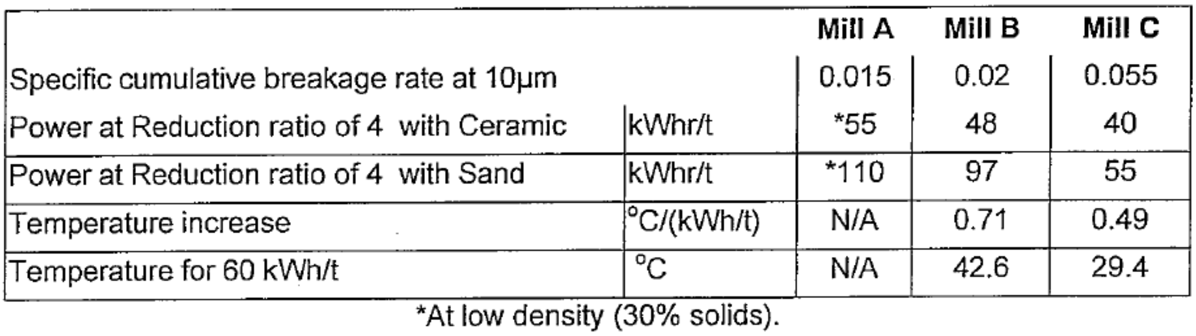


Figure 2. Relation between the grinding fineness and specific grinding work for several materials (Mannheim, 2011)

Forcing the signature plot lines through the origin in Figure 2 moves the lines from the original plots using only the mill products. While it may appear accurate at first glance, further inspection reveals that the energy predicted by the resulting line is in error by 200-300% compared to where the points actually sit. The use of the log-log plot with inclusions from 0 to 1 kWh/t and large symbols have effectively hidden this from casual observation. If samples had been taken below 100 kWh/t they would better demonstrate this inaccuracy. This approach cannot be considered sufficiently reliable for a model that will be used to size industrial machinery.

Reduction ratio type comparisons have also proven to be popular. In the example from Tati Nickel contained in Table 1, energy requirements were measured to differentiate between three types of stirred mills.

Table 1. Summary of mill performances (Nel et al, 2006)



		Mill A	**Mill B**	**Mill C**
Specific cumulative breakage rate at 10µm		0.015	0.02	0.055
Power at Reduction ratio of 4 with Ceramic	kWhr/t	*55	48	40
Power at Reduction ratio of 4 with Sand	kWhr/t	*110	97	55
Temperature increase	°C/(kWh/t)	N/A	0.71	0.49
Temperature for 60 kWh/t	°C	N/A	42.6	29.4

*At low density (30% solids).

Nel et al concluded “For the horizontal ultra fine mill with ceramic grinding media, 2.4 % of the particles bigger than 10 microns were reduced to less than 10 microns using 1 kWh/t specific energy input, thus 33.3 kWh/t specific energy was required to mill Phoenix concentrate to 80 % passing 10 microns.”

The math used is basically that it took 1 kWh/t to reduce 2.4% of the feed to under 10 microns. Therefore it must take 33.3 kWh/t to reduce 80% of the feed to below 10 microns. 80%/2.4%=33.3. This approach while being convenient is not necessarily appropriate.

When examining grinding efficiency in this way the examples with the coarsest feed and or product will always appear to be the more efficient option. For reduction ratios to be a valid option in modelling of fine grinding, the relationship between grind size and energy would have to be linear. Unfortunately this is not the case and each step finer requires relatively more energy for smaller and smaller reductions in size.

Figure 3 is another example of a reduction ratio model used to justify a process decision, in this case comparing media type and size. However if the extreme point from the test work was used, the entire graph could be populated by columns of dots. For example, grinds of 60 microns to 30 microns and 40 microns to 20 microns both have reduction ratios of 2, but the finer example will require more energy.

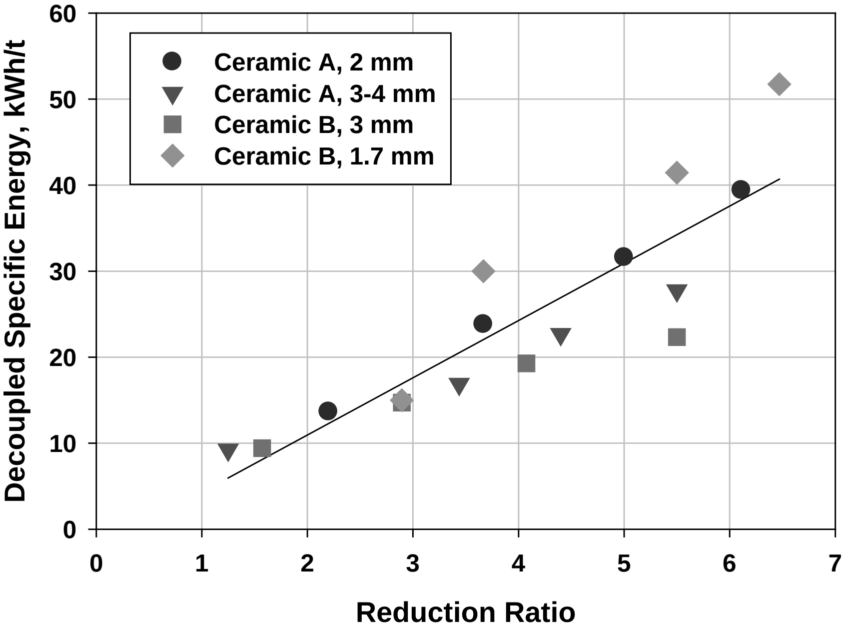


Figure 3. Comminution energy versus reduction ratio (Farber et al, 2010)

For these reasons reduction ratios are of limited use for comparison or prediction tools when dealing with fine grinding.

The Squared Function Dependence

Using the Finch McIvor method for particle production in ball mills as a guide, the first attempt at modelling the IsaMill on a size by size particle basis was made. While the relationship of energy versus percent passing of a certain size is linear for the ball mill, for the IsaMill the relationship was not linear. However it was observed from the curve produced that the relationship approximated a squared function.

Plotting these first attempts on energy versus percent passing 10 µm squared resulted in Figure 4. Not only does the squared function produce a straight line but that line passes through the feed size at zero energy.

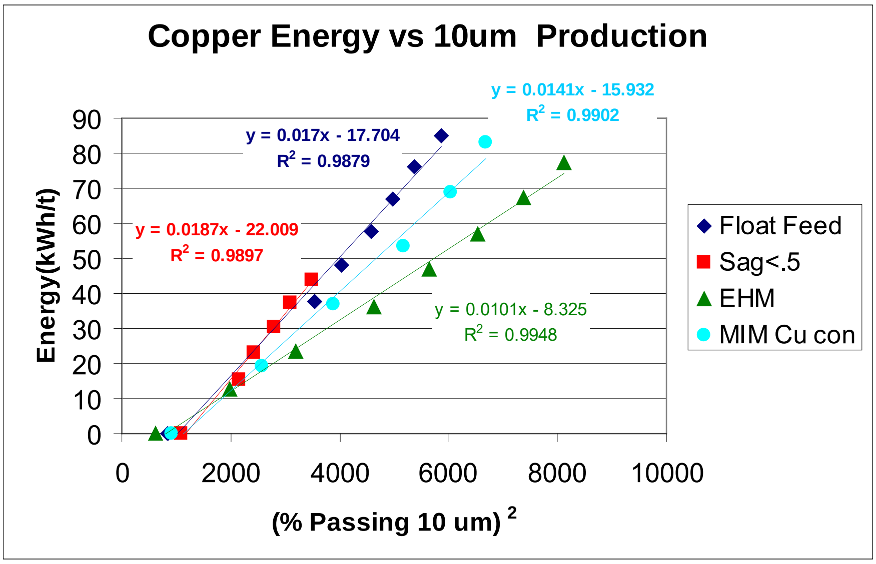


Figure 4. Fines production of various copper ores (Larson, 2013)

This shows promise that the mill is creating surface area through fines in a predictable reliable manner.

It was initially thought that by switching the axes, a simple easy to understand model could be used with the squared function as the basis. This is shown in Figure 5. A plot is developed using an original feed and energy values. Then for a new feed the new size is plotted and the line from the original feed is just moved up or down the y-axis while keeping the squared function lines parallel. The new energy to a given percent passing the size of interest can then be calculated.

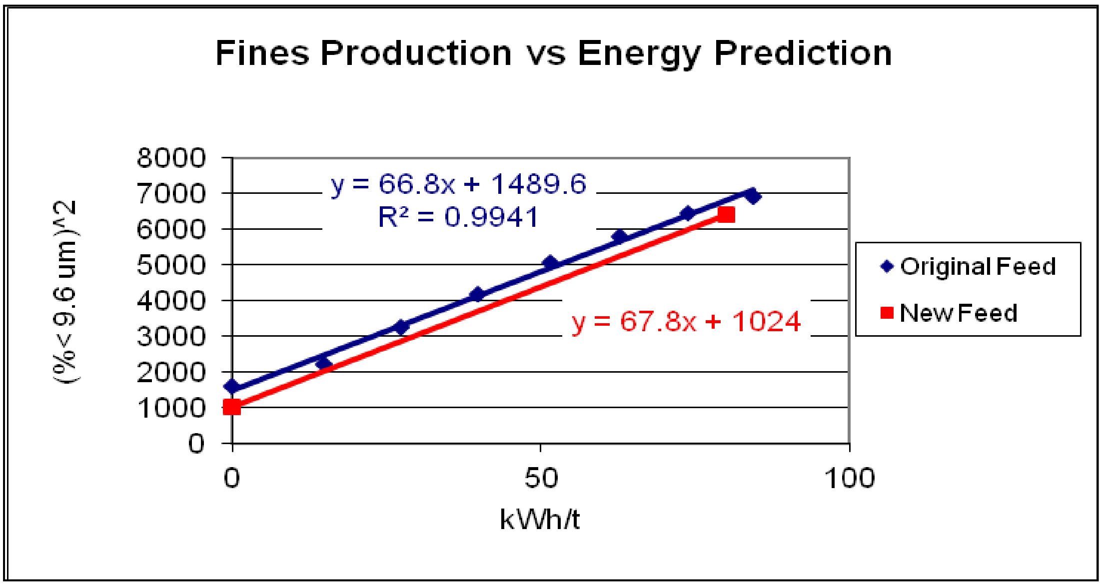


Figure 5. Graphical form of IsaMill squared function for fines production model (Larson, 2013)

Validation of Squared Dependence in MS Excel

To validate the basic idea of the squared function, the JKMRC ran a test program with multiple ores ground to varying size distributions in a rod mill. Different feed sizes were then ground in the M4 IsaMill using the same media size distribution to determine if one test could be used to predict the results of the other.

In the first case shown in Table 2, MIM run of mine copper ore was ground down to P_{80}'s of 72 and 47 microns. Bothe samples were processed through the M4 IsaMill using the same 2.5 mm top size graded ceramic media charge. The raw data for these two tests is shown below, with energy per pass, percent passing a given size and the percent passing that size squared.

Table 2. MIM Copper ROM sizing and energy per pass data (Larson, 2013)

	Energy	%<	%<^2										
		2.4 µm		5 µm		9.6 µm		13.5 µm		19 µm		27 µm	
P_{80}= 72µm	0	15.00	225.00	25.28	639.08	38.13	1453.90	45.00	2025.00	51.48	2650.19	58.11	3376.77
	9.3	18.33	335.99	31.46	989.73	48.41	2343.53	57.57	3314.30	66.38	4406.30	74.96	5619.00
	19.2	20.62	425.18	35.72	1275.92	55.58	3089.14	66.40	4408.96	76.49	5850.72	85.32	7279.50
	28.9	23.31	543.36	40.10	1608.01	62.33	3885.03	74.40	5535.36	84.58	7153.78	92.14	8489.78
	37.2	24.31	590.98	42.16	1777.47	65.77	4325.69	78.17	6110.55	88.05	7752.80	94.76	
				44.10	1944.81	69.31	4803.88	81.93	6712.52			96.91	
P_{80}= 47µm	0	16.23	263.41	28.11	790.17	43.49	1891.38	51.70	2672.89	59.58	3549.78	67.50	4556.25
	8.3	19.26	370.95	33.63	1130.98	52.71	2778.34	63.09	3980.35	72.81	5301.30	81.62	6661.82
	17.3	21.40	457.96	37.34	1394.28	58.51	3423.42	69.94	4891.60	80.24	6438.46	88.68	7864.14
	24.9	22.73	516.65	40.17	1613.63	63.30	4006.89	75.75	5738.06	86.17	7425.27	93.64	8768.45
	32.2	24.67	608.61	43.09	1856.75	67.48	4553.55	79.92	6387.21	89.43	7997.72	95.65	
	39.5	26.45	699.60	45.99	2115.08	71.83	5159.55	84.05	7064.40			97.37	

The main basis for this model is the assumption that the squared function lines for varying size fractions will be parallel for different feed size distributions and that the predictive process only has to move those lines up and down the y-axis depending on the amount of a given size present in the new feed to predict the new energy.
The results of doing this are shown in Figure 6 with individual lines for 27, 19, 13.5, 9.6, 5 and 2.4 microns plotted for both feeds tested.

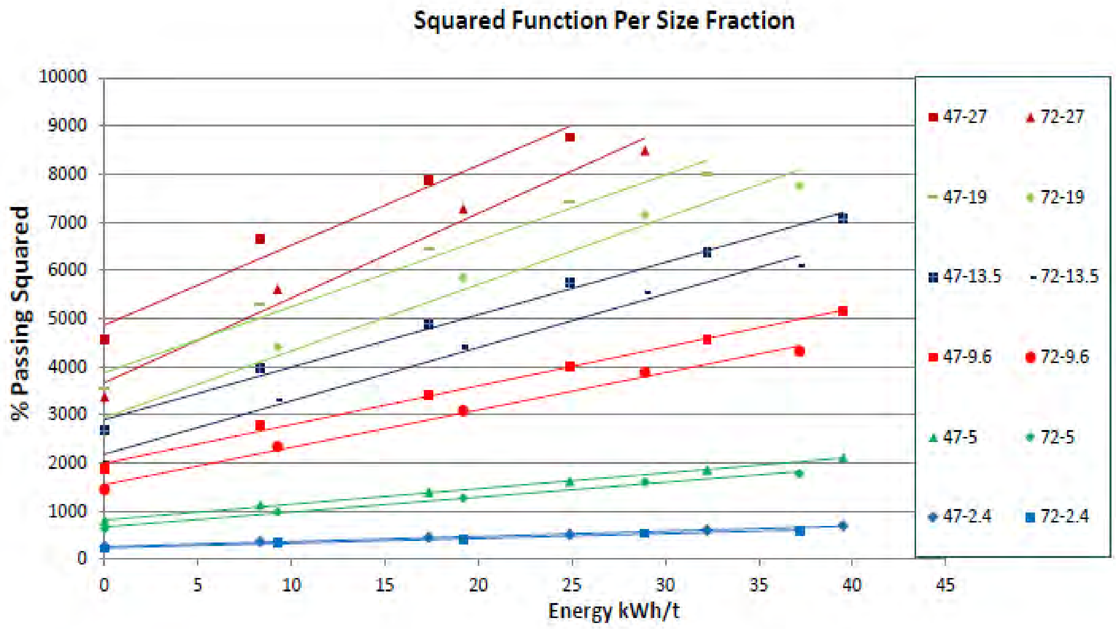


Figure 6. Complete comparison of different feed size squared function lines

It is assumed that as media size is changed to larger media, the lines for the coarsest size fraction will steepen and the finer size fraction will flatten. For smaller media, the coarsest fraction will flatten and the finer fraction will steepen as fine particles are preferentially broken compared to larger particles by the smaller media.

The actual slopes and error associated with Figure 6 are detailed in Table 3 below. The results of the 47 micron feed were plotted and used to predict the product size distribution shown in Figure 7. Given the +/- 5% error associated with the M4 signature plot test this can be considered a successful demonstration of the squared function model.

Table 3. MIM Copper ROM squared function predicted versus actual (Larson, 2013)

Size fraction (microns)	Slope 47 micron feed	Slope 72 micron feed	Error	New feed intercept	New feed predicted %passing	New Feed actual %passing
2.4	10.658	10.013	6.05%	225	24.93	24.31
5	32.6	30.85	5.37%	639.1	43.03	42.2
9.6	80.47	77.63	3.53%	1453.9	66.69	65.77
13.5	108.81	110.79	1.82%	2025	77.93	78.17
19	136.65	138.2	1.13%	2650.2	87.94	88.05
27	165.72	175.56	5.94%	3376.8	97.68	95.65

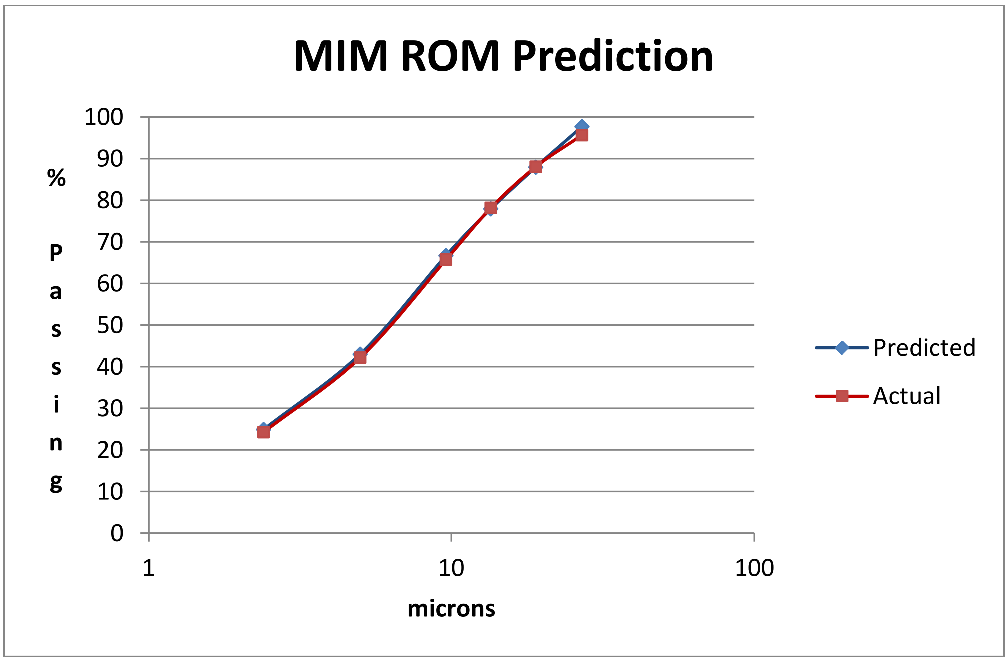


Figure 7. Predicted versus actual product size distribution MIM Copper ROM using the squared model (Larson, 2013)

A second validation example is detailed in Table 4. In this case, MIM lead/zinc run of mine ore was ground in a pilot rod mill to P_{80}'s of 131 and 68 microns. Both samples were then processed with the same 3.5 mm top size graded charge of media.

Table 4. MIM Lead Zinc ROM sizing and energy per pass data (Larson 2013)

	Energy	%<	%<^2										
P_{80}=		2.4 µm		5 µm		9.6 µm		13.5 µm		19 µm		27 µm	
131 µm	0	17.95	322.32	25.92	672.02	34.11	1163.72	38.02	1445.52	41.75	1742.78	45.71	2089.40
	7.9	27.14	736.58	40	1600.00	53.88	2903.05	60.64	3677.82	67.24	4521.89	74.36	5529.41
	15.0	31.14	969.70	46.61	2172.49	63.93	4087.68	72.9	5314.41	81.51	6643.88	89.385	7989.68
	22.0	34.22	1171.01	51.81	2684.28	72.42	5244.66	82.42	6793.06	90.44	8179.39	96.02	
	28.7	36.075	1301.41	55.365	3065.28	77.89	6066.85	87.63	7679.89	94.31		98.14	
	35.6	38.04		58.68		82.1		91.11		96.45		99.01	
P_{80}=	0	23.453	550.06	33.633	1131.20	44.29	1961.90	49.43	2443.65	54.31	2950.30	59.573	3548.98
68 µm	6.9	28.6	817.96	42.32	1790.98	57.56	3313.15	65.44	4283.05	73.37	5383.16	81.715	6677.34
	13.8	32.09	1029.45	48.265	2329.51	66.94	4481.63	76.63	5872.92	85.50	7311.11	92.82	8615.55
	20.0	34.61	1197.85	52.77	2784.67	74.36	5529.41	84.44	7130.11	92.05	8473.20	97.05	
	26.0	36.95	1365.30	56.44	3185.47	79.11	6259.18	88.72	7872.13	95.01		98.39	
	32.4	37.94		58.94		82.83		91.77		96.86		99.17	

In this test, the slopes matched up very well, with most error at the coarsest size fraction. This is shown in Table 5 and in Figure 8.

Table 5. Lead/Zinc ROM squared function prediction versus actual (Larson, 2013)

Size fraction (microns)	Slope 68 micron feed	Slope 131 micron feed	Error	New feed intercept	New feed predicted %passing	New feed actual %passing
2.4	30.88	33.63	8.18%	322.3	34.76	36.075
5	78.36	82.41	4.91%	672.02	54.05	55.37
9.6	166.08	170.33	2.50%	1163.72	77.01	77.89
13.5	210.87	218.63	3.55%	1445.52	86.59	87.64
19	276.89	294.01	5.82%	1742.78	98.31	94.31
27	367.1	393.84	6.79%	2089.4	112.35	98.14

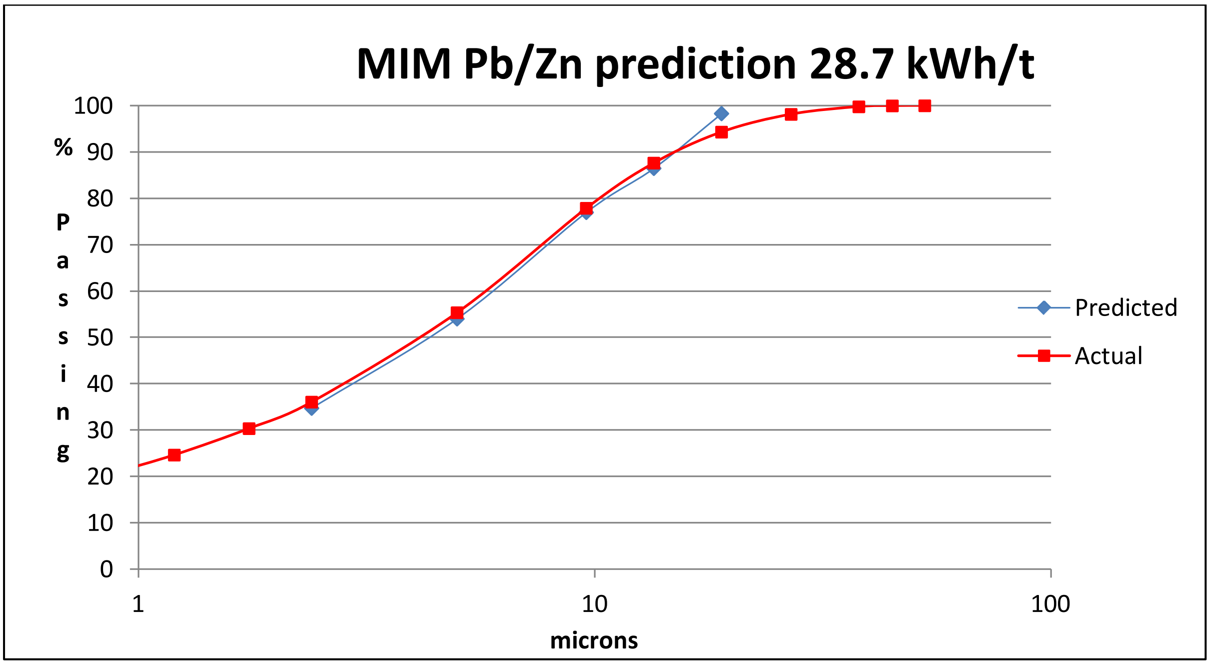


Figure 8. Predicted versus actual product size distribution MIM Lead/Zinc ROM (Larson, 2013)

The coarsest sizes tend to be more difficult to predict. As the sizes get closer to the P_{100} there is much less of the material to be created and the curve starts to flatten and give a worse prediction if the data is not cut off correctly. It is also generally a size fraction that is more likely to be held in the mill and is difficult both to sample and to measure correctly in the Malvern.

JKSimMet Model Implementation and Testing

Despite the apparent success of validating the model in MS Excel, it was still necessary to prove that JKSimMet programming could be implemented to duplicate these results without manual adjustments to the data.

The original signature plot data was provided to the JKMRC along with the feed size to a second test. Besides the new feed size distribution two energies were provided with the test from which the JKSimMet model was to predict the resulting product size distribution curves. The results comparing the model values to what was actually ground in the M4 IsaMill are shown in Figures 9 and 10 for two MIM copper examples and one MIM lead/zinc sample.

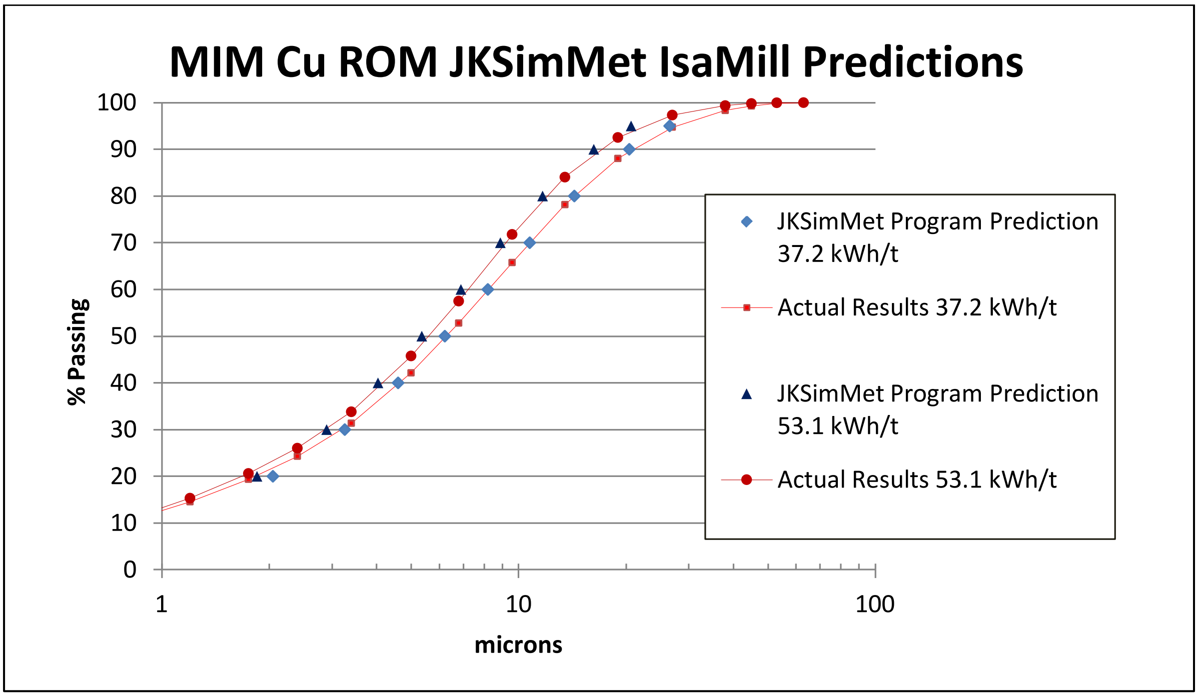


Figure 9. JKSimMet IsaMill model predictions versus actual for MIM Copper ROM

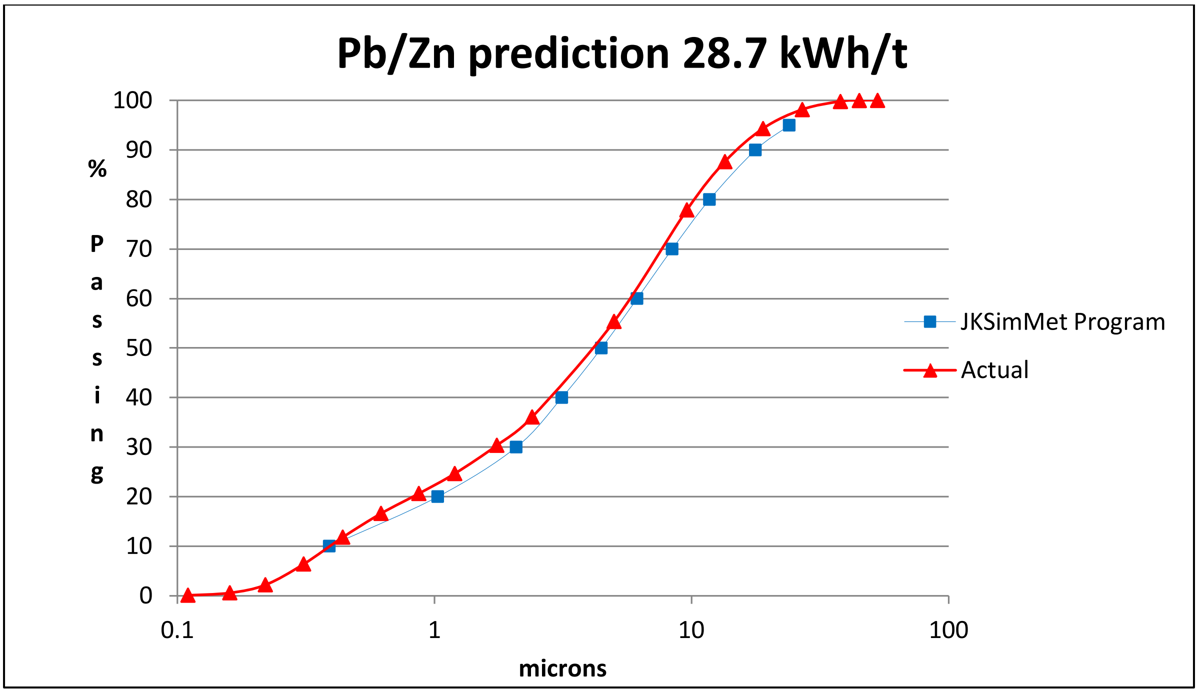


Figure 10. JKSimMet IsaMill model predictions versus actual for MIM Lead/Zinc ROM

These cases can be considered a success as they accurately match the results generated without the need for manually adjusting the settings. In this case all original data over a P_{95} is ignored automatically by the program.

Implementation into JKSimMet

A brief summary of the model implementation is included. The simplicity of the squared function allows for a relatively easy to understand model.
The IsaMill spreadsheet model was first converted into a MatLab program to allow ease of parallel testing. Once this testing was complete, the ability to predict the full product size distribution was added using spline functions.
The model was then converted into Fortran within the JKSimMet Model Developers kit.
At this point, the full calculation of the signature plot from measured data was also added to ensure a reproducible approach to interpolation and calculation.

The squared function is applied as follows:
The first step is to determine the linear relationship of energy versus the square of percentage passing 9.6um (or other size fraction that is valid for use) from the original feed. Any data from an individual pass that is coarser than the P_{95} for that pass is disregarded;
The second step is to calculate the point of new feed with zero energy input in the map of energy vs. the square of percentage passing 9.6um;
The third step is to plot a line parallel to the line in the first step from the point of new feed of the second step;
The final step is to find the value of the percentage passing 9.6um at the new input energy.
After the above method applied for a few size classes, the product size distribution can be interpolated.
The model was then tested against a range of reliable test work data.

Using the Model as an Evaluation Tool

While reduction ratios cannot be used to evaluate the efficiency of different fine grinding options, the squared function can be used to a certain extent. With just the IsaMill model, different variables including feed sizes can be tested for efficiency. In the case of only comparing IsaMills, the slope of the squared function line at a size or sizes of interest can be compared. The steeper the line the more energy efficient the grind is going to be. This is not proven to work when comparing different technologies. When comparing against a ball mill or tower mill or other high speed stirred mill, the comparison would only be valid if both received the same feed size. In those cases the squared function line could be developed for the IsaMill and just the product vs. energy point plotted for the other technology. If that point was to the right of the IsaMill squared function line, that option would be less energy efficient and if it were to fall to the left of the IsaMill line it would be more efficient. If the feed sizes were different, the analysis would be invalid as the shape of the line is yet to be established for other technologies.

Future Work

Future work to fully develop this model will focus on how a change in media size affects the individual particle size slopes along with how mineralogy and ore SG will impact on energy efficiency. It is thought that as media size increases the spread between slopes will increase. That is the finer particle lines will flatten and the coarser particle lines will steepen. These need to be able to be predicted to widen the size ranges that can be predicted where the change in feed size necessitates a change in media size.

Predicting the effects of mineralogy with regards to energy will be more difficult. There are some observations in XT test work database that a lower solid specific gravity in magnetite feed will result in a higher required energy to grind to a given target size. This may be due to the lower density material having more particles per tonne needing attrition breakage than a higher density magnetite feed along with the extra silica generally being harder. This explanation is more difficult when it comes to most copper and gold ores where the gangue material can vary more widely than just being silica. A pyrite gangue can increase the solid density but depending on the pyrite can either be very hard or quite brittle.

It is also accepted that the relationship between grinding achieved and incremental energy input becomes even more inefficient at still finer sizes. Hence, an increased exponent may be required for products very much finer than the “normal” IsaMill range.

Acknowledgements

The authors wish to thank Glencore-XT and the JKMRC for allowing this work to be published and Ming Zhao He and Paul Kay for performing the test work that went into the validation examples and improving the understanding of these breakage mechanisms.

References

Farber B, Durant B, Bedesi N, 2010. Effect of Media Size and Mechanical Properties on Milling Efficiency and Media Consumption; Comminution 10; Cape Town, South Africa

Gao M, Reemeyer L, Obeng D, Holmes R; 2007. Efficiency of the Detritor Mills at Zinifex Century Mine; AUSIMM proceedings of 9th Mill Operators’ Conference

Johnson, N W, Gao, M, Young, M F and Cronin, B, 1998. Application of the IsaMill (a horizontal stirred mill) to the lead-zinc concentrator (Mount Isa Mines Ltd) and the Mining Cycle, in Proceedings AusIMM ’98 – The Mining Cycle, pp 291-298 (The Australasia Institute of Mining and Metallurgy: Melbourne)

Larson M; 2013. Experimental Study of IsaMill Performance Leading to a Preliminary Model, Master of Philosophy thesis, University of Queensland-JKMRC (unpublished)

Mannheim V, 2011. Empirical and scale-up modeling in stirred ball mills, Chemical Engineering Research and Design 89, (2011) 405-409

Nel G, Swarts A, Donegan S, Johnson G, 2006. Comparing the milling Efficiencies of Pilot Scale Horizontal and Vertical Ultrafine Grinding Mills for Tati Sulfide Concentrate; Proceedings of the Randol Innovative Metallurgy Forum 2005, Perth, Australia

Мельницы доизмельчения: проблемы масштабирования

Майкл Ларсон[1], Грег Андерсон[1], д-р Роб Моррисон[2], Майкл Янг[1]
1- Xstrata Technology
2- Julius Kruttschnitt Mineral Research Centre

Автореферат

По мере того как руды становятся более мелкозернистыми, растет потребность в доизмельчении перед обогащением или выщелачиванием. Несмотря на это, не существует стандартной методики расчета расхода энергии на измельчение ниже 70 микрон. Стандартный тест Бонда применим к крупному измельчению в шаровой мельнице, но не подходит для измельчения до 70 микрон с перемешиванием тонкой мелющей среды. В отсутствие стандартной методики, энергопотребление нередко определяется на основании расчетов поставщика или по аналогии с другими применениями. Однако поставщики используют самые разнообразные методы масштабирования, что влечет большой разброс в результатах расчета энергопотребления. Даже для похожих мельниц расчеты могут расходиться на 100-300%. Это неизбежно приводит к серьезным ошибкам, т.е. установке мельницы сильно завышенной или сильно заниженной мощности. В данной статье эта проблема рассматривается путем сопоставления фактической производительности промышленных мельниц доизмельчения с их первоначальными проектными расчетами. Анализ подтверждает, что серьезные просчеты в масштабировании действительно допускаются, а затем тиражируются при выборе мощности "по аналогии".

Рассмотрены условия проведения испытаний, позволяющие получить точную оценку мощности мельницы. Обязательными являются следующие условия испытания: тестирование в непрерывном режиме (не серии тестов), обеспечение установившегося режима (в мельнице не должно оставаться крупной фракции), внесение поправки на классификацию, непосредственное (а не опосредованное) измерение расхода энергии, применение среды такой же крупности, как и в промышленной установке. Несоблюдение даже одного из этих условий может повлечь погрешность до 40%, а несоблюдение нескольких условий увеличивает погрешность в несколько раз.

Введение

Потребность в применении мельниц доизмельчения растет по мере перехода на более сложные и мелкозернистые руды. Однако точные методики для проектного расчета измельчения пока отсутствуют. Стандартной отраслевой методики испытаний измельчения с перемешиванием не существует. Тест Бонда и другие испытания для решения задач масштабирования, рассчитанные на более грубый помол, не применимы к измельчению с перемешиванием тонкой мелющей среды до крупностей менее 70 микрон. Использование предназначенных для более грубого помола методик применительно к более тонкому доизмельчению влечет занижение расчетного расхода энергии вследствие разницы в размере шаров, в траекториях и во взаимодействии между частицами руды и шарами и между шарами и корпусом в небольших мельницах. Видоизмененные тесты, основанные на тесте Бонда, такие как тест Левина, более точны, но нередко требуют слишком больших затрат времени и слишком большого объема пробы. В результате поставщики мельниц доизмельчения разработали собственные методики масштабирования. Однако последний опыт показывает, что такие методики приводят к расхождению 300-400% в результатах расчетов затрат энергии на одинаковые задачи. В ряде случаев расчеты для одной и той же мельницы, выполненные разными производителями, отличались более чем на 100%. Очевидно, что такая ситуация не является приемлемой. Проектировщикам необходима стандартная методика правильного расчета энергозатрат и точные данные о фактических эксплуатационных (а не проектных) показателях для выполнения сопоставительного анализа. Без такой стандартной методики велико искушение выбрать наименьшее значение энергопотребления, основанное, например, на аналогии с другими проектами, а не на эксплуатационных показателях. Такой подход приведет к бесконечному воспроизведению неправильных решений и к провалу проектов.

В настоящем исследовании фактические характеристики доизмельчения сопоставляются с проектными показателями. Неудачный опыт проектирования объясняется недостатками лабораторных методик. Рассмотрены разные методики и их интерпретации, а также описаны условия точного масштабирования любых технологий доизмельчения.

Методика масштабирования для IsaMill^{TM}

Компания MIM спроектировала IsaMill^{TM} в начале 1990-х годов вследствие потребности в тонком измельчении руды из разрабатываемых ею рудных тел. Поскольку инженеры MIM одновременно занимались эксплуатацией, им требовалось точное масштабирование. Кроме того, они имели возможность скорректировать расчеты с использованием эксплуатационных показателей мельниц. В результате они разработали методики точного масштабирования 1:1 от лабораторной до промышленной установки, используя мельницы IsaMill мощностью 1,1 МВт и объемом 3 000 л (M3000) на рудниках Джордж Фишер (MIM) и MRM (рис. 1). При использовании этой методики IsaMill^{TM} M4 (4 л, горизонтальная) работает в многопроходном режиме, при котором продукт одного непрерывного прохода становится питанием для следующего прохода, с получением характерного графика энергопотребления в зависимости от крупности продукта. Мельница заполняется на 80% мелющей средой той же крупности, что и в промышленном применении (как правило, 1-3 мм), и запускается на скорости 1500 об/мин. Конструкция и режим работы мельницы - диски, скорость вращения, крупность среды, тип среды и наполнение, а также сепаратор на выходе - очень близки к промышленной мельнице. Поэтому поступающие частицы исходного материала попадают в такие же условия измельчения и классификации, что и в промышленной установке. Поскольку объем мелющей среды относительно размера испытательной установки невелик, эффект корпуса мельницы незначителен, что подтверждается масштабированием в соотношении 1:1.

Масштабирование M3000 на MIM и MRM

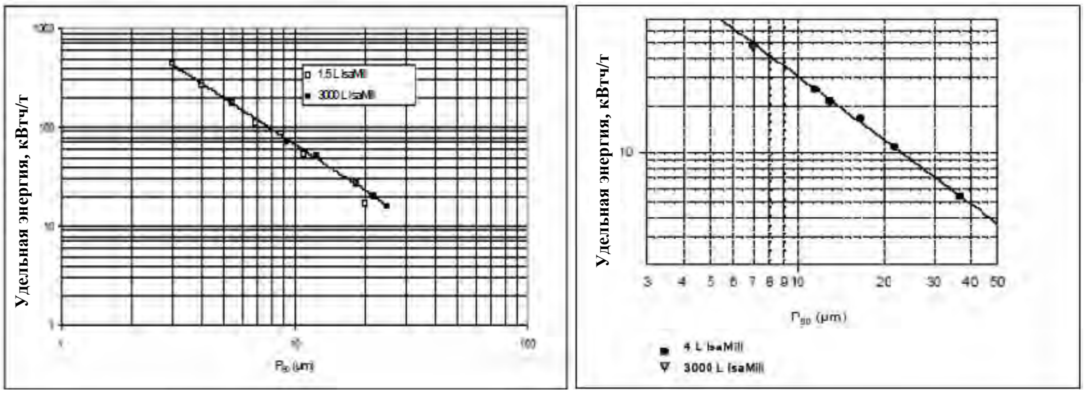


Рисунок 1. Масштабирование IsaMill M3000 на MIM (слева) и MRM (Barns K, Curry D; 2006)

Причины точного прямого масштабирования лабораторной установки IsaMills до промышленной следующие:

- Используется мелющая среда такой же крупности и такого же типа, что и в промышленной установке, - поправочного коэффициента не требуется.
- Конструкция и механизм мельницы одинаковы - горизонтальная мельница, аналогичные диски измельчения с близкой окружной скоростью, встроенный сепаратором, аналогичный промышленному. Таким образом, механизмы, скорости и физические процессы в лабораторной мельнице максимально приближены к промышленной мельнице.
- Лабораторная мельница имеет большие размеры относительно частиц материала и мелющей среды, поэтому «эффект корпуса» незначителен (расстояние между корпусом и дисками измельчения является достаточным).
- Испытания проводятся при такой же крупности и плотности подаваемой пульпы.
- Перерабатывается достаточное количество питания, чтобы установка вышла на устойчивый режим. Каждая точка характерного графика является результатом одного непрерывного прохода материала через мельницу, при котором мельница выходит на устойчивый режим. Содержимое мельницы должно замещаться несколько раз за проход, так как при

сохранении даже небольшого количества крупных частиц расчетные энергозатраты промышленной установки в устойчивом режиме оказываются существенно занижены. Это принципиально важно, и этого невозможно добиться при проведении серии тестов, потому что не происходит непрерывного поступления нового материала и выйти на устойчивый режим невозможно. Требование об устойчивом режиме является обязательным при тонком измельчении; к счастью, его можно выполнить при относительно небольших объемах проб (15 кг).

- Измерение потребляемой энергии непосредственно на валу агитатора, т.е. так, как это делается на промышленных мельницах. Это также позволяет точно вычислить мощность холостого хода, которая необходима для масштабирования лабораторной мельницы до установки любого размера.

Эта методика, разработанная для предприятий MIM, в настоящее время успешно применяется и на других установках. На рис. 2 показана характеристика масштабирования мельницы IsaMill M10000 (2,6 МВт / 3500 л.с.) компании Anglo Platinum.

Модель IsaMill	**Установленная мощность, кВт**	**Объем камеры, л**	**Удельная энергия, кВт-ч/т**	**Содержание твердого в пульпе, %**	**P98, мкм**	**P80, мкм**
M4	4	3,5	37	39	47,5	16,0
M10000	2600	10000	37	42	42,5	16,5

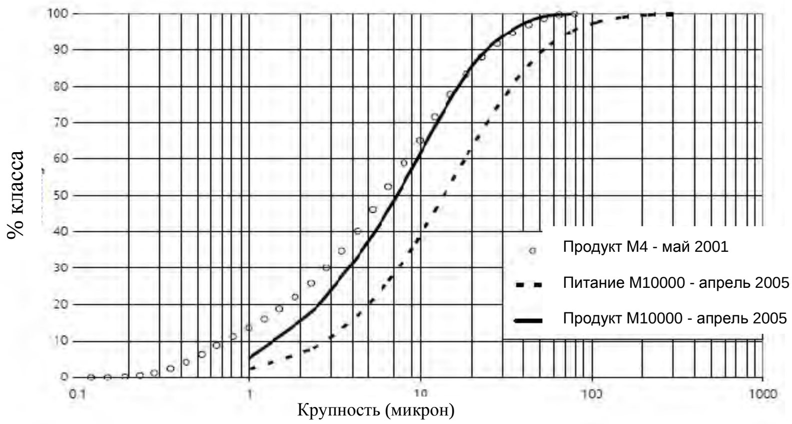


Рисунок 2. Масштабирование M10000, Вестерн Лим (Curry D, Clark L, Rule C; 2005)

Масштабирование SMD

Известны результаты масштабирования мельниц с перемешиванием мелющей среды SMD на руднике Сенчери. (Gao M, Reemeyer L, Obeng D, Holmes R; 2007). На руднике Сенчери они применяются для доизмельчения свинцово-цинковой руды, что очень близко к применению на рудниках Джордж Фишер и MRM. Тем не менее, по результатам масштабирования лабораторной мельницы 0,55 кВт до промышленной установки 355 кВт на Сенчери сообщается о погрешности 20%, т.е. промышленная установка оказалась менее эффективной, чем можно было ожидать на основании результатов испытаний. Это может быть связано с тем, что в лабораторных условиях передаваемая мощность измерялась через реактивный момент в камере измельчения, а не непосредственно (Nesset J, Radziszewski P, Hardie C, Leroux D; 2006); хотя на это была сделана поправка, тем не менее, погрешность 20% при масштабировании лабораторной мельницы 0,55 кВт до промышленной установки 355 кВт сохранилась. Это может быть связано с тем, что мельница не вышла на устойчивый режим работы.

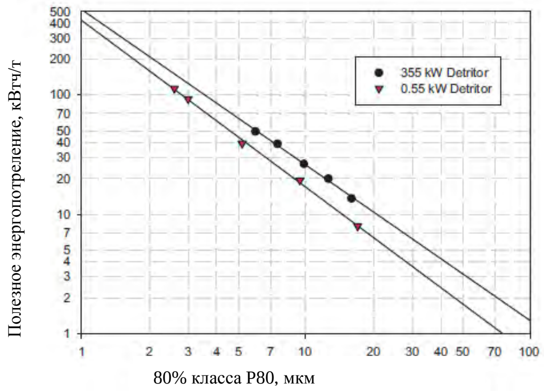


Рисунок 3. Масштабирование мельницы с перемешиванием мелющей среды SMD на руднике Сенчери (Gao et al, 2007)

Методика масштабирования, разработанная для IsaMill, была затем применена к промышленной установке SMD на руднике Сенчери. Методика IsaMill M4 позволила точно предсказать показатели промышленной установки на руднике Сенчери при P_{80}. Это интуитивно верно: хотя мельницы имеют разную конструкцию (горизонтальная и вертикальная), в них используется одинаковая среда при похожих скоростях, поэтому можно ожидать близких характеристик энергопотребления при измельчении. Единственное различие между испытаниями заключалось в меньшей крупности при P_{98} (более крутое распределение по крупности) в IsaMill, что связано с наличием встроенного сепаратора и проблемами проскальзывания в промышленной установке SMD.

Промышленная установка SMD и лабораторная IsaMill

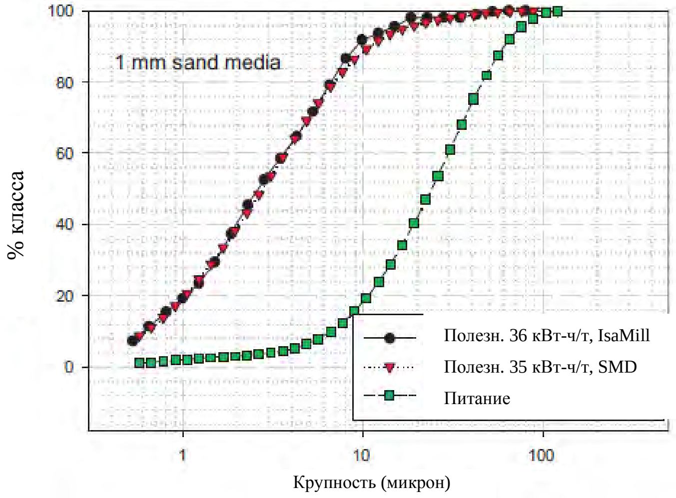


Рисунок 4. Сравнение характеристик промышленной установки SMD и IsaMill 4 л (Gao et al, 2007)

Погрешности вследствие недостаточного измельчения крупных частиц

Тестирование в непрерывном режиме, имитирующее измельчение в промышленной установке, должно обеспечивать точное масштабирование. Однако необходимо следить за тем, чтобы крупные частицы не скапливались в мельнице и чтобы они достаточно разламывались в ходе тестирования. Почему следует избегать накопления материала, будет рассмотрено ниже; использование слишком малой пробы для выхода на устойчивый режим приведет к занижению энергопотребления. В то же время, неизмельчение наиболее крупных частиц может повлечь завышение энергопотребления. Если мелющая среда имеет слишком малую крупность для размола наиболее крупных частиц, они скапливаются в мельнице и оттесняют мелющую среду. Это влечет к падению эффективности измельчения и повышению передаваемой мощности (аналогично накоплению "класса критической крупности" в мельнице полусамоизмельчения). Снижение эффективности измельчения не всегда очевидно, если частицы остаются внутри мельницы.

Можно с уверенностью утверждать, что это стало причиной схожих результатов, о которых сообщается в (Farber et al.). В этой работе сравниваются два вида среды, A и B. Более мелкая среда представлена как более эффективная, однако результаты неверны вследствие накопления крупного материала в мельнице. Если учесть крупный материал, фактическое энергопотребление намного выше. В первом графике на рис. 5 показано меньшее трение среды типа B, влекущее меньшую передаваемую мощность в воде. Естественно, что в любой среде на перемешивание менее крупных частиц нужно затратить меньше энергии, чем на перемешивание более крупных. (Легче засунуть руку в ведро, наполненное шариками диаметром 3 мм, чем в ведро, наполненное шариками 12 мм). Этого же можно ожидать при работе на пульпе, однако второй график на рис. 5 показывает обратное: в средах обоих видов расходуется больше энергии на измельчение руды с использованием мелкой среды (платиновая руда UG2). Можно с уверенностью утверждать, что это указывает на ошибочность испытаний: мелкая среда (1,7 мм тип B и 2 мм тип A) была слишком мелкой, чтобы измельчить крупные частицы, вследствие чего эффективность измельчения снизилась и в мельнице стали накапливаться крупные частицы. Поэтому построенные характерные графики (рис. 6) неверны.

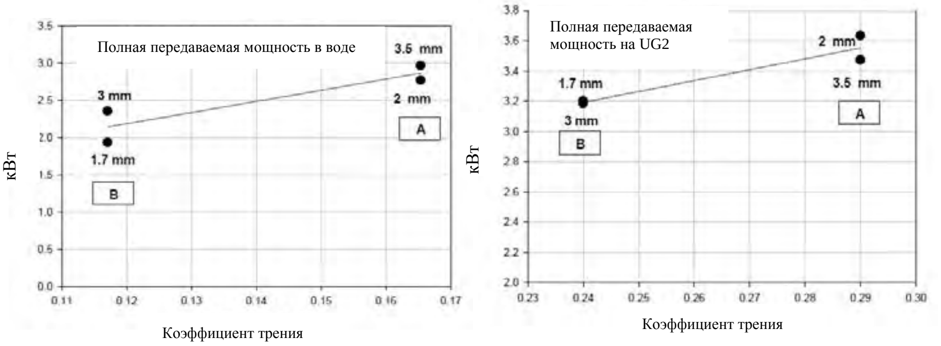


Рисунок 5. Передаваемая мощность в воде (слева) и пульпе UG2 (справа) при первом проходе тестирования с целью построения характерного графика (Farber et al; 2010)

Характерный график M4 с накоплением крупного класса

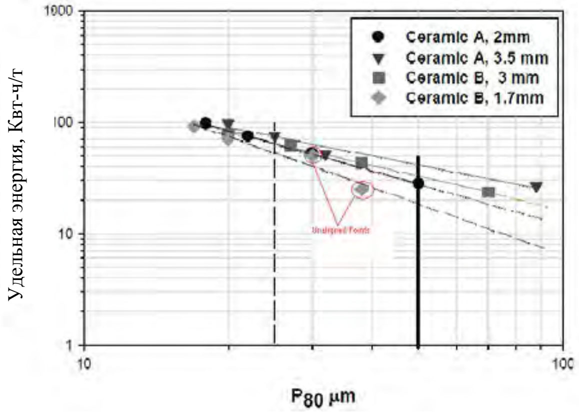


Рисунок 6. Характерный график для UG2 при мелющих средах разной крупности и удельной плотности (Farber et al, 2010)

И действительно, на рис. 6 виден признак, указывающий на ошибку: первая точка на линии "1,7 мм керамика В" не совпадает с линией, проведенной через три другие точки, - крупность меньше ожидаемой. Это служит прямым указанием на накопление крупных частиц в мельнице при первом проходе, на недостижение устойчивого режима, а следовательно, и на ошибочность результатов испытания. Необходимо отметить, что отклонение этой точки не представляется значительным на двойной логарифмической шкале мелкого масштаба, использованной на рис. 6. Однако это отклонение становится намного более заметным, если построить тот же график в более крупном масштабе (10-100 кВт/т вместо 1-1000 кВт/т). Поэтому нужно быть очень внимательным при использовании двойной логарифмической шкалы при тонком помоле.

Аналогичные исследования, но с обеспечением измельчения крупных частиц и выхода на устойчивый режим работы, были проведены Ларсоном. Были сопоставлены разные крупности и виды сред в воде и при измельчении медной руды. Это принесло ожидаемые результаты: мелкие среды более эффективны, потребляют меньше энергии как в воде (рис. 7), так и в медной руде (рис. 8), если они способны измельчить наиболее крупные частицы.

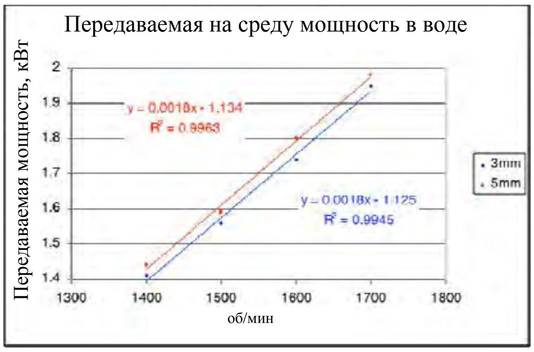


Рисунок 7. Крупность среды Hira и полная передаваемая мощность в воде (Larson M; 2010)

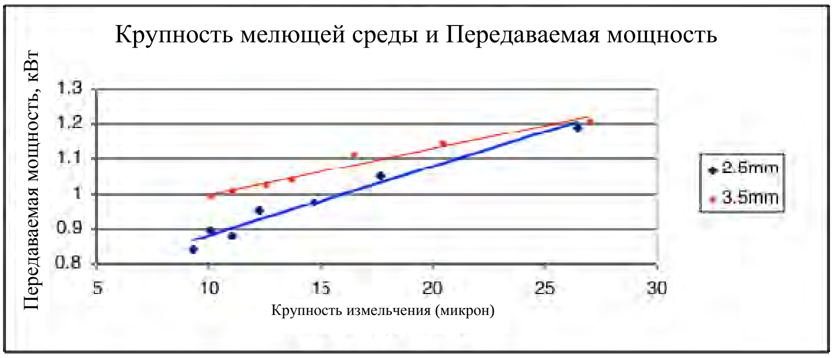


Рисунок 8. Крупность среды и передаваемая мощность на медном концентрате (Larson M; 2010)

На рис. 9 показана разница между полезной передаваемой мощностью при аналогичной среде, но при разной крупности помола. Данные получены в ходе отдельных испытаний для построения характерных графиков для одной и той же руды в одинаковых условиях. Обе среды были способны измельчить верхний класс крупности. При использовании среды 5 мм отношение P_{98}/P_{80} немного меньше: 2,6 против 2,7 при среде 3 мм.

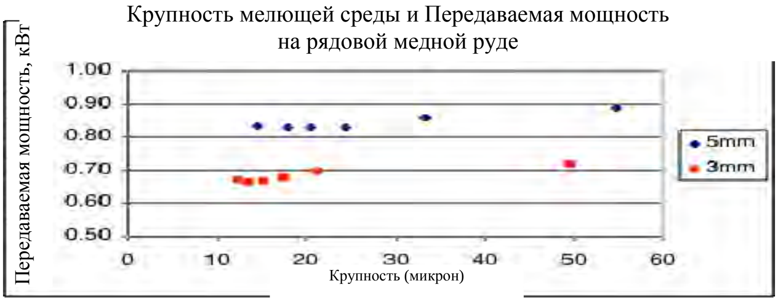


Рисунок 9. Крупность среды Hira и Передаваемая мощность на медной рядовой руде (He M; 2010)

Как показано в табл. 1 и на рис. 8 и 9, менее крупная среда MT1 уменьшает передаваемую мощность при одинаковых условиях пульпы. Более того, по мере измельчения пульпы передаваемая мощность уменьшается. На рис. 8 первые точки, соответствующие 1,2 кВт, имеют почти одинаковую передаваемую мощность. В этих точках среда 2,5 мм позволяет получить P_{98} = 95 мкм, а среда 3,5 мм - P_{98} = 63 мкм. Точки с меньшей крупностью имеют меньшие отношения, поскольку среда 2,5 мм более способна измельчить верхний класс в данных проходах.

Полезная мощность мельницы при разных крупностях среды MT1			
Крупность среды	1,5 мм	2,5 мм	3,5 мм
Полезная мощность	0,80 кВт	0,98 кВт	1,07 кВт

Таблица 1. Средняя полезная передаваемая мощность мельницы при измельчении медного концентрата при разных крупностях среды (Larson M; 2010)

Масштабирование M10000 на руднике Моренчи

Конструкция мельницы M10000 на руднике Моренчи описана ниже и в таблице 3. Хотя испытание с использованием M4 IsaMill точно предсказало энергопотребление для достижения требуемого P_{80}, были выявлены и другие важные факторы, необходимые для точного прогнозирования P_{98} и рассмотренные ниже (Cole J, Wilmot J; 2009).

В проекте было заложено P_{80} = 7 микрон и P_{98} = 15 микрон. Фактическая крупность исходного материала составляла P_{80} = 11 микрон и P_{98} = 34 микрон при переработке концентратов Моренчи. Концентраты Моренчи имели изменчивое содержание пирита, который хуже поддается измельчению, чем относительно чистый халькопирит. Учитывая минералогический состав концентратов Сьеррита, содержащих примерно 85% халькопирита, было решено перейти на переработку концентратов Сьеррита. Эти концентраты имели более постоянный состав и были менее твердыми по сравнению с концентратами Моренчи. Концентраты Сьеррита также близки к концентратам Багдад, которые были использованы для расчета установки IsaMill. P_{80} и P_{98} у концентратов Сьеррита были тоньше, чем у концентратов Моренчи (7,4 микрон и 25 микрон, соответственно), но все же не соответствовали проектным характеристикам.

Проектное содержание пирита в концентрате составляло 22,5%, фактическое же содержание в концентратах Моренчи имело разброс от 10-15% вплоть до 55%.

Критерии проектирования комплекса выщелачивания

Morenci MT-DEW-SX	
Критерии проектирования	
Производительность, мт/ч	26.4
Сверхтонкий помол	
Производительность, м т/ч	32.0
Р80, мкм	7.0
Р98, мкм	15.0
кВт-ч/т	68.0

Таблица 2. Критерии проектирования М10000 на руднике Моренчи (Cole J, Wilmot J; 2009)

Даже при разном минералогическом составе руд, невозможность достижения проектного отношения P_{98}/P_{80}, основанного на первоначальных испытаниях, указывает на то, что крупные твердые частицы не перемалывались полностью и задерживались в испытательной установке. Хотя на данный момент это единственный случай, в котором P_{80} масштабируется, а P_{98} - нет, это означает, что если цель испытаний заключается в точном прогнозировании P_{98}, стандартной пробы 15 кг может быть недостаточно для мельницы М4.

На руднике Кумтор компании Centerra Gold имеется аналогичное применение с использованием мельницы М10000 IsaMill для измельчения золотопиритного сырья крупностью 20 мкм до P_{80} = 10 мкм перед цианидным выщелачиванием. На рис. 10 показано, что это масштабирование было выполнено успешно, включая расчет P_{98} (в действительности промышленная мельница позволяет получать гранулометрический состав, который немного лучше расчетного).

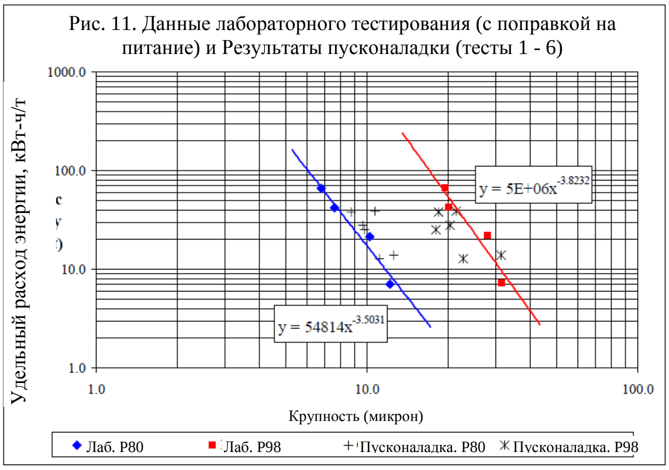


Рисунок 10. Масштабирование М10000 на руднике Кумтор (Kazakoff J, Mortimore A, Smith S, Curry D; 06)

Методики для башенных мельниц

Условия проведения тестирования башенной мельницы Nippon Eirich показаны на рис. 11. Эти методики вызывают ряд вопросов при сопоставлении с описанными выше принципами. Во-первых, рециркуляция разгрузки мельницы непосредственно на насос питания мельницы означает, что устойчивый режим никогда не будет достигнут, - питание мельницы смешивается с мелкой фракцией от первого прохода (напротив, в тестировании на M4 IsaMill проба непрерывно подается из емкости питания, проходит через мельницу и поступает в отдельную емкость продукта, воспроизводя таким образом процесс полномасштабное измельчение).

Grindability test 易磨性实验

Модель: башенная мельница NE008
Емкость: 8 л
Проба: ~ 10 кг / тест
Длительность тестирования: 1 неделя
Цель: предварительная проверка измельчаемости

Модель: башенная мельница KM-5
Емкость: 120 л
Проба: примерно 150 кг / тест
Длительность тестирования: 2 недели
Цель: расчет размера мельницы для технологической гарантии

March, 2009
Page 25

NIPPON EIRICH CO., LTD.

EIRICH

Рисунок 11. Брошюра Nippon Eirich, март 2009 года

Во-вторых, в этих тестах, скорее всего, используется проба слишком малого объема относительно размера лабораторной мельницы. Внутренний объем мельницы должен замещаться не менее 3-4 раз, чтобы установка вышла на устойчивый режим и чтобы предотвратить накопление крупных частиц (что влечет значительное занижение энергопотребления). В таблице 3 показано, что в данном испытании башенной мельницы используется примерно лишь половина от объема материала, необходимого для достижения устойчивого режима.

Питание для тестирования и свободной объем мельницы в зависимости от типа

Тип мельницы	Объем мельницы (л)	Объем твердой фазы (л)	Отношение (рекомендуется не ниже 3)
M4 IsaMill	1.35	5.0	3.70
Башенная мельница NE008	2.35	3.33	1.42
Башенная мельница KM-5	35.2	50.0	1.42
Удельная плотность твердой фазы 3,0, коэффициент заполнения средой 0,66, рабочий объем мельницы (по геометрии) 0,88			

Таблица 3. Сравнение соотношения твердой фазы и свободного объема мельницы при тестировании на измельчение

Последствия слишком малого объема подаваемого материала показаны на рис. 12 (по результатам тестирования пилотной башенной мельницы). Характерная кривая существенно отличается при 50 л материала (0,7 свободного объема мельницы) и 150 л материала (2 свободных объема мельницы). Очевидно, что при небольшом объеме пробы мельница не находится в устойчивом режиме. В данном случае расчет энергии, необходимой для измельчения до P_{80} = 70 мкм, увеличивается на 30% при большем объеме материала. На самом же деле, объем материала должен быть еще больше, т.е. действительная погрешность еще выше.

Слишком малый объем материала в сочетании с разбавлением нового поступающего материала продуктом измельчения увеличивают погрешность еще более и могут повлечь значительное занижение расчетного энергопотребления.

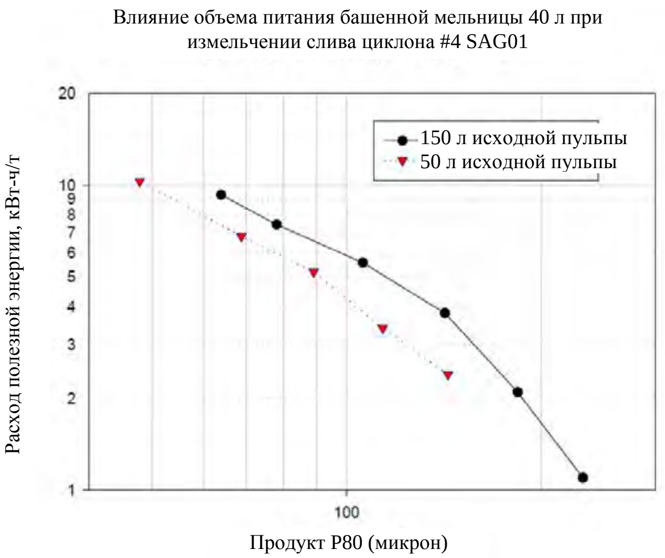


Рисунок 12. Влияние изменения объема поступающей в пилотную башенную мельницу твердой фазы (Gao M)

Были проведены расширенные тестирования, чтобы исключить эту погрешность при стандартных испытаниях на мельнице IsaMill M4 с пробой 15 кг. Ниже в качестве примера приведено сравнение характерных графиков для проб стандартного железного концентрата весом 15 и 35 кг. В обоих испытаниях магнетитовый концентрат (F_{80} = 65 микрон) измельчался в среде 2,5 мм. Среда крупностью 3,5 мм была выбрана исходя из соображений энергоэффективности и лучшего измельчения верхнего класса по сравнению со средами меньшей крупности. Полученная крутая кривая распределения по классам крупности свидетельствует о повышении содержания в конечном концентрате на 2-3% в зависимости от требуемого помола. Кривые P_{80} и P_{98} находятся в пределах погрешности, допустимой для испытания на M4. Железная руда имела удельную плотность твердой фазы 4,2. Этот показатель немного выше обычного. Руда меньшей удельной плотности займет больший объем, поэтому выводы из данного испытания применимы и к таким пробам.

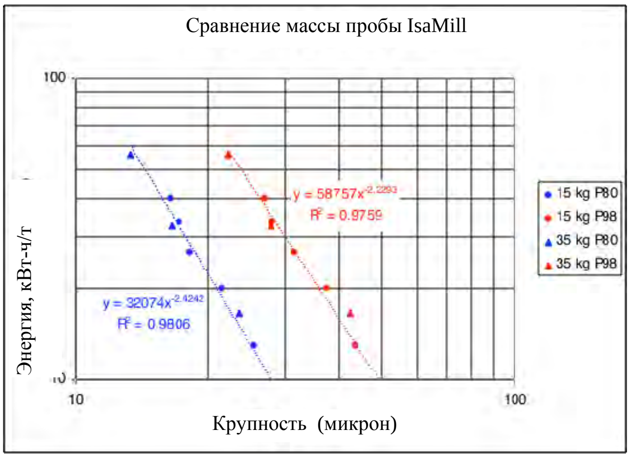


Рисунок 13. Влияние изменения веса материала на характерную кривую в мельнице M4 IsaMill (Larson M, Villadolid V; 2010)

В этих двух тестированиях P_{80} = 16,5 и 16,7 мкм соответствует энергопотребление немного выше 40 кВт-ч/т. Две кривые распределения по классам крупности сравниваются на графике ниже. Их форма почти идентична. Это также означает, что при обеих массах пробы мельница работает в устойчивом режиме и производит продукт без эффекта сегрегации.

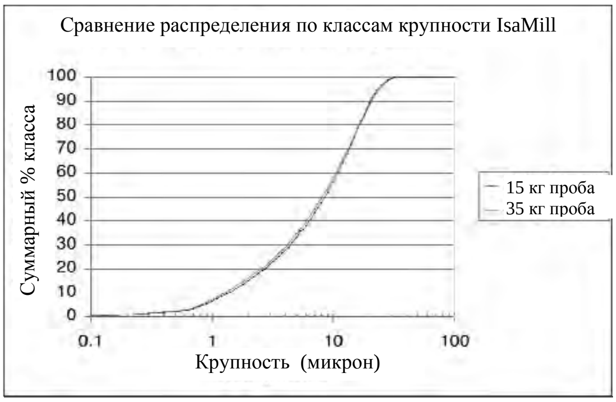


Рисунок 14. Влияние массы питания на гранулометрический состав продукта M4 IsaMill (Larson M, Villadolid V; 2010)

Сравнительный анализ доизмельчения (Nesset et al.)

В работе 2006 года Nesset и другие сравнивают эффективность измельчения в шаровой мельнице, IsaMill, SMD и Vertimill. Авторы приходят к выводу, что методика измерения мощности в SMD по реактивному моменту занижает энергопотребление на 32% и 39% в двух испытанных типах сред. Если мощность в SMD измеряется на двигателе, разница в энергопотреблении между IsaMill и SMD при P_{80} практически отсутствует, хотя при P_{98} разница есть. Это подтверждает выводы, сделанные выше. Однако в работе Nesset также сделан вывод, что лабораторная башенная мельница на 43% более эффективна, чем IsaMill и SMD, при использовании самой мелкой из сред, испытанных в башенной мельнице. Поскольку IsaMill и SMD проектировались как более эффективные в сравнении с башенными мельницами при тонком помоле и это было неоднократно подтверждено промышленной эксплуатацией, данный вывод требует проверки. Он указывает на еще один возможный источник погрешности в лабораторных испытаниях - использование среды с крупностью, отличной от промышленной установки.

Крупность среды имеет важнейшее значение для эффективности измельчения - потребность в использовании более мелких сред для повышения энергоэффективности стала причиной разработки как IsaMill, так и SMD. В испытаниях башенных мельниц в работе (Nesset et al.) использовалась среда 5 мм, которая никогда не используется в промышленной эксплуатации, а более крупные среды, которые также были испытаны в башенной мельнице, показали намного более низкую энергоэффективность. Самая мелкая из сред, промышленно используемых в башенных мельницах, - это 12 мм (в противном случае мелющая среда выносится из мельницы). В пересчете на кубический метр среда 5 мм имеет в 2,4 раза большую площадь поверхности по сравнению со средой 12 мм - это огромная разница для измельчения истиранием в башенной мельнице. Наиболее эффективные крупности среды для разных установок согласно Nesset et al. показаны в таблице 4. Хотя среда 5 мм наиболее эффективна в шаровой и башенной мельницах, ключевой момент состоит в том, что ее невозможно использовать в полномасштабных установках из соображений удержания среды и энергоемкости (размер установки). Именно поэтому были разработаны высокоскоростные мельницы с перемешиванием мелющей среды (IsaMill и SMD) - чтобы на практике использовать преимущества энергоэффективности, связанные с мелкими мелющими средами.

Эффективная крупность среды

Технология	Тип мелющей среды	Крупность среды (мм)
IsaMill	Керамические шары	2,2
SMD	Керамические шары	2,2
VertiMill	Стальная дробь	5
Лабораторная шаровая мельница	Стальная дробь	5
Мельница доизмельчения цинка Brunswick	Мелкая дробь, сталь	16

Таблица 4. Наиболее эффективные крупности среды (Nesset et al; 2006)

Даже неподходящая крупность среды не объясняет полностью необычайную эффективность башенной мельницы, о которой сообщается в статье Nesset. Вероятно, что тестирование башенной мельницы также не смогли выйти на устойчивый режим и в мельнице накапливались крупные частицы. Подтверждающие пилотные данные отсутствуют, но исходя из сведений о размере пробы и количестве испытаний, можно заключить, что объем материала, подаваемого в башенную мельницу, превышал объем мельницы всего в 1,5 раза, а этого недостаточно для выхода на устойчивый режим. Это предположение основано на графике (рис. 15 ниже) гранулометрического состава продукта при трех разных крупностях среды. График показывает, что среда 5 мм улучшила показатели P_{80} и P_{98} относительно более крупных сред. Это неожиданный результат: как правило, крупная среда измельчает крупные частицы более эффективно (особенно в мельнице с перемешиванием мелющей среды). Такой результат, противоречащий интуитивным ожиданиям, убедительно свидетельствует о том, что крупные частицы накапливались в башенной мельнице во время испытания, что повлекло значительное занижение энергопотребления (см. выше).

Влияние мелющей среды среды в Vertimill на распределение по классам крупности

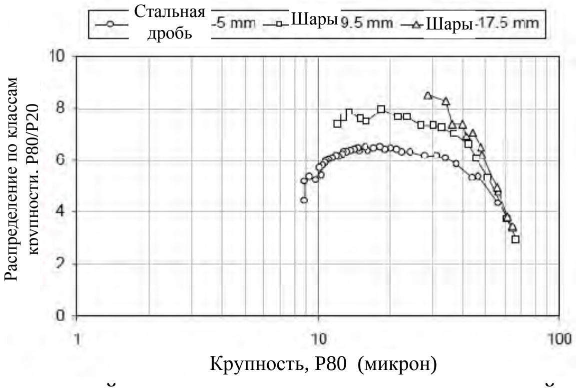


Рисунок 15. Гранулометрический состав продукта и крупность мелющей среды в Vertimill (Nesset et al; 2006)

Сравнительный анализ проектных и эксплуатационных показателей башенных мельниц

В таблице 5 сравниваются некоторые проектные и фактические показатели башенных мельниц. При составлении таблицы были использованы общедоступные источники информации (референц-лист поставщиков), проекты AMIRA P336/P9 и собственные эксплуатационные данные Xstrata.

Проектные и эксплуатационные показатели башенных мельниц

Применение	Энергопотребление для требуемой крупности продукта		Проектные/фактические показатели
	Проектный расчет поставщика	Факт*	
Доизмельчение Cu	6 кВт-ч/т при 45 мкм	13 кВт-ч/т при 45 мкм	48%
II измельчение Pb	4,4 кВт-ч/т при 63 мкм	7,9 кВт-ч/т при 63 мкм	57%
III измельчение Pb	7,1 кВт-ч/т при 45 мкм	12 кВт-ч/т при 45 мкм	63%
Доизмельчение Zn	5,7 кВт-ч/т при 30 мкм	9,4 кВт-ч/т при 30 мкм	61%
Доизмельчение Zn	19,4 кВт-ч/т при 20 мкм	25,5 кВт-ч/т при 20 мкм	76%
Доизмельчение Pb	16,7 кВт-ч/т при 20 мкм	31 кВт-ч/т при 20 мкм	54%
Доизмельчение Ni	11,7 кВт-ч/т при 60 мкм	13,5 кВт-ч/т при 60 мкм	84%
Доизмельчение Fe	9,4 кВт-ч/т при 30 мкм	13,8 кВт-ч/т при 30 мкм	68%

* Фактическое энергопотребление было рассчитано с использованием эксплуатационного рабочего индекса по результатам исследований на фабриках

Таблица 5. Проектные и фактические показатели башенных мельниц

Эксплуатационный индекс был также рассчитан на основании всех имеющихся данных о промышленных мельницах и представлен на рис. 16. Проектные показатели не представлены, но этот график наглядно показывает, на что мельницы в действительности способны. Недавние проектные расчеты доизмельчения, показанные в нижней части графика, значительно занижают фактические эксплуатационные показатели мельниц.

Эксплуатационные показатели башенных мельниц

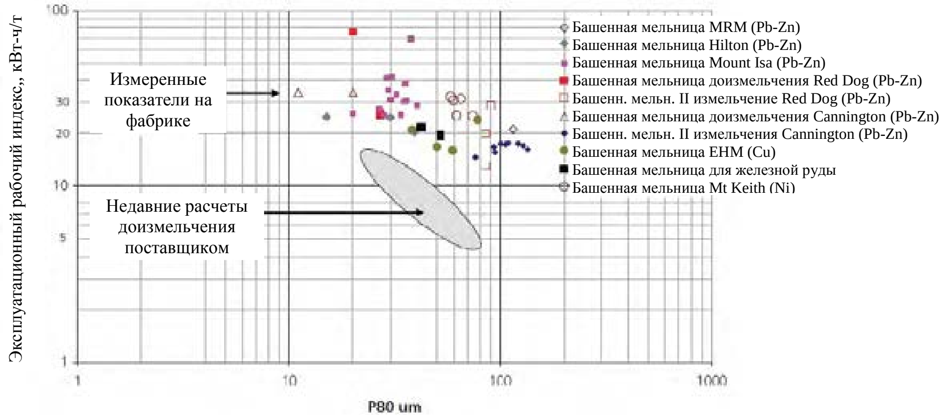


Рисунок 16. Обзор эксплуатационных рабочих индексов башенных мельниц (Pease J; 2010)

Видоизмененные тесты Бонда и Левина

Представляется, что тест Левина в недостаточной мере используется при расчете шаровых мельниц доизмельчения. Некоторые результаты представлены в таблице 6. Требование к весу пробы (20-30 кг) может являться препятствием к использованию. Тест Левина представляет собой видоизмененный тест на измельчение в шаровой мельнице Бонда: в первом используется меньший размер ячейки сит и более мелкий исходный материал. В тесте Левина лабораторная мельница работает в течение разных промежутков времени (разное энергопотребление). После каждого промежутка содержимое мельницы полностью выгружается и просеивается с использованием номинала, соответствующего данному промежутку. Нижний продукт заменяется свежим верхним классом для поддержания постоянного объема. Эта процедура выполняется с 75, 53, 45 и 38 микрон. Поскольку проводится 4 теста, для каждого из которых необходимо примерно 2 л материала, потребность в материале может превышать максимальный возможный объем пилотной партии, тем более что последняя, как правило, представлена концентратом основной флотации. Тем не менее, если проба необходимой массы доступна, представляется, что этот тест дает наиболее точные результаты по сравнению с другими методиками.

Сравнительный анализ эксплуатационных данных и результатов
тестов на измельчение

Материал и происхождение	кВт-ч/т	
	Фабрика	Тест на измельчаемость тонкого материала
Золотая руда, East Driefontein	14.4	14.8
Золотая руда, Libanon	10.8	11.3
Золотая руда, Western Deep Levels, pVCR Reef	14.7	13.8
Золотая руда, Western Deep Levels, Carbon Leader	17.6	16.4
Флуорит, Chemspar	4.3	3.3
Медно-свинцово-цинковая руда, Блэк Маунтин	12.8	11.1
Медно-цинковая руда, Priска	13.6	12.8
Песчаные хвосты, Crown Mines	9.3	9.3

Таблица 6. Результаты теста Левина и показатели промышленных шаровых мельниц (Levin J; 1989)

Условия эксплуатации

Результаты технологических испытаний на продукте определенной технологии не могут быть распространены на другую технологию измельчения. Каждая технология дает собственную кривую распределения по классам крупности при заданном значении P_{80}. Отношение P_{98} к P_{80} необходимо рассматривать в каждом случае отдельно.

Это имеет большое значение для извлечения при выщелачивании. Чем больше верхнего класса, тем ниже извлечение. Материалы, измельченные до требуемого значения P_{80}, могут различаться по содержанию частиц максимальной крупности в три раза. Это также влияет на содержание в конечном концентрате при флотации, поскольку невысвобожденные крупные частицы будут флотироваться, что может повлечь повышение содержания пустой породы в концентрате.

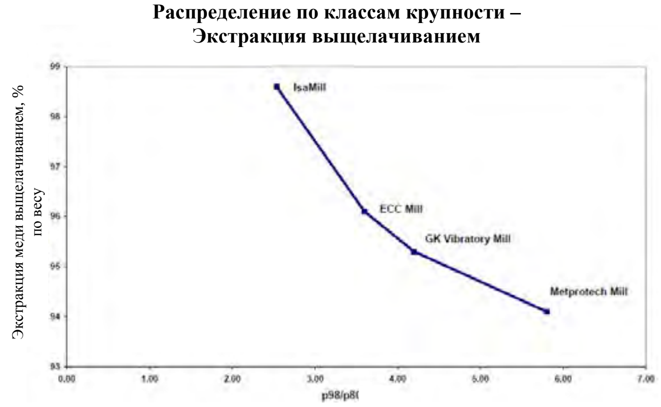


Рисунок 17. Влияние четкого распределения по классам крупности на извлечение при выщелачивании (Pease J, Young M, Curry D; 2007)

Кроме того, необходимо использовать правильную плотность при проведении тестирования для целей масштабирования. Эффект вязкости становится более выраженным по мере приближения к тонким классам, свойственным для доизмельчения. В определенный момент в зависимости от образованной площади поверхности пульпа начнет нести мелющую среду, вместо того чтобы перемешивать ее и измельчат. В настоящее время Xstrata Technology рекомендует использовать при тестировании воронку Марша для контроля плотности и оптимизации влияния вязкости на энергоэффективность. Воронка Марша представляет собой простую воронку, в которой измеряется время, необходимое для прохождения через нее одной кварты пульпы. Воронка Марша не заменяет комплексного исследования реологии, но является удобным средством измерения, идеальным для использования в лабораторных исследованиях или тестировании на площадке.

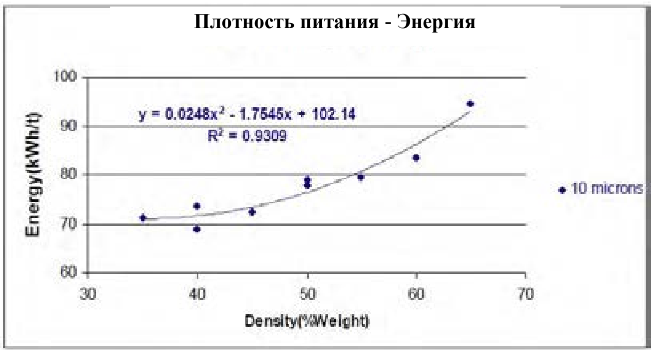


Рисунок 18. Влияние плотности питания на эффективность тонкого помола (Larson M, Morrison R, Young M; 2008)

Перспективы

Xstrata Technology также работает над совершенствованием масштабирования IsaMill посредством проведения спонсируемых и собственных исследований. Одна из последних разработок - это модель IsaMill JKSimMet. Эта модель основана на относительно простой функции, прогнозирующей образование мелких фракций. Стандартная характерная кривая имеет один недостаток, который заключается в том, что, как правило, кривая не проходит через точку, соответствующую подаче материала при нулевой энергии. Это ограничивает возможности использования характерной кривой в части прогнозирования энергопотребления в большем, нежели фактический, диапазоне крупностей. Кривую можно экстраполировать, но чтобы иметь уверенность в результатах всегда лучше охватить необходимый диапазон непосредственно самим тестированием.

Квадратичная функция получения тонкого класса также описывает линейное отношение, но всегда проходит через точку, соответствующую нулевой энергии. Эта модель является продолжением работы Макайвора (McIvor R and Finch J; 2007), показавшей, что новый продукт тоньше определенной требуемой крупности имеет характеристику, близкую к линейной, для шаровых и стержневых мельниц. Для IsaMill оказалось невозможным определить зависимость между процентной долей пропускаемого материала в каждом классе измельчения и расходом энергии. Однако было установлено, что если возвести класс измельчения в квадрат, можно провести прямую линию, соединяющую все точки, включая питание, как показано на рис. 19 применительно к разным медным рудам. Единственное ограничение состоит в том, что крупность должна быть измерима во всех пробах. Это, скорее всего, потребует введения фракции между F_{10} питания и P_{90} продукта.

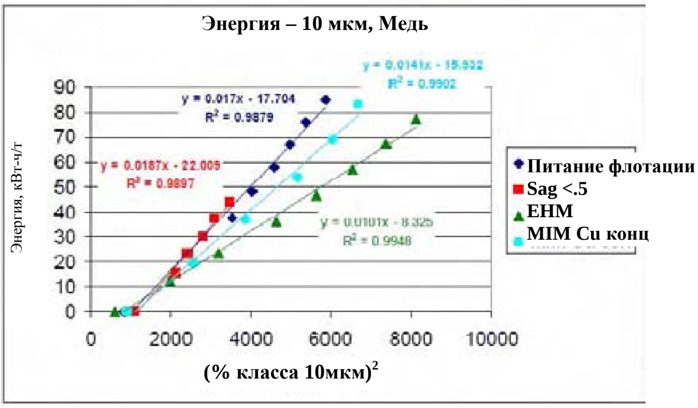


Рисунок 19. Квадратичная функция получения тонких классов применительно к различным медным рудам (Larson et al; 2008)

Этот метод также позволяет более правильно рассчитать типоразмер мельницы, если расчетная крупность питания меняется после тестирования. Сохраняя параллельность линии и изменяя крупность питания, можно определить новое значение энергопотребления. На рис. 19 характеристика показана как квадратичная функция. Новый более крупный исходный материал можно нанести, рассчитав новое квадратичное значение и проведя новую линию графика зависимости энергопотребления и крупности через такую точку слева от первоначальной линии. Сохраняя новую линию параллельной первоначальной, можно определить энергопотребление для требуемого продукта.

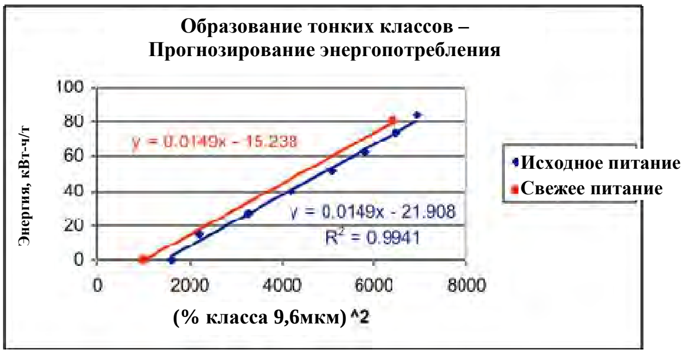


Рисунок 20. Квадратичная функция для прогнозирования энергопотребления при получении тонких классов из более крупного питания (Larson; 2010)

Скорее всего, квадратичная функция точно описывает процесс по той причине, что измельчение истиранием увеличивает площадь поверхности равномерным образом. В IsaMill частицы также становятся более округлыми. Поскольку это верно и для других технологий измельчения истиранием, существует большая вероятность того, что квадратичная функция или иная, похожая на нее функция применима и к этим процессам.

Заключение

В настоящее время для расчетов доизмельчения применяется целый ряд тестов, а прогнозируемые ими показатели энергопотребления имеют огромный разброс. Очевидно, что причиной являются серьезные ошибки, влекущие либо завышение, либо занижение установленной мощности и неоптимальные эксплуатационные показатели установок. Отрасль нуждается к общепринятой стандартной методике достоверного прогнозирования расхода энергии на доизмельчение. Для этого необходимо понять, как фактические эксплуатационные показатели доизмельчения соотносятся с расчетными. Экспертиза технологий тонкого помола JKMRC призвана дать ответ на этот вопрос. Перерабатывающим, проектным и инженерным компаниям следует поддержать тщательное и независимое экспертное исследование с целью изучения эксплуатационных данных и разработки отраслевого стандарта.

Пока такой стандарт отсутствует, существующие методики масштабирования необходимо подвергнуть тщательной проверке, с тем, чтобы фактические показатели мельниц доизмельчения при полномасштабной эксплуатации не отличались от проектных. Инженерам следует детально проверять все расчеты, предоставляемые поставщиками, но не в сравнении с конструкциями других компаний, а на соответствие фактическим эксплуатационным показателям других внедренных установок.

Благодарность

Авторы выражают признательность Xstrata Technology и JKMRC за данные исследований, использованных в этой работе, и за помощь в ее написании.

Список использованной литературы

1. Barns K, Curry D; 2006; Stirring the Pot: A New Direction for IsaMilling; Ultrafine Grinding 06
2. Cole J, Wilmot J; 2009; MORENCI CONCENTRATE LEACH PLANT FIRST YEAR REVIEW; SME 2009 Annual Meeting; Denver USA
3. Curry D, Clark L, Rule C; 2005; Collaborative Technology Development-Design and Operation of the World’s Largest Stirred Mill; Proceedings 2005 Randol Conference; Perth Australia
4. Farber B, Durant B, Bedesi N; 2010; Effect of Media Size and Mechanical Properties on Milling Efficiency and Media Consumption; Comminution 10; Cape Town, South Africa
5. Gao M, Reemeyer L, Obeng D, Holmes R; 2007; Efficiency of the Detritor Mills at Zinifex Century Mine; SAIMM
6. Gao M; CSIRO report
7. He M; 2010; JKMRC private report to Xstrata Technology
8. Kazakoff J, Mortimore A, Smith S, Curry D; 2006; Introduction of IsaMill Technology into a Major Gold Flotation and Leaching Operation in Central Asia; Comminution 06; Perth Australia.
9. Larson M; 2010; Development of a Simulation Model of an IsaMill; MPhil Thesis, University of Queensland JKMRC
10. Larson M, Morrison R, Young M; 2008; Improving Grinding Efficiency With The IsaMill; Comminution 08, Falmouth England
11. Larson M, Villadolid V, Xstrata Technology Internal Report-Iron Ore, May 2010
12. Levin J; 1989; Observations on the Bond standard grindability test, and a proposal for a standard grindability test for fine materials; SAIMM Journal, January 1989
13. McIvor R and Finch J; 2007; The Finch-McIvor Functional Performance Based Grinding Circuit Modeling System; SME Annual Meeting
14. Nesset J, Radziszewski P, Hardie C, Leroux D; 2006; ASSESSING the PERFORMANCE and EFFICIENCY of FINE GRINDING TECHNOLOGIES; CMP; Ottawa Canada
15. Nippon Eirich Brochure; 2009
16. Pease J; Elephant in the Mill; 2010; XXV IMPC Proceedings; Brisbane Australia
17. Pease J, Young M, Curry D; 2007; Fine Grinding as Enabling Technology- The IsaMill

Модернизация обогатительной фабрики Телфер - установка дополнительных мощностей на перечистной флотации и доизмельчении медных и пиритных концентратов

Д.Р. Симэн[1], Ф. Бернс[2], Б. Адамсон[3], Б.А. Симэн[4], П. Мэнтон[5]

Автореферат

Обогатительная фабрика Телфер, расположенная в Большой Песчаной пустыне, шт. Западная Австралия, перерабатывает на двух технологических линиях золото-медную руду, поступающую в настоящее время из одного подземного рудника и двух карьеров, с разным минералогическим составом. Цикл флотации в составе каждой из линий спроектирован с возможностью эксплуатации в нескольких режимах в зависимости от минералогии питания. Большая часть руды, добываемой на фабрике Телфер, перерабатывается в последовательном режиме, при котором содержащие медь минералы флотируются с получением коммерческого медного концентрата, после которой следует флотация золотосодержащего пиритного концентрата, который перерабатывается на гидрометаллургическим способом (с использованием цикла CIL). Золото на фабрике извлекают в форме гравитационного продукта в цикле первичного измельчения в медный концентрат и, в меньшей степени, в цикле CIL.

С момента повторного пуска в 2004 году после установки новой обогатительной фабрики предприятие Телфер испытывает проблемы, связанные с низким содержанием меди в концентрате, отчасти вследствие избыточного вовлечения несульфидных безрудных минералов в цикл флотации меди, а в последнее время вследствие образования композитных частиц меди при переработке руды из соседнего рудника, которая ранее не поступала на новую обогатительную фабрику Телфер. Извлечение золота в цикле CIL также ниже обычного для отрасли.

В данной статье рассматриваются последние изменения, внесенные в технологический цикл, с целью повышения эффективности. Переоборудование цикла включало установку следующих единиц оборудования: две мельницы IsaMill ™ для ультрамтонкого измельчения (одна из них в медном цикле, другая - в пиритном цикле), две флотомашины Джеймсон для повышения эффективности сепарации тонкой пустой породы и батарея из пяти установок Outotec TC30 для извлечения меди и золота из потока доизмельченного пирита. Оборудование было закуплено непосредственно у производителей; проектные и монтажные работы по переоборудованию и модернизации предприятия были поручены инженерной компании.

Введение

Золото- и мededобывающее предприятие Телфер расположено в регионе Пилбара Западной Австралии. Открытая добыча руды (рудник Мейн Доум) была возобновлена в 2003 году; затем в середине 2006 года была начата подземная добыча (рудник Телфер Дипс). Медь в руде находится в разных формах: от преимущественно халькоцита в карьере до

---

[1] член AusIMM, главный металлург, Newcrest Mining Limited, 193 Great Eastern Highway, Belmont WA 6104. Эл. почта: david.seaman@newcrest.com.au

[2] кандидат в члены AusIMM, металлург завода, Newcrest Mining Limited. Эл. почта: fraser.burns@newcrest.com.au

[3] металлург завода, Newcrest Mining Limited. Эл. почта: brooke.adamson@newcrest.com.au

[4] член AusIMM, старший металлург-консультант, Newcrest Mining Limited, 193 Great Eastern Highway, Belmont WA 6104. Эл. почта: brigitte.seaman@newcrest.com.au

[5] руководитель проекта модернизации фабрики Телфер, Newcrest Mining Limited, 400 Georges Street, Brisbane Qld 4000. Эл. почта: peter.manton@newcrest.com.au

преимущественно халькопирита в подземном руднике. В обоих случаях золото присутствует в форме свободного золота и гранул золота, заключенных в сульфиде меди и пирите.

Обогатительная фабрика имеет две параллельные технологические линии (линия 1 и линия 2), в настоящее время работающие с производительностью 24 млн. т в год, из которых примерно 6 млн. т поступает из подземного рудника. На линию 1 поступает смесь руды открытой и подземной добычи, а на линию 2 поступает исключительно руда из карьера. Более подробная геологическая и минералогическая характеристика рудника, исходные критерии проектирования обогатительной фабрики и методика эксплуатации содержатся в ранее опубликованных работах Goulsbra et al. (2003) и Benson et al. (2007).

На обеих технологических линиях для обогащения руды используются различные схемы. Основной схемой является последовательная флотация, при которой медьсодержащие минералы извлекаются в коммерческий медный концентрат с последующей реактивацией и флотацией пирита, выщелачиваемого цианидом с целью извлечения золота. Большая часть свободного золота, не извлеченного на гравитационной стадии при измельчении, извлекается в медный концентрат.

В последнее время Телфер также перерабатывает руды из соседнего месторождения Вест Доум. Если руда из Мейн Доум, основного источника руды на предприятии Телфер, как правило, поддается хорошему высвобождению после измельчения до требуемого номинального класса крупности P_{80} 120 мкм, то руда месторождения Вест Доум имеет существенно другую минералогию. В частности, руда Вест Доум имеет значительно более высокое содержание серы, а минералы меди, как правило, обрамляют или вкраплены в пирит в форме жил.

В Zheng et al. (2010) сообщается о первоначальных технологических изменениях на предприятии Телфер для решения проблемы перегрузки цикла перечистки меди на линии 1, а также об основных аспектах переоборудования технологического цикла, проведенной на предприятии. В настоящей статье описаны три основных изменения на обогатительной фабрике за последние 12 месяцев, которые были внесены, чтобы повысить технологические показатели фабрики. Изменения представлены в таблице 1.

Таблица 1. Переоборудование технологического цикла на предприятии Телфер с учетом различных факторов, влияющих на эффективность производства

Фактор/возможность	**Переоборудование фабрики**
Высокое содержание высвобожденной несульфидной (не представляющей ценности) пустой породы в медном концентрате, препятствующее повышенному извлечению золотосодержащего пирита и золотосодержащих композитных частиц	Снятие головки - две флотомашины Джеймсон (E3432/8) в составе цикла перечистки, установленные перед существующим двухстадийным циклом перечистки
Низкое содержание медного концентрата при переработке руды Вест Доум - из-за худшего высвобождения медных сульфидов по сравнению с рудой из основного месторождения Мейн Доум	Доизмельчение медной руды - мельница доизмельчения медной руды и цикл предварительной классификации при доизмельчении медного концентрата основной флотации (IsaMill M3000 - 1,1 мВт)

Фактор/возможность	Переоборудование фабрики
Пониженное извлечение и высокая эксплуатационная себестоимость извлечения золота в цикле выщелачивания пирита (CIL)	Доизмельчение и повторная перечистка пирита - мельница доизмельчения пирита и машины перечистной флотации для высвобождения и извлечения золота и меди из пиритного концентрата перед цианированием (IsaMill M 5000 - 1,5 мВт и 5 машин Outotec емкостью 30 куб.м со статорами с высоким сдвиговым усилием)

На рис. 1 и 2 показана технологическая схема Телфер до и после переоборудования, описываемого в настоящей статье.

На момент написания настоящей статьи первая очередь переоборудования (установка флотомашин Джеймсон для скальпирования перечистной флотации) была завершена и сдана в эксплуатацию, а последующие два очереди близились к завершению, и их пуск ожидался в течение нескольких месяцев после завершения настоящей статьи. На рис. 3 представлена трехмерная модель размещения основного оборудования после технологического переоборудования. Оборудование было сосредоточено на одном участке из соображений удобства обслуживания и сведения к минимуму затрат и вмешательства в работу предприятия в ходе монтажных и пусконаладочных работ. Мельницы доизмельчения установлены на общей платформе, оборудованной портальным краном грузоподъемностью 10 т для нужд техобслуживания.

Основное оборудование было закуплено компанией Newcrest Mining непосредственно у производителей, чтобы сократить сроки монтажа и разбить капитальные затраты на этапы. Разработка технологии и основные проектные расчеты были выполнены специалистами Newcrest Mining (авторами настоящей статьи). Монтаж оборудования был поручен на основании договора EPC (проектирование-закупки-строительство) с фиксированной ценой сторонней инженерной компании (GR Engineering Services Limited). Xstrata Technology поставила комплекс оборудования, состоящий из мельницы доизмельчения пирита (мельница доизмельчения медной руды была приобретена не новой, но не бывшей в эксплуатации у стороннего поставщика), платформы для установки обеих мельниц, питающих и разгрузочных бункеров, системы подачи среды, сопутствующего оборудования и металлоконструкций. Для доизмельчения была выбрана технология IsaMill ™ благодаря ее проверенной энергоэффективности, а также инертной среде, препятствующей пассивации поверхностей сульфидов (Pease et al., 2006). В конфигурации мельниц IsaMill также были использованы некоторые из усовершенствований, предложенных в (Rule, de Waal, 2011). Компания Outotec поставила для цикла доизмельчения пирита флотомашины (5 машин OT30), оборудованные механизмом флотации Outotec (Coleman, Rinne, 2011), а также статорами с высоким сдвиговым усилием (Bilney, MacKinnon, Kok, 2006), чтобы создать оптимальные условия флотации тонкой фракции.

Срок выполнения строительно-монтажных работ на объекте составит примерно 12 месяцев к моменту завершения всех трех очередей с учетом того, что заказы на поставку оборудования были размещены примерно за шесть месяцев до начала работ. Бюджет проекта переоборудования технологического цикла был одобрен в июле 2011 года, а завершение пусконаладки всего оборудования ожидается к середине августа 2012 года.

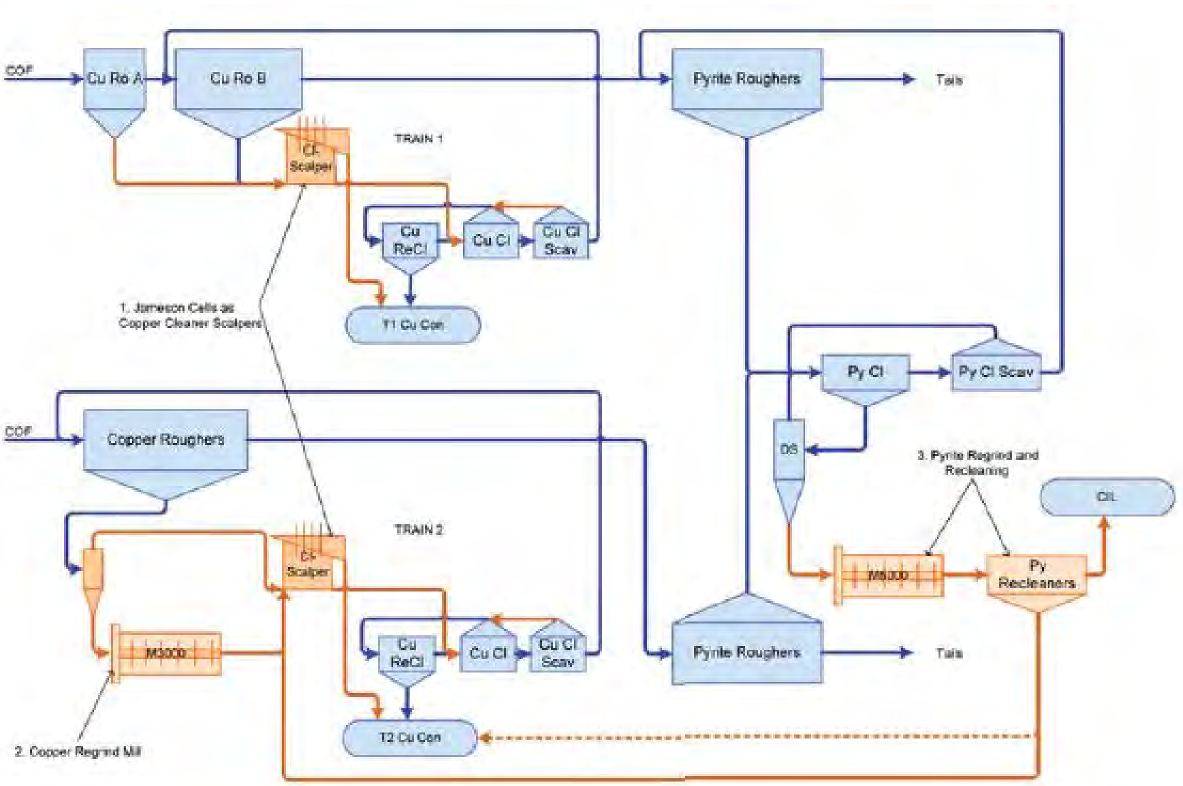


Рис. 1. Технологическая схема предприятия Телфер до переоборудования, описанного в настоящей статье (Примечание: цикл измельчения с гравитационным извлечением и скоростной флотацией не показан)

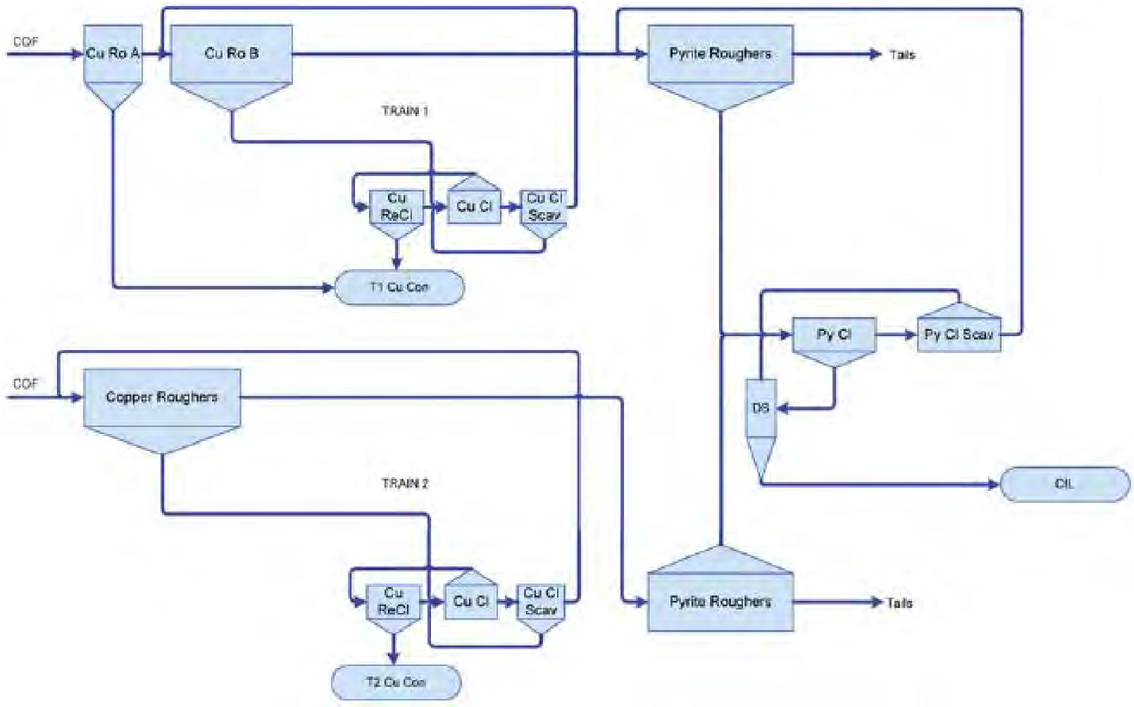


Рис. 2. Технологическая схема предприятия Телфер после переоборудования, описанного в настоящей статье; изменения выделены оранжевым цветом (Примечание: цикл измельчения с гравитационным извлечением и скоростной флотацией не показан)

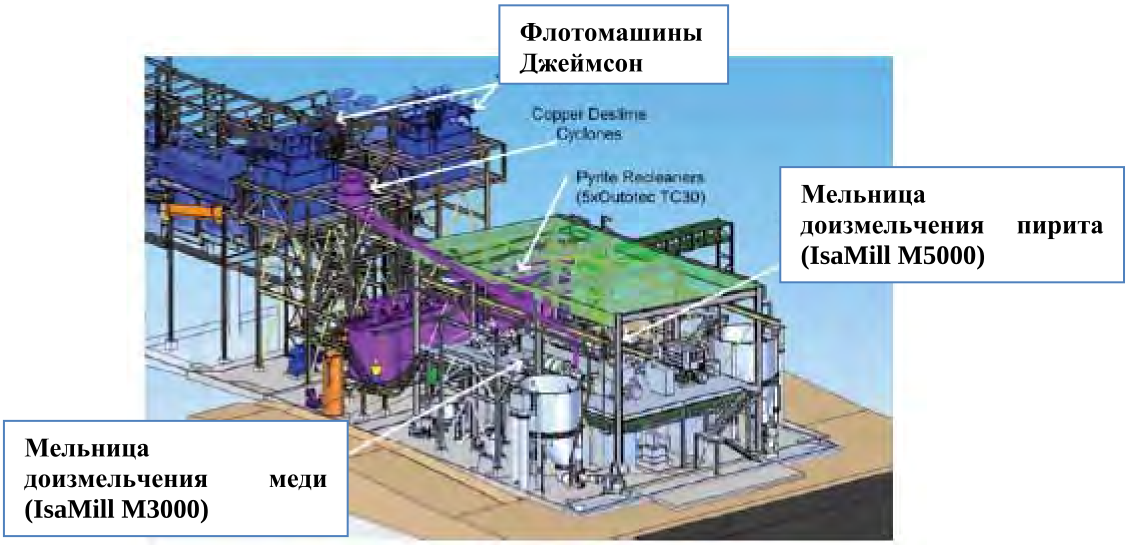


Рис. 3. Модель размещения оборудования на предприятии Телфер после переоборудования

Снятие головки перечистной флотации

На основании анализа исторических данных по медному концентрату в Seaman, Manton, Griffin (2011) сделан вывод, что большая доля высвобожденной несульфидной нефлотируемой пустой породы извлекалась в медный концентрат посредством механического выноса.

В Zweng, Crawford, Manton (2009) содержится подробная информации о переоборудовании линии 1, выполненном в 2009 году с целью сепарации пустой породы и устранения узкого места в цикле перечистки. Хотя результаты этого переоборудования были положительны, потребовалось и было выполнено дальнейшее переоборудование, описанное в Seaman, Manton, Griffin (2011).

Новые флотомашины Джеймсон, установленные для снятия головки перечистной флотации, были пущены в эксплуатацию в ноябре 2011 года. Каждая из флотомашин была установлена так, чтобы разгрузка хвостов и концентрата осуществлялась самотеком. Для этого флотомашины были установлены на металлическую платформу высотой 15 м над уровнем поверхности, а насосы рециркуляции были установлены на уровне поверхности для удобства обслуживания. Дополнительные затраты на подъем платформы компенсируются сокращением затрат на эксплуатацию и обслуживание за счет отсутствия необходимости в установке насосов для хвостов и концентрата и бункеров. Каждая из флотомашин оборудована восемью аэраторами и работает от насоса Warman 10/8 мощностью 75 кВт с производительностью с учетом рециркуляции примерно 700 куб.м/ч. Производительность по свежему питанию каждой флотомашины Джеймсон составляет около 175-350 куб.м/ч (или примерно 20-60 т/ч по твердому). Система промывочной воды рассчитана на расход до 100 куб.м/ч на флотомашину.

Флотомашина Джеймсон, описанная в Evans, Atkinson, Jameson (1995), является высокоэффективной флотомашиной, более компактна по сравнению с традиционными механическими флотомашинами и позволяет эффективно использовать промывку пены для лучшей сепарации пустой породы. Схема флотомашины Джеймсон последней модели представлена в Young et al. (2006).

На рис. 4 показана фотография установки. Работы по данной очереди проекта переоборудования предприятия Телфер были начаты в октябре 2010 года, в мае 2011 года было заказано основное оборудование, а в ноябре 2011 года флотомашины были пущены в эксплуатацию.

Рис. 4. Фотография флотомашин Джеймсон (скальпирование перед перечисткой) на обогатительной фабрике Телфер

Эффективность цикла скальпирования перечистки

Преимущество установки флотомашин Джеймсон заключается в сепарации несульфидной пустой породы, что позволяет извлекать медленно флотируемые ценные минералы (как композитную, так и высвобожденную тонкую фракцию), а также замещать высвобожденную несульфидную пустую породу золотосодержащим пиритом. В ходе разработки проекта была проведена серия флотационных тестов, и полученные кривые селективности были использованы в модели флотации, прогнозирующей конечную эффективность установки (Seaman, Manton, Griffin, 2011). Предполагалось, что во флотомашинах Джеймсон среднее извлечение меди составит не менее 50%.

На рис. 5 показана селективность, полученная при выполнении серии флотационных тестов, относительно фактических эксплуатационных показателей установки, полученных при опробовании и собранных за четыре месяца после пуска флотомашин Джеймсон.

На рис. 5 видно, что фактические показатели флотомашин Джеймсон достаточно хорошо соответствуют модели в технологической линии 1, в то время как линия 2 имеет более низкую эффективность. Кроме того, в большинстве случаев показатель постадиального извлечения во флотомашинах значительно превышает прогнозный показатель 50%.

Из эксплуатационных данных также следует, что при более высоком извлечении металла (более 80%) флотомашины Джеймсон теряют селективность. Поэтому предприятие Телфер контролирует и поддерживает постадиальное извлечение металла на уровне ниже 80%.

До установки флотомашин Джеймсон машины второй перечистной флотации меди были перегружены, вплоть до перелива пены в некоторых случаях (см. рис. 6). Эта проблема была устранена установкой флотомашин Джеймсон, поскольку теперь в механические флотомашины поступает намного меньше флотируемого материала.

Флотомашины Джеймсон показали себя простыми в эксплуатации и не требующими значительного внимания со стороны операторов при изменении условий эксплуатации. Из-за отложений в предшествующих насосах и емкостях во флотомашинах нередко забиваются аэраторы.

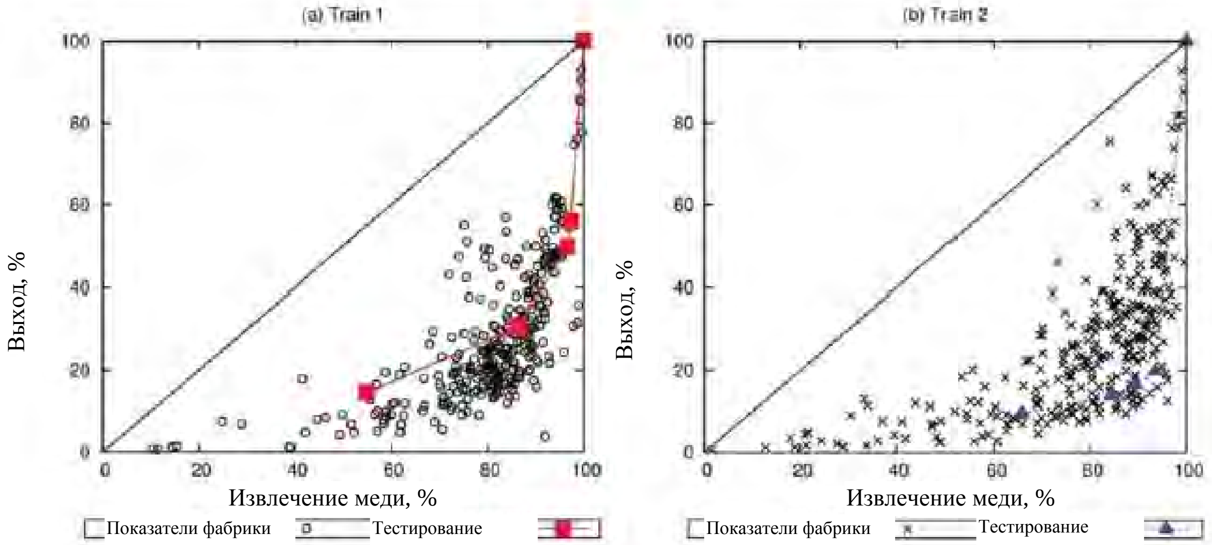


Рис. 5. Выход и селективность меди на (А) линии 1 и (В) линии 2. Сопоставление данных точечных проб (ноябрь 2011 - март 2012) с данными флотационных испытаний, проведенных при разработке проекта

Рис. 6. Фотография перегруженных машин вторичной перечистной флотации технологической линии 2 до установки флотомашин Джеймсон

Количественный анализ по классам крупности

В отношении цикла перечистки технологической линии 2 было предпринято изучение количественного анализа по классам крупности и составлен масс-баланс, чтобы установить эффективность цикла в зависимости от класса крупности.

На рис. 7 показано извлечение различных элементов в зависимости от крупности во флотомашине Джеймсон, а также содержание пустой породы в зависимости от класса крупности в концентратах флотомашины Джеймсон (скальпирование перечистки) и вторичной перечистки (традиционные флотомашины).

Можно заметить, что показатели извлечения ценных элементов (золота и меди) во флотомашине Джеймсон особенно высоки во всем диапазоне крупностей с небольшим снижением на отрезках максимальной и минимальной крупности. Содержание несульфидной пустой породы (НПП) в зависимости от крупности указывает на то, что флотомашина Джеймсон сепарирует НПП намного лучше механической флотомашины благодаря использованию в первой промывочной воды. Испытания с подачей промывочной воды в механические флотомашины показали, что возможно улучшение показателей сепарации НПП. Постоянная схема подачи промывочной воды в механические флотомашины будет, как ожидается, реализована в течение нескольких месяцев после написания настоящей статьи.

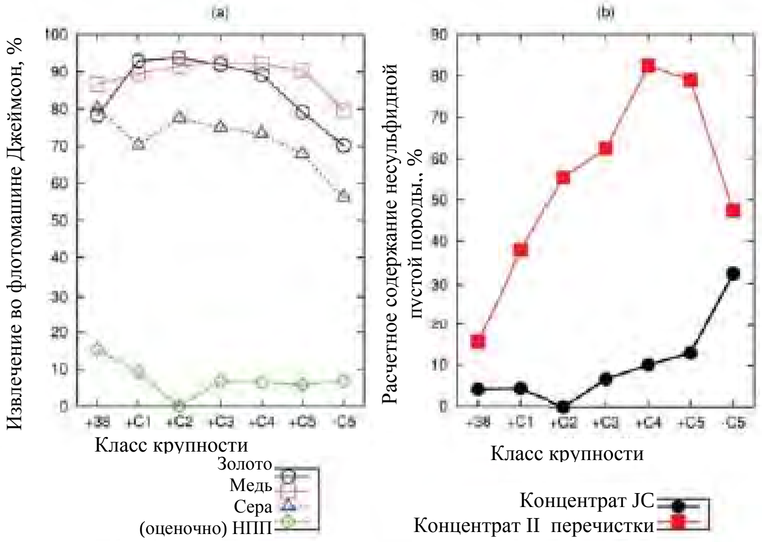


Рис. 7. Зависимость между крупностью и извлечением во флотомашине Джеймсон и расчетное содержание несульфидной пустой породы (НПП) по классам крупности в концентрате флотомашины Джеймсон и после вторичной перечистной флотации в традиционных флотомашинах в зависимости от фракции грохота/классификатора циклонного типа (НПП по данным количественного анализа)

Технологическое исследование блока перечистной флотации

Технологические исследования были выполнены в течение двух месяцев перед пуском флотомашин Джеймсон. На основе собранных данных был выполнен масс-баланс. Полученные показатели эффективности перечистной флотации представлены ниже в таблице 2.

Таблица 2. Результаты исследования цикла перечистной флотации линии 1 (Т1) и линии 2 (Т2)[a] до установки флотомашин Джеймсон

Извлечение в цикле перечистной флотации					
	Выход	**Cu**	**Au**	**Fe**	**S**
	11,3	76,2	86,8	27,2	31,3
	31,9	90,3	89,0	50,7	46,7
	21,7	89,4	88,9	62,3	69,4
	24,1	93,2	94,6	57,6	74,1
Среднее, Т1	22,2	87,3	89,8	49,4	55,4

Извлечение в цикле перечистной флотации					
	Выход	**Cu**	**Au**	**Fe**	**S**
	26,7	50,0	80,1	51,5	61,6
	29,5	88,1	94,2	74,8	85,2
Среднее, Т2	28,1	69,0	87,2	63,1	73,4

[a]. Блок цикла перечистной флотации представлен медным концентратом основной флотации в качестве питания, медным концентратом второй перечистной флотации и хвостами скальпирования перечистки меди в качестве продукта.

Можно заметить, что до установки флотомашин Джеймсон эффективность блока перечистной флотации была непостоянной и в среднем составляла менее 90% на обеих технологических линиях как для меди, так и для золота. Пониженное извлечение железа и серы является следствием намеренной сепарации пирита в цикле перечистной флотации.
В таблице 3 представлены обобщенные данные по извлечению в блоке перечистной флотации после установки флотомашин Джеймсон. Необходимо обратить внимание на то, что на технологической линии 1 поток концентрата основной флотации А теперь проходит через блок перечистной флотации, хотя ранее этот поток направлялся в обход этого блока.

Таблица 3. Результаты исследования цикла перечистной флотации после установки флотомашин Джеймсон

Извлечение в цикле перечистной флотации (после флотомашин Джеймсон)					
	Выход	**Cu**	**Au**	**Fe**	**S**
	89,5	99,5	99,6	98,2	95,4
	43,6	92,4	93,4	45,7	47,3
	25,1	95,2	95,3	55,0	57,3
	11,0	87,4		27,1	37,9
	26,7	95,3		47,2	55,7
	33,2	97,5		56,5	65,6
	20,8	95,3		37,5	48,0
	57,5	99,4		84,0	91,5
	64,0	99,6		84,2	93,0
Среднее, Т1	52,7	95,7	96,1	66,3	66,7
	68,6	94,3	97,2	82,6	90,0
	25,5	97,4	98,0	67,9	83,8

Извлечение в цикле перечистной флотации (после флотомашин Джеймсон)					
	Выход	**Cu**	**Au**	**Fe**	**S**
Среднее, Т2	47,0	95,8	97,6	75,3	86,9

Из сравнения показателей эффективности блока перечистной флотации до и после установки флотомашин Джеймсон следует, что общая эффективность блока значительно повысилась, а извлечение золота и меди в цикле перечистной флотации превысило 95%.
На более эффективную работу блока перечистной флотации также указывает содержание меди и золота в хвостах скальпирования перечистной флотации, которое значительно снизилось в обеих технологических линиях (таблица 4).

Таблица 4. Содержание меди и золота в хвостах скальпирования перечистки до и после установки флотомашин Джеймсон (средний дневной совокупный показатель за два месяца до и после запуска)

	Содержание меди в хвостах скальпирования перечистки, %		**Содержание золота в хвостах скальпирования перечистки, г/т**	
	Линия 1	**Линия 2**	**Линия 1**	**Линия 2**
До	0,46	0,46	2,00	2,79
После	0,39	0,31	1,78	1,38
Значимость, %	84	100	75	100

Эксплуатационные данные за два месяца до и после установки флотомашин Джеймсон был подвергнуты анализу с целью определить, насколько повысилось извлечение в цикле после установки флотомашин. После внесения поправки на известные факторы, влиявшие на извлечение в течение этого времени (содержание в питании, минералогия руды, производительность установки, содержание в концентрате и т.д.), было установлено, что период окупаемости скальпирования перечистки составляет от двух до семи месяцев.

Повышение содержания в медном концентрате

В течение последних 12 месяцев, пока идет подготовка к увеличению добычи в карьере Мейн Доум, на обогатительной фабрике Телфер, помимо основной руды из Мейн Доум, также перерабатывается руда Вест Доум. Исторические данные испытаний и экспериментальной эксплуатации показывают, что медный концентрат, полученный из руды Вест Доум, как правило, имеет недостаточное для коммерческой реализации содержание. Это связано с рядом факторов, среди которых следующие:

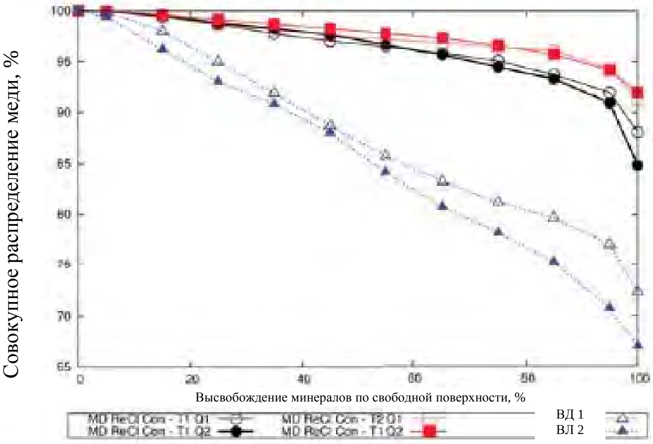
- высокое содержание пирита в руде Вест Доум,
- минералогия руды Вест Доум: медь преимущественно представлена вторичными медными сульфидами (халькоцит, борнит и т.д.) меньшей крупности по сравнению с медно-сульфидными минералами руды Мейн Доум,
- в среднем медно-сульфидные минералы высвобождаются хуже, чем в руде Мейн Доум.

Введение ограничений на количество руды Вест Доум в питании позволило преодолеть некоторые из сдерживающих факторов, однако возникали и будут возникать в будущем ситуации, в которых количество руды Вест Доум в питании превышает допустимый уровень.

На рис. 8 показано высвобождение по свободной поверхности медно-сульфидных минералов из руды Мейн Доум при второй перечистной флотации на технологической линии 1 и 2 в двух кварталах (кв. 1 и 2 2010 финансового года) и высвобождение медно-сульфидных минералов из руды Вест Доум при второй перечистной флотации в лабораторных условиях. График наглядно показывает, что сульфиды меди хуже высвобождаются из руды Вест Доум. Этому соответствуют данные высвобождения в медном концентрате основной флотации в промышленных условиях, представленные в таблице 5. Из проведенных минералогических исследований известно, что медно-сульфидные минералы наиболее часто ассоциированы с пиритом в форме жильной либо поверхностной минерализации. На рис. 9 показаны два оптических изображения типичной ассоциации минералов меди и пирита в руде Вест Доум.

Таблица 5. Характеристика высвобождения концентрата основной флотации Вест Доум, полученная при экспериментальной эксплуатации

Минерал	**Высвобождение, % (>95%)**
Пирит	96,6
Халькопирит	69,8
Другие сульфиды меди	64,9
Другие минералы	97,3



Концентрат второй перечистной флотации МД, Т1, кв. 1
Концентрат второй перечистной флотации МД, Т1, кв. 2

Концентрат второй перечистной флотации МД, Т2, кв. 1
Концентрат второй перечистной флотации МД, Т2, кв. 2

Рис. 8. Сравнение высвобождения по свободной поверхности минералов меди в концентратах второй перечистной флотации из руды Мейн Доум в кв. 1 и 2 2010 года на технологической линии 1 и 2 (Т1 и Т2) и в концентратах второй перечистной флотации из руды Вест Доум в лабораторных условиях (ВД1 и ВД2)

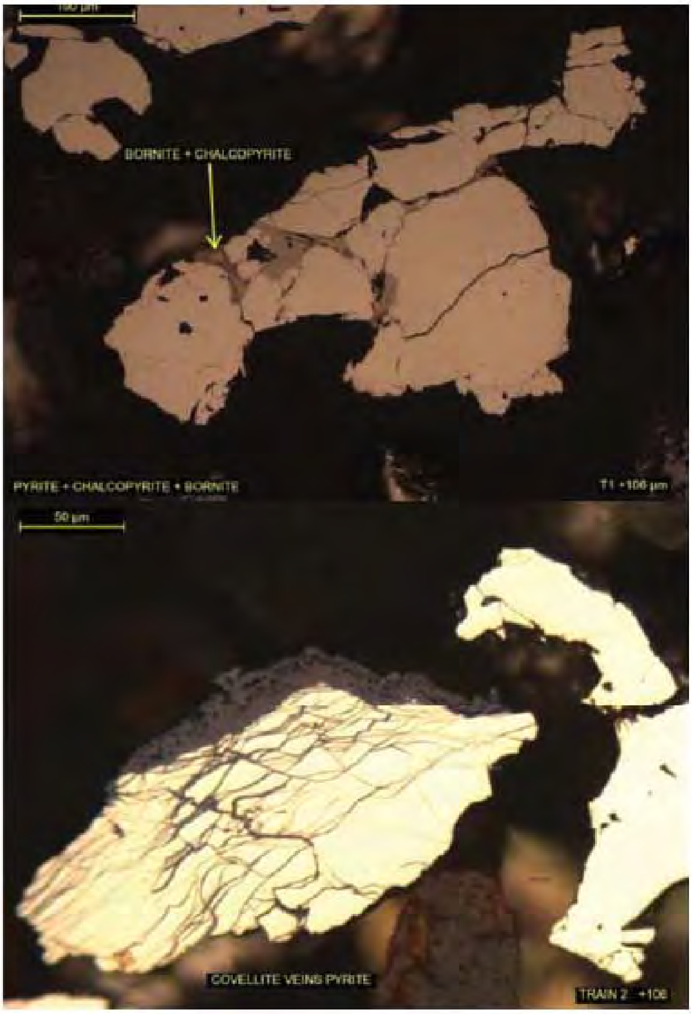


Рис. 9. Примеры включений сульфида меди в частицах пирита

Учитывая описанные проблемы высвобождения, неудивительно, что предпринятые ранее попытки повысить содержание в медном концентрате посредством применения депрессоров, селективных коллекторов или хелатирования (активирующих медь ионов в растворе) были в целом безуспешны.

По результатам перспективных лабораторных тестов с применением доизмельчения между основной флотацией и перечистной флотацией меди, на предприятии Телфер были проведены испытания пилотной установки IsaMill ™ M20 (см. рис. 10) при переработке руды Вест Доум. В пилотную мельницу подавался порциями медный концентрат основной флотации, который проходил измельчение в IsaMill перед флотацией измельченного продукта в лабораторной флотомашине с целью установить, повысит ли доизмельчение медного концентрата основной флотации селективность пирита и минералов меди при флотации. Пилотная мельница также была использована для определения энергопотребления при промышленной эксплуатации посредством построения характерных кривых для медного концентрата основной флотации. Более подробно о пилотной установке и расчетах измельчения, выполненных на другом предприятии, см. в Seaman et al. (2007).

На рис. 11 показана селективность меди и серы при флотации в лабораторных условиях медного концентрата основной флотации технологической линии 2 до и после доизмельчения в пилотной мельнице ISAMill ™.

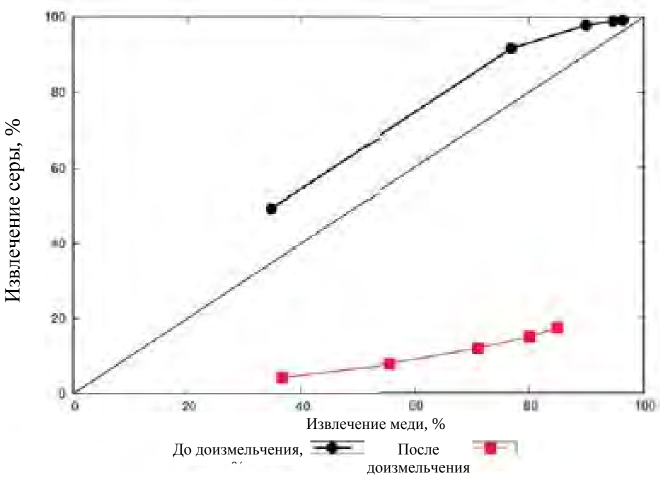


Рис. 11. Селективность меди и серы с доизмельчением и без доизмельчения медного концентрата основной флотации технологической линии 2 при переработке руды Вест Доум

Рис. 10. Для проектирования и масштабирования цикла доизмельчения меди и пирита была использована пилотная установка IsaMill M20

После доизмельчения в пилотной мельнице селективность меди и серы значительно повысилась. Измельченный продукт имел более медленную кинетику флотации по сравнению с неизмельченным продуктом, что, скорее всего, связано с меньшей крупностью частиц на выходе из мельницы. В данном случае класс измельчения P_{80} питания и продукта доизмельчения составил 40 и 20 мкм, соответственно.

В связи с этим в настоящее время устанавливается мельница доизмельчения для измельчения медного концентрата основной флотации на технологической линии 2 при переработке материала с высоким содержанием руды Вест Доум. Установка IsaMill ™ в цикле показана на рис. 2. Медный концентрат основной флотации будет поступать в блок гидроциклонов, нижний продукт будет доизмельчаться в мельнице IsaMill ™ M3000 (мощность 1 мВт), работающей в открытом цикле. Слив гидроциклонов будут поступать под действием силы тяжести в разгрузочный бункер мельницы и вместе с продуктом мельницы будет подаваться насосом в существующий цикл перечистной флотации. Будет установлен бункер разгрузки большой емкости с воздухораспределителями, обеспечивающий достаточное для прохождения флотации окисление пульпы после уменьшения класса крупности в мельнице IsaMill ™.

Повышение извлечения золота из пиритного концентрата

Исторические показатели эффективности цикла CIL на фабрике Телфер ниже нормальных для отрасли, а в настоящее время среднее извлечение золота, содержащегося в пиритном концентрате, составляет 75,2%. Детальный диагностический анализ факторов, влияющих на эффективность цикла CIL, содержится в Burns et al. (2012). По результатам анализа крупностей, диагностических тестов выщелачивания и подробного минералогического исследования в указанной работе установлено, что неудовлетворительное высвобождение зерен золота в крупных частицах пирита было основной причиной недостаточной эффективности выщелачивания пиритного концентрата на фабрике Телфер. Кроме того, было обнаружено, что высокий расход реагентов в цикле вызван проникновением в цикл CIL большого количества растворимой в цианиде меди (преимущественно в виде композитных с пиритом частиц).

В Burns et al. (2012) показано, что оптимальным подходом к повышению эффективности цикла CIL (повышение извлечения и сокращение эксплуатационных расходов) является установка мельницы доизмельчения в существующем потоке питания цикла CIL с последующим удалением высвобожденных частиц золота и меди флотацией (вторая перечистная флотация пирита) перед выщелачиванием хвостов флотации. При проведении лабораторного и пилотного тестирования извлечение золота повысилось примерно до 90% от общего извлечения золота при более высоком извлечении меди и сокращении примерно на 25% расхода цианида в цикле CIL вследствие снижения концентрации растворимой в цианиде меди в питании CIL.

Улучшение показателей связано с высвобождением мелких зерен золота (~ 5-10 мкм), заключенных в более крупных частицах пирита. Большая часть этого золота (~ 50-75%) будет извлечена флотацией и объединена с коммерческим медным концентратом перед выщелачиванием хвостов флотации. На рис. 12 показаны микрофотографии типичных включений золота в пирит перед доизмельчением.

Эффективность такой конфигурации цикла и критерии масштабирования были рассчитаны с использованием пилотных и лабораторных методик. Для доизмельчения питания CIL до различных крупностей при подготовке к лабораторной серии тестов по флотации и диагностическому контролю выщелачивания в бутылях была использована пилотная мельница IsaMill ™ M20 мощностью 1,5 кВт (см. рис. 10). Помимо подготовки питания для лабораторных тестов, пилотная мельница также позволила построить кривые энергопотребления, которые были впоследствии использованы для масштабирования мощности промышленной установки.

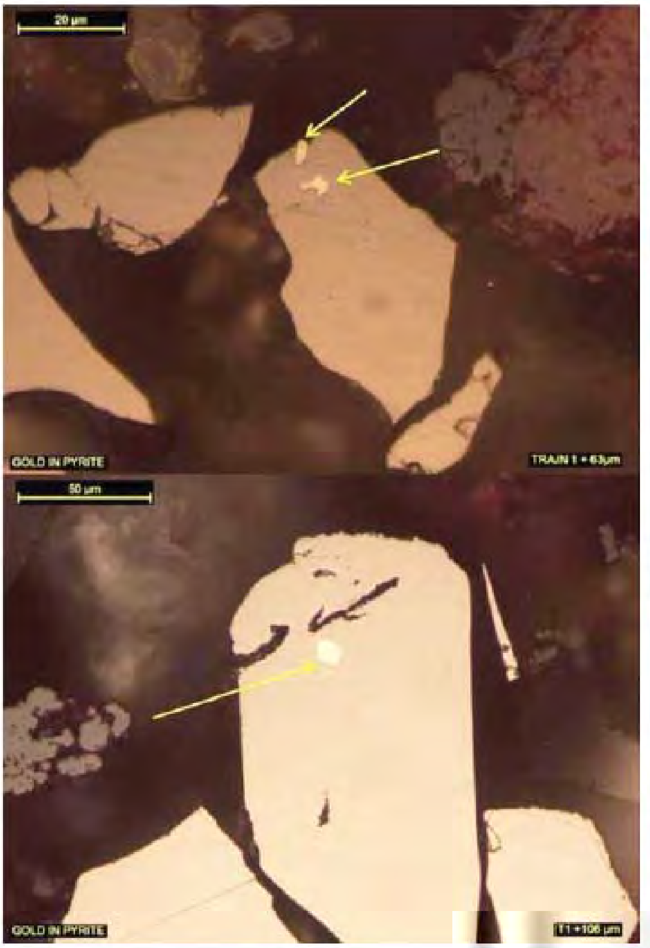


Рис. 12. Примеры включений золота в более крупных частицах пирита.

На рис. 13 показано извлечение золота после доизмельчения до различных крупностей при пилотном тестировании (каждая последовательность данных соответствует одному дню проведения тестов). При требуемой крупности доизмельчения 25 мкм флотационное извлечение в некоторых из тестов было приближено к текущему извлечению золота в цикле CIL (~ 75%). Ожидается, что эксплуатационные расходы на доизмельчение составят примерно 10% в пересчете на унцию от эксплуатационных расходов цикла CIL. Если высокое флотационное извлечение золота будет стабильно достигаться при промышленной эксплуатации, это может стать достаточным основанием для демонтажа цикла CIL в перспективе, когда себестоимость выщелачивания станет необоснованно высокой. Неясно, почему извлечение золота значительно снижалось при измельчении до класса P_{80} менее 20-25 мкм. Авторы предполагают, что это может быть связано с переизмельчением золота / композитных частиц золота с более высоким содержанием сульфидной пустой породы по сравнению с пиритом с более низким содержанием сульфидной пустой породы вследствие классификации внутри мельницы IsaMill ™. Эта проблема будет изучена дополнительно после запуска цикла в эксплуатацию.

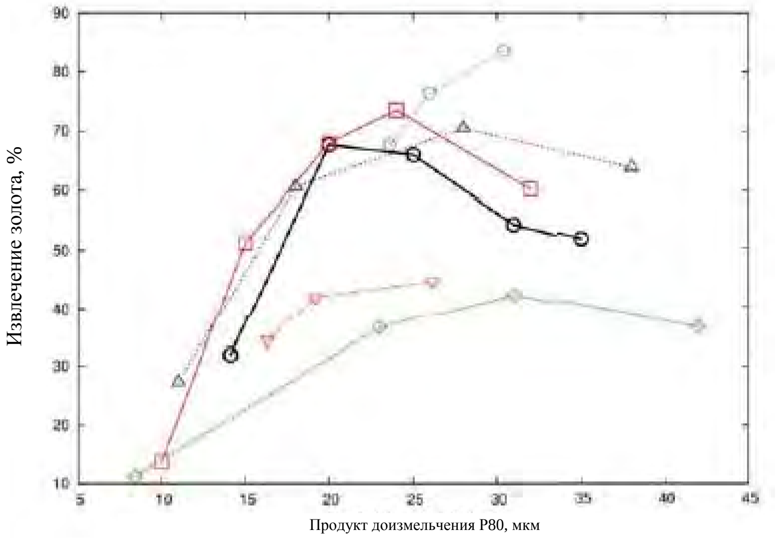


Рис. 13. Флотационное извлечение золота на стадии второй перечистной флотации пирита в зависимости от класса доизмельчения (P_{80}) в разные дни

Мельница IsaMill™ (M5000) мощностью 1,5 мВт устанавливается для доизмельчения пиритного концентрата до требуемого класса крупности P_{80} 25 мкм при максимальной проектной производительности 90 т/ч. В качестве питания в мельницу будут поступать пески действующего двухстадийного цикла обесшламливания с гидроциклонами. Продукт мельницы будет разгружаться в бункер увеличенной емкости (рассчитанный примерно на пять минут пребывания), в котором технологический воздух будет распределяться по пульпе для повышения уровня растворенного кислорода в пульпе для лучшей депрессии пирита и флотации золота. После перекачивания в батарею из пяти флотомашин Outotec емкостью 30 куб.м (вторая перечистная флотация пирита) пульпа будет разбавляться. Флотомашины будут оборудованы статорами с большими сдвиговыми усилиями (Bilney et al., 2006) и механизмами флотации (Coleman, Rinne, 2011) для создания оптимальных условий флотации тонкого класса. После цикла концентрат второй перечистной флотации пирита сможет подаваться в емкости для конечного медного концентрата или в цикл перечистной флотации меди, если будет необходима дополнительная сепарация пирита. Хвосты второй перечистной флотации пирита будут направляться в существующий цикл выщелачивания для дальнейшего извлечения золота, а также смогут быть направлены в обход цикла CIL как отвальные хвосты.

Заключение

Первая очередь (установка скальпирования / снятия головки перечистной флотации меди, флотомашины Джеймсон) текущего проекта переоборудования обогатительной фабрики Телфер выполнена успешно и в соответствии с ожиданиями. Вторая и третья очередь переоборудования находились в процессе выполнения на момент написания настоящей статьи, но должны быть завершены до представления настоящей работы на конференции.

Мельница доизмельчения меди позволит повысить содержание меди в концентрате при переработке руды Вест Доум посредством высвобождения минералов меди из медно-пиритных бинарных частиц (преимущественно халькопирит и халькоцит). Такое переоборудование цикла необходимо с учетом того, что фабрика Телфер приступает к переработке руды Вест Доум.

Цикл доизмельчения пирита позволит повысить извлечение золота из пирита и сократить эксплуатационные расходы существующего цикла CIL.

Проект будет выполнен в течение 12 месяцев с момента начала работ на объекте и в течение 14 месяцев с момента утверждения бюджета всех очередей проекта компанией Newcrest Mining.

Благодарности

Авторы выражают благодарность Newcrest Mining Limited за разрешение опубликовать настоящую работу. В лабораторных исследованиях, пилотных тестах, разработке технологии, а также в непосредственном выполнении работ на обогатительной фабрике Телфер участвовало множество специалистов. В частности, Крейг Чейз-Данлоп сыграл важную роль в координации строительных работ на объекте, а Барри Гринсилл осуществлял квалифицированное руководство проектом. Авторы выражают глубокую признательность всем тем сотрудникам фабрики Телфер, которые помогли успешно выполнить настоящий проект.

Список использованной литературы

Benson, M A, Headley, C R, Hille, S O and Jacob, N P, 2007. Differential flotation commissioning at Telfer, in Proceedings Ninth Mill Operators' Conference, pp 17-24 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Bilney, T, MacKinnon, S and Kok, J. 2006. Assessment of high shear stator performance at Kanowna Belle gold mine, in Proceedings MetPlant 2006, pp 312-320 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Burns, F, Seaman, D, Peng, Y and Bradshaw, D, 2012. Development of pyrite regrind at Telfer gold mine, in Proceedings 11th Mill Operators' Conference, pp 45-52 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Coleman, R, and Rinne, A, 2011. Flotation mechanism design for improved metallurgical and energy performance, in Proceedings of MetPlant 2011, pp 405-418. (The Australasian Institute of Mining and Metallurgy: Melbourne).

Evans, G,M, Atkinson, B,W and Jameson, G J, 1995. The Jameson cell, Flotation Science and Technology (ed: K A Matis), pp 331-363 (Marcel Dekker Inc: New York).

Goulsbra, A, Dunne, R, Lane, G, Dreisinger, D and Hart, S, 2003. Telfer project process plant design, in Proceedings Eighth Mill Operators' Conference, pp 103-113 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Pease, J D, Curry, D C, Barns, K E, Young, M F and Rule, C, 2006. Transforming flowsheet design with inert grinding – The Isa Mill, in Proceedings of the 38th Annual Canadian Mineral Processors (Canadian Institute of Mining, Metallurgy and Petroleum: Montreal).

Rule, C, and de Waal, H, 2011. IsaMillTM Design improvements and operational performance at Anglo Platinum, in Proceedings MetPlant 2011, pp 176-192 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Seaman, D R, Barns, K E, Sharman, P J, Thomas, S A and Stokes, K J, 2007. Process design of a regrind facility at the Leinster Nickel Operations to improve concentrator recovery, in Proceedings Ninth Mill Operators' Conference, pp 131-138 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Seaman, D R, Manton, P and Griffi n, 2011. Separation efficiency improvement of a low-grade copper-gold flotation circuit, in Proceedings Procemin 2011, pp 263-272 (The Mining Engineering Department of Universidad de Chile and Gecamin: Santiago)

Young, M, Barns, K, Anderson, G and Pease, J, 2006. Jameson cell: The 'comeback' in base metals applications using improved design and flowsheet, in Proceedings 38th Annual Meeting of the Canadian Mineral Processors, pp 311—322 (Canadian Institute of Mining, Metallurgy and Petroleum: Montreal).

Zheng, X, Crawford, A and Manton, P, 2009. Telfer Train 1 copper flotation circuit reconfiguration, in Proceedings Tenth Mill Operators' Conference, pp 233-243 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Zheng, X, Manton, P, Burns, F, Crawford, A, and Griffin, P, 2010. Operating strategies to maximise gold recovery at Telfer, Minerals Engineering, 23(15):1159-1166.

Оптимизация схемы переработки магнетитовых руд месторождения в Западной Австралии

D David[1], M Larson[2] and M Li[3]

1. FAusIMM, Process Consultant, AMEC Minproc, Level 14, 140 St Georges Terrace, Perth WA 6000. Email: dean.david@amec.com
2. Senior Process Engineer, Xstrata Technology, 5th Floor, 509 Richards Street, Vancouver BC, V6B 2Z6, Canada. Email: mlarson@xstratatech.com
3. Senior Metallurgy Manager, Grange Resources, Level 11, 200 St Georges Terrace, Perth WA 6000. Email: Michelle.Li@grangeresources.com.au

АННОТАЦИЯ

Разработка магнетитовых месторождений в Западной Австралии привело к проектированию нескольких самых крупных мельниц и фабрик в мире. Один из проектов демонстрирует рост эффективности, который возможен благодаря разработке простой, но всеобъемлющей программе испытаний технологической схемы. На основании опыта существующих предприятий, перерабатывающих магнетитовую руду в Австралии, и месторождений Месаби и Маркетт в США, была разработана основная технологическая схема. Благодаря расширенным испытаниям с мельницами самоизмельчения, шаровыми и смесительными мельницами, была оптимизирована технологическая схема, с целью использовать сильные стороны мельниц каждого типа для получения необходимой конечной крупности помола. Лабораторные работы были подтверждены на демонстрационной установке, с целью оптимизировать КПД энергии каждой стадии измельчения с обеспечением достаточного высвобождения на каждой стации для достаточной отбраковки пустой породы. При использовании трехстадийного помола, шаровая мельница может лучше всего гарантировать высвобождение и выведение всей пустой породы верхнего продукта на второй стадии магнитного обогащения. Включение мельницы IsaMill в качестве третьей стадии помола с ее характерным точным гранулометрическим составом продукта, гарантировало получение максимального качества и упростило процесс переработки после установки, при этом обеспечивая дальнейшие улучшения в части общих капитальных и эксплуатационных затрат на измельчение продукта. В этом случае сочетание двух технологий после мельницы самоизмельчения является более эффективным, чем установка этих мельниц по отдельности при снижении общей установленной мощности на 1/3 и годовых затрат на мелющие тела на 2/3.

ВВЕДЕНИЕ

За последнее десятилетие интерес к железорудным месторождениям Австралии сместился в сторону включения огромных магнетитовых месторождений, разбросанных по Западной Австралии. Заинтересованность в этих месторождениях на данном этапе проявилась в многочисленных фабриках для обогащения магнетитовых руд, которые находятся на стадии проектирования и строительства. Более ранние обогатительные фабрики для магнетитовых руд в Австралии были ограничены небольшими производствами, такими как Savage River и OneSteel’s Project Magnet.

С производительностью питания 3600 т/ч и конечным помолом P_{80} = 34 мкм, количество помола, необходимого для данного проекта, будет большим. Ожидается более низкое конечное качество по сравнению с типичным, т.е. ниже 3% SiO_2. К тому же, в руде присутствует небольшое количество магнитного колчедана, который необходимо обогащать флотацией, чтобы соответствовать требованиям по сере. В этом случае проверили всю технологическую схему, тестируя ранее проверенные технологии существующих заводов с целью оптимизировать все стадии процесса. Лабораторные испытания сочетали с пилотными, чтобы гарантировать максимальную экономическую эффективность при сохранении качества продукта.

ПРОЕКТИРОВАНИЕ ТИПИЧНОЙ ОБОГАТИТЕЛЬНОЙ ФАБРИКИ ДЛЯ МАГНЕТИТОВЫХ РУД

Ниже изображены пять технологических схем обогатительных фабрик для магнетитовых руд из Миннесоты (рис. 1, 4 и 5), Мичигана (рис. 3) и Австралии (рис. 2). Хотя в этих схемах присутствует смесь стержневых, шаровых мельниц и мельниц самоизмельчения (в зависимости от того, какая технология была наиболее распространена или что было в наличии на момент проектирования этих фабрик), необходимо выделить два этапа в этих процессах.

Использование гидросепараторов является общепринятым на фабриках с обратной флотацией кремнезема или без нее. Выведение шламов абсолютно необходимо перед стадией флотации кремнезема, это также помогает выводить тонкодисперсный кремнезем, который вероятнее всего перешел бы в магнитный концентрат.

Независимо от того, называется ли он гидросепаратором, гидроклассификатором, классификатором сифонного типа или классификатором-сгустителем, основной принцип одинаков. “Сгуститель” используется, но с добавлением воды восходящим потоком. Более крупные и тяжелые частицы магнетита осаждается на дне, а восходящий поток воды выносит тонкодисперсный кремнезем в верхний слив. Также для повышения скорости осаждения и минимизации потребления химического флокулянта можно добавлять в питание магнитный флокулянт.

Грохоты для тонкой классификации необходимы, чтобы перенаправить на доизмельчение материал класса крупности +105 мкм, присутствие которого отрицательно сказалось бы на качестве конечного концентрата, а в случае обогатительной фабрики Savage River это вылилось в недопустимый износ трубопровода для концентрата. В случае руд восточной гряды Месаби, например, Миннтак, концентрат, измельченный до класса крупности P_{80} = 325 меш (44мкм), обычно имеет 8% SiO_2, половина этого кремнезема в 10% массы составляет фракция крупностью +140 меш (105 мкм). При использовании грохотов для тонкой классификации эту фракцию можно вывести, и снизить конечное содержание кремнезема в концентрате до 4-5%.

Среди всех представленных производств только в технологической схеме Empire Mine не применяют грохоты тонкой классификации. В результате измельчения на фабрике Empire Mine получали конечный продукт класса P_{80} = 20 мкм, что делает использование грохотов тонкой классификации не практичным и не нужным. Также, как выяснилось, только на фабрике Empire Mine применяют обратную флотацию в качестве конечной стадии обогащения.

Обратная флотация кремнезема используется в крайнем случае при обогащении магнетита, так как часть магнетита будет потеряна с хвостами. Магнитные сепараторы тонкой классификации отбракуют весь высвободившийся кремнезем (исключая кремнеземистые шламы, которые присутствуют в процессе вместе с водой) перед флотацией. Следовательно, практически весь крупный кремнезем флотации будет с прилипшими частицами магнетита. Флотация кремнезема с использованием амина также выявляет проблемы со шламами и является очень восприимчивой к химическому составу воды. Хотя гидросепараторы должны быть установлены несмотря ни на что, чтобы удалять шламы перед стадией магнитной флотации, дополнительное преимущество гидросепаратора в виде сокращения растворимых солей является преимуществом только в том случае, когда выполняется стадия флотации кремнезема. По мнению Ивасаки (1983) флотация железных руд жирными кислотами довольно чувствительна к ионам магния и кальция в растворах пульпы. Для оптимальной работы уровень этих ионов необходимо удерживать ниже 100 ppm.

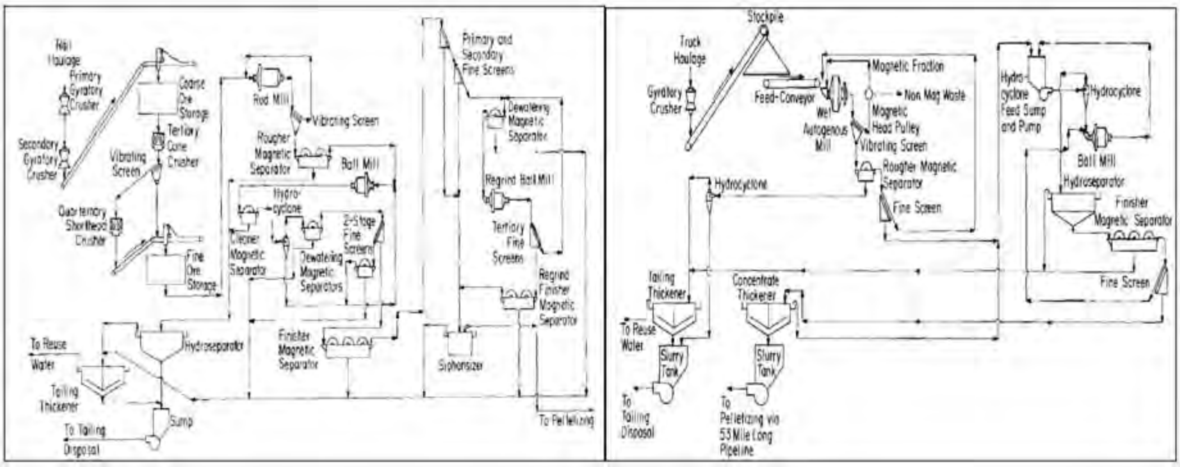


Рис. 1 и 2: Технологические схемы Erie Magnetite и Savage River (Девани, 1985)

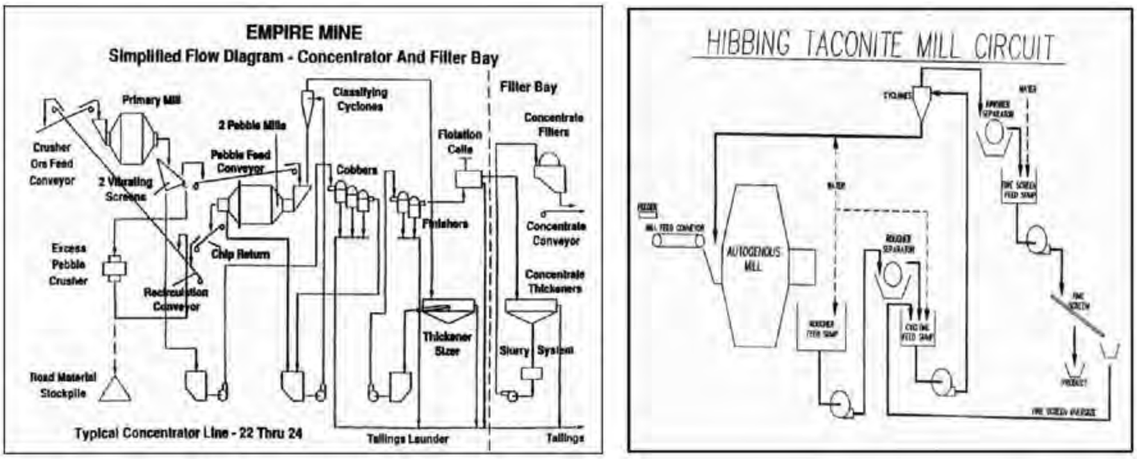


Рис. 3 и 4:Технологические схемы Empire Mine и Hibbing Taconite (МасАйвор и Гринвуд, 1996)

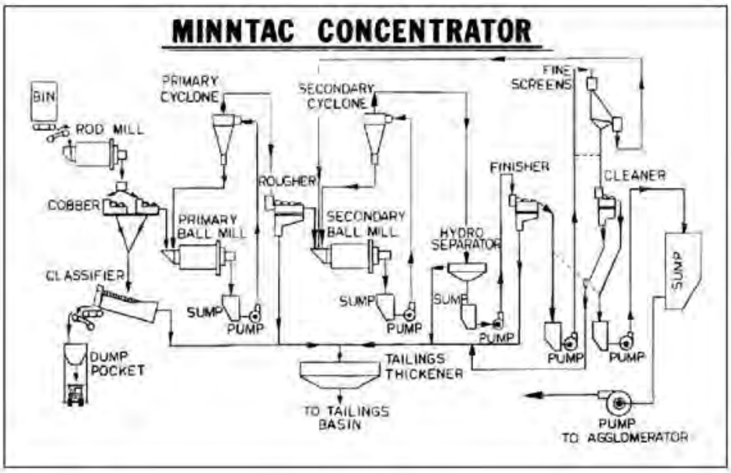


Рис. 5: Технологическая схема Minntac (Девани)

ПРОГРАММА ПРОВЕДЕНИЯ ИСПЫТАТЕЛЬНЫХ РАБОТ

Испытательные работы, проведенные компанией AMMTEC, для данного месторождения магнетитовых руд в Западной Австралии состояли из:

- Пилотного испытания мельницы первичного самоизмельчения,
- Лабораторных испытаний (тест Левина) и пилотного испытания шаровой мельницы вторичного измельчения,
- Лабораторных и ограниченных непрерывных испытаний мельницы IsaMill вторичного измельчения,
- Лабораторного испытания третичного измельчения, ограниченных непрерывных и пилотных испытаний IsaMill,
- Испытаний трубки Дэвиса и пилотных испытаний магнитного обогащения на различных промежуточных и конечных продуктах,
- Испытаний гидросепарации на конечном магнитном концентрате IsaMill,
- Испытаний флотации сульфида на конечном магнитном концентрате, и
- Тестирования классификации конечного концентрата, выполняется поставщиками.

Мельница первичного самоизмельчения

Вся руда перемалывалась на демонстрационной мельнице самоизмельчения AMMTEC (1,74 м – диаметр внутренней футеровки и 0,46м – эффективная длина измельчения) и затем прогонялась через магнитный сепаратор. Для защиты магнитного сепаратора мельницу самоизмельчения закрывали 1 мм экраном. Предполагается, что промышленная мельница самоизмельчения будет закрыта экраном более крупного размера, до 4 мм максимум. На промышленной установке требуется перерабатывать 3600 т/ч питания, в идеале через две линии измельчения, каждая из которых оснащена большой мельницей самоизмельчения с двойной шестерней или с безредукторным приводом. С энергией, поступающей от демонстрационной мельницы самоизмельчения при заполнении 25% , класс крупности питания магнитного сепаратора P_{80} (-1мм) составлял примерно 330 мкм. Магнитный концентрат был значительно крупнее, чем немагнитный с классом крупности P_{80} = 420 мкм и 200 мкм соответственно. На первой стадии магнитного обогащения выводили 40% от общей массы в виде бедных хвостов. Оставшиеся 2200 т/ч подают на вторую стадию измельчения. Дополнительный анализ позволяет предположить, что P_{80} магнитного концентрата увеличиться с 420 мкм до 770 мкм при использовании 3 мм экрана на промышленной установке по сравнению с 1 мм экраном демонстрационной установки.

Результаты испытания шаровой мельницы

Перед проведением испытаний на демонстрационной шаровой мельнице, выполнили один тест Левина на питании вторичного измельчения, результаты представлены на рис. 6. Тест Левина является заменой стандартного теста на определение индекса Бонда с тонкодисперсным питанием, в нем используют последовательные проходы измельчения в открытом цикле с применением известной энергии при каждом проходе. Результаты этого теста с классом крупности P_{80} в микронах, нанесенные на кривую удельного потребления энергии (кВтч/т), показаны на рис. 6.

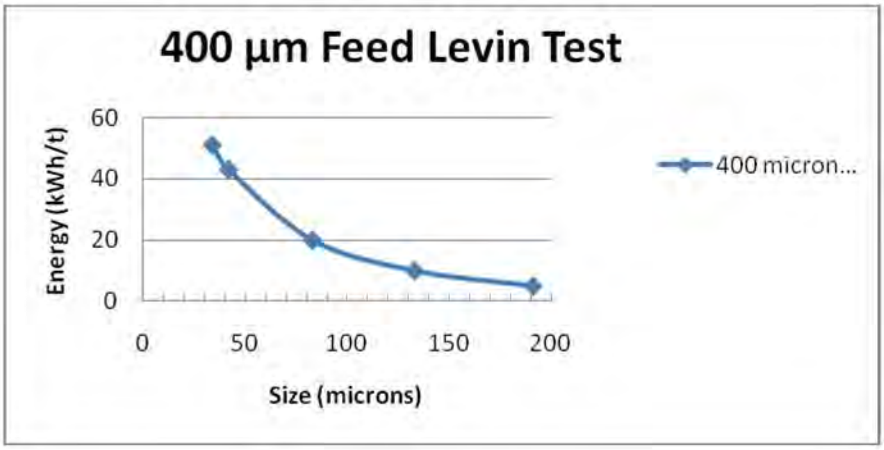


Рис. 6: Тест Левина для шаровой мельницы

Тест Левина для тонкодисперсного продукта шаровой мельницы прогнозирует удельное потребление энергии 52,5 кВтч/т для измельчения питания класса F_{80} ~400 мкм (продукт мельницы самоизмельчения/магнитного сепаратора) до продукта класса P_{80} 34 мкм. Дополнительные 2,3 кВтч/т будут необходимы в случае снижения крупности питания P_{80} 770 мкм до питания, используемого при проведении теста, P_{80} 420 мкм, тем самым требование к удельному потреблению энергии составляет 54,8 кВтч/т. Однако, если в тест Левина не будет включен этап классификации для выведения мелочи, энергия расходуется без надобности на измельчение материала, который уже достиг конечной целевой крупности помола или даже мельче. Следовательно, ожидается, что тест даст преувеличенную оценку удельного потребления энергии, особенно для широких диапазонов крупности. Действительные результаты, полученные при использовании демонстрационной шаровой мельницы AMMTEC 6’ с мелющими телами 25 мм, представлены классом крупности P_{80} 37 мкм, полученным с использованием энергии 40 - 45 кВтч/т. При подаче смешенного питания 2200 т/ч на линии 1 и 2 и вводом 2,3 кВтч/т для понижения крупности с 770 мкм до 420 мкм при однократной проходке через шаровую мельницу для снижения крупности с 770 мкм до 34 мкм потребуется около 114 МВт энергии шаровой мельницы за исключением энергии, связанной с питающими насосами циклона. Это равно 6 самым крупным шаровым мельницам, 3 на линию. К тому же около $86 миллионов будут уходить на закупку стальных мелющих тел ежегодно. Высокий расход мелющих тел обусловлен высоким средним индексом абразивности Бонда 0,44, вызванным присутствием силикатов и гранатов пустой породы. У типичной магнетитовой руды будет Ai (индекс абразивности) 0,25 или меньше. Требование использовать стальные шары 25 мм для достижения требуемой тонины помола усиливает скорость износа среды. Также потребуются капитальные затраты в размере $15 миллионов в случае цикла с шаровой мельницей имеющего целью получение продукта 34 мкм на разгрузочные грохоты, которые необходимы для минимального снижения кремнезема и защиты трубопровода концентрата. Комбинация шаровой мельницы и циклона не может гарантировать выведения частиц крупностью +100 мкм из питания трубопровода.

Вторичное измельчение на IsaMill

То же питание, что было испытано в предыдущем тесте Левина, пропустили через мельницу M4 IsaMill для построения кривой энергоемкости. Для обеспечения размола верхнего продукта использовали 5 мм керамические мелющие тела. Тест на энергоемкость состоит из двух емкостей, одна – для питания мельницы, и другая используемая в качестве емкости для выгрузки. В середине каждого прохода отбирают пробу для определения крупности и плотности. Расход, время и чистая затрата энергии на проход регистрировались, а крупность отмечалась на кривой удельного потребления энергоемкости (кВтч/т). Затем положение клапанов меняется, чтобы включить баки питания и разгрузки, этот процесс повторяется много раз, при каждом прохождении повышая потребляемую энергию. На основании чего была построена кривая энергопотребления. Эта кривая представлена прямой линией гранулометрического состава частиц на логарифмической сетке в сравнении с кривой энергопотребления, которая масштабируется напрямую на промышленные мельницы IsaMill. Результат теста на определение энергоемкости изображен на рис. 7.

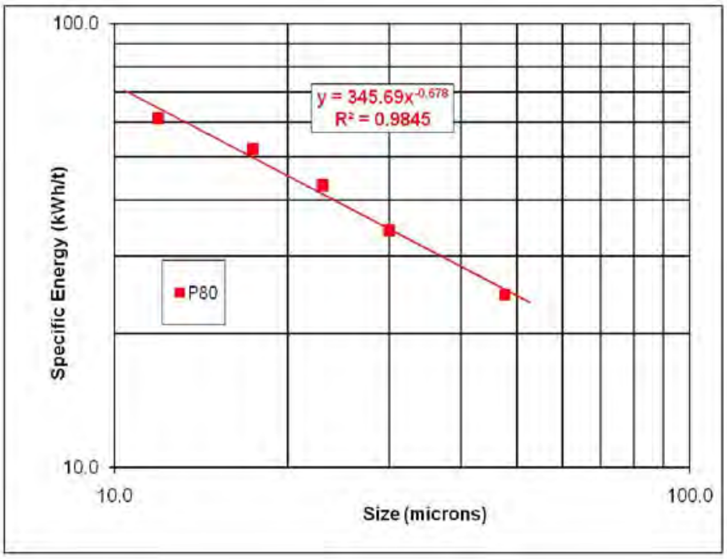


Рис. 7: Кривая энергоемкости для IsaMill, 400 мкм

На мельнице IsaMill можно было сократить питание вторичного измельчения крупностью 400 мкм до 34 мкм с использованием 31,7 кВтч/т. Общая прогнозируемая энергия IsaMill составляет 34 кВтч/т с 2,3 кВтч/т необходимыми для сокращения крупности от 770 мкм до 420 мкм (в предыдущем испытании шаровой мельницы) (обратите внимание, что данный расчет является теоретическим, так как питание магнетитовой руды F_{80} 770 мкм для IsaMill нецелесообразно, и неэффективность IsaMill на материале такой крупности означает, что значение 2,3 кВтч/т является оптимистичным). Это явное улучшение удельного энергопотребления на 30% по сравнению с шаровой мельницей (для данной теоретического более крупного питания). Это привело бы к установленной мощности 78 МВт, если бы использовали мельницы IsaMill в сравнении с 114 МВт которые требуются для шаровой мельницы. К тому же годовые затраты на мелющие тела снизятся с $86М за стальные шары до $57М за керамическую мелющую среду IsaMill. Помимо высокой абразивности руды, гальваническое действие магнитного колчедана, присутствующего в руде, может способствовать высокому уровню износа стальных мелющих тел (Ивасаки, 1999г). Эта опасность отсутствует в случае использования керамических тел.

Даже в случае такого радикального улучшения, как в случае с шаровой мельницей, все равно остаются проблемы с использованием IsaMill а одностадийном процессе для данной цели. Из-за присутствия крупного и прочного гранита и жильного силиката, а также из-за трения, создаваемого при перемешивании этих частиц, износ мелющих тел был выше, чем ожидалось. Несмотря на улучшенные показатели по износу по отношению к стальным мелющим телам шаровой мельницы, износ керамической среды IsaMill был в два раза выше предполагаемого уровня. Энергию, затраченную на измельчение этого гранита до 34 мкм, можно считать потраченной впустую. Изучаемый гранит представляет собой прочный глиноземный силикат и действительно имеет свойства близкие к свойствам типичной мелющей среды IsaMill. Было бы намного эффективнее вывести это утильсырье по возможности сразу перед высвобождением. На рис. 8 крупным планом изображены кристаллы гранита, полностью высвободившиеся в диапазоне крупности 125-150 мкм.

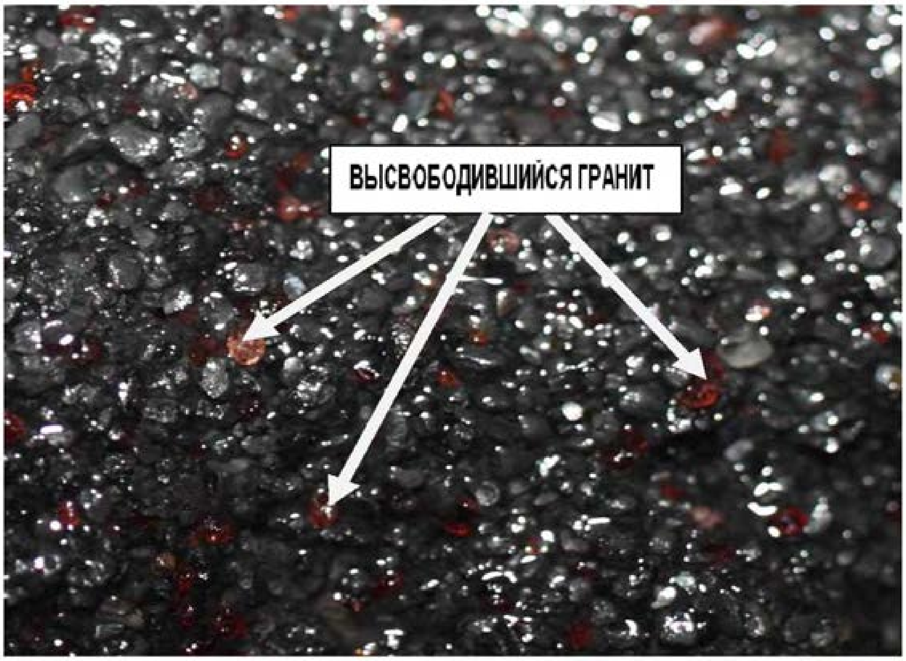


Рис. 8: Высвободившиеся частицы гранита после измельчения 150 мкм

Отображая результаты теста Левина и вторичного измельчения IsaMill вместе (рис. 9) можно увидеть, что рост эффективности при использовании мельницы IsaMill для определенного типа руды начинается при классе крупности продукта P_{80} приблизительно 100 мкм. При крупности продукта выше данной точки, шаровая мельница является более эффективной. Это в основном функция крупности мелющих тел. Мелющая среда крупнее 5мм, используемая в мельнице IsaMill, обеспечила бы более эффективное сокращение крупности до 100 мкм, но не была бы настолько эффективная при целевой конечной крупности 34 мкм. При таком снижении крупности, применяемом к прочной породе, измельчение всего потока в одной стадии никогда не достигнет оптимальной эффективности для всех классов крупности при использовании любой технологии измельчения.

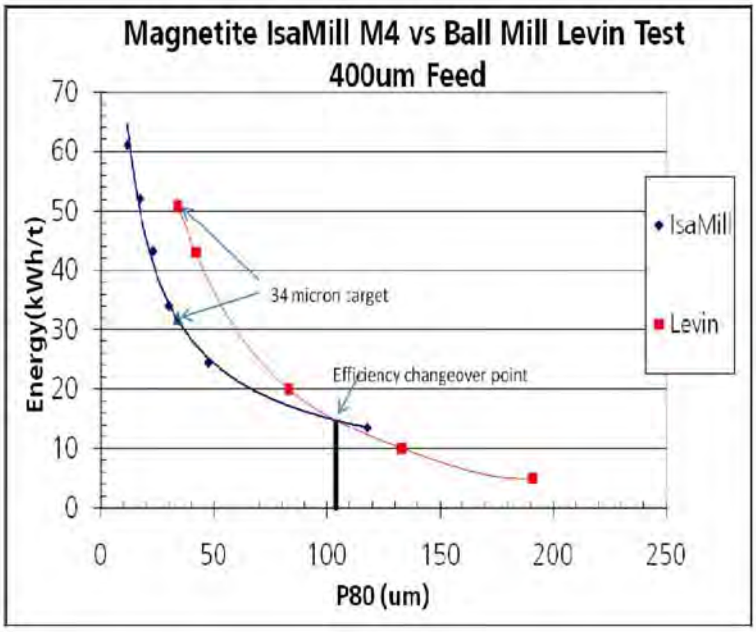


Рис. 9: Сравнение результатов тестирования IsaMill/Левина

Улучшение эффективности IsaMill при тонкодисперсных классах крупности можно объяснить механизмом измельчения и используемой мелющей средой. При тонкодисперсных продуктах крупностью ниже 70-100 мкм, истирание становится главным механизмом измельчения. 5 мм мелющая среда дает не только большую площадь поверхность по сравнению с 25 мм мелющей средой шаровой мельницы, но и более высокую вероятность ударов, чем в случае более крупной стальной мелющей среды, также она является более плотной, чем стальная среда.

Более эластичная стальная среда не переносит энергию до частиц руды так же эффективно, как керамическая среда. Твёрдость по Виккерсу для используемой керамики около 1000, в то время как для стальной среды она обычно находится в диапазоне 500-600.

Изменение эффективности также видно на рис. 10 (Бурфорд и Нива, 2008г), сравнение IsaMill с башенной мельницей на магнетитовом питании Ernest Henry, взятом из хвостов медной обогатительной фабрики. В случае крупности ниже 70 мкм более мелкая и прочная среда IsaMill считается более эффективной, чем 12 мм стальные тела, используемые в башенной мельнице. Это тестирование проводили с использованием 3,5 мм керамической среды. При использовании 5 мм среды это точка пересечения будет сдвинута в сторону более крупную.

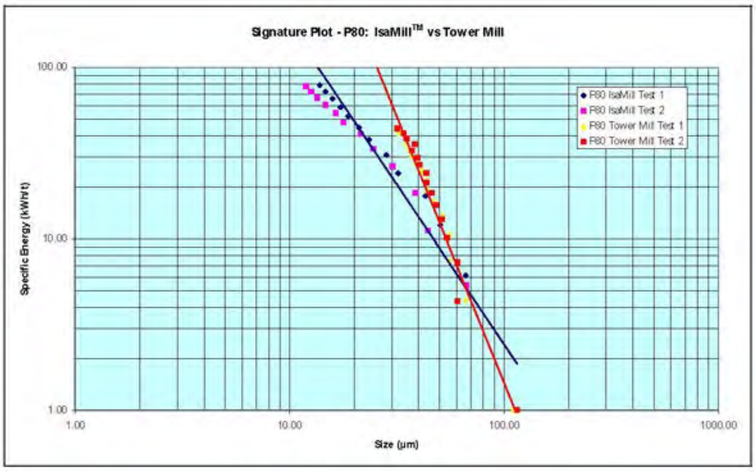


Рис. 10: Сравнение измельчения магнетитовой руды Ernest Henry на мельнице IsaMill и на башенной мельнице

Из рисунков 9 и 10 ясно видно, что измельчение в шаровой или башенной мельнице с использованием типичной среды более эффективно, когда применяется к производству “крупного” продукта в диапазоне +80 мкм. Следовательно, измельчение на шаровой мельнице выбрали, чтобы перекрыть «пробел» в эффективности, который имеется между измельчением в мельницах самоизмельчения с данной рудой до около 400 мкм P_{80} и применением IsaMill для крупности питания 100 мкм и тоньше. Согласно тесту Левина, для измельчения питания 400 мкм P_{80} до продукта 100 мкм P_{80} требуется 15 кВмч/т. При пилотном испытании было достигнуто измельчение в шаровой мельнице (с мелющими телами 32 мм) с 420 мкм до 78 мкм при удельном энергопотреблении 11,6 кВтч/т, что значительно более продуктивно, чем прогноз теста Левина. При переводе измельчения в полномасштабное производство с мощностью 2200 т/ч с крупности 770 мкм до 100 мкм, требование по установленной мощности шаровой мельницы составляет 34 МВт, это достижимо при использовании двух больших шаровых мельниц с двойной шестерней (одна на линии).

Только при необходимости получить продукт шаровой мельницы крупностью 80 – 100 мкм можно получить дополнительную выгоду в виде перехода со стальных шаров 25 или 32 мм, которые использовались при тестировании, на 40мм шары. Использование 40 мм шаров обеспечивает достаточное сокращение верхнего продукта при переработке питания 770 мкм F_{80} и значительно снижает износ стальных мелющих тел.

Третичное измельчение на IsaMill

Продукт промежуточной пилотной шаровой мельницы затем передавали на стадию магнитного обогащения, где выводили еще 20% по массе, большая была представлена кварцем, и небольшая часть гранитом. В результате на стадию измельчения IsaMill подавали менее абразивное питание крупностью 76 мкм, этот материал перерабатывали на мельнице M4 IsaMill, применяя стандартную кривую энергоемкости и керамическую мелющую среду 5мм. Результаты, показанные на рисунке 11, говорят о том, что для измельчения питания F_{80} 76,5микрон до продукта P_{80} 34 мкм требуемая энергия IsaMill составляет 12,4 кВтч/т.

Испытание IsaMill на более крупном материале (рис. 7) показало, что 13,4 кВтч/t будет необходимо для измельчения с F_{80} 76,5микрон до P_{80} 34микрон. Эта разница в 7,5% может быть результатом более мягкого питания благодаря дополнительной стадии магнитного обогащения. Она также очень близка к 5% пределу погрешности, относящемуся к этим испытаниям.

Эта кривая энергопотребления сравнима с другими испытаниями IsaMill на железной руде, которые выполнялись на похожем питании железной руды F_{80} и с энергией, приспособленной к 76,5 мкм, как показано на рисунке 12 (Ларсон, 2011).

Тест Левина показывает, что шаровая мельница требует 30 кВтч/т для подобного снижения крупности. Однако, эта разница двух испытаний на демонстрационной шаровой мельнице (420 мкм до 34 мкм против 420 мкм до 100 мкм) составила только 22 кВтч/т, необходима более тщательная оценка мощности шаровой мельницы для достижения конечного измельчения. Еще раз нужно отметить, что тест Левина дал преувеличенную оценку требования по питанию шаровой мельницы. Самое лучшее из имеющихся сравнение двух испытанных «соперников» тонкого измельчения – 22 кВтч/т для шаровой мельницы против 12 кВтч/т для IsaMill, Преимущество удельного энергопотребления в 45% только для IsaMill на данном этапе. К тому же реализовалась экономия на установленной мощности и капитальных затратах за счет отсутствия необходимости в циклонах в цикле IsaMill, в то время как они необходимы для шаровой мельницы.

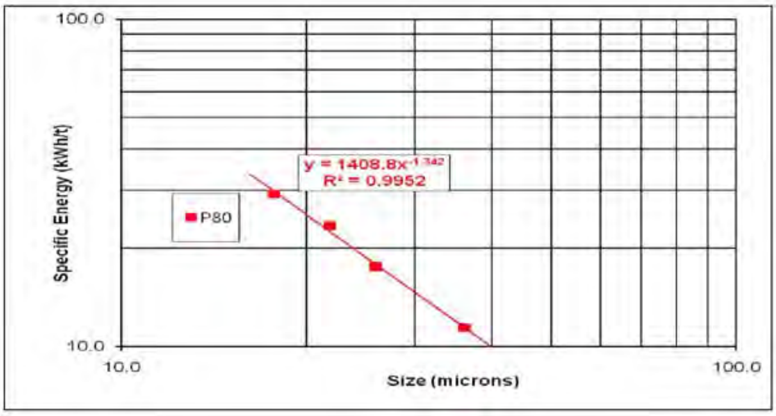


Рис. 11: Кривая энергоемкости для тонкодисперсного питания магнетитовой руды IsaMill

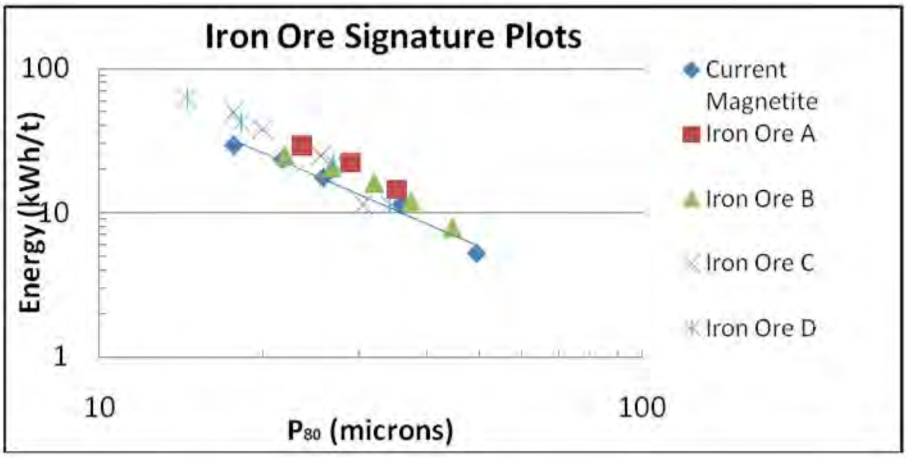


Рис. 12: Кривая энергоемкости для обычной железной руды

Более тонкое питание F_{80} 76,5 микрон также пропускали через мельницу M4 IsaMill в ходе короткого непрерывного испытания 180 кг и затем непрерывно как часть демонстрационной установки. Энергия 12-13 кВтч/т была намечена при помощи регулирования скорости мельницы. Целевую крупность P_{80} 34 мкм поддерживали при проведении этих тестов, включая снижение крупности верхнего продукта.

Также было обнаружено еще одно преимущество в износе керамической мелющей среды при измельчении этого более тонкого обогащенного материала. Износ среды был сокращен на 35% на кВтч по сравнению с вторичным питанием IsaMill (420 мкм F_{80}).

При условии, что в верхнем продукте может иметься примесь силиката с примесью железа, важно рассмотреть полностью распределение продукта по крупности, полученное при доизмельчении. Этот гранулометрический состав в зависимости от энергопотребления, который является наиболее близким к целевому значению 34 мкм, показан на рисунке 13.

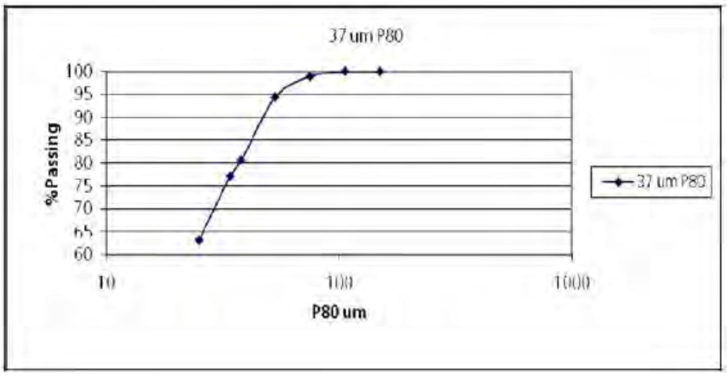


Рис. 13: Кривая гранулометрического состава продукта IsaMill

Гранулометрический состав продукта IsaMill составляет 98,9% класса 75 мкм и 100% класса 105 мкм при P_{80} 37 мкм. Это устраняет необходимость в более тонких грохотах, что экономит до $15 миллионов капзатрат вместе с эксплуатационными затратами, связанными с потенциальным техобслуживанием высокопроизводительного оборудования. Обычно 15 дюймовые циклоны будут давать соотношение P_{98}/P_{80} около 3-4 и это определенно требует использования нижних грохотов. В данном случае соотношение IsaMill ниже 2.

Кроме того природное улучшение качества концентрата без надобности в тонких грохотах и более лучший гранулометрический состав имеют дополнительное преимущество при получении еще более сверхтонких частиц. На промежуточном этапе магнитного обогащения после шаровой мельницы выводили 80% кремнезема, присутствующего в этом потоке. Это в результате приводит к меньшему количеству кремнезема, который необходимо перемалывать до 34 мкм. Это сокращение материала, который можно превратить в кремнеземистые шламы, вместе с ранее упомянутым более точным гранулометрическим составом, может привести к проектированию менее крупного гидросепаратора, что снижает занимаемую площадь одной из самых крупных единиц оборудования в технологической схеме. Также преимущества можно найти в фильтровании концентрата. Это испытание все еще продолжается на момент написания данной статьи.

Конечные результаты

Для определения наиболее эффективного средства сокращения первично раздробленной магнетитовой руды до целевой крупности P80 около 34мкм использовали сочетание лабораторных и полупромышленных испытаний. Эффективность была увеличена использованием специальных машин для различного помола в сочетании с отбраковкой пустой породы по возможности на более ранних этапах технологической схемы.

Было выяснено, что на стадии помола в мельнице самоизмельчения создается относительно крупный продукт в полупромышленном масштабе и этот поток будет более крупный в промышленном масштабе. Такая крупность первичного магнетитового концентрата делает его непригодным для подачи на устройство тонкого измельчения, такое как мельница IsaMill, но является идеальным для питания шаровой мельницы. Однако, как выяснилось шаровая мельница не подходит для приема руды из разгрузки мельницы самоизмельчения для конечного целевого измельчения на одном этапе, так как она становится относительно неэффективной для помола тоньше 80 мкм. Испытательные работы показали, что уместно использовать шаровую мельницу для помола до около 100 мкм P_{80} с последующим магнитным обогащением и с завершающим измельчением до 34 мкм в мельнице IsaMill. Экономится около 60 МВт мощности (40% от общей мощности и ~50% мощности для этапа измельчения) по сравнению с одностадийным циклом шаровой мельницы и $62 млн. ежегодно на затратах на мелющую среду как показано в Таблице 1. Необходимо включить в технологическую схему дополнительную стадию магнитного обогащения, а тонкие грохоты можно исключить, а также снизить требование к гидросепаратору. В мельнице IsaMill имеется внутренний центробежный классификатор, поэтому нет необходимости в установке циклонов на стадии третичного помола. Там, где это возможно, будет использоваться самотек, чтобы избежать установки дополнительных насосов. Например, верхний слив циклона шаровой мельницы будет самотеком стекать в промежуточный распределитель магнитного сепаратора, и сами магнитные сепараторы будут расположены таким образом, чтобы магнетит стекал непосредственно в питающие баки IsaMill.

Таблица 1. Сравнение установленной мощности и годовых затрат на мелющую среду

	Скорость подачи питания (т/ч)	Удельное потребление энергии (кВтч/т)	Установленная мощность (МВт)	Оценка по приблизительным годовым затратам на мелющие тела
Мельница самоизмельчения, продукт 770 мкм	3800	8,5	40	$0
Одностадийная шаровая мельница, продукт 34 мкм	2200	47	114	$86 млн
Одностадийная мельница IsaMill, продукт 34 мкм	2200	34	78	$57 млн
Шаровая мельница, продукт 100 мкм	2200	12	34	$13 млн
Мельница IsaMill, продукт 34 мкм	1720	13	24 58	$11 млн $24 млн

Хотя будет установлена мощность 58 МВт шаровой мельницы и мельницы IsaMill, только около 50 МВт рабочей мощности будет необходимо при средних эксплуатационных условиях. Установка восьми мельниц IsaMill M10 000 мощностью 3,0 МВт (четыре на линии) также оставляет некоторое место для расширения в будущем. Хотя каждая мельница запитывается от двигателя мощностью 3 МВт, в среднем необходимо только 2,7-2,8 МВт на мельницу. В таком случае, при остановке мельницы IsaMill на техобслуживание, можно активизировать другие мельницы до полной мощности, не воздействуя на производительность.

Добавление гидросепаратора в этот цикл является критичным для достижения содержания кремнезема менее 2%. Нижние магнитные грохоты не способны полностью вывести шламы, присутствующие и захватываемые при конечной крупности помола. Как выяснилось, гидросепараторы снижают содержание кремнезема в концентрате магнитного сепаратора на примерно 1%.

После гидросепаратора и магнитного обогащения имеется стадия флотации сульфида. В то время как пирит выводится в хвосты на стадиях магнитного обогащения, магнитный колчедан, присутствующий в руде, является магнитным и будет переходить в конечный концентрат. Содержание серы в конечном магнитном концентрате составляет 0,6%, а это неприемлемо для установок гранулирования. После стадии флотации сульфида содержание серы в конечном продукте снижается до менее 0,1%, что приемлемо.

Таким образом, конечная технологическая схема получения концентрата для данного проекта следующая:

Две мельницы самоизмельчения мощностью 20 МВт с 3мм грохотами с последующей стадией магнитного обогащения. Этот концентрат является питанием двух шаровых мельниц мощностью 17 МВт с гидроциклонами, верхний слив которых является питанием второго комплекта магнитных сепараторов. Магнитный концентрат является питанием для восьми мельниц M10 000 IsaMill мощностью 3 МВт. Продукт IsaMill имеет конечную крупность помола без дальнейшей классификации и последовательно загружается в гидроциклоны, с последующей стадией конечного магнитного обогащения и, наконец, флотацией сульфида. Нижний продукт флотации сульфида откачивается на фильтрационную установку, которая находится в стороне. На каждой стадии магнитного обогащения, гидроциклонов и флотации получают отвальные хвосты без их повторной циркуляции.

ВЫВОДЫ

Данное исследование можно считать успешным, так как были достигнуты значительные улучшения по сравнению с предварительным проектом в отношении капитальных и эксплуатационных затрат. К тому же были выполнены строгие требования по содержанию кремнезема и серы в конечном концентрате при этом снижении затрат. Еще более впечатляющим эти результаты делают короткие временные рамки выполнения этой работы и завершения проекта. Кампания полупромышленных испытаний на разгрузке мельницы самоизмельчения продлились одну неделю. Она включала работу двух разных шаровых мельниц с добавлением мельницы M4 IsaMill, для последовательной работы в полупромышленном масштабе. Экономия +56 МВт мощности помола и более $60млн. ежегодно на мелющей среде свидетельствует об эффективности простого, но в то же время хорошо продуманного плана.

БЛАГОДАРНОСТИ

Авторы хотели бы выразить свою благодарность работникам AMMTEC за испытания, проведенные в ходе данной кампании полупромышленной установки.

СПИСОК ИСПОЛЬЗУЕМОЙ ЛИТЕРАТУРЫ

Burford B, Niva E, 2008. Comparing Energy Efficiency in Grinding Mills, *in Metallurgical Plant Design and Operating Strategies (MetPlant 2008,)* pp 45-64, (AUSIMM, Melbourne).

Devaney F, 1985. Section 20, Iron Ore, in *SME Mineral Processing Handbook (ed: NL Weiss),* Society of Mining Engineers of AIME.

Iwasaki I, 1983. Iron Ore Flotation, Theory and Practice, Gaudin Lecture, 1982 Annual AIME Meeting, *AIME Transactions Volume 274* pp 622-631.

Iwasaki I, 1999. Iron Ore Flotation-Historical Perspective and Future Prospects, Nonsulfide Minerals, Section 3: in *Advances in Flotation Technology* pp 231-243.

Larson M, 2011. Xstrata Technology Internal Iron Ore Report.

McIvor R, Greenwood B, 1996. Pebble Use and Treatment at Cleveland-Cliffs’ Autogenous Milling Operations, *in SAG 1996*, pp 1129-1141 (Mining and Mineral Processing Engineering, UBC, Vancouver).

IsaMill- 1:1 Direct Scaleup from Ultrafine to Coarse Grinding

Larson M., Anderson G., Barns K., Villadolid V.
Xstrata Technology
307 Queen Street, Level 4
Brisbane, QLD 4000 Australia

mlarson@xstratatech.com

Abstract

The IsaMill has been used commercially in concentrator plants for over 15 years. Improvements in ceramic grinding media, mill design and wear components have advanced the IsaMill to the point where it can readily accept F_{80}'s of +300 microns. One thing that has not changed since the early days of development is the robust 1:1 scaleup of the mill from the laboratory to the mine site. This paper examines Xstrata Technology's efforts to both improve the grinding capability of the IsaMill and the work that has gone into ensuring the accuracy and precision of independent laboratories across the world that perform IsaMill signature plot scaleup work. Common issues encountered in design testwork are discussed in an effort to promote proper scaleup among all suppliers.

Introduction

Each year independent mineral processing laboratories around the world perform over 100 signature plots to provide an energy scaleup number for Xstrata Technology's IsaMill. 8 laboratories around the world are now certified every two years to ensure the reliability of their work and the robustness of their technique. This is a time consuming process to gather and split the concentrate sample, run replicate tests and ship out concentrate and media to the individual laboratories around the world. Once the testwork has been done any non-conforming laboratories must be inspected to determine any deviation from normal. The total cost of the program exceeds A$50,000. However, the results provide a level of confidence in the testwork being done that ensures every test can result in a process guarantee without Xstrata Technology being involved in the actual testwork. Through this partnership with independent laboratories improvements have been made to the signature plot procedure to reduce error wherever possible and the range of media sizes tested has increased with improved knowledge of the inner workings of the M4 4 litre IsaMill.

Section 1. IsaMill scaleup testwork

Historically IsaMill testwork was done in-house by MIM and then Xstrata Technology. However, as testwork is not a core business and the interest in signature plot testwork increased it became desirable to have external independent laboratories take on this workload.

The testwork itself is relatively simple and repeatable as long as the standard procedure is followed. This repeatability increases with practice.

Figure 1 and Table 1 represent the repeatability possible by someone with limited experience running the M4 (Larson 2012). The margin of error for these five tests is about 3.2% from the average energy to 10 microns. Each individual error from average is shown in the table.

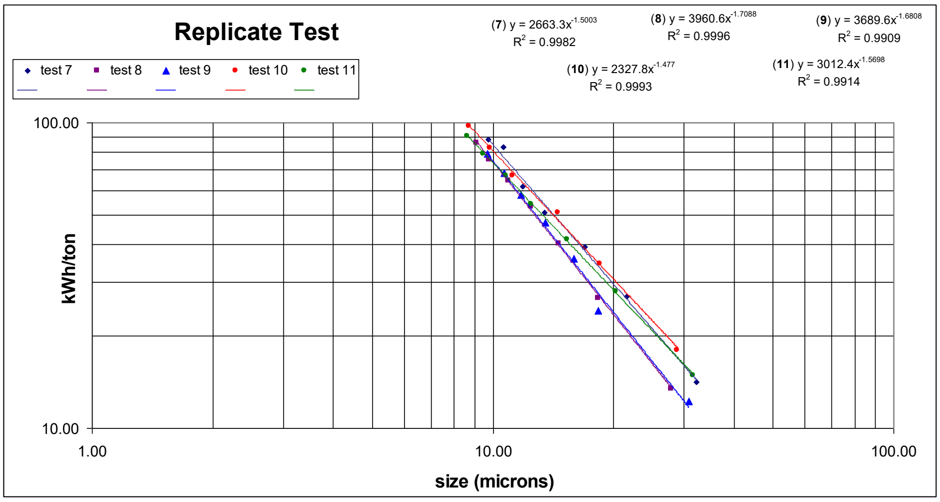


Figure 1. Copper concentrate replicate test

Energy to 10 microns (kWh/ton)				
Test 7	Test 8	Test 9	Test 10	Test 11
84.16	77.44	76.94	77.62	81.12
+5.91%	-2.54%	-3.17%	-2.31%	+2.09%
Mean 79.46 kWh/ton				
Standard Deviation 3.11				

Table 1. Copper concentrate replicate test energy to 10 microns

Figure 2 and Table 2 represents the repeatability of the first standard sample, with a margin of error of 6.8% from the average of the repeatability tests. This was done for the 2009/2010 standard sample.

In this round of testwork 7 different labs were certified. The criterion was to fall within 5% of the average energy of 42.8 kWh/t to a P_{80} of 15 microns from the XT replicate tests. Of the 7 laboratories certified, five were completed with their first try, one on their second and one on their third. In the case of the latter, there were issues with their Malvern laser sizer sub-sampling that was fixed after a visit to audit their procedure. The energies reported by the seven labs averaged 41.5 kWh/t.

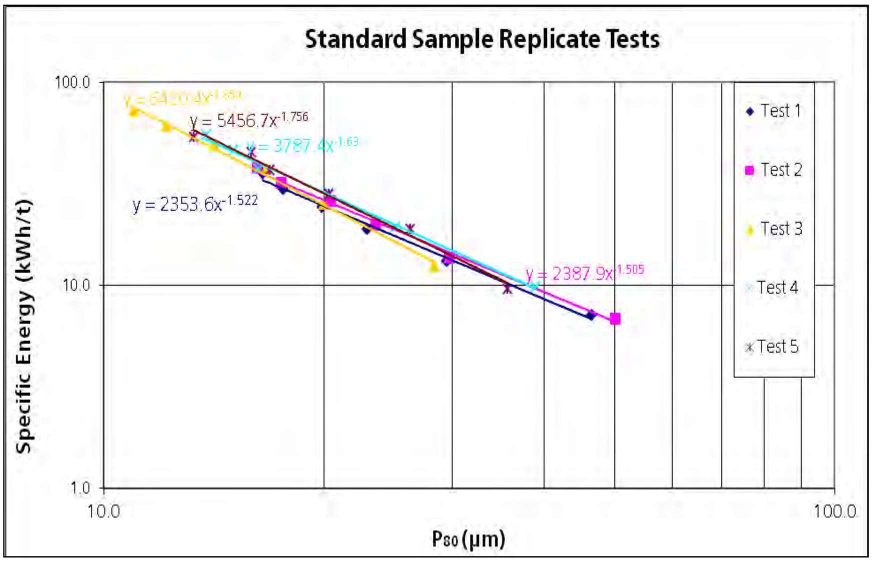


Figure 2. 2009/2010 standard sample replicate test (Villadolid 2009)

	Test 1	Test 2	Test 3	Test 4	Test 5	Average
Energy to 15 microns(kWh/t)	38.14	40.58	42.38	45.81	47.01	42.78
Error from average energy	10.9%	5.2%	.9%	7.1%	9.9%	6.8%

Table 2. 2009/2010 standard sample replicate test target energy and error

Figure 3 and Table 3 are the replicate tests for the most recent standard sample done in 2011/2012. This has a margin of error of 2.1 % to 12 microns and is performed by an XT engineer with the most recent practice of anyone performing signature plot tests. In this case the standard for the labs to maintain certification was adjusted to +/- 5% of the average energy rather than the +/- 2.1% shown possible by the replicate tests.

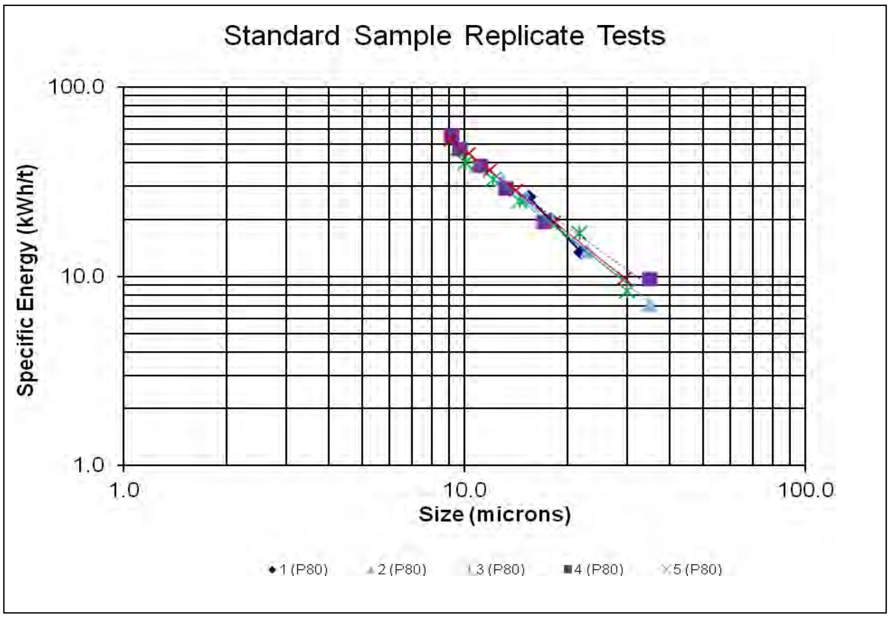


Figure 3. 2011/2012 standard sample replicate test (Villadolid 2011)

	Test 1	Test 2	Test 3	Test 4	Test 5	Average
Energy to 12 microns(kWh/t)	35.71	35.12	33.15	35.15	35.71	34.97
Error from average energy	2.1%	.4%	5.2%	.5%	2.1%	2.1%

Table 3. 2011/2012 standard sample replicate test target energy and error

Section 2. Common sources of error

The IsaMill signature plot test will have an average margin of error of about +/- 5% under ideal conditions to common IsaMill product sizes. Standard techniques have been set in an effort to minimize this error.

The energy meter will have an error of about +/- 1% according to the common manufacturers of energy meters in use. The slurry flow measurement will have an error of about .5 seconds, or 1-1.5% depending on flow. The density measurement will have an error of about 1%. The actual timing of each pass will have a margin of error of about 1%. As each pass requires about 6-7 minutes to complete, an error of 2-3 seconds on calling the tank empty is a minimal error. The remaining measurement error is due to any sizing or sampling inconsistencies in the Malvern laser sizer.

These errors are almost inevitable, improper technique will only serve to amplify them. Some errors encountered while doing laboratory audits include the following (this is not to single out any particular lab for criticism but to promote proper techniques among all labs).

2.1 Flow measurement

Improper flow measurement can greatly increase the margin of error of a test. Flow can either be measured with a 1 litre Marcy scale cup or with a suitably large graduated cylinder. With the Marcy scale cup it is relatively straight forward as the time is measured for slurry- not froth- to discharge from the holes marking the 1 litre point. In this case time/volume is actually being measured.

With a graduated cylinder time/volume cannot be used as the froth will obscure where the slurry actually reaches the line for the volume of interest. In the case of using a graduated cylinder to measure slurry flow the slurry must be directed into the cylinder for a given amount of time and then the level analysed when the sample has settled. In this case the measurement is in volume/time.

2.2 Density measurement

The density sample is taken in the middle of each pass. This requires that the feed sample is properly mixed to avoid feed density segregation between the beginning and end of the pass. For this reason all tanks used for IsaMill signature plots must have internal baffles and a mixer capable of suspending .5mm material of an SG of 4-5. Failure to use baffles can cause segregation and also centrifuging of the sample as it nears the end of the pass, starving the feed pump of slurry and artificially extending the pass and increasing the measured energy consumption.

The slurry density sample is never to be calculated by using the Marcy scale, but rather by a wet/dry mass. The Marcy scale can overestimate the slurry density if flow is allowed to continue into the cup even after overflowing. It can also underestimate the slurry density if material is allowed to spill and less than 1 litre of slurry is measured.

2.3 Sample sizing

The sizing of the individual pass samples can be wrought with errors that are sometimes hard to detect. For this reason Xstrata Technology has general guidelines for Malvern use during scaleup tests. All subsamples should be mixed in a baffled beaker and samples taken with an adjustable micro-pipette. Use of magnetic stirrers should be avoided as the magnetic bar can be interfered with by the sample pipette. Baffles must be used to prevent segregation between the middle and outside of the sample container. The adjustable pipette is needed to ensure that all slurry drawn is actually put into the Malvern. In the case of magnetite testwork screen sizing is preferred to avoid the possibility of magnetic agglomeration affecting the laser sizer results.

In one case a labs standard sample P_{80} energy was coming out higher than acceptable but the line for the P_{98} signature plot energy matched what would be expected from the replicate test, as did the mill net power draw. After examination it was determined that faulty sub-sampling for the Malvern was the cause. In this case the Malvern was located across the room from the sample mixer and a bulb pipette was used to take the Malvern sample. As the obscuration level of the Malvern determines how much sample is discharged, the combination of excess sample and segregation in the bulb resulted in the top size being segregated to the bottom and ultimately biasing the Malvern sample. By using an adjustable micro-pipette only the amount of slurry actually needed is sampled and all of it discharged into the Malvern, preventing any segregation biases.

The Malvern itself is basically a black box poorly understood by much of industry. Without proper care readings can be sporadic and inaccurate without the user knowing. There are specific settings for minerals but the one setting that is most commonly wrong is the absorption value. This can commonly range from .01-1 depending on the Malvern model. Xstrata Technology generally recommends that this setting be placed at .1 so that particles of .3-1 microns are measured and accounted for. By setting the absorption improperly sub 1 micron particles will not be measured. For most applications this may not be a major problem but for some IsaMill applications that size range can make up 10-20% of an ultrafine size distribution.

2.4 Viscosity

Excess viscosity can also have detrimental effects to energy efficiency if not fully understood and appreciated. Xstrata Technology has started to provide all new operating sites and M4 laboratories with a Marsh funnel. The Marsh funnel is a simple yet effective tool for quickly determining the viscosity of slurry without the need to stop a test to perform more complex rheology measurements. It consists of just a funnel and a one quart (.946 Litres) container. One quart of water takes about 28-30 seconds through the funnel.

Typically in a M4 signature plot test the net power draw will drop from pass to pass. This is simply an effect of the material being ground finer and thus being easier to mix. It was found over the course of several M4 and M20 test programs that a set of conditions existed where this power draw would start to increase. Comparing to Marsh funnel readings taken during these same passes it was found that this increase in power draw correlated to Marsh funnel times of about +38 seconds for the one quart of slurry to pass through the funnel. This time limit will change with media size, as more of a void space between larger media won't be as sensitive to viscosity changes but this aspect has not yet been fully investigated.

Figure 4. Marsh Funnel

(Wikipedia, 2008, http://en.wikipedia.org/wiki/Marsh_funnel)
The Marsh funnel is not a rheometer, because it only provides one measurement under one flow condition. However the effective viscosity can be determined from following simple formula.

$\mu = \rho\,(t - 25)$

where μ = effective viscosity in centipoise
ρ = density in g/cm^3
t = quart funnel time in seconds

This is by no means meant to be a precision measurement. It does though act as an invaluable tool when each pass through the IsaMill only allows for 6-7 minutes for all measurements and it must be decided quickly if water is to be added to dilute the next pass.

Each pass through the IsaMill is separately accounted for with an individual energy so each pass at a different density does not affect the others.

2.5 Sample mass and segregation

The standard for most materials is to provide 15 kg of dry solids for each M4 signature plot. The correct amount of solids is critical to ensure that steady state is reached and a representative discharge is sampled without coarse solids being segregated and held in the mill. This requirement will be the same for all stirred mill tests where a continuous discharge sample is collected. An example of this is shown below in Figure 5 by Gao testing different feed masses in a 40 litre pilot Tower Mill. The smaller sample mass shows less energy required to reach equivalent particle sizes. Although this result may look good it is not realistic compared to the test done with more sample. For this reason Xstrata Technology recommends that to ensure proper scaleup, testwork is done with 3-4 x solid volume than mill void volume. For common sulphide ores this results in 15 kg of total mass. For something like magnetite with a higher solid SG more material is required. If a very coarse product and low energy is desired more sample still may be required to ensure steady state is reached in the first pass and that there is enough time per pass for the operator to reliably take all measurements with the high flowrate. If screen sizings are to be done where more mass is removed during each pass more starting mass will be required.

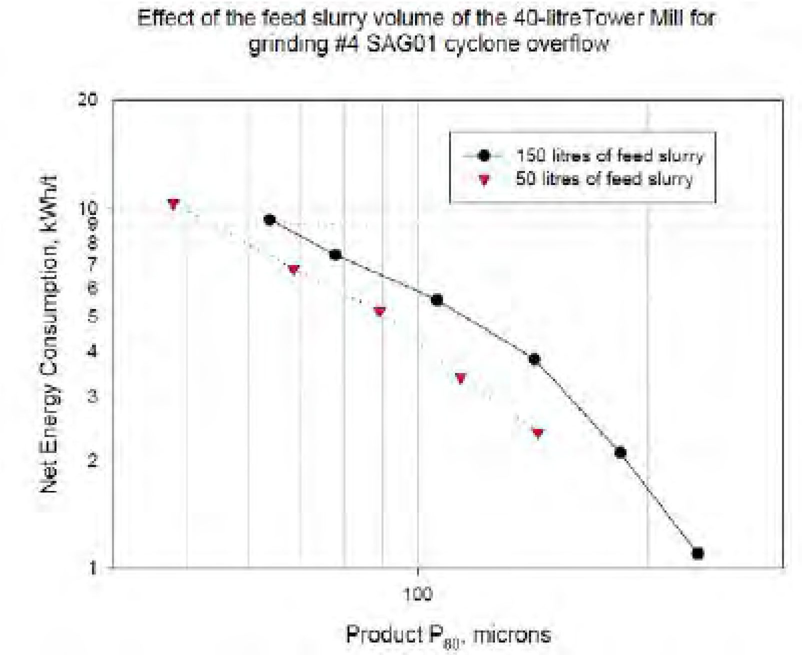


Figure 5. Effect of varying feed solids mass in a pilot tower mill (Gao)

Not all stirred mill test programs follow these guidelines however. Published conditions below compared to the IsaMill signature plot standard show recommended solid volumes of less than half of that used to size an IsaMill and barely enough solids to equal the total mill void space. (Nippon Eirich 2009 and Rahal, Erasmus and Major 2011)

Mill Type	**Mill Open Volume(L)**	**Sample Mass**	**Solid Volume(L)**	**Ratio**
M4 IsaMill(4L)	**1.35**	**15kg**	**5.00**	**3.70**
Nippon-EirichNE008(8L)	**~2.35**	**10kg**	**3.33**	**1.42**
Nippon-EirichKM-5(120L)	**~35.20**	**150kg**	**50.00**	**1.42**
Knelson-Deswik10(10L)	**~4.72**	**20kg**	**6.67**	**1.41**

Table 4. Regrind mill test volume ratios

Section 3. Larger grinding media

One of the most significant advances in IsaMill technology over the last 5 years has not come from Xstrata Technology itself but from media suppliers around the world. The original UFG IsaMills were run on either sand or slag, a major limitation to the coarseness of the feed that can be processed. Even the first widely used ceramic media in an IsaMill was limited to an effective size of 3.5mm. While progressing up the feed coarseness scale, there were still limitations as to what this was capable of grinding. In recent years 5 to 6.5mm high quality ceramic media has become available that has greatly increased the efficiency of grinding coarser material in the IsaMill. This media has also resulted in product size distribution curves that are much sharper and more efficient to process downstream. An example of this from Anderson, et al, is shown in Figure 6 for MRM ore. At an equivalent energy and feed size the 5-6mm media gives a much sharper size distribution curve with the top size completely ground.

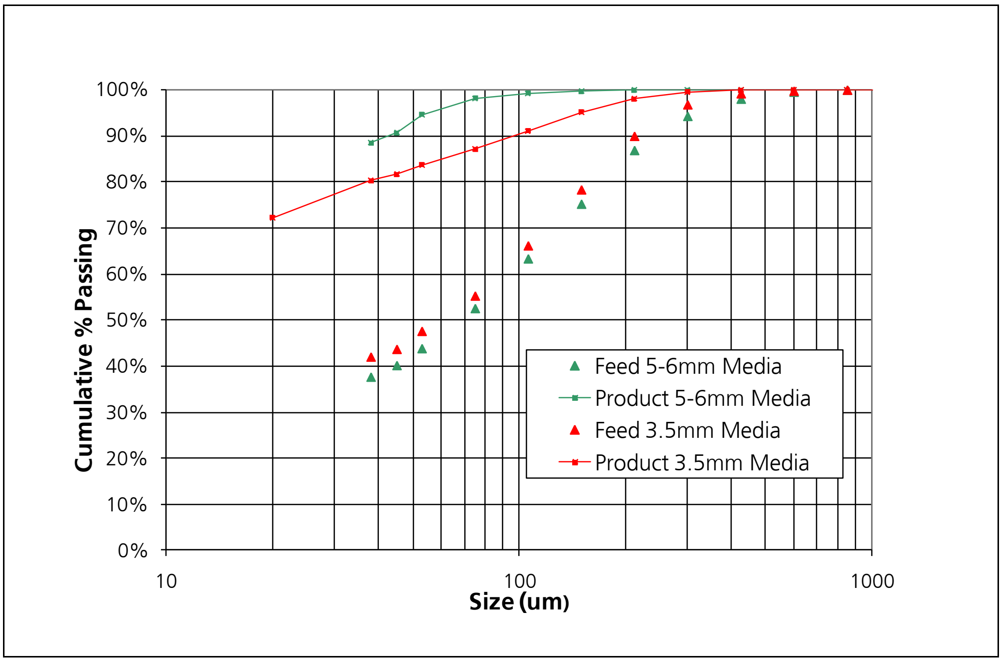


Figure 6. Performance of 3.5 mm Media at 16 kWhr/t and 5-6 mm at 16 kWhr/t

In the case of the Ernest Henry Mine magnetite circuit the mill feed during startup was regularly between an F_{80} of 300-350 microns with a top size approaching 1mm. Due to the coarseness of this stream a 6.5mm ceramic media manufactured by Cenotec was chosen to ensure top size breakage. Signature plot work was done on the cyclone underflow of the rougher magnetic separator at AMMTEC prior to starting this circuit. The comparison between the AMMTEC signature plot (blue) and the EHM commissioning surveys (red) are shown below in Figure 7. In this case the test ore and commissioning ore were taken months apart and the media top size is 6mm for the AMMTEC test and 6.5mm for the EHM M10,000 but the scaleup to the mill target of 45 microns still falls within the 5% margin of error associated with the test. In this case the net energy for the M10,000 to grind to a P80 of 42 microns is 29.9 kWh/t versus 28.8 kWh/t.

It is currently unknown how large of a media can be tested in a standard M4 IsaMill and still scale to full size mills. At some point the wall effect will become apparent and the power draw will increase due to media shearing between the disc tip and mill shell. This has not happened at 6mm.

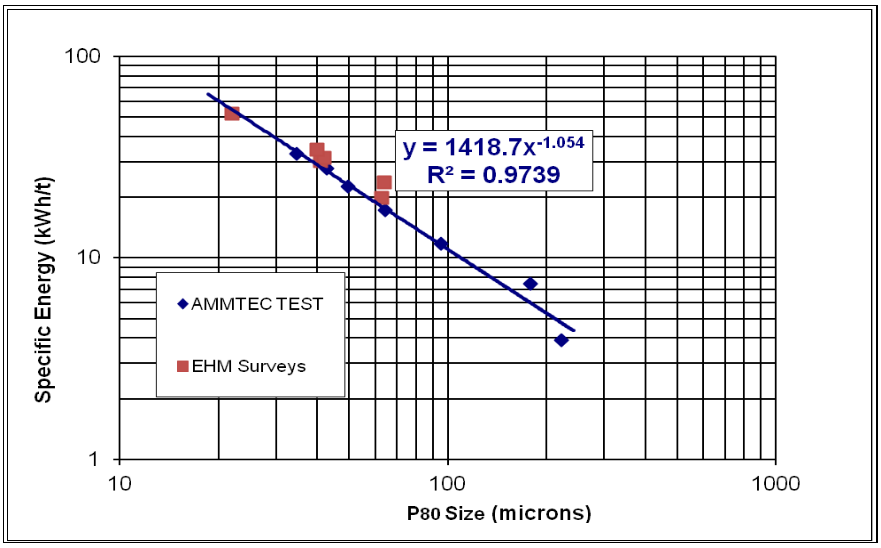


Figure 7. EHM M10,000 vs. AMMTEC M4 scaleup (Larson 2011)

Despite being a 6.5mm top size and needing to grind coarse abrasive magnetite slurry the Cenotec media at EHM has remained very round and worn at a rate of 8-9 g/kWh. Figure 8 shows this media after being emptied into the media bin one month into operation. At this point the media would have turned over slightly more than 1 time.

Figure 8. Cenotec 6.5mm ceramic media charge at EHM after 1 month (Larson 2011)

Advances in the size of media available and the confidence that it will scale from a standard M4 test has led to other test programs to take advantage of IsaMill technology in new grinding duties. Work performed by AMMTEC from David, et al, shown below in Figure 9 demonstrates the energy efficiency of the IsaMill when taking a coarser magnetite feed and grinding to a target of 34 microns. This efficiency gain over the ball mill increases as the grind goes finer. This would further promote the possibility of running the IsaMill either alone at a small scale or in series with a ball mill with 300-400 micron F80's and being efficient to grind products below 60-70 microns.

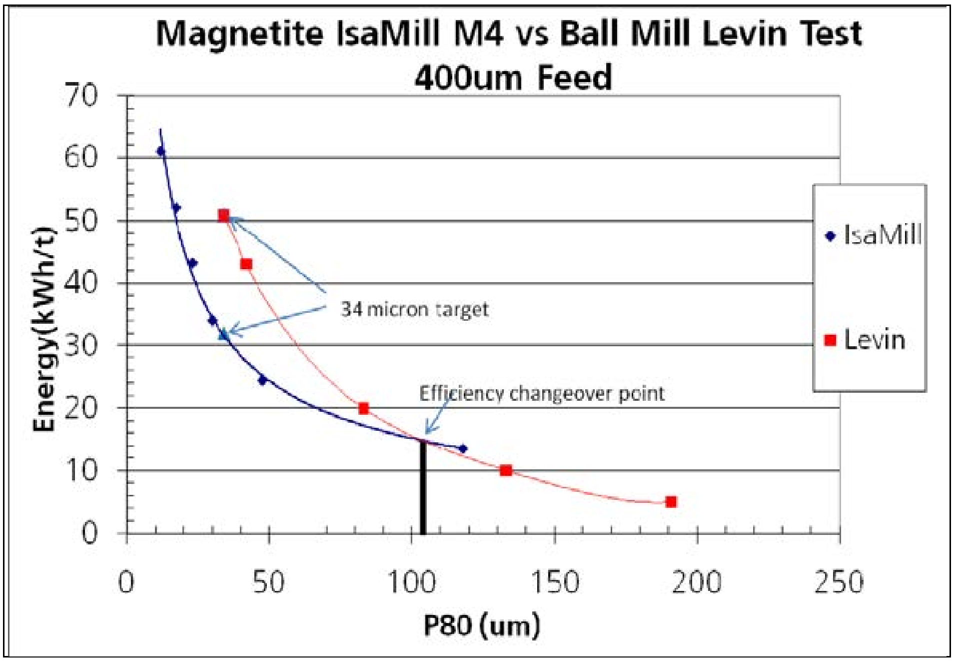


Figure 9. IsaMill/Levin test comparison (David, Larson, Le 2011)

Similar to the MRM results with the sharper product size distribution the magnetite results (Figure 10) showed no material over 106 microns in a 37 micron P_{80} and very little over 75 microns. The P_{98}/P_{80} ratio in the IsaMill is typically a positive factor even with smaller grinding media but the use of 5mm media on feeds typically served by 3.5mm media can improve this further. In the case of magnetite this results in less oversize middlings and an improved concentrate grade.

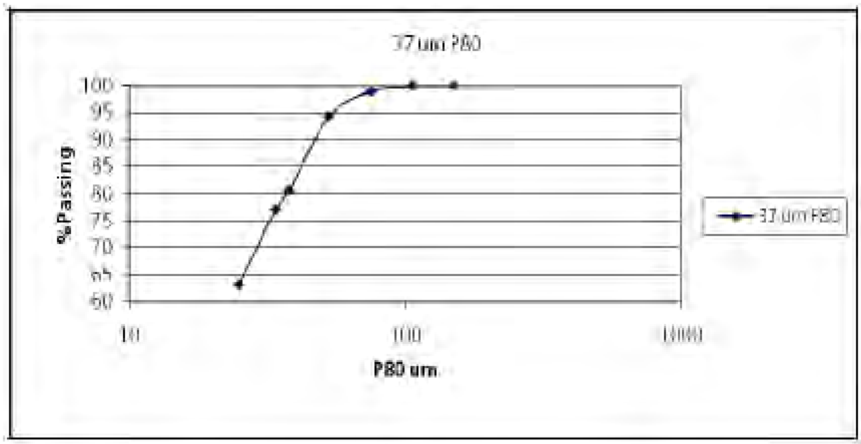


Figure 10. IsaMill magnetite product size distribution curve (David, et al)

In this program the M4 was also run continuously in the pilot plant to confirm that the energy and product size distribution were not a result of coarse material segregation and to produce feed for the downstream processes. The M4 is ideally suited for this application as at a common magnetite regrind energy of 10-15 kWh/t about 100 kg/hour can be processed through the mill.

The transition to 5mm media may also improve the energy efficiency of a process depending on the target P_{80}. In all cases whether comparing different technologies or just media size for a particular technology there is a cross-over point where one becomes more efficient than the other. In the cases in Figure 11 the larger media is more efficient to a coarser product size. When an application requires a finer grind and more energy there is enough residence time for the 3.5mm media to break down the coarsest incoming feed at the same rate it comes in. Compared to 5mm media it will remove smaller pieces at a time so needs more time to grind a coarser feed. This is ideal for a finer grind and high energy application. At a coarser product and lower energy the high flowrate may overwhelm the 3.5mm media and fill up the mill with coarse material.

This is shown in Figure 11 for two different platinum applications. At product sizes above 30 microns the 5mm media will be more efficient than the 3.5mm media. In this case the smallest media size capable of breaking the incoming feed is not necessarily the most efficient option.

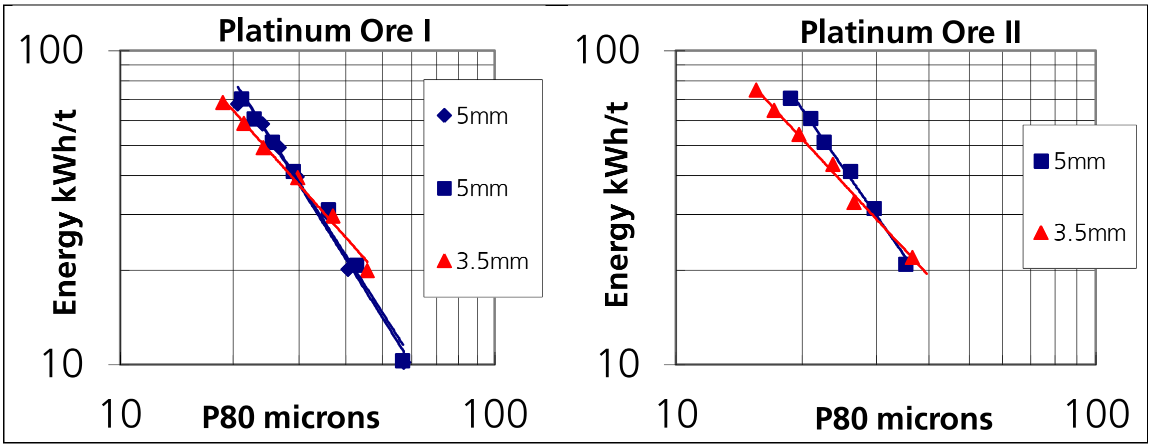


Figure 11. Comparison of 5mm and 3.5mm media for platinum ore (Larson 2010)

Use of +5mm media has the potential to improve energy efficiency and product size distribution for existing operations while increasing the range of feed sizes the IsaMill can treat in future applications.

Conclusions

By treating the independent mineral processing laboratories around the world as partners and involving them in advances in IsaMill technology Xstrata Technology has been able to maintain rigorous standards in the quality of scaleup work done by these labs. Advances in the ability to treat coarser feeds have not impaired the ability of large scales mills to still be accurately scaled from the standard 4 litre IsaMill. Large improvements in energy efficeincy and product size distributions can be realized by utilizing the 4 litre mill in the lab or pilot plant setting.

Acknowledgements

The authors would like to thank Xstrata Technology for permission to publish this work and assistance in writing it. We would also like to thank Xstrata Copper's Ernest Henry Mine for permission to publish the EHM magnetite IsaMill scaleup work and AMMTEC for performing this lab work.

References

Anderson G, Smith D, Strohmayr S; (2011), IsaMill™ Technology in the Primary Grinding Circuit; Proceedings of the 2011 SAG Conference, Vancouver Canada

David D, Larson M, Le M; (2011) Optimising Western Australia Magnetite Concentrator Design, Proceedings of the 2011 MetPlant Conference, Perth Australia

Gao M; CSIRO report

Larson M; (2010) Internal Xstrata Technology media size memo, October 2010.

Larson M.; (2011) EHM Magnetite report. December 2011.

Larson M; (2012) Experimental Study of IsaMill Performance Leading to a Preliminary Model, Master of Philosophy thesis, University of Queensland-JKMRC

Nippon Eirich Brochure; March 2009

Rahal D, Erasmus D, Major K; (2011), Knelson-Deswik Milling Technology: Bridging the Gap between Low and high Speed Stirred Mills; Canadian Mineral Processors 43rd Annual Operators Conference, Ottawa, Canada.

Villadolid V; (2009), 2009 Standard sample internal report

Villadolid V; (2011), 2011 Standard sample internal report

Wikipedia; (2008) http://en.wikipedia.org/wiki/Marsh_funnel accessed 12/3/2008

INTERNATIONAL
AUTOGENOUS GRINDING
SEMIAUTOGENOUS GRINDING
AND
HIGH PRESSURE GRINDING ROLL
TECHNOLOGY
2011

EDITORS:

Ken Major
Mineral Processing Consultant
KWM Consulting Inc.
Maple Ridge, B.C.

Brian C. Flintoff
Senior Vice President, Technology
Metso Minerals Canada Inc.
Kelowna, B.C.

Bern Klein
Associate Professor and Department Head
Norman B. Keevil Institute of Mining Engineering
University of British Columbia

Kelly McLeod
Process Consultant

Proceedings of an International Conference on Autogenous Grinding, Semiautogenous Grinding and High Pressure Grinding Roll Technology held September 25 – 28, 2011in Vancouver, B.C., Canada. This conference is made possible by the support from the Norman B. Keevil Institute of Mining Engineering of the University of British Columbia, the Canadian Mineral Processors of Canadian institute of Mining, Metallurgy and Petroleum and the Canadian Mining Industry Research Organization.

A PILOT-SCALE EXAMINATION OF A HIGH PRESSURE GRINDING ROLL / STIRRED MILL COMMINUTION CIRCUIT

*J.A. Drozdiak[1] and B. Klein[1], S. Nadolski[2], A. Bamber[3]

[1]*NBK Institute of Mining*
517 – 6350 Stores Road
Vancouver, Canada V6T 1Z4
*(*Corresponding author: jeffatmining@yahoo.com)*

[2]*Koeppern Machinery Australia*
73 Pavers Circle
Malaga, Australia

[3]*BC Mining Research*
122 – 1857 West 4th Avenue
Vancouver, Canada V6J 1M4

A PILOT-SCALE EXAMINATION OF A HIGH PRESSURE GRINDING ROLL / STIRRED MILL COMMINUTION CIRCUIT

ABSTRACT

In this paper we examine, through pilot-scale testing, the possibility of operating a circuit comprised of two stages of HPGR comminution, followed by grinding through a horizontal stirred mill. In order to assess whether the novel circuit design could achieve the reduced energy requirements indicated in the literature, two more-established circuits, a cone crusher / ball mill and an HPGR / ball mill, were examined using a combination of testing and flowsheet simulation. The results showed that, based solely on the specific energy requirements for comminution, the HPGR / stirred mill circuit achieved a reduction of 9.2% and 16.7% over the HPGR / ball mill and cone crusher / ball mill circuits, respectively.

KEYWORDS

High pressure grinding roll, stirred mill, energy, comminution, flowsheet development

INTRODUCTION

The mining industry will be faced with new challenges in the years ahead. The exponentially-increasing global population has resulted in an increased demand for raw resources. With the known rich, coarse-grained deposits being depleted, attention has turned to development of low-grade deposits requiring increased tonnages to achieve adequate metal production. This increased tonnage has resulted in an increased energy demand associated with metal extraction. Coupled with this, society is becoming increasingly conscious of their footprint on the environment, and serious attempts have begun to reduce carbon emissions and increase energy efficiency (Norgate & Haque, 2010). To adapt to this changing landscape, the mining industry must begin to accept and adapt new, more energy-efficient technologies and begin to focus on developing flowsheets capable of addressing the above issues.

Comminution, the process of crushing and grinding ore to liberate valuable minerals, is the most energy-intensive part of the processing flowsheet, and accounts for upwards of 75-80% of the overall energy consumption of the processing plant (Abouzeid & Fuerstenau, 2009). In addition, unit operations such as tumbling mills are as low as 1% efficient (Fuerstenau & Abouzeid, 2002). Currently, the main comminution circuits employed in the mining industry to process hard-rock, low-grade deposits include some form of tumbling mill. This equipment utilizes steel balls (ball mills), competent ore (Autogenous Grinding (AG) mills), or the combination of the two (Semi-Autogenous Grinding (SAG) mills) to fracture rock using the breakage mechanisms of impact and abrasion. The rotation of these large, cylindrical mills, coupled with the low probability of ball – particle collisions, results in a high demand for energy in the grinding process. Although their established circuit design and ability to process high tonnages is a huge benefit, the increased energy demand and inability to efficiently grind to liberation sizes below 45 µm (Shi, Morrison, Cervellin, Burns, & Musa, 2009) could slowly decrease their role in flowsheet designs of the future, especially as the increased demand for raw resources results in an increase in the development of finer-grained deposits.

In the past 20 years, new, more energy-efficient technologies have been developed and adapted for hard-rock mining comminution. The High Pressure Grinding Roll (HPGR), an innovative technology adapted from the cement and briquetting industries, has begun to be considered for more base metal projects now that roll surfaces have been developed to treat hard, abrasive ores (Dunne, 2006). Operating with two counter-rotating rolls, HPGRs create a compressive bed of particles between the rolls, utilizing the process of inter-particle breakage. This form of breakage results in improved comminution performance with a decreased demand on energy (Klymowsky, Patzelt, Knecht, & Burchardt, 2006). Additionally, unlike tumbling mills, which require steel balls to act as an energy transfer medium, HPGRs transfer energy directly from the rolls to the bed of material, resulting in an increase in energy efficiency (Fuerstenau & Kapur, 1995). Another technology, known as a horizontal stirred mill or IsaMill^{TM}, was adapted from the pharmaceutical and related industries in the early 1990s to help effectively process fine-grained ore bodies (Johnson, Gao, Young, & Cronin, 1998). The IsaMill^{TM} consists of a cylindrical tube with a centrally-rotating shaft, mounted with evenly-spaced grinding discs. Loaded with small ceramic grinding media (2 - 6 mm) and operated at high speeds, the equipment utilizes high-intensity attrition breakage to reduce particles in size. The rotation of a central shaft, as opposed to the entire grinding chamber (tumbling mills), results in decreased energy requirements, while the combination of small, hard (ceramic) grinding media and increased media velocity has been shown to improve the energy efficiency of grinding in fine particle sizes (Burford & Clark, 2007).

In this paper, we examine the possibility of incorporating the above-mentioned energy-efficient equipment into a single flowsheet, and eliminating the need for a tumbling mill. The biggest obstacle surrounding this research was that the proposed circuit would be operating both pieces of equipment outside of their normal operating range. As HPGRs began being adapted to the hard-rock mining sector, they found the most functionality in a tertiary crushing role, preparing feed for the ball mill (Morley, 2006). Therefore, the process envelope for an HPGR operating in hard-rock circuits typically has feed sizes of up to 70 mm, and products normally no finer than 4 mm (Gruendken, Matthies, & van der Meer, 2010). At the same time, horizontal stirred mill technologies such as the IsaMill^{TM} have begun to be well-established in grinding applications as a regrind mill, providing a more energy-efficient alternative for processing rougher concentrates (Burford & Clark, 2007).

The combination of an HPGR and a stirred mill into a single flowsheet has been discussed several times in the literature. Valery and Jankovic (2002) proposed the first concept of a combination HPGR / stirred mill circuit in a study examining the need for a reduction in the energy requirements of comminution. Simulating results for a more energy efficient circuit, a high-intensity blasting, two stage HPGR / Vertimill$^{®}$ circuit was compared to a conventional blasting, SAG / ball mill circuit. The simulation results predicted an energy savings of 45%, but no actual testwork was conducted. Pease (2007) presented the concept of an HPGR / IsaMill^{TM} circuit in his discussion of coarse stirred milling at McArthur River. No testing was carried out, but Pease predicted that this circuit could be an example of comminution flowsheet design of the future. Ayers, Knopjes, and Rule (2008) described the first operation of an HPGR / IsaMill^{TM} circuit using pilot-scale equipment. The authors documented Anglo Platinum's research into applying the IsaMill^{TM} to coarser feed applications. A continuously operating circuit was established using an HPGR in closed circuit with a dry screen, followed by wet screening of the undersize, at a cut size of 850 µm. The screen product was fed to an M250 IsaMill^{TM} operating with 3.5 mm MT1 ceramic grinding media. With an f80 of 300 µm and a product p80 of 45 µm, the IsaMill^{TM} circuit achieved 1.3 tph, with a specific energy consumption of 75 kWh/t and a total circuit energy consumption of 80 kWh/t.

Using pilot-scale equipment, we developed an appropriate circuit layout and determined the potential specific energy for comminution required to process a copper-nickel sulphide ore from an f80 of 21 mm to a p80 of 75 µm. To determine whether the novel circuit arrangement could reduce the energy demand for comminution indicated in the literature, specific energy requirements for two established circuit layouts, a cone crusher / ball mill and an HPGR / ball mill, were explored using a combination of laboratory testing and circuit simulation. The ability to improve comminution efficiency, while providing the flexibility of grinding efficiently to fine product sizes, could help make the HPGR / stirred mill circuit an attractive alternative for future comminution flowsheets.

EXPERIMENTAL PROGRAM

Three circuits were examined for this energy comparison study, a cone crusher / ball mill circuit, an HPGR / ball mill circuit, and the novel HPGR / stirred mill circuit. The feed size to each circuit was fixed at an f80 of 21 mm, and a product p80 of 75 µm was chosen as a suitable feed size for flotation. The circuits were evaluated solely on the power consumed per tonne of material, in order to achieve an equivalently-sized product from an equivalently-sized feed. Energy requirements of material-handling equipment, such as conveyors, pumps and screens, were not taken into account.

Comminution Circuits

What follows is a discussion of the three comminution circuits examined in this paper. The approach in all cases was to determine an appropriate set of design criteria for each flowsheet and to calculate the specific work index for each stage of comminution, based on the work index determined and the transfer sizes selected.

Cone Crusher / Ball Mill Circuit

The first circuit we examined was a cone crusher / ball mill circuit, typically found in a three-stage crushing flowsheet. This circuit was the industry standard for hard-rock comminution prior to the establishment of SAG mill technology. The circuit is comprised of a cone crusher in closed circuit with a screen, followed by a ball mill in closed circuit with a cyclone. The flowsheet of this circuit is shown in Figure 1. Data for the circuit was generated from a combination of Bond grindability testing and flowsheet simulation using JK SimMet® software.

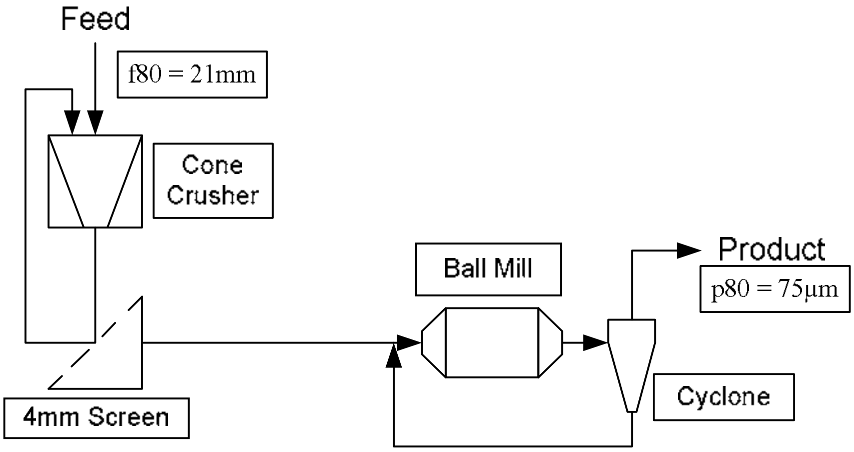


Figure 1 – Cone crusher / ball mill flowsheet

HPGR / Ball Mill Circuit

The second circuit we examined was an HPGR / ball mill circuit. This circuit mimics the standard HPGR comminution flowsheet currently being used in the hard-rock mining sector at operations such as Cerro Verde in Peru (Vanderbeek, Linde, Brack, & Marsden, 2006). The circuit is comprised of a high pressure grinding roll in closed circuit with a screen followed by a ball mill in closed circuit with a cyclone (refer to Figure 2). Data for this circuit was generated using a combination of HPGR pilot-scale testing, Bond grindability testing, and simulation using JK SimMet® software.

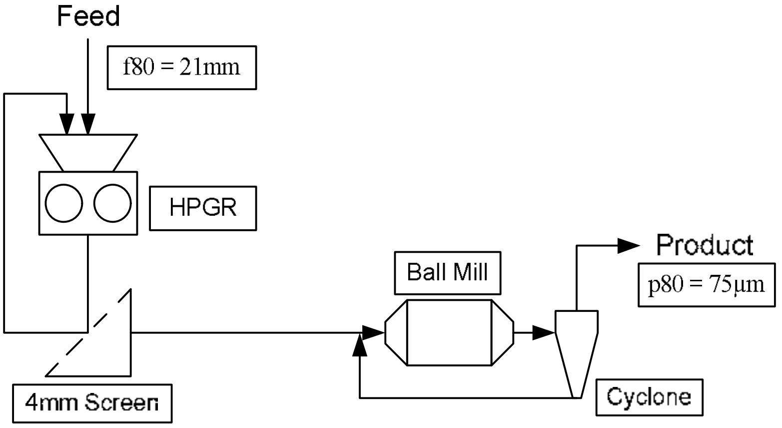


Figure 2 – HPGR / ball mill flowsheet

For HPGR pilot-scale evaluation, tests were carried out to assess the influence of different process parameters on comminution performance. These tests included the variation of specific pressing force, as well as closed-circuit testing with a 4 mm screen. Data from this study was entered into JK SimMet® to model fit an appropriate HPGR model. The T10H and HPGR power coefficient model parameters were fitted using the procedure outlined by Daniel and Morrell (2004). After calibration of the HPGR model, simulation was carried out for the HPGR / ball mill circuit to determine the appropriate transfer size between the HPGR and the ball mill.

HPGR / Stirred Mill Circuit

The final circuit we examined was the HPGR / stirred mill circuit. Daniel (2007b) determined from HPGR pilot-scale testing that two consecutive passes through the HPGR produced the highest size reduction ratios and further passes through the rolls resulted in diminishing size reduction and lower energy efficiency. Therefore, we determined that the novel HPGR / stirred mill circuit would incorporate two stages of high pressure grinding to prepare the feed for stirred milling. Drozdiak, Nadolski, Bamber, Klein, and Wilson (2010) demonstrated through pilot-scale testing that an appropriate transfer size between a two-stage HPGR circuit and a stirred mill circuit would be 710 µm. Using this data, two different HPGR / stirred mill circuits were examined. Circuit A comprised of the first-stage HPGR in open circuit feeding the second-stage HPGR in closed circuit with a 710 µm screen, and the undersize passing through a stirred mill in open circuit (refer to Figure 3), while Circuit B comprised of the first-stage HPGR in closed circuit with a 4 mm screen and the undersize feeding the same circuit layout as Circuit A (refer to Figure 4). Since no small-scale tests are available to determine the specific energy requirements for this equipment, pilot-scale testing was performed for the entire circuit.

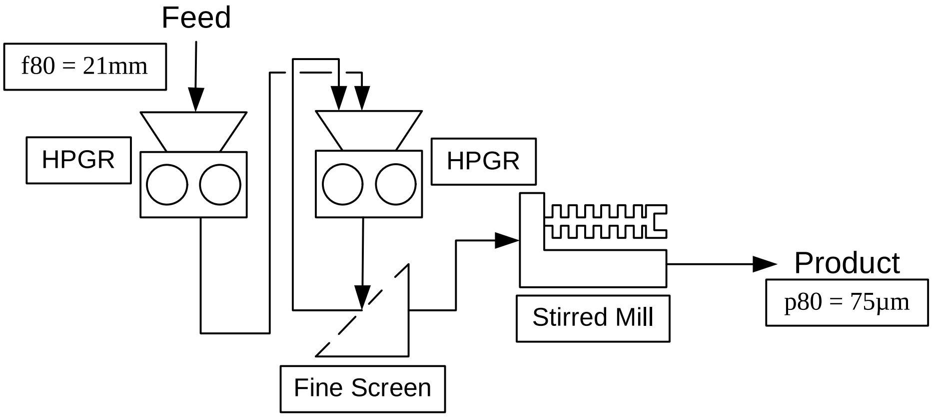


Figure 3 – HPGR / stirred mill flowsheet (Circuit A)

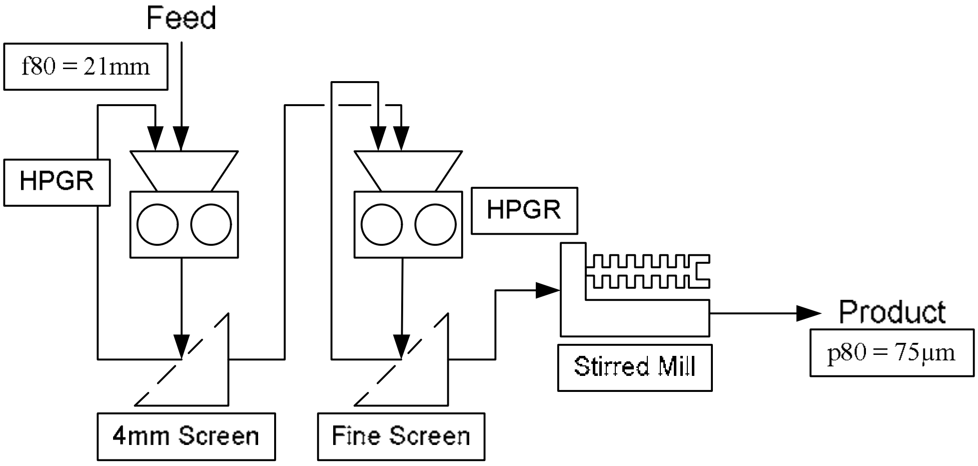


Figure 4 – HPGR / stirred mill flowsheet (Circuit B)

Sample Description

The sample used for this study came from Teck Limited's Mesaba copper-nickel deposit located in the Mesabi Range of the Duluth intrusive complex, situated in North-eastern Minnesota. This complex is comprised of mafic volcanics (tholeiitic basalt) with layered intrusions of primarily a gabbro-troctolite composite (Minnesota Geological Survey, 2010). Mineralogy of the Mesaba deposit comprises mainly of massive and disseminated sulphides with the main minerals of interest being chalcopyrite (copper), cubanite (copper), and pentlandite (nickel). The inferred resource stands at 700 Mt, with a grade of 0.46% Cu and 0.12% Ni (Infomine, 2001).

Approximately 5 tonnes of sample, at nominally 100% minus 100 mm, was shipped to UBC. We screened and crushed the material in a laboratory jaw crusher to 100% minus 32 mm, and homogenized and split the sample into sixteen 45-gallon drums using a rotary sample splitter. A representative sample was taken for size distribution, bulk density and moisture content determination. A moisture content of 1% and a bulk density of 2.16 t/m^3 with a Specific Gravity (SG) of 3.0 were established for the ore. The particle size distribution of the sample is shown in Figure 5.

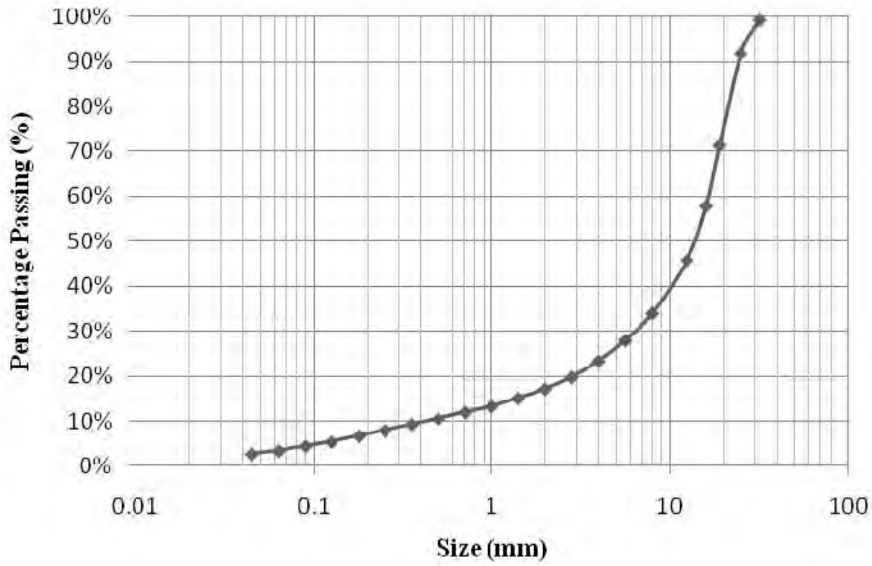


Figure 5 – Feed particle size distribution of Mesaba ore

Equipment

The following section describes the main pieces of test equipment used and the methodology used for calculating specific energy consumption.

High Pressure Grinding Roll

HPGR testing was conducted using a pilot-scale unit manufactured by Koeppern Machinery Australia. The pilot unit is custom-made for obtaining design information for sizing and selection of industrial-scale units. Table 1 summarizes the technical data provided by Koeppern for the machine. Experimental data was recorded every 200 ms through the programmable logic controller (PLC) data logger and downloaded to a laptop. The computer system measures time, roller gap (left and right), pressing force (left and right), and power draw. A picture of the HPGR pilot unit is shown in Figure 6.

Table 1 – HPGR machine specifics

Roller Diameter	750 mm
Roller Width	220 mm
Press Drive	Dual Output Shaft Gear Reducer
Feed System	Gravity
Wear Surface	Hexadur® WTII
Installed Power	200 kW
Maximum Pressing Force	1800 kN
Maximum Specific Pressing Force	8.5 N/mm^2
Variable Speed Drive	up to 40 RPM (1.55 m/s)

Figure 6 – Pilot-scale HPGR installation at the NBK Institute of Mining

A pilot test with the HPGR comprises the crushing of one 45-gallon drum of material (~375 kg). The material is loaded into a feed hopper with the use of an overhead crane and drum tipper. Once the machine conditions are stabilized, the slide gate of the feed hopper is opened and the test begins. The material flows with the aid of gravity through the HPGR rollers and drops on to the product conveyor located below the rolls. Once the test is complete, specific throughput and specific energy consumption are determined for the test using the power draw off the main motor and the throughput recorded during the stable operating period.

Since the HPGR does not grind uniformly across the roller width, a splitter gate is installed on the end of the product conveyor to separate the product into centre, edge and waste streams. The centre portion is finer than the edge portion and, during testing, a particle size distribution is performed on each to accurately predict size distributions for full-scale operations. For square rollers found in industrial units, where roll diameter is equal to roll width, the proportion of centre and edge product is observed to be approximately 85% centre and 15% edge. All of the HPGR product size distributions presented in this paper account for this through scaling of centre and edge size distributions at a ratio of 85:15. Material collected during unstable operation, initial response, and material run-out periods, was designated as waste material and only material which had been crushed during stable press operation was collected for analysis.

Horizontal Stirred Mill

Stirred mill testing was carried out using Netzsch's M20 horizontal stirred mill. The mill has a capacity of 20 litres and is installed with an 18.6 kW motor. The mill was fitted with the current IsaMill^{TM} internal grinding configuration and a Variable Frequency Drive (VFD). The installation of the VFD allowed for direct readings of mill power and mill speed. To monitor the mill, sensors were installed for feed pressure, and both feed and product temperature. A PLC interface and data logger was also installed to control the mill settings and record all important mill parameters during testing. The mill configuration, including grinding disc design, was based on recommendations from Xstrata Technology and allows for the ability to scale-up results to what would be expected for industrial IsaMills^{TM}.

A Watson-Marlow and Bredel SPX 25 hose pump and corresponding VFD were used to feed the mill. The pump has a capacity of 25 L/min and was designed to handle viscous slurries. The installation of a VFD for the 1.5 kW pump motor allowed for accurate monitoring and control of mill flow rate. The corresponding mixing system was comprised of two 180 L-capacity mix tanks with corresponding 250 W variable speed agitators and was designed to mix slurries at upwards of 60% solids with a particle top size as coarse as 1.2 mm. The piping system for the circuit was set up so that each mix tank could easily be switched from product to feed with minimal delay. The final setup is shown in Figure 7.

Figure 7 – M20 stirred mill installation at the NBK Institute of Mining

For testing of the stirred mill energy requirements, a graph of specific energy consumption and p80 grind size was generated. This graph, known as a signature plot, is the common method used in the industry for accurate sizing of full-scale IsaMills^{TM} and has a scale-up ratio of 1:1 (Gao, Weller, & Allum, 1999). The procedure entails running the material through the mill a select number of times and recording the energy requirements and product size after each pass. The passes are carried out consecutively in order to observe the energy consumption as the size of the product decreases. The results provide a series of points plotted on a log – log graph that shows the relationship between energy input and product size (p80).

Particle sizing for this work was done using a Malvern Mastersizer 2000. This laser sizing equipment utilizes the principle that grains of different sizes diffract light at different angles; a decrease in size produces an increase in diffracted angle. This equipment has become the standard for analyzing size ranges unrealistic for screening (Larson, Morrison, & Pietersen, 2008).

Vibrating Screen

All screening work carried out for HPGR closed-circuit testing was performed using a SWECO® Vibro-Energy® Separator. This vibrating screen, model ZS40, is equipped with a 373 W motor and a counterweight system to produce both vertical and horizontal vibrating motion. The screener is equipped with 1 m diameter wire mesh screens.

Bond Test Ball Mill

Energy requirements for ball mill grinding were determined using Bond ball mill work indices for cone crusher and HPGR product. Representative samples were screened at minus 3.35 mm and processed through a standard Bond ball mill measuring 305 mm in length and 305 mm in diameter, with a 285 ball charge weighing 20 125 g. Testing was carried out using the standard Bond Ball Mill Grindability Test procedure developed by Bond (1961). For the crushing work index, insufficiently sized material was available to perform impact testing. Therefore, a traditional approach was taken and the Bond work index was used. The resulting indices were then used with the Bond equation to calculate specific energy consumption for both crushing and grinding.

RESULTS

Cone Crusher / Ball Mill Circuit

The specific energy consumption of comminution for the circuit was determined with a flowsheet developed using JK SimMet® software. The circuit was designed for 250 tph capacity and equipment was sized based on a product p80 of 75 µm. Table 2 summarizes the equipment sized for the circuit.

Table 2 – Equipment selection for the cone crusher / ball mill circuit

Cone Crusher	
Closed Side Setting	2.8 mm
Re-circulating Load	~30%
Product Screen	
Aperture Size	4 mm
Ball Mill	
Diameter	5 m
Length	10 m
Critical Speed	70%
Media Charge	40%
Media Top Size	35 mm
Re-circulating Load	~250%
Hydrocyclones	
Quantity	6
Cyclone Diameter	420 mm
Inlet Diameter	175 mm
Vortex Finder Diameter	150 mm
Apex (Spigot) Diameter	113 mm
Length	500 mm
Cone Angle	20°

Simulation of the flowsheet predicted that the appropriate transfer size between the crushing circuit and the ball mill circuit would be 80% passing 2.12 mm. To calculate the overall specific energy consumption for the cone crusher and ball mill, work indices were determined for the material. In the case of the cone crusher, no material was available for the size requirements, 50-75 mm, necessary to perform impact testing. Therefore, a traditional approach was taken and the Bond ball mill work index was used. Locked-cycle testing was performed using two sieve sizes (106 µm and 150 µm), to allow for the comparison of different product sizes. The results for the work indices of the circuit are shown in Table 3.

Table 3 – Bond work indices for the cone crusher / ball mill circuit

Locked Cycle Screen Size (µm)	Feed f80 (µm)	Product p80 (µm)	Bond Work Index (kWh/t)
150	2,133	119	15.9
106	2,241	80	16.5

Using the Bond work indices and the transfer size determined from flowsheet simulation, the theoretical energy requirements for the circuit are calculated using the Bond equation (Bond, 1961). Calculation of the cone crusher energy requirements will use the Bond work index at a sieve size of 150 µm. This coarser screen size provides a lower estimate for the energy requirements of a cone crusher and provides a best-case scenario for the crushing circuit. Calculation of the ball mill energy requirements will use the Bond work index at a sieve size of 106 µm. The final product for the circuit was set at a p80 of 75 µm and the Bond work index, at a sieve size of 106 µm, better reflects the energy requirements to grind to this finer particle size. A summary of the resulting energy requirements for the circuit are shown in Table 4.

Table 4 – Summary of the cone crusher / ball mill circuit energy requirements

Unit Operation	Feed f80 (mm)	Product p80 (mm)	Specific Energy Consumption (kWh/t)
Cone Crusher	21	2.12	2.36
Ball Mill	2.12	0.075	15.47
TOTAL			17.83

HPGR / Ball Mill Circuit

The HPGR Circuit

To evaluate the energy requirements for the HPGR / ball mill circuit, HPGR pilot-scale testing was performed to determine the ideal specific pressing force for the Mesaba material and assess how operating the HPGR in closed circuit with a 4 mm screen would affect comminution performance. Four initial tests were done to determine the effect of specific pressing force. Pressures of 2 N/mm^2, 3 N/mm^2, 4 N/mm^2, and 5 N/mm^2 were chosen and comparisons were made with respect to product size, net specific energy consumption, and specific throughput (m-dot). All tests were performed at a roller speed of 0.75 m/s and feed moisture content by weight of 2.5%.

The comparison of product particle size at different specific pressing forces is shown in Figure 8. As the pressing force increased, both the p50 and p80 decreased, although the effect on p80 was more pronounced than the effect on p50. This result is due to an increased force being exerted on the particles as they flow through the rolls. An increased force promotes increased breakage and the effect is more pronounced on larger sized particles, hence the steeper trend for the p80.

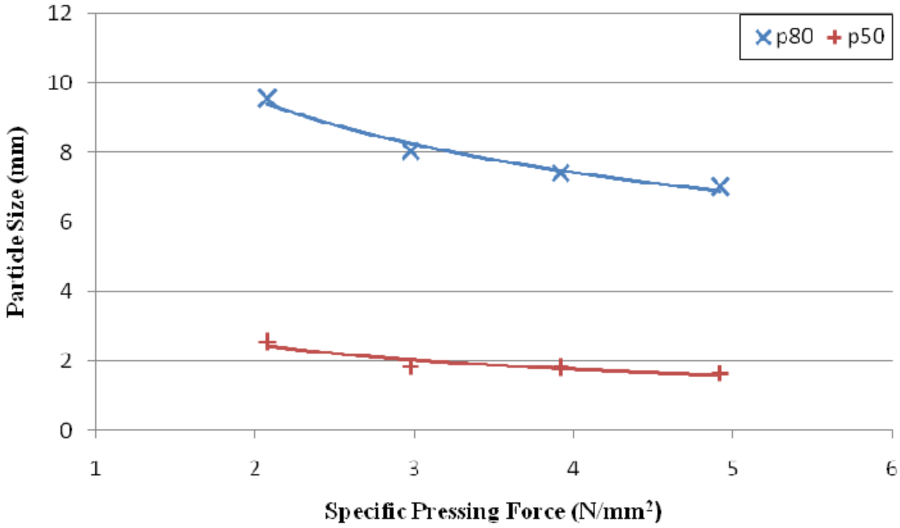


Figure 8 – Comparison of specific pressing force and product size

The comparison of specific throughput (m-dot) at different specific pressing forces is shown in Figure 9. As the pressing force increased, the specific throughput decreased. This trend is due to the gap between the rollers decreasing slightly with increasing pressing forces, resulting in the reduction of throughput in the machine.

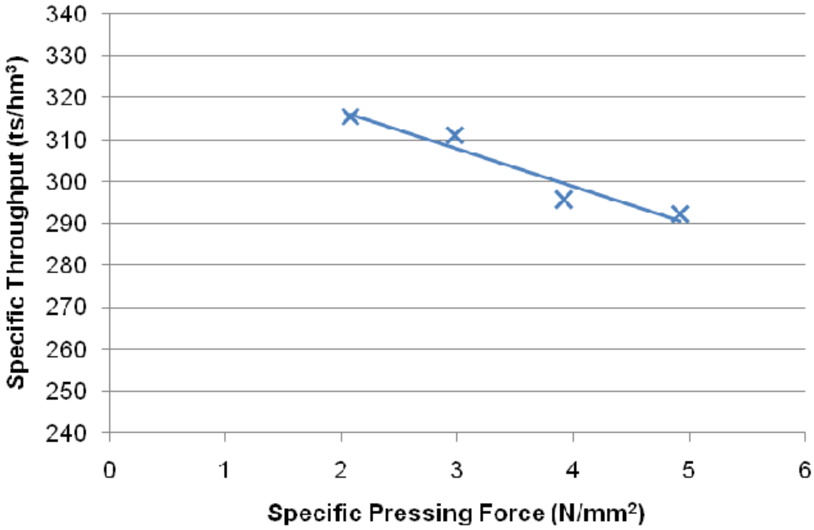


Figure 9 – Comparison of specific pressing force and specific throughput

The comparison of specific energy consumption at different specific pressing forces is summarized in Figure 10. As the pressing force increased, the energy consumption also increased. This is typical of the process because more energy is being transmitted into the material at higher pressures.

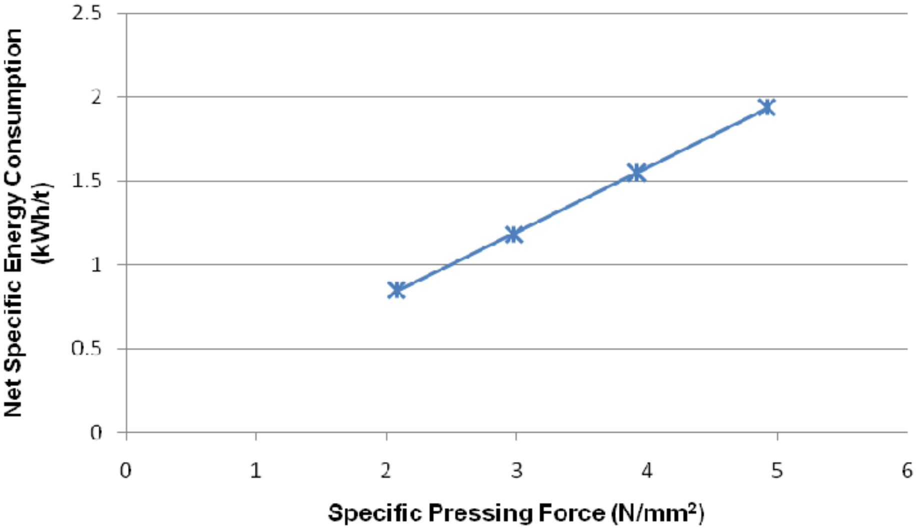


Figure 10 – Comparison of specific pressing force and specific energy consumption

A specific pressing force of 4 N/mm^2 was selected for the remainder of pilot-scale testing. The results indicated that a pressing force of 4 N/mm^2 provided a fine balance between energy consumption and size reduction without a significant change in specific throughput.

To test the effect of closed-circuit operation, locked-cycle testing was conducted using a 4 mm screen. Material was processed through the HPGR at 4 N/mm^2 and the product screened at 4 mm using the SWECO® vibrating screen. Using the product size distributions from testing, the percentage of plus 4 mm was calculated (at 90% screening efficiency) and then combined with fresh feed and re-run through the HPGR. This process was repeated two more times to simulate closed-circuit operation. Results were generated to determine size reduction, specific throughput, and specific energy consumption for each cycle. The resulting product size for each cycle is shown in Figure 11. The chart shows that the introduction of a re-circulating load decreased the product size and began to stabilize by cycle four.

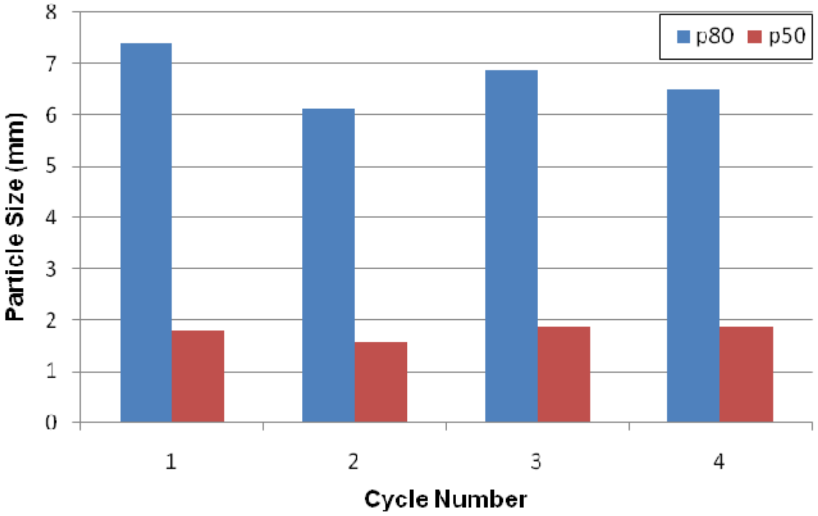


Figure 11 – Product size for closed-circuit testing

The effect on specific throughput for closed-circuit testing is shown in Figure 12. Closed-circuit operation had little effect on specific throughput. The variation between each cycle can probably be attributed to testing error.

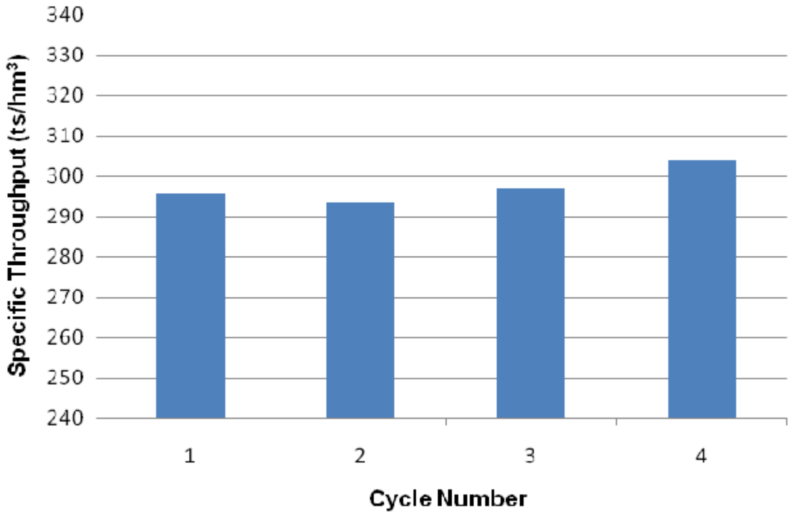


Figure 12 – Specific throughput for closed-circuit testing

The results for specific energy consumption are displayed in Figure 13. The closed-circuit testing had little to no effect on specific energy consumption.

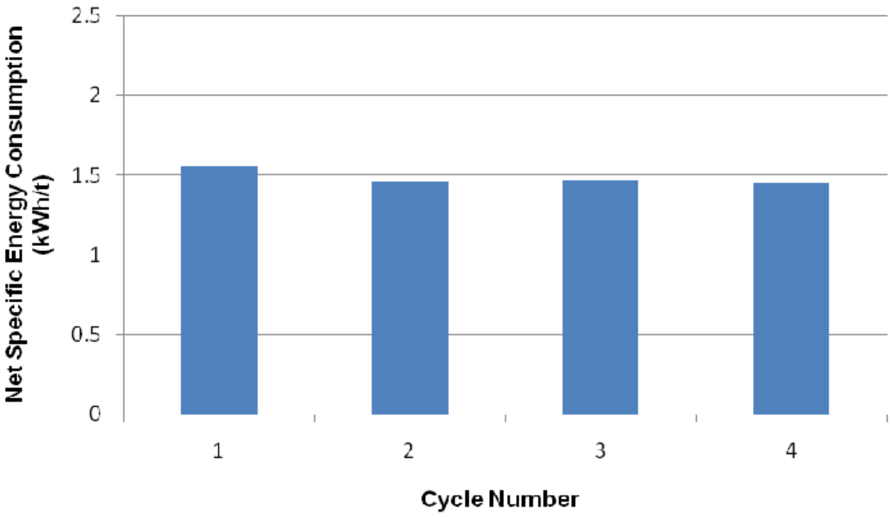


Figure 13 – Specific energy consumption for closed-circuit testing

Results from the last cycle of testing are summarized in Table 5. These results will be used for energy calculations, as well as the experimental data required for model fitting with JK SimMet®.

Table 5 – Results for the last cycle of closed-circuit testing

f80	21.77
f50	13.38 mm
p80	6.61 mm
p50	1.91 mm
Percentage Passing -4 mm	67.4%
Net Specific Energy Consumption	1.45 kWh/t
(-4 mm) Net Specific Energy Consumption	2.15 kWh/t
Specific Throughput	304 ts/hm^3

Flowsheet Development and Circuit Energy Calculations

Using the results from the last cycle of HPGR closed-circuit testing, model fitting of an HPGR circuit was performed using JK SimMet®. The T10H and HPGR power coefficient model parameters were fitted using the model fit tool in JK SimMet®. This tool uses an iterative function to fit experimental data to simulated data by adjusting model parameters until a correlation can be achieved. The T10h and HPGR power coefficient parameters relate to the breakage mechanisms in the compression zone of the HPGR and the product size for closed-circuit testing was used as the experimental data. The procedure used for calibrating the HPGR model was outlined by Daniel and Morrell (2004). The resulting model fit was able to simulate a product size distribution similar to the one generated experimentally. Once an HPGR model was calibrated for use with Mesaba ore, a flowsheet was designed for 250 tph capacity with a product p80 of 75 µm. The equipment sized for the flowsheet is summarized in Table 6.

Table 6 – Equipment selection for the HPGR / ball mill circuit

High Pressure Grinding Roll	
Roller Diameter	1,200 mm
Roller Width	1,000 mm
Re-circulating Load	~45%
Product Screen	
Aperture Size	4 mm
Ball Mill	
Diameter	5 m
Length	9.1 m
Critical Speed	70%
Media Charge	40%
Media Top Size	27.5 mm
Re-circulating Load	~250%
Hydrocyclones	
Quantity	7
Cyclone Diameter	350 mm
Inlet Diameter	175 mm
Vortex Finder Diameter	150 mm
Apex (Spigot) Diameter	113 mm
Length	450 mm
Cone Angle	20°

Simulation of the flowsheet predicted that the appropriate transfer size between the HPGR circuit and the ball mill circuit would be 80% passing 1.6 mm. Several publications have indicated that HPGR comminution and the presence of micro-cracks in the product, results in a decrease in the Bond work index when compared with conventionally crushed product (Daniel, 2007a; Muranda, 2009; Rule, Smit, Cope, & Humphries, 2008). To confirm this advantage, Bond ball mill work indices were determined for HPGR product at different specific pressing forces. Samples were taken from HPGR centre product and screened at 3.35 mm with no additional crushing. As with Bond work indices for cone crusher product, two separate screen sizes (106 µm and 150 µm) were tested to allow for the comparison of different product sizes. The results, including cone crusher product for comparison, are summarized in Table 7.

Table 7 – Summary of Bond work indices

Locked Cycle Screen Size (µm)	Feed Preparation Method	Feed f80 (µm)	Product p80 (µm)	Bond Work Index (kWh/t)
150	Cone Crusher	2,133	119	15.9
	HPGR - 3 N/mm^2	1,854	125	14.5
	HPGR - 4 N/mm^2	1,849	124	14.5
	HPGR - 5 N/mm^2	1,497	118	13.3
106	Cone Crusher	2,241	80	16.5
	HPGR - 3 N/mm^2	1,765	81	15.8
	HPGR - 4 N/mm^2	1,764	81	15.7
	HPGR - 5 N/mm^2	1,682	79	15.7

A reduction in Bond work index was achieved between cone crusher and HPGR product. However, the reduction went from 8.8% to 4.8% with a decrease in screen size. Results also showed a reduction in Bond work index with increasing specific pressing force, although this effect may be attributed to the finer feed size. The reduction in screen size may have caused a decrease in the effectiveness of product micro-cracking and a relatively higher amount of energy was required to produce the finer product. Using the specific energy results obtained from Table 5, coupled with the Bond work index for 4 N/mm^2 at a screen size of 106 µm and the transfer size determined from JK SimMet® simulation, the specific energy requirements for the HPGR / ball mill circuit can be summarized in Table 8.

Table 8 – Summary of the HPGR / ball mill energy requirements

Unit Operation	Feed f80 (mm)	Product p80 (mm)	Specific Energy Consumption (kWh/t)
HPGR	21	1.6	2.15
Ball Mill	1.6	0.075	14.2
TOTAL			16.35

HPGR / Stirred Mill Circuit

The HPGR / stirred mill circuit required considerably more pilot-scale testing than the previous two circuits, since very few operating examples could be found in the literature. The results of pilot-scale testing determined the appropriate layout for the two-stage HPGR circuit and provided the corresponding specific energy consumption for circuit energy summation.

The HPGR Circuit

Pilot-scale testing was conducted to produce suitable data for the HPGR section of the HPGR / stirred mill circuit. Since size reduction is limited with one stage of HPGR comminution (Daniel, 2007b), design of an HPGR / stirred mill circuit required at least two consecutive stages of HPGR comminution to produce a particle size acceptable for stirred milling. With the transfer size between the second-stage HPGR and the stirred mill established by Drozdiak et al. (2010), work was done to determine the appropriate transfer size between each stage of HPGR crushing. Two options were examined to find the appropriate circuit layout. In Circuit A, the first-stage HPGR was placed in closed circuit with a 4 mm screen, while in Circuit B, the first stage remained open circuit and the second stage accepted product directly from stage one.

For Circuit A, closed-circuit testing product from the HPGR / ball mill circuit was processed again through the HPGR at the same roller speed (0.75 m/s) and specific pressing force (4 N/mm^2). The use of the same specific pressing force for second-stage HPGR crushing stems from work performed by Rule et al. (2008), in which they found that no difference was observed when changing the specific pressing force in the second stage of two-stage HPGR crushing. For Circuit B, fresh feed was processed through two consecutive stages of HPGR comminution using the same roller speed and specific pressing force as Circuit A. The results for both options are summarized in Table 9. The size distributions for Circuits A and B are presented in Figure 14 and Figure 15, respectively.

Table 9 – Summary of results for the first-stage HPGR operating in open (Circuit A) and closed (Circuit B) circuit

		Circuit A	Circuit B
HPGR Stage 1	f80	21.77 mm	21.54 mm
	f50	13.38 mm	13.7 mm
	HPGR p80	6.61 mm	7.68 mm
	HPGR p50	1.91 mm	1.88 mm
	Circuit p80	1.86 mm	7.68 mm
	Circuit p50	489 µm	1.88 mm
	Circuit Reduction Ratio	11.7	2.8
	Net Specific Energy Consumption	1.45 kWh/t	1.54 kWh/t
	Percentage Passing 4 mm	67.4%	
	(-4 mm) Net Specific Energy Consumption	2.15 kWh/t	
	Specific Throughput	304 ts/hm^3	307 ts/hm^3
HPGR Stage 2	f80	1.86 mm	7.68 mm
	f50	489 µm	1.88 mm
	HPGR p80	1.12 mm	2.79 mm
	HPGR p50	222 µm	462 µm
	Circuit p80	332 µm	339 µm
	Circuit p50	124 µm	142 µm
	Circuit Reduction Ratio	5.6	22.6
	Net Specific Energy Consumption	1.2 kWh/t	1.23 kWh/t
	Percentage Passing 0.71 mm	71.3%	56.5%
	(-0.71 mm) Net Specific Energy Consumption	1.68 kWh/t	2.18 kWh/t
	Specific Throughput	235.81 ts/hm^3	311 ts/hm^3
	TOTAL SPECIFIC ENERGY CONSUMPTION	**3.83 kWh/t**	**3.72 kWh/t**

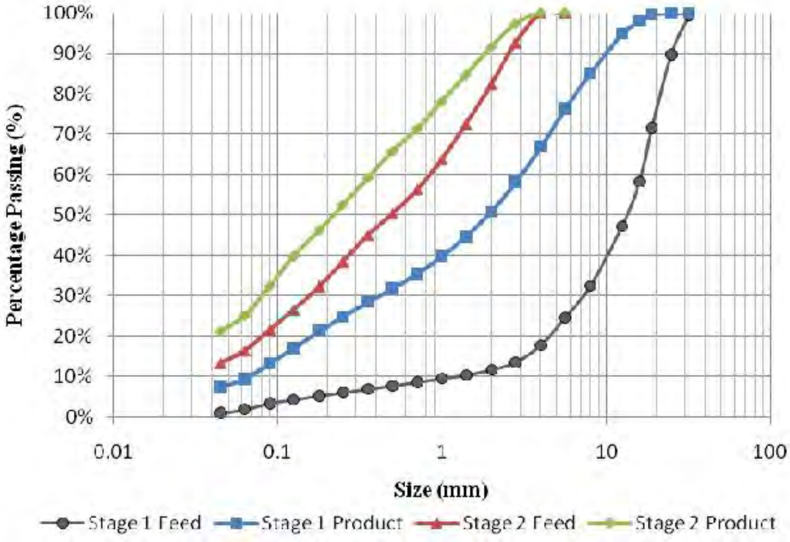


Figure 14 – Particle Size Distributions for Circuit A

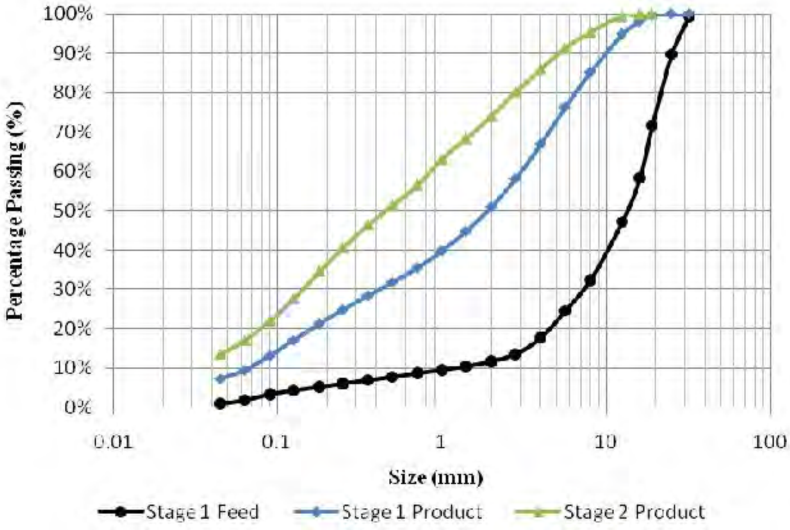


Figure 15 – Particle Size Distributions for Circuit B

Operating the first stage of HPGR crushing in open circuit required less energy compared with operating in closed circuit with a screen. If looked at strictly from an energy perspective, very little difference is gained choosing one circuit over the other, but if design and operating factors are considered, the choice of operating the first stage in open circuit becomes the better option. The ability to operate the circuit without a screen allows for the elimination of extra auxiliary equipment such as screens and conveyors, while the absence of an additional stage of wet screening would help to reduce the adverse effects that increased moisture content would have on HPGR performance (Fuerstenau & Abouzeid, 2007). Although the increased re-circulating load resulting in the second stage would require an increase in tonnage and machine size, this would be countered by the decreased machine size required for stage one. Overall, the reduced complexity offered by open circuit configuration led to us selecting this configuration for further testing.

Once the open circuit configuration was selected for stage one, additional pilot-scale testing was performed to evaluate how comminution performance would be affected by operating the second stage in closed circuit with a 710 µm screen. Testing was conducted in a similar manner to the locked cycle evaluation method used for the HPGR / ball mill circuit. Product from Circuit B was screened at 710 µm and a calculated split of oversize was mixed with fresh product from stage one and processed through the HPGR. This procedure was repeated two more times in order to simulate closed-circuit operation. The resulting product size for each cycle is shown in Figure 16. The product size increased slightly with the introduction of a re-circulating load. This is in contrast to the results for the HPGR / ball mill circuit, where the introduction of a re-circulating load caused a decrease in product size. This increase may have been the result of a finer re-circulating load reducing the breakage within the compressive bed.

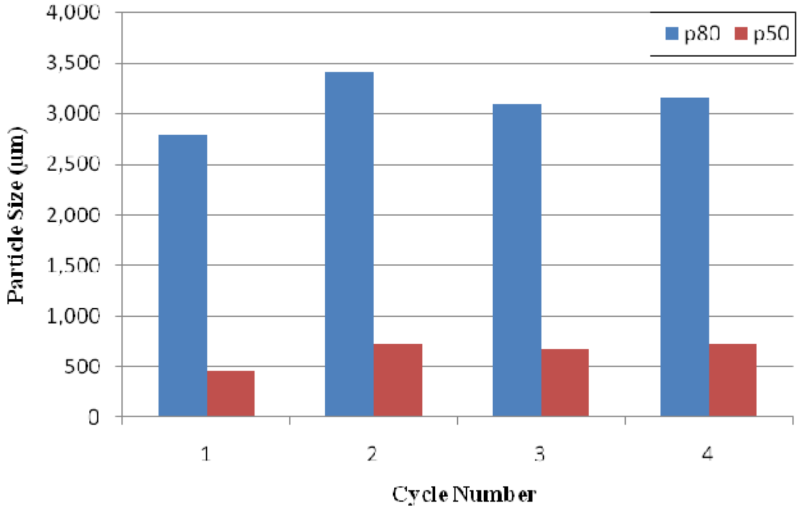


Figure 16 – Product size for second stage closed circuit testing

The results for the effect of closed-circuit operation on specific throughput are displayed in Figure 17. The introduction of a re-circulating load had no effect on specific throughput.

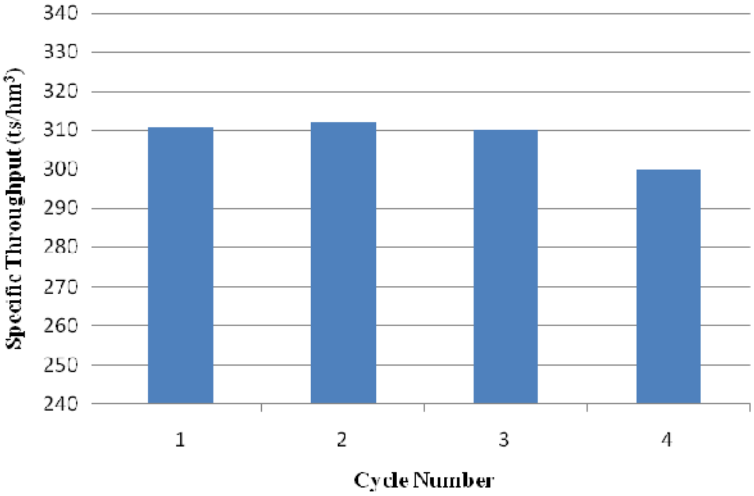


Figure 17 – Specific throughput for second stage closed circuit testing

The results for the effect of closed-circuit operation on specific energy consumption are summarized in Figure 18. As with specific throughput, the introduction of a re-circulating load had no effect on specific energy consumption.

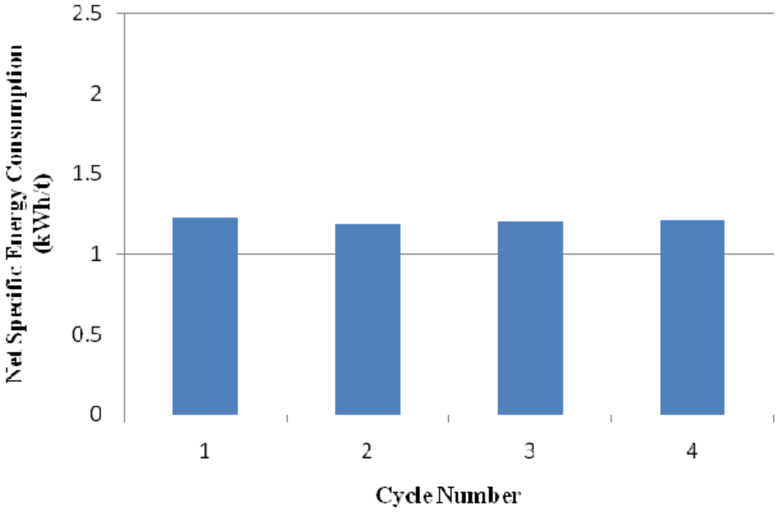


Figure 18 – Specific energy consumption for second stage closed circuit testing

To achieve efficient screening at 710 µm for an industrial operation, the practice of wet screening is necessary. Fuerstenau and Abouzeid (2007) found that the introduction of moisture to an HPGR circuit leads to adverse effects on throughput and energy consumption. The effect of moisture on second-stage HPGR crushing was tested using product from the final closed-circuit cycle. We wet screened the sample over a 710 µm screen to determine the potential moisture content for oversize in a closed-circuit operation. The saturated oversize, with a measured moisture content of 10.5%, was then used to run an additional closed-circuit cycle. A summary of the results is presented in Table 10. To allow for a direct comparison of the effects of wet screening, the results from cycle four (dry) are presented as well. As expected, the results show an adverse effect on throughput and energy consumption, although the product size became considerably finer. The data generated for the wet screening cycle represents the worst-case scenario, and thus will be used for the energy calculations for the circuit.

Table 10 – Comparison of wet and dry screening for second stage closed circuit operation

	Dry Cycle	Wet Cycle
Feed Moisture Content	2.4%	5.8%
f80	5.69 mm	6.41 mm
f50	1.79 mm	1.95 mm
p80	3.16 mm	2.88 mm
p50	718 µm	523 µm
Percentage Passing -710µm	49.8%	54.8%
Net Specific Energy Consumption	1.45 kWh/t	1.96 kWh/t
(-710 µm) Net Specific Energy Consumption	2.91 kWh/t	3.58 kWh/t
Specific Throughput	304 ts/hm^3	232 ts/hm^3

The Stirred Mill Circuit

To determine the specific energy requirements for the stirred mill circuit, two signature plots were performed using the minus 710 µm undersize from the second stage of closed-circuit testing. We chose operating conditions to target a specific energy input of 7-9 kWh/t per pass through the mill. This energy input would create an evenly-spaced set of data points on the signature plot and allow for accurate prediction of energy requirements for coarse grind sizes. Table 11 summarizes the operating conditions used for testing and Figure 19 shows the resulting signature plots. Testing showed that an average specific energy consumption of 9.73 kWh/t was required to grind to a p80 of 75 µm. This value was selected to be used for energy calculations of the HPGR / stirred mill circuit.

Table 11 – Summary of stirred mill operating conditions

f80	340 µm
Feed Weight	100 kg
Percent Solids by Weight	57%
Percent Solids by Volume	31%
Flow Rate	20.4 L/min
Mill Speed	1,169 RPM
Grinding Media Top Size	6 mm
Grinding Media Type	Ceramic (Manufactured by Cenotec)

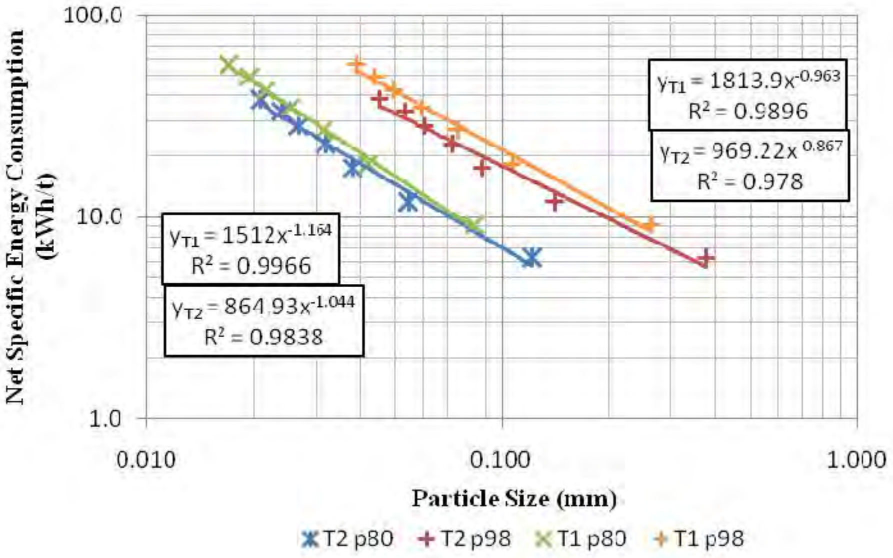


Figure 19 – Stirred mill signature plot results

The size measurements used to generate the signature plots in Figure 19 were performed using a Malvern Mastersizer 2000. For a comparison, pass one product was also sized using screens. Since Malvern sizing is based on volume, while screening is based on weight, results will not be identical. All other testwork performed for this flowsheet relied on size results from screening. Therefore, a comparison should be made. Malvern and screening comparisons for T1 and T2 are shown in Figure 20 and Figure 21, respectively. The screening results indicated a finer product than the Malvern results. These results show that the signature plots generated using Malvern sizing, can be considered a conservative estimate for energy consumption, since size results may have been finer using screens. Unfortunately, screening is impractical below 38 µm (the product size after pass two), so Malvern sizing was used for all stirred mill products in order to remain consistent.

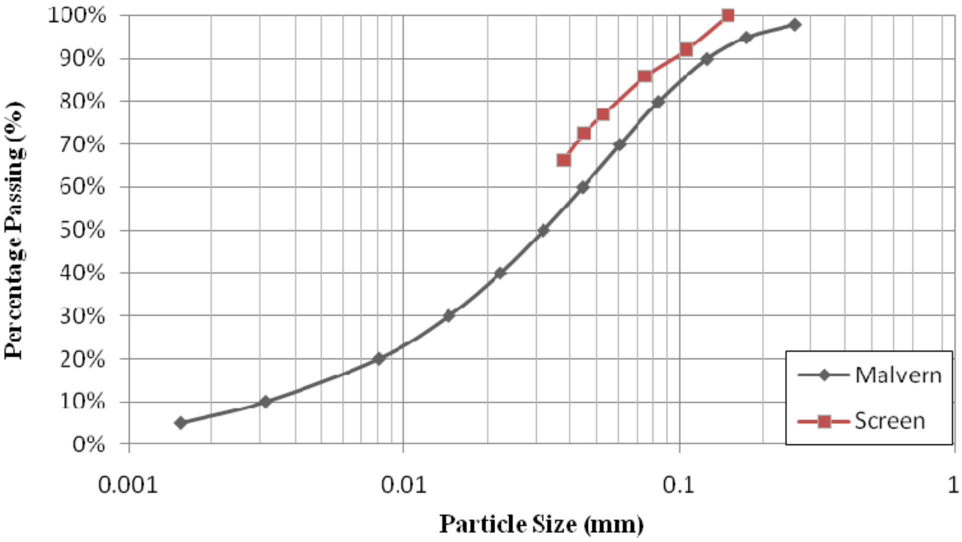


Figure 20 – Malvern and screen comparison for T1

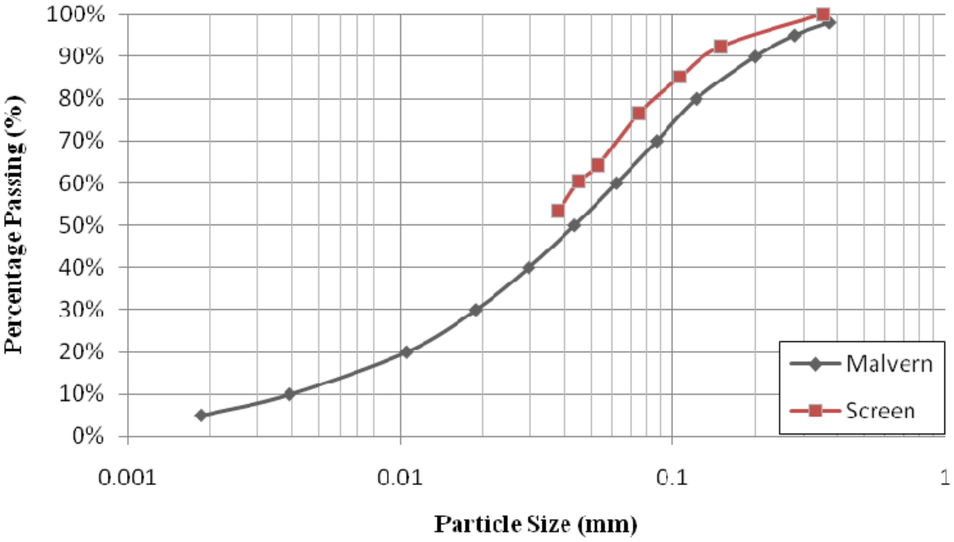


Figure 21 – Malvern and screen comparison for T2

Circuit Energy Summary

The resulting specific energy requirements obtained from pilot-scale testing are summarized in Table 12. For a comparison, results from dry and wet screening for second-stage HPGR are included. Results show that the implementation of wet screening would result in a 4.7% increase in specific energy consumption for the circuit.

Table 12 – Summary of the HPGR / stirred mill energy requirements

Unit Operation	Feed f80 (mm)	Product p80 (mm)	Specific Energy Consumption with Dry Screening (kWh/t)	Specific Energy Consumption with Wet Screening (kWh/t)
First Stage HPGR	21	7.68	1.54	1.54
Second Stage HPGR	7.68	0.34	2.91	3.58
Stirred Mill	0.34	0.075	9.73	9.73
TOTAL			14.18	14.85

DISCUSSION

Using the energy requirements determined for each circuit, a bar graph is generated to summarize the specific energy consumption for each stage of comminution (refer to Figure 22). The graph shows that the HPGR / stirred mill circuit required the lowest specific energy consumption and achieved a reduction of 9.2% and 16.7% over the HPGR / ball mill and cone crusher / ball mill circuits, respectively.

The results presented in this paper were obtained from pilot-scale testing on a single test for each operating variable. Since pilot-scale testing required a significant quantity of material per test, 350 kg for the HPGR and 100 kg for the stirred mill, the reproducibility and standard deviation could not be determined for each changing variable. However, using five homogenized drums, we did perform some repeatability testing using the pilot-scale HPGR. Results showed that specific energy consumption had a standard deviation of 0.0167 and specific throughput, a standard deviation of 11.43. For stirred mill testing, since only two signature plots were generated at similar conditions, the standard deviation could not be calculated and instead the median of 0.23 was considered. The energy figures associated with Bond grindability testing were found to have a standard deviation of 0.0548, when comparing the three results of HPGR product at a screen size of 106 µm. With these results, testing errors were calculated for each circuit at a 95% confidence interval. Table 13 summarizes the statistics related to each circuit energy result. The error values show that the HPGR / stirred mill circuit contained the most potential for a variation in reported results. With the inclusion of testing error, the HPGR / stirred mill circuit still required the lowest specific energy consumption for comminution. The testing error presented here can be considered only an approximation because the results are based on only a few tests.

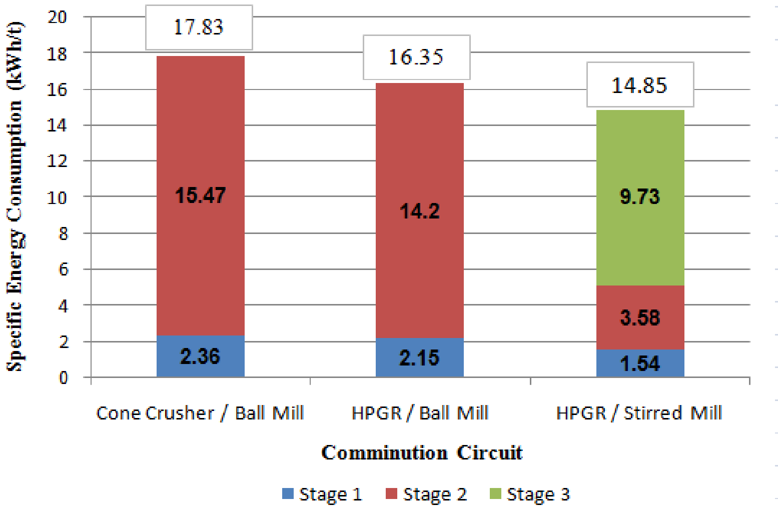


Figure 22 – Summary of specific energy consumption for each circuit

Table 13 – Statistics summary of circuit energy values

	Sample Set	Mean (kWh/t)	Standard Deviation	Standard Deviation of the Mean	95% Confidence Interval	Upper Limit	Lower Limit
Cone Crusher Specific Energy Value	3	2.36	0.0548	0.0316	0.0620	2.42	2.30
Ball Mill Specific Energy Value	3	15.47	0.0548	0.0316	0.0620	15.53	15.41
HPGR Energy Value	5	2.15	0.0167	0.00747	0.0146	2.16	2.14
Ball Mill Energy Value	3	14.2	0.0548	0.0316	0.0620	14.26	14.14
Stage 1 HPGR Energy Value	5	1.54	0.0167	0.00747	0.0146	1.55	1.53
Stage 2 HPGR Energy Value	5	3.58	0.0167	0.00747	0.0146	3.59	3.57
Stirred Mill Energy Value*	2	9.73	0.23	0.163	0.319	10.05	9.41

*Median used instead of standard deviation

Total Specific Energy Consumption With 95% Confidence Interval	Cone Crusher / Ball Mill Circuit	17.83 +/- 0.09
	HPGR / Ball Mill Circuit	16.35 +/- 0.06
	HPGR / Stirred Mill Circuit	14.85 +/- 0.32

The energy values determined in this paper did not take into account any auxiliary equipment for the circuits. Each circuit would require additional energy requirements for feeders, conveyors, screens, pumps and cyclones. Additional energy requirements for the cone crusher / ball mill circuit would result from screens and conveyors for the crushing circuit and pumps and cyclones for the ball mill circuit. For the HPGR / ball mill circuit, increased energy requirements would result from screens and conveyors in the HPGR circuit and pumps and cyclones in the ball mill circuit. The energy requirements for the HPGR / stirred mill circuit would increase with a feed conveyor for first-stage HPGR crushing, screens and conveyors for second-stage HPGR crushing, and pumps for the stirred mill circuit. The extra energy required for the increased quantity of conveyors would be counteracted by the reduction in energy related to an open-circuit grinding configuration. The energy requirements for a de-agglomerator were not necessary for Mesaba ore, due to a low flake competency. However, this energy requirement would need to be considered for an ore that produced more competent flake. Overall, the increased energy requirements for all three circuits, when incorporating auxiliary equipment, should not affect the results significantly.

A preliminary design of an HPGR / stirred mill circuit is shown in Figure 23. The flowsheet design incorporates a closed-circuit crusher prior to HPGR processing. This step prevents oversized material from entering the HPGR circuit, ensuring the prevention of single particle breakage and damage of the metal studs. The design of the secondary crushing circuit is similar to the configuration installed at Cerro Verde (Vanderbeek et al., 2006). A cone crusher is placed in a reversed closed-circuit arrangement, which would reduce throughput and improve crushing efficiency for the cone crusher by screening out fine particles from the feed. A metal detector is placed on a conveyor prior to entering the HPGR circuit, to protect the HPGR roller lining from tramp metal. The two stages of HPGR comminution are designed with the first stage operating in open circuit to reduce materials handling requirement and decrease the amount of water (wet screening) entering the circuit. To achieve efficient screening for the second stage, inclusion of both de-agglomeration and wet screening steps were incorporated to handle competent flake. Alternatively, the possibility arises to implement an air classifier instead of wet screening for fine size classification, eliminating the addition of moisture inherent with a wet screening circuit. The use of an air classifier for fines production in an HPGR circuit has been shown to operate effectively in the cement industry (Aydogan, Ergun, & Benzer, 2006) and with careful design could be implemented in hard-rock circuits. To ensure optimal feed density for efficient stirred milling, undersize from the HPGR circuit would be fed to a mixing tank, where water would be added to control pulp density. Larson, Morrison, Shi, and Young (2008) stated that ideal operating conditions for stirred milling require a solids content of 40-50%, depending on viscosity of the slurry. This could be achieved with a simple process control loop installed between the water addition tank and the IsaMill^{TM} feed tank. For operation of the IsaMill^{TM} circuit, the simplicity available with the dynamic classifier and open-circuit configuration would eliminate any need for a recycle stream.

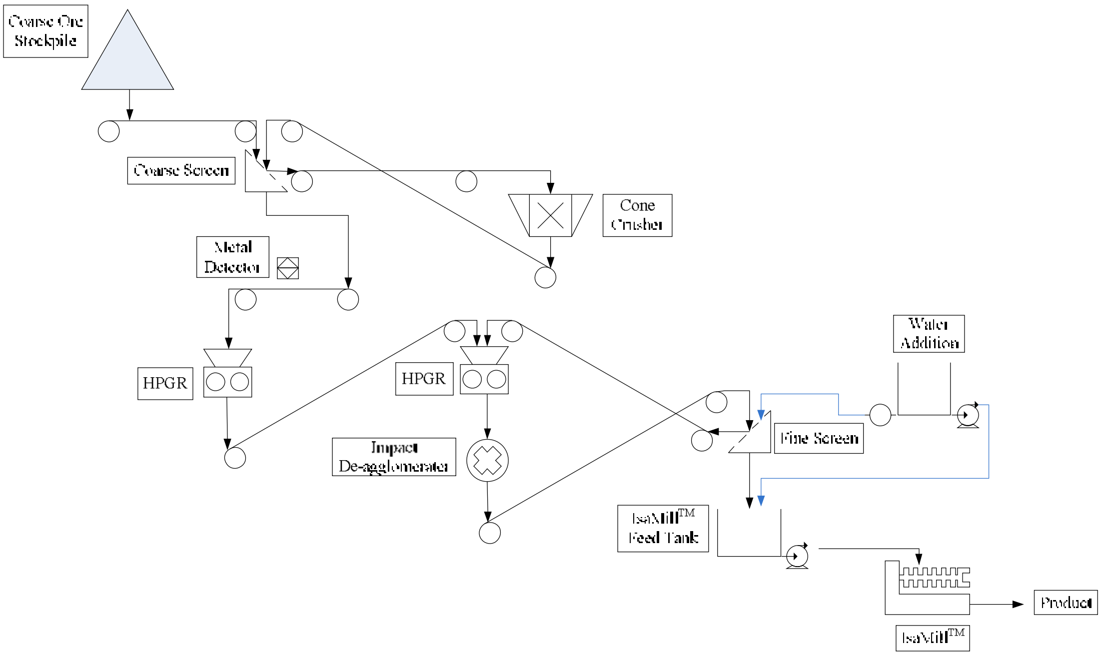


Figure 23 – Proposed layout for an HPGR / stirred mill circuit

CONCLUSIONS

In this paper we demonstrated that an HPGR / stirred mill circuit is technically feasible and that, based solely on the specific energy requirements for comminution, the novel circuit could achieve a reduction of 9.2% and 16.7% over the HPGR / ball mill and cone crusher / ball mill circuits, respectively. Although this paper did not take into account the energy requirements for auxiliary equipment, the findings documented here provide an incentive to further explore the concept of an HPGR / stirred mill circuit. With the mining industry requiring a reduction in the energy demand associated with its processes, the design and implementation of novel flowsheets, such as the HPGR / stirred mill circuit presented here, should help keep the mining industry sustainable for future generations.

ACKNOWLEDGEMENTS

We would like to thank Xstrata Technology for providing technical expertise and funding in support of this project, as well as Teck Limited for supplying Mesaba ore for testing.

REFERENCES

Abouzeid, A. M., & Fuerstenau, D. W. (2009). Grinding of mineral mixtures in high-pressure grinding rolls. *International Journal of Mineral Processing*, *93*(1), 59-65.

Aydogan, N. A., Ergun, L., & Benzer, H. (2006). High pressure grinding rolls (HPGR) applications in the cement industry. *Minerals Engineering*, *19*(2), 130-139.

Ayers, C., Knopjes, L., & Rule, C. M. (2008). Coarser feed applications of MIG IsaMilling. *MEI Comminution'08 Conference,* Falmouth, Cornwall, UK.

Bond, F. C. (1961). Crushing and grinding calculations part I and II. *British Chemical Engineering, 6*(6,8)

Burford, B. D., & Clark, L. W. (2007). IsaMill™ technology used in efficient grinding circuits. *VI11 International Conference on Non-Ferrous Ore Processing,* Wroclaw, Poland. 21-23.

Daniel, M. (2007a). *Energy efficient mineral liberation using HPGR technology.* (Doctor of Philosophy, The University of Queensland).

Daniel, M. (2007b). Triple pass HPGR circuit concept. *CSRP'07 Delivering Sustainable Solutions to the Minerals and Metals Industries,* Melbourne, Australia.

Daniel, M., & Morrell, S., (2004) HPGR model verification and scale-up. *Minerals Engineering*, *17*(11-12), 1149-1161.

Drozdiak, J. A., Nadolski, S., Bamber, A., Klein, B., & Wilson, S. (2010). A comparison of the energy requirements of an HPGR / stirred mill circuit and conventional grinding circuits for the comminution of mesaba Ore . *Proceedings of the 42nd Annual Conference of the Canadian Mineral Processors,* Ottawa, ON, Canada.

Dunne, R. (2006). HPGR - the journey from soft to competent and abrasive. *International Autogenous and Semiautogenous Grinding Technology 2006,* Vancouver, BC, Canada. *4* 190-205.

Fuerstenau, D. W., & Abouzeid, A. M. (2002). The energy efficiency of ball milling in comminution. *International Journal of Mineral Processing, 67*(1-4), 161-185.

Fuerstenau, D. W., & Abouzeid, A. M. (2007). Role of feed moisture in high-pressure roll mill comminution. *International Journal of Mineral Processing, 82*(4), 203-210.

Fuerstenau, D. W., & Kapur, P. C. (1995). Newer energy-efficient approach to particle production by comminution. *Powder Technology, 82*(1), 51-51.

Gao, M., Weller, K. R., & Allum, P. (1999). Scaling-up horizontal stirred mills from a 4-litre test mill to a 4000-litre "IsaMill". *Powder Technology Symposium*, Pennsylvania State University, Pennsylvania, USA.

Gruendken, A., Matthies, E., & van der Meer, F. P. (2010). Flowsheet considerations for optimal use of high pressure grinding rolls. *Minerals Engineering, 23*(9), 663-669.

Infomine. (2001). *Mesaba property profile.* Retrieved February 17, 2011, from http://www.infomine.com/companies-properties/reports/propertyreport.aspx?pid=14754

Johnson, N. W., Gao, M., Young, M. F., & Cronin, B. (1998). Application of the ISAMILL (a horizontal stirred mill) to the lead-zinc concentrator (mount isa mines ltd) and the mining cycle. In *Proceedings of the 1998 annual conference on mining cycle, AusIMM, apr 19 - 23 1998* (pp. 291-297). Mount Isa, Australia.

Klymowsky, R., Patzelt, N., Knecht, J., & Burchardt, E. (2006). An overview of HPGR technology. *International Autogenous and Semiautogenous Grinding Technology 2006,* Vancouver, BC, Canada. *4* 11-26.

Larson, M., Morrison, R., & Pietersen, K. (2008). Measurement and comparison of very fine size distributions. *MEI Comminution'08 Conference,* Falmouth, Cornwall, UK.

Larson, M., Morrison, R., Shi, F. N., & Young, M. F. (2008). Improving grinding efficiency with the IsaMill. *CSRP'08 Delivering Sustainable Solutions to the Minerals and Metals Industries*, Brisbane, QLD, Australia.
Minnesota Geological Survey. (2010). *Mineral potential and geology of the duluth complex.* Retrieved February 17, 2011, from http://www.geo.umn.edu/mgs/mnpot/dcmplx.html

Morley, C. (2006). Flowsheets for HPGR. *International Autogenous and Semiautogenous Grinding Technology 2006*, Vancouver, BC, Canada. *4* 172-189.

Muranda, R. (2009). A laboratory scale ball mill grindability comparison of the HPGR and conventional crusher products. *Proceedings of the VI Internatonal Mineral Processing Seminar*, Santiago, Chile.

Norgate, T. E., & Haque, N. (2010). Energy and greenhouse gas impacts of mining and mineral processing operations. *Journal of Cleaner Production*, *18*(3), 266-274.

Pease, J. D. (2007). *Case study coarse IsaMilling at McArhur river.* Retrieved January 28, 2011, from http://www.isamill.com/downloads/Joe%20Pease2.pdf

Rule, C. M., Smit, I., Cope, A. J., & Humphries, G. A. (2008). Commissioning of the polycom 2.2/1.6 5.6 MW HPGR at anglo Platinum’s new mogalakwena north concentrator. *MEI Comminution’08 Conference*, Falmouth, Cornwall, UK.

Shi, F. N., Morrison, R., Cervellin, A., Burns, F., & Musa, F. (2009). Comparison of energy efficiency between ball mills and stirred mills in coarse grinding. *Minerals Engineering*, *22*(7-8), 673-680.

Valery, W., & Jankovic, A. (2002). The future of comminution. *34th International October Conference on Mining and Metallurgy - Proceedings*, Bor Lake, Yugoslavia. 287-298.

Vanderbeek, J. L., Linde, T. B., Brack, W. S., & Marsden, J. D. (2006). HPGR implementation at Cerro Verde. *International Autogenous and Semiautogenous Grinding Technology 2006*, Vancouver, BC, Canada. *4* 45-61.

INTERNATIONAL
AUTOGENOUS GRINDING
SEMIAUTOGENOUS GRINDING
AND
HIGH PRESSURE GRINDING ROLL
TECHNOLOGY
2011

EDITORS:

Ken Major
Mineral Processing Consultant
KWM Consulting Inc.
Maple Ridge, B.C.

Brian C. Flintoff
Senior Vice President, Technology
Metso Minerals Canada Inc.
Kelowna, B.C.

Bern Klein
Associate Professor and Department Head
Norman B. Keevil Institute of Mining Engineering
University of British Columbia

Kelly McLeod
Process Consultant

Proceedings of an International Conference on Autogenous Grinding, Semiautogenous Grinding and High Pressure Grinding Roll Technology held September 25 – 28, 2011in Vancouver, B.C., Canada. This conference is made possible by the support from the Norman B. Keevil Institute of Mining Engineering of the University of British Columbia, the Canadian Mineral Processors of Canadian institute of Mining, Metallurgy and Petroleum and the Canadian Mining Industry Research Organization.

STIRRED MILLING AT ANGLO AMERICAN PLATINUM

Chris Rule*

[1]*Anglo American Platinum,*
55 Marshall Street, PO Box 62179, Marshalltown, Johannesburg, South Africa,

*(*Corresponding author: ChrisR@angloplat.com)*

STIRRED MILLING AT ANGLO AMERICAN PLATINUM

ABSTRACT

This paper tells the story of the rapid commercial application of stirred milling technology at Anglo American Platinum's many Concentrators. The first stirred mill – a 10 000 litre IsaMill^{TM}; the largest at the time, was commissioned in late 2003. The project development work was presented at SAG 2001. This operation was "proof of concept" and vindicated the decision to scale up the technology; allowing a rapid and extensive roll-out of mainstream regrind mills as well as several further flotation concentrate regrind mills, over the next six years.

Currently 23 stirred mills are in operation with 64 MW of installed drive capacity – this paper describes the projects' roll-out and the significant results achieved to date. Many operational problems have been solved along the way, some of these will be described. The introduction of the technology has seen a step change in metallurgical results and progressively decreasing operating costs.

KEYWORDS

Process Mineralogy, PGM, Liberation, Plant Optimisation; UG2 reef, Merensky reef; Platreef; Stirred milling; Mainstream Inert grinding, MIG; Ultra Fine Grinding - UFG , Fine Grinding - FG

INTRODUCTION

The ore profile being mined at Anglo American Platinum has changed dramatically in the last two decades. Concurrently the production of PGMs has grown markedly. The ore sources have changed from being wholly based on the Merensky reef, in the 1980s, to the current split where UG2 makes up over 50% of ore treated in the fourteen managed Concentrator plants and is forecast at 16.7 million tonnes for 2011. Platreef at .just under 11 million tonnes now ranks second with Merensky now only 4.5 million tonnes; total ore processed including other surface material is in excess of 41 million tonnes for 2011. See figure 1, which shows the profile change in the last 10 years.

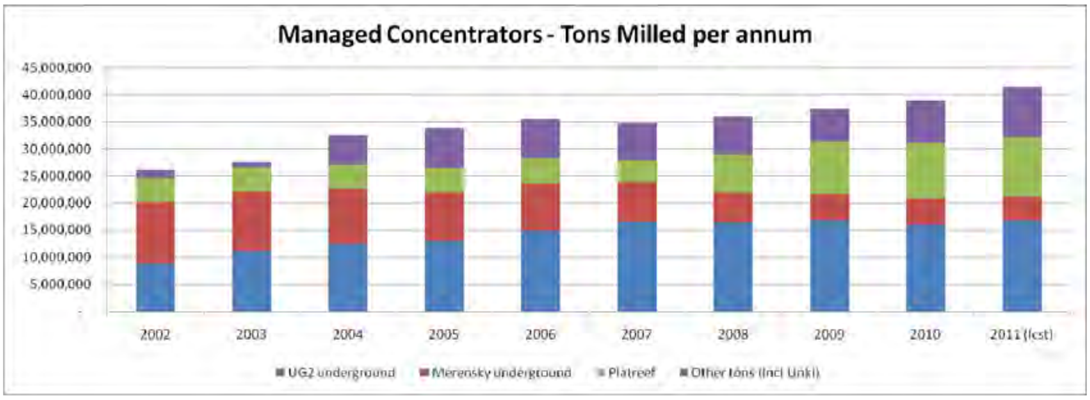


Figure 1: Ore milled over the last 10 years at Anglo American Platinum Managed Concentrators

The ore types have very different mineralogical characters which largely determine the metal extraction efficiencies. Unfortunately, the UG2 and Platreef are more mineralogically complex. They have smaller sized PGMs minerals and increasingly higher dissociation of PGM minerals from the larger and hence more easily flotable sulphide minerals; chalcopyrite, pentlandite and pyrrhotite. To compound the plant extraction challenge the head grades delivered have declined over the same time period. The average feed grade has dropped from excess of 5.5 gpt combined PGMs, in 1999; to approximately 3.2 gpt in 2010.

Historically, AAP mining operations were mainly on the western Bushveld complex with the major operations at the Rustenburg, Amandelbult and Union section mining complexes. The Merensky reef mined there delivered grades often in excess of 6 gpt combined PGMs. See figure 2, for the location of the Anglo Platinum mining operations on the Bushveld complex and inset, a diagram of the Great Dyke in Zimbabwe.

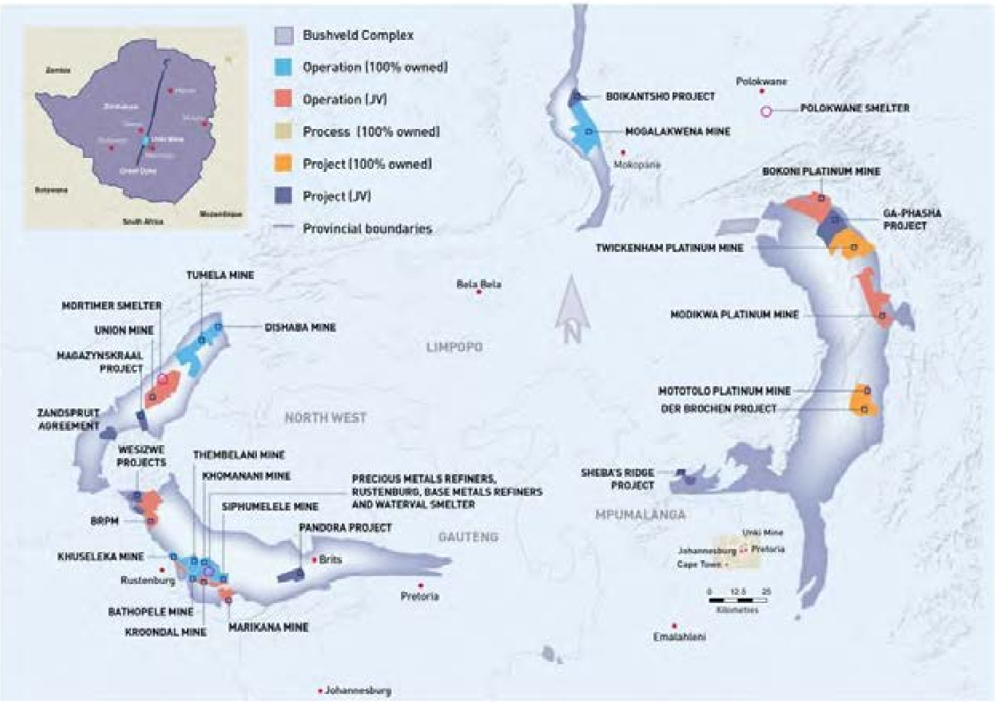


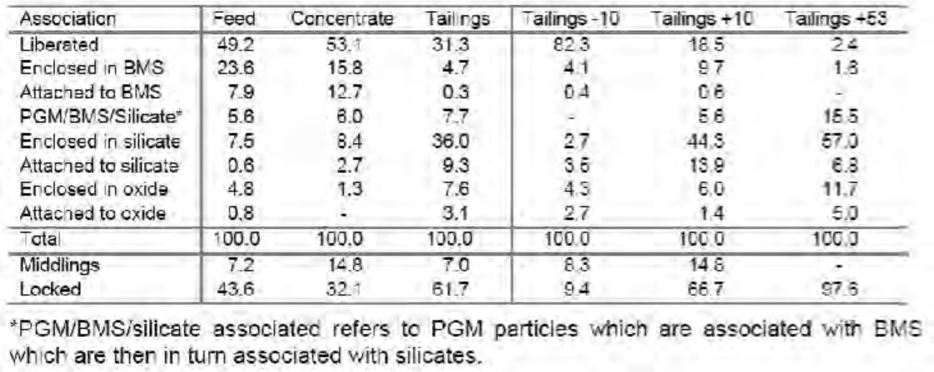
Figure 2: Bushveld Complex – this diagram shows the footprint of Anglo American Platinum's operations on the Bushveld Complex in the northern provinces of South Africa and on the Great Dyke in Zimbabwe

Process mineralogy was recognized a fundamental capability that was required to facilitate a planned programme to address this declining trend in metals recovery. Investment in Anglo Research's, (previously Anglo Platinum's Research mineralogy section); has been considerable. (Schouwstra 2004)

In 2004, AAP's Concentrator Technology function motivated to have a comprehensive programme for each plant's monthly composites to be analysed by size fraction and further to submit selected samples to be analysed mineralogically. The volume of mineralogical work for twenty or so plants includes the non managed Concentrators; was enormous, so a priority process selected the reduced submissions for mineralogy. This priority rating was dictated by the plant's relative ounce contribution and an ongoing assessment of the plant's operational results at any time.

Anglo American Platinum has always had the capacity and desire to investigate potential beneficial new technologies. Investment in process mineralogy has been made bringing in the latest technology and hardware in the preceding years. The knowledge gained on the orebodies grew in line with this investment and the intensive exploration programmes on the tenements held by the company on the Bushveld and Great Dyke PGM provinces. Plant troubleshooting studies had already shown the loss profile of the UG2 and Platreef operations was principally due to incomplete liberation and to losses of sub 10 micron liberated PGM minerals. The following illustrations; taken from the monthly composite programme results, illustrates this clearly.

Table 1: Mineralogical association data for typical monthly composites for MF2 UG2 plant samples, size fraction analyses shows potential for further liberation



Association	Feed	Concentrate	Tailings	Tailings -10	Tailings +10	Tailings +53
Liberated	49.2	53.1	31.3	82.3	18.5	2.4
Enclosed in BMS	23.6	15.8	4.7	4.1	9.7	1.6
Attached to BMS	7.9	12.7	0.3	0.4	0.6	-
PGM/BMS/Silicate*	5.6	6.0	7.7	-	5.6	15.5
Enclosed in silicate	7.5	8.4	36.0	2.7	44.3	57.0
Attached to silicate	0.6	2.7	9.3	3.5	13.9	6.8
Enclosed in oxide	4.8	1.3	7.6	4.3	6.0	11.7
Attached to oxide	0.8	-	3.1	2.7	1.4	5.0
Total	100.0	100.0	100.0	100.0	100.0	100.0
Middlings	7.2	14.8	7.0	8.3	14.6	-
Locked	43.6	32.1	61.7	9.4	66.7	97.6

*PGM/BMS/silicate associated refers to PGM particles which are associated with BMS which are then in turn associated with silicates.

Table 1 shows the potential for futher grinding clearly, with the majority of losses of PGMs due to incomplete liberation, both locked and middlings particles in both the +10 micron and especially in the +53 micron fractions. The following illustrations shows the incomplete liberation in tabular form and in the false colour two dimensional images sourced form the MLA analyser.

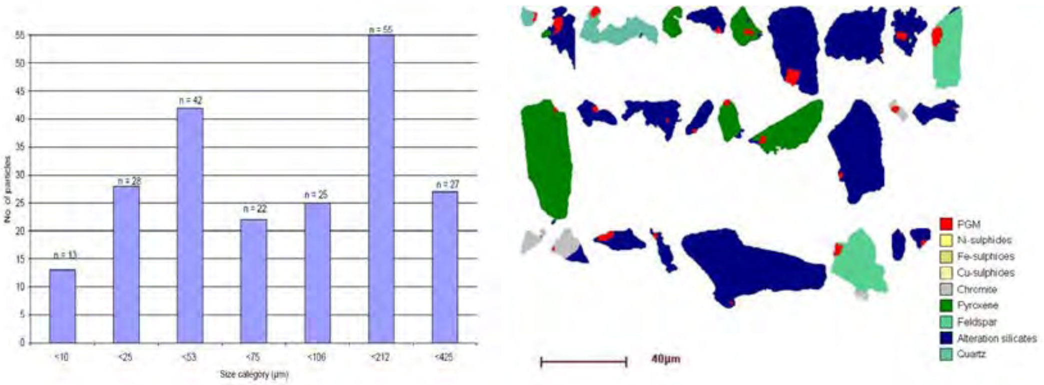
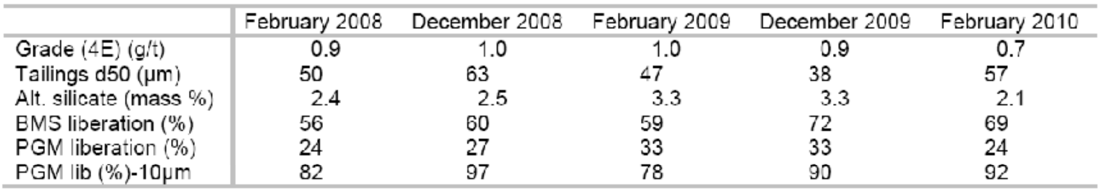


Figure 3: Number of composite particles containing PGMs by size fraction in a typical UG2 tailings sample. Particle map (false colour), showing the association of PGMs, (red), with mainly gangue minerals in a final tailing sample analysed from the plant composite programme

A history for each plant is built up for the sampled streams; in table 2 the average PGM liberation in tailings is seen at low levels of 20-35% and generally these are in the sub 10 micron fraction which are more difficult to float.

Clearly then an economic comminution technology that could improve PGM particle liberation economically and without generating large amounts of “super-fines” would potentially be a key to reaching much higher PGM recoveries. Stirred milling potentially was that technology.

Table 2: Typical UG2 Concentrator monthly composite plant tailings data – assay and mineralogy



	February 2008	December 2008	February 2009	December 2009	February 2010
Grade (4E) (g/t)	0.9	1.0	1.0	0.9	0.7
Tailings d50 (µm)	50	63	47	38	57
Alt. silicate (mass %)	2.4	2.5	3.3	3.3	2.1
BMS liberation (%)	56	60	59	72	69
PGM liberation (%)	24	27	33	33	24
PGM lib (%)-10µm	82	97	78	90	92

At the beginning of 2000, with the recent commercial availability of IsaMill^{TM} stirred mill technology; a 4 litre bench scale mill was assessed at AAP's research and development facilities along with other vertical stirred milling technology units. The three primary requirements for further regrinding in the PGM circuits were:

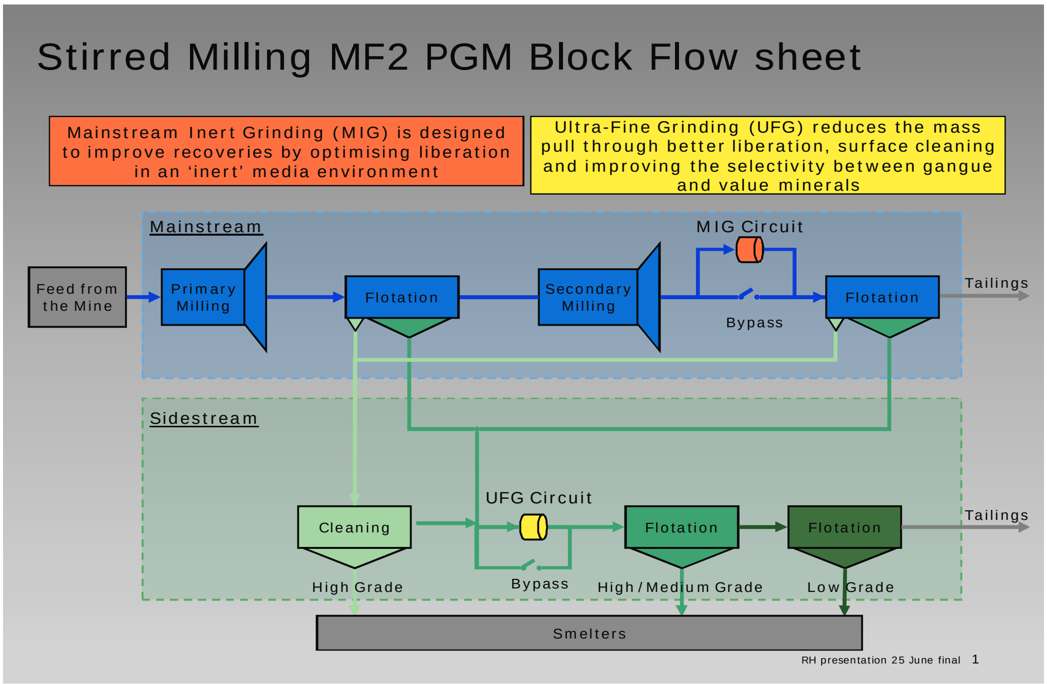
- Due to mainstream tonnages a large stirred mill was targeted; at that time the largest mills available were 1.1 MW 3000 litre for the IsaMill^{TM} technology and for vertical stirred milling the 355 kW SMD. Multiple small units were not favoured,
- Inert or non steel media grinding was the preference due to potential surface chemistry and slurry chemistry issues with iron hydroxide precipitation and coatings on value minerals,
- The technology should be energy efficient and obviously must be economically viable with an expected recovery improvement potential of a further 2-5% PGM recovery.

Promising results lead to acquisition of larger units that could be used in pilot plant studies. Results at pilot scale led to the decision to use IsaMill^{TM} technology in the new tailings re-treatment project, "WLTR" project, near Rustenburg. [5] The significant scale up decision, from a 3000 litre unit to a 10000 litre unit was made at that time by the author, thus facilitating the economic use of the technology for future mainstream applications and once a suitable ceramic media had been identified or developed. The initial mill was developed in a three party collaboration with Netzsch Feinmetal GmbH and Xstrata Technology; then known as MIM Technology. To mitigate the scale up risk a variable speed drive was chosen – the drive train was 2.6 MW due to the initial media being locally mined crushed and 2-5mm, screened silica sand

THE ISAMILL^{TM} TECHNOLOGY ROLL-OUT PROGRAMME

The circuit applications identified as being the optimal circuit interventions are illustrated in the following block flow sheets, figure 4 and figure 5. The mainstream application – given the name "MIG" or mainstream inert grinding was added on regrind ball mill product ahead of scavenger rougher flotation; resulting in better mainstream PGM extraction. The "UFG" or ultra fine grinding application was targeted at modifying the typical plant grade/recovery relationships in the cleaning circuits for the mainstream flotation products and results in better product grades and higher circuit PGM recovery by regrinding composites and applying intensive surface attritioning. The two acronyms were created to prevent confusion and recognized the essential features; MIG - inert grinding using non steel media in mainstream applications, typically targeting 80% -53 micron products and UFG, concentrate regrind targeting 80% less than 20 microns. It would be more correct to call this application "FG" or fine grinding to differentiate from the very fine regrinding targeting products that are sub 10 micron common in lead/zinc production and in leaching applications.

This typical PGM Concentrator, "MF2" flow sheet shown in figure 4, incorporating the stirred milling technology is applicable to the Platreef and Merensky plant circuits.



Stirred Milling MF2 PGM Block Flow sheet

Mainstream Inert Grinding (MIG) is designed to improve recoveries by optimising liberation in an 'inert' media environment

Ultra-Fine Grinding (UFG) reduces the mass pull through better liberation, surface cleaning and improving the selectivity between gangue and value minerals

Figure 4: Process flow diagram of the application of stirred milling as 1) mainstream or MIG and 2) concentrate regrind or UFG is a typical PGM industry stage grinding and flotation, or MF2, circuit

For UG2 reef, due to its essentially bimodal form – a mix of silicate minerals and chrome spinel minerals in the chromitite reef; a variant of this MF2 circuit had been developed. The new circuit applied MIG stirred milling technology in the split regrind circuit. The primary rougher flotation tailings are split with a hydrocyclone cluster prior to further processing. This produces two streams; an underflow rich in higher specific gravity and coarse chromite spinel and the overflow enriched in lower specific gravity silicates and fines. The majority of PGMs follow the silicates; their occurrence in the in-situ ore is predominantly in the silicate minerals in between the chromite spinel crystals. Either the MIG IsaMill^{TM} follows a regrind ball mill or it treats the cyclone overflow stream directly; dependant on the available milling units at the plant.

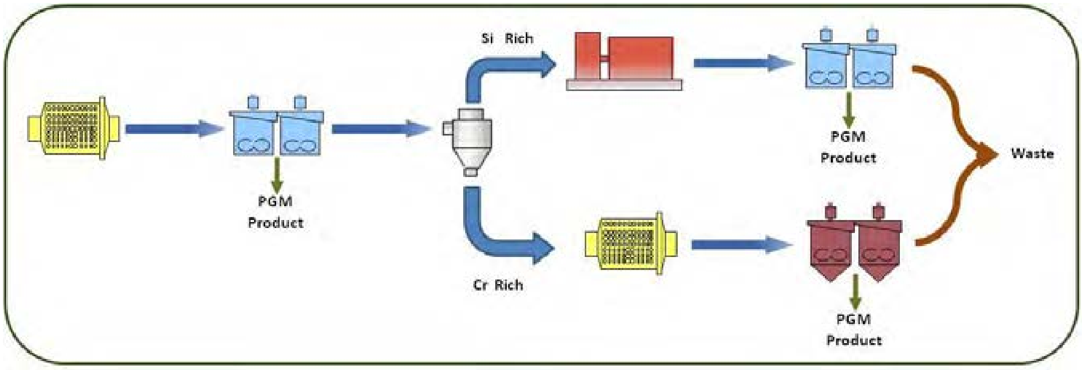


Figure 5: Process flow diagram for a split regrind MF2 circuit for UG2 ore treatment – shows the MIG IsaMill^{TM} on the cyclone overflow and the open circuit ball mill on the cyclone underflow

The adoption of IsaMill^{TM} stirred milling was a relatively fast adoption of a new technology. Key steps in this process can be listed:

- 2000/2001; scoping work carried out at AAP research facilities, "ARC", at Germiston, east of Johannesburg and at the operational support centre, "DML" and pilot plant at Rustenburg. The pilot studies confirmed the potential for both mainstream stirred milling, "MIG" and concentrate regrind, "UFG" in mainstream flotation product cleaning,
- 2001; decision taken to use IsaMill^{TM} stirred milling technology and to develope the first M10000 litre unit; development within a tripartite collaboration with Xstrata Technology, (then known as Mt Isa Mines Technology) and Netzsch Feinmetal GmbH. Key was the choice of scaled up unit; this was to allow future mainstream applications to be realized. Silica sand chosen as the grinding media in the absence of a proven ceramic media.
- 2001/2003; inclusion of UFG concentrate regrind M10000 IsaMill^{TM} with variable speed 2.6 MW drive to mitigate design scale-up inaccuracy; in the tailings retreatment project, WLTR; at Rustenburg. Successfully commissioned in early 2004; proof of concept established. (Buys *et al.* 2004)

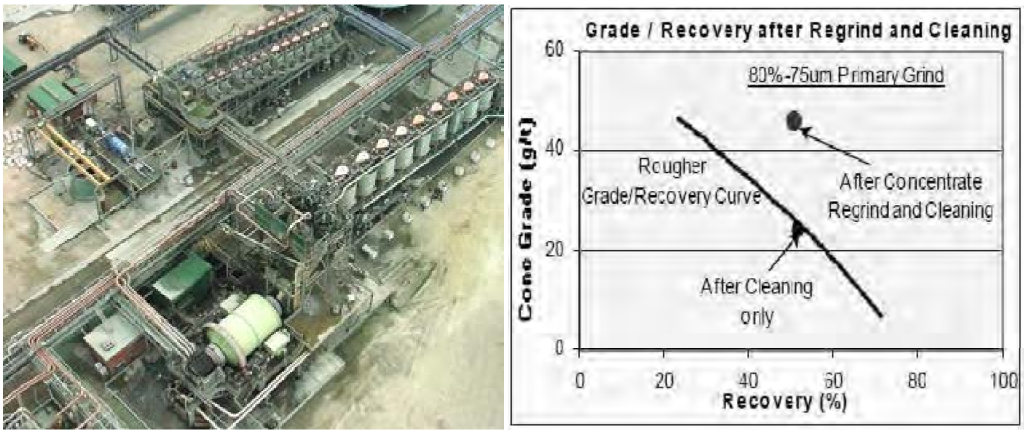


Figure 6: Aerial photograph showing the WLTR site and a graph showing the change in PGM grade/recovery potential by applying UFG stirred milling of mainstream flotation concentrates

- 2004/2005; initiation of a comprehensive off-site and on-site pilot programme with on-site piloting of various stream feeds using a M20 IsaMill^{TM} unit and associated "FCTR" mini flotation rig. Initiation of the routine monthly composite plant sample programme. Fractional analyses and mineralogical analyses on plant feeds, plant flotation product and plant tailings. Realization of the economic viability and then definition of the priority for establishment of a wide scale roll-out of IsaMill^{TM} technology for both mainstream MIG and concentre regrind UFG applications at major AAP Concentrators. Identification of an economically viable ceramic grinding media for MIG applications. (Rule *et al.,* 2008)

Figure 7: Photographs of the test rig used to evaluate the potential of stirred milling – shown here at Mogalakwena South Concentrator in late 2004

- 2005/2006; the initiation of the ceramic media development programme. Silica sand, furnace slag and other natural medias could not be used in MIG applications due to the relative hardness of the ore and media. With the approval of the first MIG project at Mogalakwena South; the approving AAP body required that a multi-media ceramic supply base be put in place for the technology.

Figure 8: Grinding media used in IsaMillTM applications; in the left hand side photograph samples of furnace slag ex Mt Isa, Colorado River sand, screened ore particles and steel shot and in the right hand side photograph two samples of ceramic media

This was achieved by setting up a testing facility which included utilising both the 4 litre and 100 litre IsaMills at the Divisional Metallurgical Laboratory operational support facility in Rustenburg. A comprehensive test programme was established to allow assessment of the ceramic medias available in the world market. Simultaneously major suppliers were contacted and engaged with. Later in 2007, further QA/QC capability was established in order that ceramic media deliveries could be checked for adherence to specification. The programme to date has tested some 200 different medias from all the major ceramic media manufacturers and has led to the development of several improved formulations with various suppliers. A ceramic media manufacturing facility in South Africa was established by a third party in 2007. This proactive involvement has led to significant operating costs benefits and indeed the media consumable cost on a US$ per kWhr consumed by the IsaMills has dropped by more than $2/3^{rds}$ from 2008 to 2011. (Bedesi et al, 2008)

- 2006; commissioning of the first mainstream or MIG IsaMill^{TM} at Mogalakwena South Concentrator – C section project utilizing a 3MW M10000 unit; note the use of higher specific gravity media resulted in increase in drive train capacity. This mill installation proved the MIG concept was a workable technology and removed any risk constraint for the full roll-out of the technology throughout AAP operations.

- 2007/2008/2009; motivation, approval for execution and installation of a further 20 IsaMill^{TM} units in MIG, (17 M1000 units) and UFG, (3 units, 2 M1000 and 1 M3000) duties at Bafokeng Rasimone, Waterval UG2, Waterval, Amandelbult Merensky, Amandelbult UG2 $^{\#}$1 and UG2 $^{\#}$2, Mogalakwena South A and B and Mogalakwena North Concentrators; (Rule *et al*. 2010)

Figure 9: IsaMill^{TM} layout photographs, Waterval UG2, Amandelbult UG2#2 and Waterval Concentrators

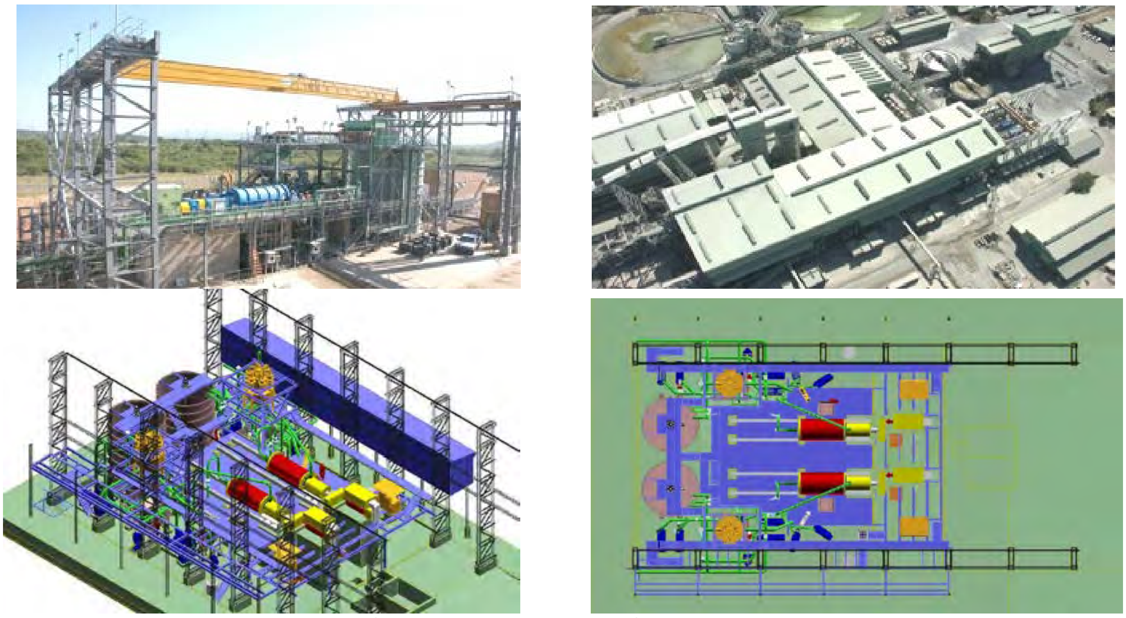


Figure 10: MIG installations at Bafokeng Rasimone, top left, shows the typical "wrap around" equipment in the circuit and on top the right hand side; Mogalakwena South Concentrator; A and B mills installed on far left hand side. Below these are the model drawings for A and B section IsaMills

OPERATING EXPERIENCE DURING THE EARLY PERIOD

There were remarkably few problems with the first M10000 installed at the WLTR Project in 2003. The scale up resulted in operational performance that was almost exactly as predicted; hence the variable speed drive unit is/was seldom utilized. Test work for future MIG application on tailings material in 2008 to 2010 was conducted and some work was done at variable speed regimes with varying media loads etc.

Following the first MIG milling application start up of the Mogalakwena C section mill in late 2006; the success resulted in the rapid approval of the next four MIG mills, two each at Mogalakwena South; A and B and at Waterval UG2 with commissioning by the end of 2007.

A major difficulty identified initially was the management of the position of the media load. Variability in mill feed volume and slurry density resulted in incidents of media compression at low flows and media loss at high flows. The media compression incidents resulted in the early modification of the internal wear discs configuration, with the disc in position 1, i.e. nearest the feed inlet; being removed totally, effectively creating a larger first milling compartment. Modifications to the dimensions of the feed end discs have been made with reduced diameter discs being successfully trialled. Too high feed slurry flows result in significant media loss – seen visually downstream and also measured in higher than expected media losses.

Figure 11: Photographs of grinding discs made of different rubber compounds and experimental ceramic tile shell liners after use

A number of materials of construction test runs have been conducted; figure 10 shows the mill internals of an M10000 and some ceramic shell liners using tiles that were unsuccessfully trialled.

A summary of issues that have been addressed to date between 2008 and 2009 follows:

i) an inordinate number of drive train bearing over temperature trips were encountered on some of the mills – the bearing lubrication system was modified to a circulated/cooled oil rather than grease based system, before the modifications were completed the mills were run at lower power levels to reduce mill stoppage events

ii) The incidence of damage from tramp material became a very significant cost and operating downtime issue for the MIG installations particularly from the middle of 2008; this included premature severe damage to internals from repeated ingress of steel ball fragments, nuts and bolts, ceramic wear tiles, (from upstream launders/still boxes etc) and coarse oversize excursions from upstream ball mills ahead of the IsaMill^{TM}.

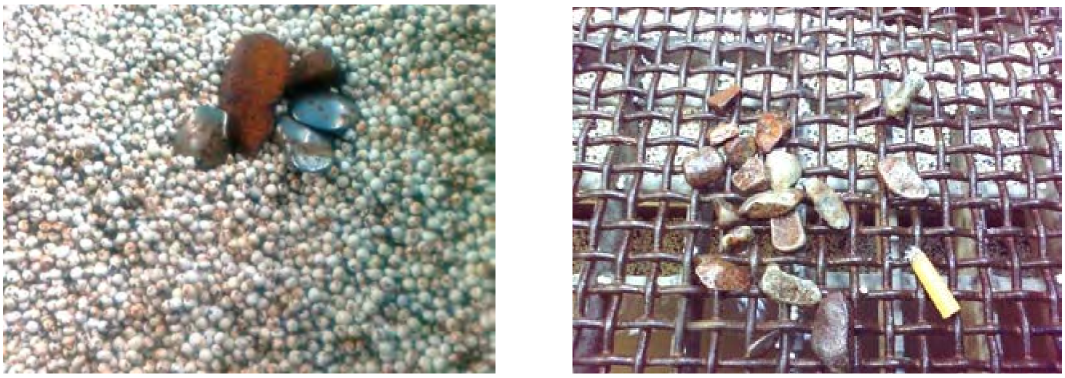


Figure 12: Photographs showing metal steel ball scats and “rust” on surface of some of the 3.5mm top size media in the dumped mill media and resting on the grating below the mill – damage to the rubber and polyurethane wear parts in the mill was severe and rapid!

iii) At times poor judgement on the wear life of grinding discs led to disc collapse and severe damage to mill internals from the broken pieces.

Whilst thee problems were analysed the MIG mills especially were run at lower loads and in the extreme case where multiple internal damage incidents occurred the mills were taken off-line at Waterval UG2 for 6 months at the end of 2008.

Figure 13: Internals of the IsaMill^{TM}; shows the grinding discs and separator with mill shell pulled back on the sliding rails – access position during maintenance

A rapid identification of root cause and development of solutions phase followed and in 2009 when the next 16 mills were commissioned the design fixes were auctioned during the installation and before commissioning; fixes were actioned including to the first 6 mills which were taken offline for modifications:

a) Media position within the mill was controlled by modifying the existing slurry recycle stream allowing fixed volume flow set-point to be implemented in the control system; slurry density was controlled by optimizing the cyclone parameters upstream and optimization of the IsaMill^{TM} feed circuit surge tank buffer control system, the 7 disc configuration was retained,

b) A linear screen was installed on the circuit feed system ahead of the surge tank feeding the mill density controlling cyclone clusters – a physical barrier to oversize material greater than 2 mm was thus imposed on the circuit. An interesting finding during the analyses exercise at one of the sites was the occurrence of copious fragments of steel in the mill contents – this was discerned form analyses of the dumped media where “rusting” as well as large fragments “spalled” from the steel ball charge of the primary mill was seen. How large fragments >20mm were able to pass through the upstream primary circuit closing vibrating screens with decks panels of 850 microns aperture, travel through the primary rougher flotation and then enter the MIG IsaMill^{TM} circuit via the densifying cyclone cluster fed from the surge tank pumps begs the question around operating and maintenance vigilance and control!

c) Where ore quality processed varied another issue was apparent and is shown in the two photographs following, figure 14. The Mogalakwena South Concentrator occasionally treats extremely hard ore and during these periods the grind coarsens appreciably – exceeding the capacity of the FAG and ball mills ahead of the two A and B MIG IsaMills to provide a fine enough feed. Critical size build up results! The problem has been addressed by bringing back into the circuit the two redundant tertiary ball mills which now provide a further comminution step ahead of the MIGs.

Figure 14: Photographs show severe critical size build up in the IsaMill^{TM} at Mogalakwena South A and B sections; the media is white and the coarse feed particles are generally dark coloured

d) Further more rigid operating recipe driven operation was rolled out across the sites – resulting in tighter control over all circuit parameters not just those impacting on "spikes" in volume and density flow through the circuit. A higher level of operating behaviour is now common place throughout the group Concentrators with benefits seen performance in all unit processes.
e) The application of a comprehensive data collection from both the DCS system, remotely, to the group control centre in Rustenburg and maintenance supervision programme on all mills was put in place; data on media consumption and stock levels is also collated and reported. This comprehensive data set and reports has prevented the incidence of wear failures to be almost eliminated by 2010. A monthly report is generated to all sites and support personnel and in parallel the routine wear data is collected at each maintenance stop for all mills and analysed and reported co-currently. Note all the wear supervision is completed by Xstrata Technology personnel and is used routinely and for testing for mill component and circuit optimization

OPERATING RESULTS

The change in performance post MIG IsaMilling is well illustrated at the Mogalakwena South Concentrator, Waterval Concentrators and the Amandelbult Concentrators; the post commissioning period results at the Rustenburg and Amandelbult sites have been significantly better than predicted.

The initial evaluation for the Mogalakwena South Concentrator comparing the results post commissioning of AAP's first MIG IsaMill^{TM} on the "C" section ore processing line against the A and B section ore processing lines led to the rapid approval of further MIG mills at Waterval UG2 and for "A" and "B" sections. The statistical analyses conducted on the pre and post commissioning period concluded a PGM delta recovery increase of between 3.5 and 4%; due to MIG stirred milling in the mainstream circuit, refer figure 15.

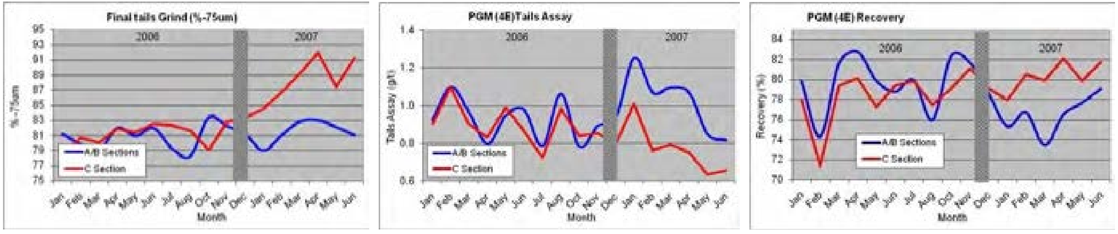


Figure 15: Results from Mogalakwena South after commissioning of the first MIG mill on "C" section compared to the other two lines, "A" and "B" operating with MIG IsaMilling – statistical analyses showed a 3-4% delta recovery improvement for the technology

Further assessment of the UG2 ore impact at the Waterval UG2 plant post commissioning in 2007 and through the early part of 2008 led to approval for the roll-out of the next phase of the technology's application on UG2 and Merensky/UG2 mixed ore Concentrators – taking the IsaMill^{TM} fleet deployed to 22 units and 64 MW of installed drive train installed power.

Rustenburg and Amandelbult mining operations are AAP's biggest production units with roughly 2/3^{rds} of the 2010 total managed Concentrators' platinum output of 1.7 million troy ounces. Roughly 2/3^{rds} of the mined and processed tonnage in 2010, 10 million tonnes; was sourced from the UG2 reef.

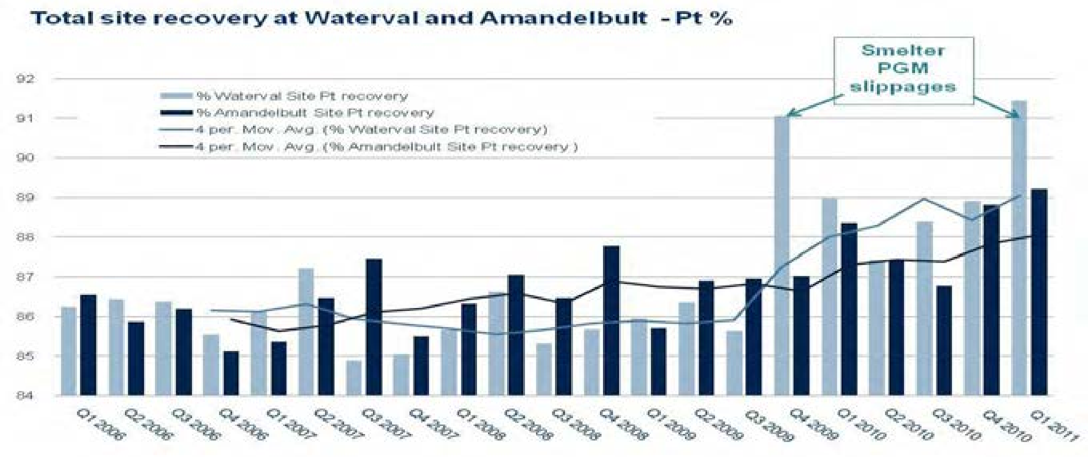


Figure 16: Waterval and Amandelbult Concentrator sites – platinum recovery including tailings scavenging operations, quarterly Q1 2006 to Q1 2011 MIG IsaMill^{TM} commissioning 2Q to 4Q 2009

Figure 16 shows the impact on metals recovery since the commissioning period – an upward trend at both sites for platinum recovery. Tailings flotation scavenging achieves further metal recovery by allowing a lower grade product to be extracted from each site's combined tailings; before disposal to surface tailings dams.

Analyses of the performance of the two UG2 plants at Amandelbult shows the reduction in tailings grades from the two plants over time; the dramatic step reduction since MIG IsaMill^{TM} commissioning stands out. Tailings grades are now at the lowest levels since Amandelbult mining operations began in the 1970s.

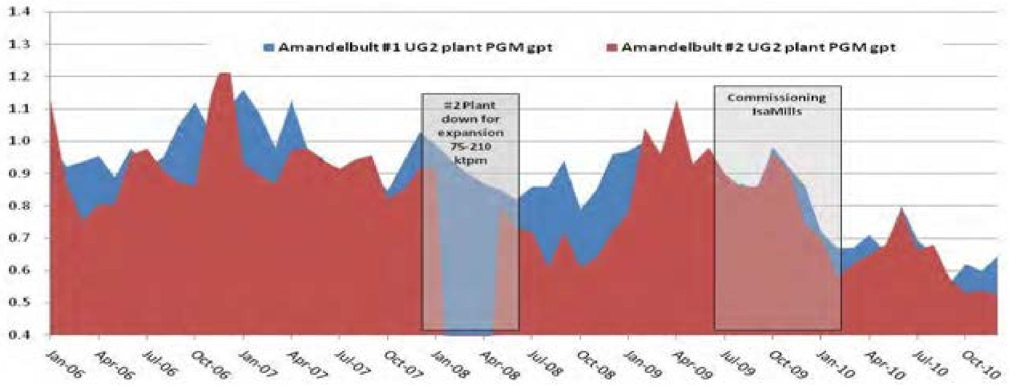


Figure 17: Comparison of the tailings values for the two Amandelbult UG2 Concentrators from 2006 to 2010; MIG IsaMills were commissioned in mid 2009 – a step change occurred with lowest ever tailings grades being achieved since the beginning of 2010

The graph shown in Figure 16 shows the comparison of PGM recovery values achieved for UG2 processing at the two mining sites at Amandelbult and its neighbour Union, some 30 kilometres away. Historically PGM metal recoveries at the Mortimer UG2 plant at Union have always been similar to that achieved at Amandelbult and averages around 80-82% extraction of PGMs. Since MIG IsaMill^{TM} commissioning a strong divergent and upward trend is seen for the Amandelbult operation.

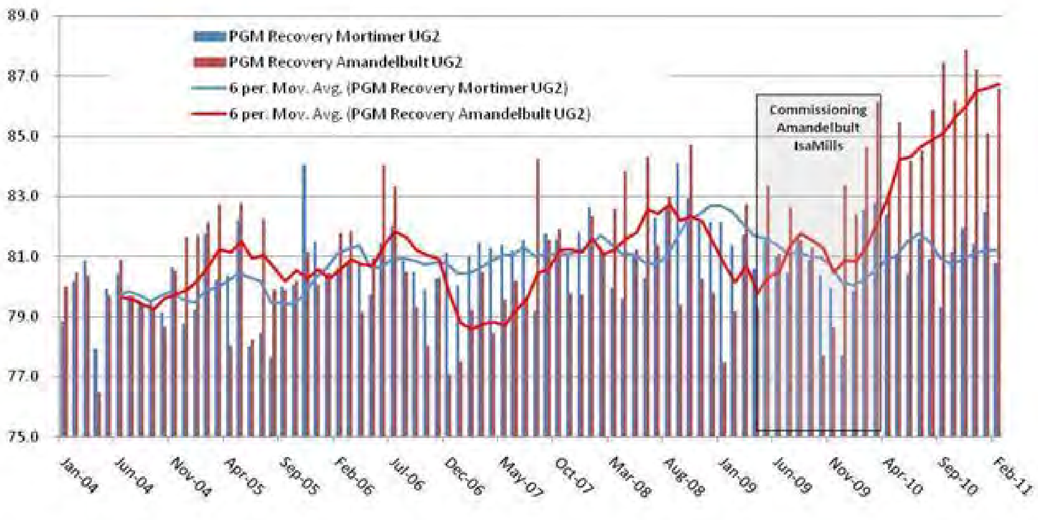


Figure 16: PGM recovery for the two UG2 Concentrator operations of the Amandelbult and Union mining operations – these were historically similar but post MIG commissioning at Amandelbult performance is now significantly superior

CONCLUSIONS

The journey to fully install stirred milling technology in AAP's Concentrator operations represents a very fast introduction of a new technology, by industry standards.

The economics of the recovery improvement potential at the plants is an overwhelming proposition due to the increasing value, with time, of the contained metal in the ROM ore. This justifies the level of investment in process mineralogy and a new comminution technology already made and continuing to be made at Anglo American Platinum. To illustrate this it is a sobering fact that an improvement of 1% in PGM recovery in the Concentrators at Anglo American Platinum is equivalent to >75 million US$ for each year at today's metal prices and exchange rates.

Currently, further stirred mill installations are being motivated. Test work and circuit analyses has identified that further upside potential at existing operations remains to be achieved as the operations are optimised further.

REFERENCES

Schouwstra, RP (2004), Application of mineralogical techniques to ore samples and plant products. *In* Ashwal, L.D. (ed), *Geoscience Africa 2004 Abstracts*. University of the Witwatersrand, Johannesburg, pp 583

Buys S, Rule C.M, Curry D (2004), The application of large scale stirred milling to the re-treatment of Merensky platinum tailings; CIM 2004, Montreal, Canada

Rule C.M, Knopjes L, Jones R.A (2008), The development of main stream inert grinding or "MIG" IsaMill technology at Anglo Platinum; Comminution 2008, Falmouth, Cornwall, England

Bedesi N, Rule C.M, Knopjes L (2008), Anglo Platinum's selection of a ceramic grinding media protocol for MIG and UFG IsaMilling, Comminution 2008, Falmouth, Cornwall, England

Rule, C.M (2010), Stirred milling – new comminution technology in the PGM industry, SAIMM Conference Platinum in Transition, Boom or Bust and in SAIMM Journal February 2011

INTERNATIONAL
AUTOGENOUS GRINDING
SEMIAUTOGENOUS GRINDING
AND
HIGH PRESSURE GRINDING ROLL
TECHNOLOGY
2011

EDITORS:

Ken Major
Mineral Processing Consultant
KWM Consulting Inc.
Maple Ridge, B.C.

Brian C. Flintoff
Senior Vice President, Technology
Metso Minerals Canada Inc.
Kelowna, B.C.

Bern Klein
Associate Professor and Department Head
Norman B. Keevil Institute of Mining Engineering
University of British Columbia

Kelly McLeod
Process Consultant

Proceedings of an International Conference on Autogenous Grinding, Semiautogenous Grinding and High Pressure Grinding Roll Technology held September 25 – 28, 2011in Vancouver, B.C., Canada. This conference is made possible by the support from the Norman B. Keevil Institute of Mining Engineering of the University of British Columbia, the Canadian Mineral Processors of Canadian institute of Mining, Metallurgy and Petroleum and the Canadian Mining Industry Research Organization.

IsaMill™ TECHNOLOGY IN THE PRIMARY GRINDING CIRCUIT

*G.S. Anderson[1], D.T. Smith[2], S.J. Strohmayr[2]

[1]Xstrata Technology
Level 4, 307 Queen Street
Brisbane, Australia, 4000
*(*Corresponding author: GAnderson@xstratatech.com)*

[2]Xstrata Zinc
McArthur River Mine
Northern Territory, Australia

IsaMill™ TECHNOLOGY IN THE PRIMARY GRINDING CIRCUIT

ABSTRACT

Originally developed as a step change in ultrafine grinding efficiency, the IsaMill™ has made inroads into conventional ball mill and tower mill mainstream grinding duties during the last 5 years - beginning with the Anglo Platinum mainstream grinding projects in South Africa. At the 2006 SAG Conference, plant testwork on the incorporation of IsaMill™ technology into the McArthur River (Northern Territory, Australia) Pb/Zn primary grinding circuit was reported. The IsaMill™ has now become an integral component of that circuit, treating up to 1mm top size feed and directly producing P_{80} 45 µm rougher flotation feed. Two 1.1 MW IsaMills™ are now used in this duty contributing to a 27% increase in overall plant throughput and a reduction in rougher feed size distribution. In this paper, we report on the current state of development and performance of the IsaMill™ in the primary grinding duty at McArthur River and also consider future directions.

KEYWORDS

IsaMill™, McArthur River, efficient, energy, ceramic, inert, primary grinding, stirred milling.

INTRODUCTION

IsaMill™ technology was developed to address the ultra fine grinding (UFG) requirements of Mount Isa Mines (MIM, now Xstrata), particularly for the McArthur River lead/zinc deposit where a 7 µm grind was required, under efficient and inert conditions, in order to make a saleable concentrate (Logan, Leung & Karelse, 1993). The 1.1 MW, 3000 litre IsaMill™ (M3000) became the enabling technology for McArthur River in 1995 and remains critical to its ongoing viability today. The history of McArthur River, the development of the IsaMill™ - between MIM and Netzsch Feinmaltechnik of Germany, as well as the operating characteristics of the IsaMill™ have been well documented elsewhere (Nihill, Stewart & Bowen, 1998; Enderle, Woodall, Duffy & Johnson, 1997; Pease, Anderson, Curry, Kazakoff, Musa, Shi & Rule, 2006).

Since its development and use for UFG duties, the IsaMill™ was commercialised and has now moved into coarser grinding, mainstream applications; allowing the key benefits of improved energy efficiency, small footprint and inert grinding environment to be applied to a range of different mineral types (Anderson & Burford, 2006). Key to this development was both the scaleup of the IsaMill™ from 3,000 litres (1.1 MW, M3000) to 10,000 litres (3 MW, M10,000) (Curry, Clark & Rule, 2005) and the development/use of ceramic media (Curry & Clermont, 2005).

Anglo Platinum lead the way in the rollout of IsaMills™ into mainstream duties based on the successful commissioning and operation of the first 10,000 litre IsaMill™ (in an ultra fine grinding application) in 2003 (Buys, Rule & Curry, 2005) and the significant evidence generated by Anglo Platinum's own mineralogical studies, suggesting that finer grinding was required to improve PGM recoveries, particularly across the Platreef and UG2 orebody types (Rule, 2010). Between 2006 and 2009, 54 MW of MIG (Mainstream Inert Grinding) IsaMills™ were commissioned across the Anglo Group in South Africa. Despite some early commissioning and operational issues, the IsaMills™ have proved their value adding in excess of 3% to platinum recoveries (Rule & de Waal, 2011). Anglo Platinum currently has a total of 18 IsaMills™ in MIG duties and four IsaMills™ in UFG duties.

At McArthur River, life of mine optimisation studies determined that conversion from underground to open cut mining was required, with a subsequent increase in plant throughput requiring additional grinding capacity. Given their history of development and operational familiarity with the IsaMill™, McArthur River initiated their own testwork into coarse grinding applications around the same time the Anglo projects were proceeding to full scale, with a view to investigating the potential of the IsaMill™ to contribute to the expansion of the primary grinding circuit at McArthur River.

MRM PRIMARY GRINDING CIRCUIT EVOLUTION

When commissioned in 1995 with nameplate capacity of 1.5 mtpa, the MRM primary grinding flowsheet consisted of a single stage 3.5 MW SAG mill. Mill discharge was screened over a 1.8 mm screen with the oversize recycled back to the SAG mill feed chute. Screen undersize was pumped to a set of cyclones that produced an overflow at around P_{80} 45 µm for rougher flotation. Cyclone underflow returned to the SAG mill feed chute.

Over time, ore sourced from the initially targetted high grade number 2 orebody depleted. Additional ore was sourced from the number 3 and 4 orebodies with interbed inclusions. This material was harder and at lower head grade. As a result, numerous changes were implemented to the primary grinding circuit to increase throughput, including additional crushing stages ahead of the SAG mill, increased SAG ball loads, uprating the SAG Mill to 4 MW and the addition of a locally sourced second hand tower mill. This allowed plant throughput to be increased to 1.8 mtpa by 2006 but also resulted in the primary grind size increasing from P_{80} of 45 to 70 µm. The 2006 flowsheet (after crushing) is shown in Figure 1.

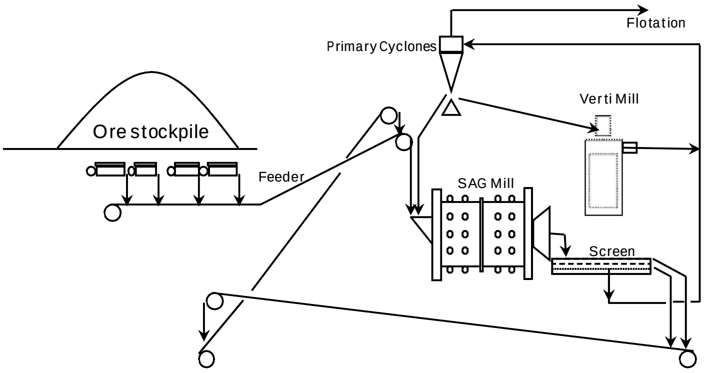


Figure 1 - McArthur River Primary Grinding Circuit 2006

In 2005, a change from an underground operation to an open cut operation was under consideration, requiring the plant to process up to 2.5 mtpa - at further reduced head grade.

IsaMill™ LAB AND PILOT WORK

A testwork program commenced to determine the suitability of using the IsaMill™ to increase the plant throughput at McArthur River. Initial scoping work was carried out at laboratory scale in 2005, followed by two separate on site pilot plant testwork campaigns in 2006 using an 18 kW 20 litre (M20) IsaMill™ - this was reported at SAG2006 (Pease et al, 2006). The on-site pilot work, treating a bleed stream from the SAG mill classification circuit, predicted that a product size P_{80} of 45 µm could be produced at approximately 10 kWhr/t using the IsaMill™.

FULL SCALE 1.1MW M3000 IsaMill™ TRIAL

Following the success of the pilot scale work, one of the six 1.1 MW M3000 UFG IsaMill™ units was reconfigured to operate in the MRM primary grinding circuit to confirm the results from the pilot work. The circuit was configured as per Figure 2 with the bleed stream originating from a modified SAG cyclone. A magnetic separator was included on the feed stream to protect the IsaMill™ from tramp SAG Mill steel - which had been identified as a potential issue during the pilot work. The IsaMill™ operated in open circuit with the product joining the SAG circuit cyclone overflow as feed to the rougher flotation circuit. The full scale trial commenced in May 2007 using 3.5 mm ceramic media with a feed size of F_{80} of 180-200 µm. Figure 3 shows the 6 x M3000 IsaMills™ at McArthur River.

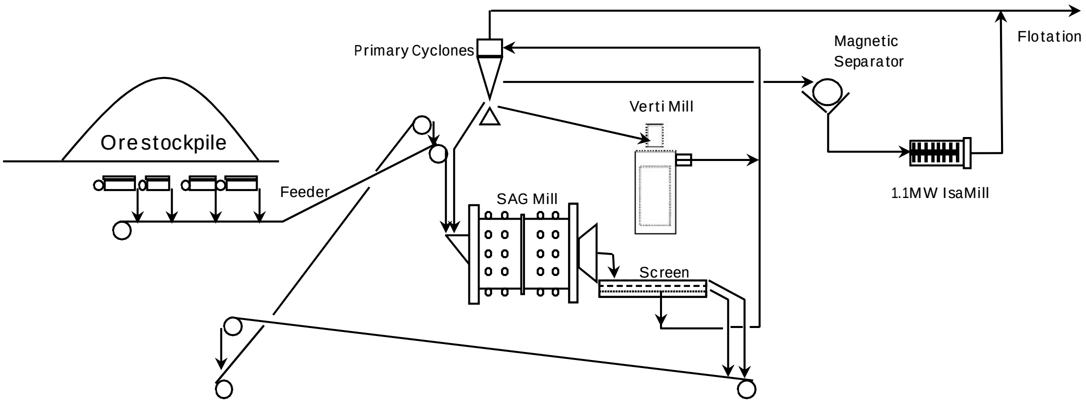


Figure 2 - Circuit Configuration for Full Scale IsaMill™ Trial

Figure 3 - 6 x M3000 IsaMills™ at McArthur River

This application stepped well outside the boundaries of previous IsaMill™ experience and a development phase was needed to optimise the grinding conditions to produce a satisfactory result. As for SAG or ball milling, in stirred milling it is also crucial to break the top size particles fast enough to avoid critical size buildup within the mill. The initial mill internal configuration and 3.5 mm media was unable to do this efficiently or quickly enough, resulting in the IsaMill™ becoming unstable after several hours of operation due to a critical buildup of unbroken coarse material at the front end of the mill. This was confirmed during crash stop inspections of the mill which revealed a large volume of unbroken compacted and dewatered coarse material at the front of the mill and internal wear patterns indicating the incoming feed bypassing this built up stationary area. The buildup of stationary media and unbroken coarse material caused a 30% drop in power draw and resulted in the product from the mill becoming unacceptably coarse due to an effective reduction in the specific energy applied. This was initially overcome by changing to a coarser 5-6 mm top size media and the derating of the internal product separator rotor, which effectively decreased the degree of media compression at the front of the mill and allowed the media and fresh feed to agitate correctly, allowing the coarse feed particles to be ground.

Following several months of development and optimisation, a mill configuration and operating parameter regime was arrived at that resulted in the IsaMill™ being able to treat an F_{80} of 200-250 µm to produce a P_{80} of 35-45 µm at 10-15 kWhr/t. When the IsaMill™ was operating at 800-900 kW, treating approximately 60-70 tph of material, that would otherwise have been returned to the SAG Mill circuit, the overall plant throughput was able to be increased by 12%, from 260 tph to 290 tph, at an overall combined grind size P_{80} of 55-65 µm.

Due to the success of the trial and the impact on the overall plant production, it was continued as a permanent addition to the primary grinding circuit. Eventually a second M3000 IsaMill™ was converted to operate in the primary grinding duty allowing a further increase in overall plant throughput.

GRINDING SURVEY DATA

Numerous surveys across the IsaMill™ circuit were taken under varying operating conditions including changes in internal disc configuration and wear compounds, media type and size, feed size distribution and ore types. Figure 4 illustrates some of the survey data collected over the last two years – all using 5-6 mm ceramic media, grouped by feed size. This is plotted against the graph originally published in SAG 2006 illustrating the testwork data and the predicted operating point for the IsaMill™.

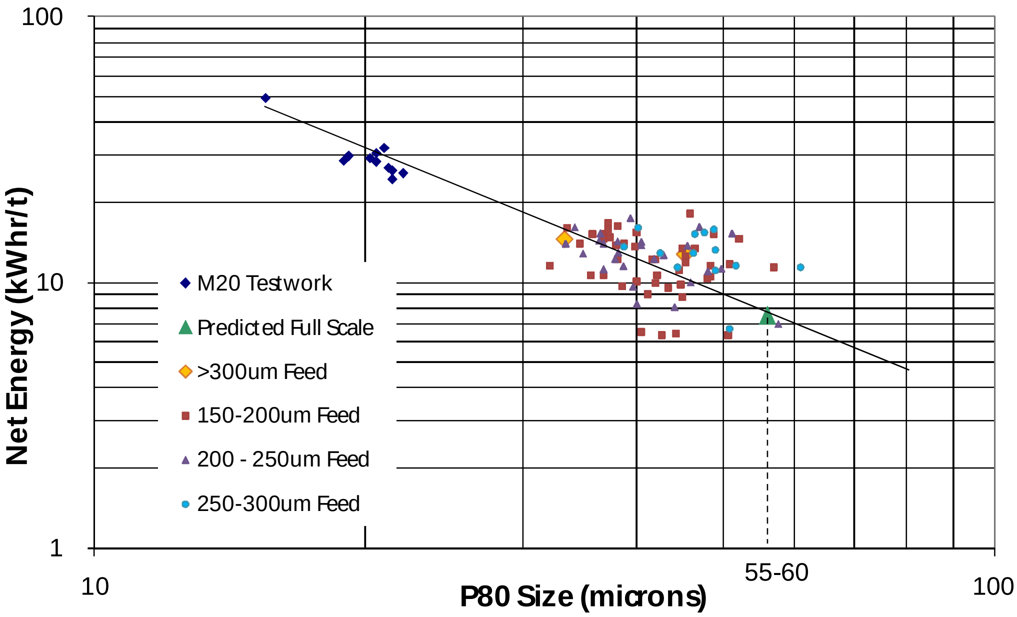


Figure 4 - IsaMill™ Operating Signature Plot

It is evident that, while there is a fair degree of scatter, the points are generally in agreement with the size-energy relationship predicted from the pilot testwork. The different feed sizes show a fair degree of overlap – indicating that the feed size did not have a noticeable impact on the product size–energy relationship. Of note are the two samples where the feed size was F_{80}>300 µm which are within the grouping of data generated from the feed sizes that were F_{80}<200 µm. This suggests that, of the energy consumed in grinding to the product size, very little is required to reduce the top end of the distribution. This is however contingent on the fact that the media is large enough to break the particles at the top end of the distribution at an adequate rate.

EFFECT OF MEDIA SIZE

Ceramic media was trialled at two different sizes in the IsaMill™. Initially the same 3.5 mm ceramic used during the pilot plant work was used. This was changed out to a 5-6 mm top size ceramic during the early stages when there were problems with achieving the grind size and mill operational stability. Once the issues had been resolved, the 3.5 mm media was retrialled. It was found that both media sizes were able to grind the IsaMill™ feed without critical buildup of unbroken coarse material occurring, provided the density was controlled below 40% solids at the targetted specific energies. However, significant differences in the breakage rates and resultant product size distributions of the two media sizes were noted.

Figure 5 compares the 3.5 mm top size media and the 5-6 mm top size media where the same product P_{80} size of 37 µm was produced. The feed size distribution was slightly coarser for the 5-6 mm media case. The 3.5 mm media consumed 16 kWhr/t to produce the P_{80} of 37 µm and P_{98} of 220 µm. The 5-6 mm media required only 11 kWhr/t to produce the same P_{80} of 37 µm but a much finer P_{98} of 110 µm. Therefore, the 5-6 mm media was able to produce the same P_{80} as the 3.5 mm but consumed less energy and produced a tighter size distribution through more effective breakage of particles at the coarse end of the feed distribution.

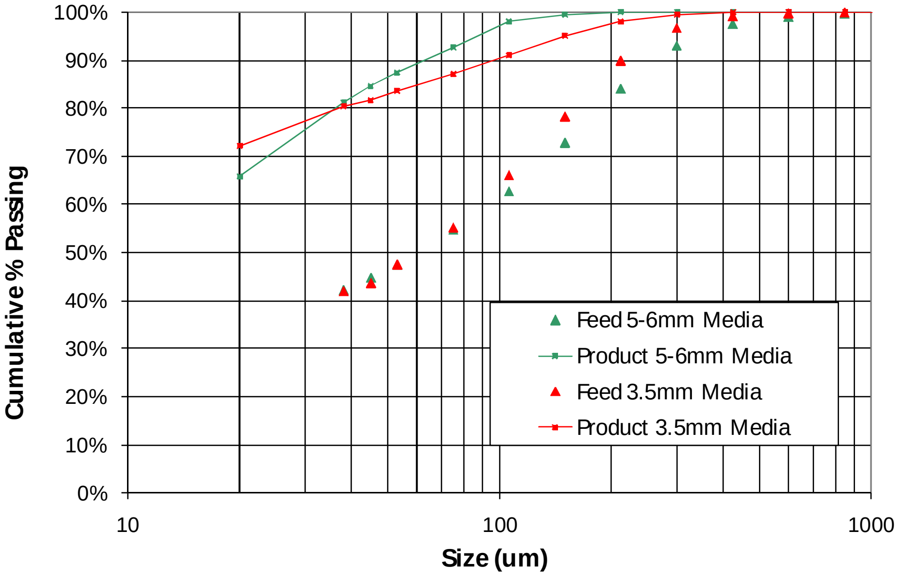


Figure 5 - Performance of 3.5 mm Media at 16 kWhr/t and 5-6 mm at 11 kWhr/t

The 5-6 mm media was able to successfully break the top size of the feed due to its ability to impart greater stress intensity as a result of its increased size (stress intensity is proportional to the cube of the media diameter). As a result, the 5-6 mm media was more efficient at achieving a given product target size. This reinforces one of the key points of stirred milling at coarser sizes – that the top size of the media must be chosen to adequately break the top size of the feed. If the top size is not broken quickly enough, it will remain in the mill leading to higher wear rates, coarser and wider product distributions, lower power efficiency and in worse case scenarios, sanding of the mill.

A second example, in Figure 6, shows the same curve at 16 kWhr/t for the 3.5 mm top size media with the energy input for the 5-6 mm top size media increased to 16 kWhr/t as well. Again, the feed size distribution for the 5-6 mm media was slightly coarser than that for the 3.5 mm media. The product size distribution from the 5-6 mm media (P_{80} of 29 µm and a P_{98} of 75 µm) was superior to that produced by the 3.5 mm media (P_{80} of 37 µm and P_{98} of 220 µm). The improvement was driven by the increased breakage rates at the top end of the distribution.

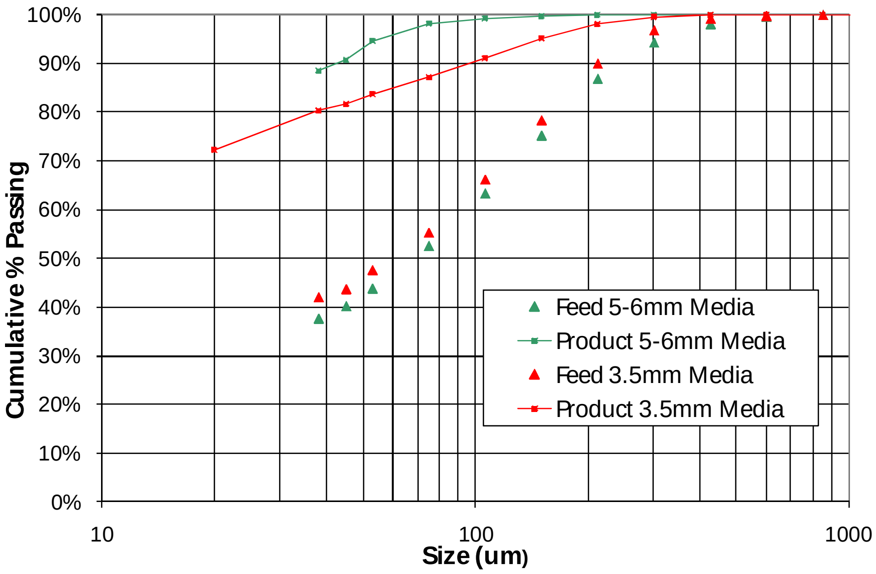


Figure 6 - Performance of 3.5 mm Media at 16 kWhr/t and 5-6 mm at 16 kWhr/t

Based on this data and the fact that it was aimed to increase the target feed size distribution to a P_{80} of +250 µm, 5-6 mm top size media became the standard for this project. This size media has proved adequate in handling the range of feed size distributions fed to the IsaMill™ up to an F_{80} of 350 µm.

The best example of coarse particle breakage was achieved when the feed to the IsaMill™ was significantly coarsened for a two day trial period through adjustment of the sample preparation cyclone. This produced feed distributions with P_{80} values >350 µm. Results of the two surveys taken during this period are included in the Figure 4 data. The distribution curves for one of the data points is shown in Figure 7 at a specific energy of 12.8 kWhr/t.

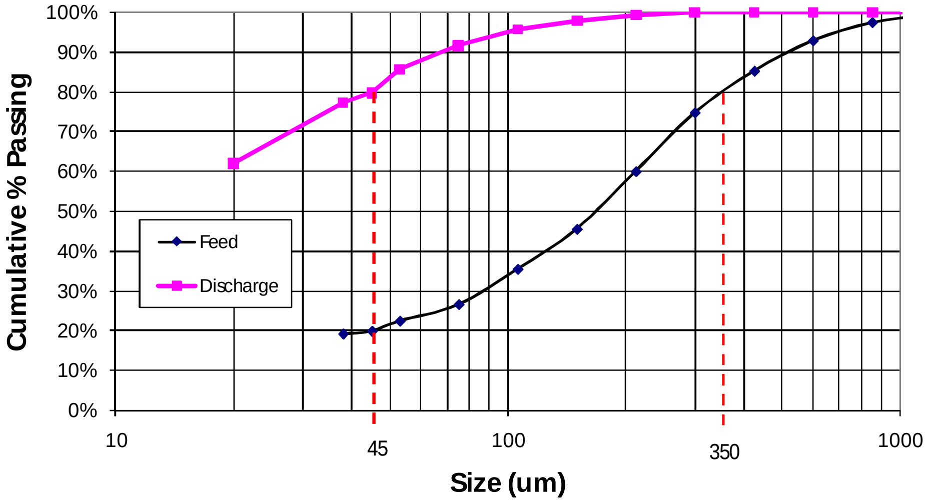


Figure 7 - Size Distributions for F_{80} 350 µm Feed at 12.8 kWhr/t using 5-6 mm media

This can also be compared to the 11 kWhr/t grind for 5-6 mm media (used in Figure 5), for a much finer feed size distribution. This reiterates the point made in the discussion of Figure 4 where the feed size distribution seems to have little impact on the energy required to reach a certain grind size. For a given media size, and provided that size is adequate, the overall energy consumption is largely determined by the target product size, rather than the feed size distribution.

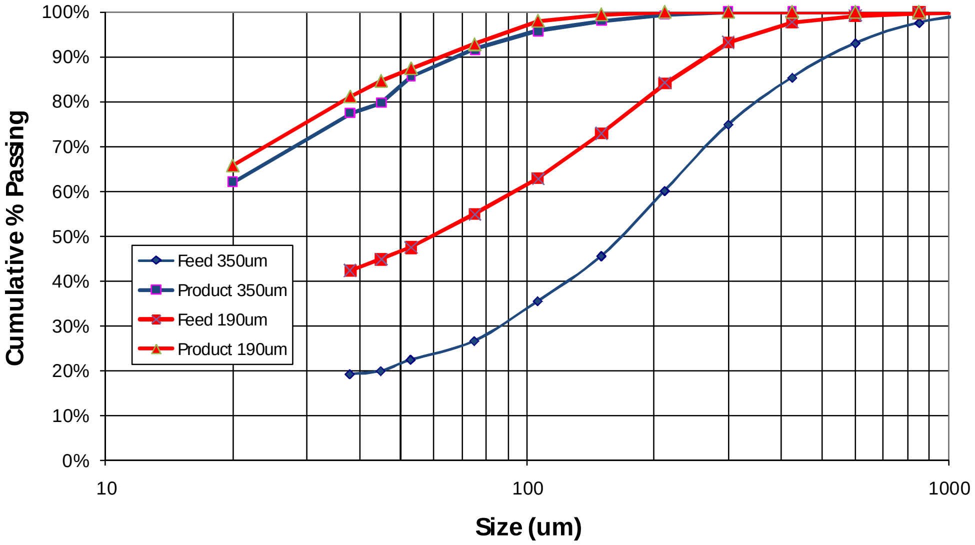


Figure 8 - Size Distributions for F_{80} 350 µm Feed at 12.8 kWhr/t and F_{80} 190 µm Feed at 11kWhr/t

MEDIA DATA

Since the project was commenced a number of ceramic medias from different suppliers were trialled with the aim of optimising performance on a supplier cost/consumption basis. Media wear rate in terms of g/kWh was monitored on a continual basis to allow the different media types to be compared on a consumption basis. Table 1 summaries the data for the 5-6 mm media top size.

Table 1 - Media Performance Data

	Media A	Media B	Media C	Media D	Media E	Media F
Cost $(AUD)/t	$ 4.60	$ 2.70	$ 4.60	$ 4.60	$ 4.60	$ 4.00
Consumption g/kWh	11-14	11-16	11-14	11-14	7-8	10-17
Max $/kWh	$ 0.064	$ 0.043	$ 0.064	$ 0.064	$ 0.037	$ 0.068
Relative Max Cost	1.00	0.67	1.00	1.00	0.57	1.06

The data gathered has not been able to discern any quantifiable information with regards to the effect of the different media types on grinding efficiency or mill internal wear. It is likely that these factors do exist but the changes have not been significant to be highlighted within the noise of the data obtained here and at the relatively coarse grind sizes. Xstrata Technology has previously conducted numerous controlled laboratory scale tests that highlight the effect of different media types on grinding efficiency. These effects are also more pronounced at finer (<20 µm P_{80} target) sizes than were targetted here.

COMPONENT WEAR RATES

In the ultra fine grinding duty at McArthur River, IsaMill™ inspections are typically at 2,000 hours. Typically a shell liner lasts 2,000 hours, feed flange lasts 10,000 hours, discharge flange lasts 15,000 hours and 3-4 of the 8 discs are replaced each inspection. While typical component lifetimes are longer - eg shell liner life is 6,000+ hours at Mt Isa – it is lower at MRM due to the high operating temperatures involved and the inability to rotate the shell liner due to the old style shells used. Newer IsaMills™ have a shell liner that can be rotated to maximise the rubber wear, therefore extending the useable life of the liner.

Initial wear rates of the IsaMill™ internal components in the primary grinding duty were significantly higher than UFG duties. The change in wear rates is a function of the feed size distribution and the presence of more gangue mineralogy - particularly silicates. In this case the feed size distribution increased from an F_{80} of approximately 45 µm to 200-350 µm and the silicate content of the ore is about 26-35% compared to 17-20% in the UFG duty. It should be noted that the use of coarser grinding media, whilst impacting the liners with a higher stress intensity due to its larger size, actually assists to reduce the amount of time the coarsest particles spend in the mill and therefore limit their influence on wear. This was demonstrated during the change from 3.5 mm to 5-6 mm media where it was assumed that the wear rates would increase due to the coarser media presence, however the overall wear rate remained the same due to the reduction in time the coarsest particles spent in the mill. Figure 9 shows some data from another IsaMill™ treating two separate feed types. The media size was the same in both cases but the change in feed size from 80% -75 µm to 40% -75 µm resulted in a nearly 5 fold increase in disc consumption. It is likely that larger media would have reduced the overall wear rate in the mill at the coarser feed size distribution.

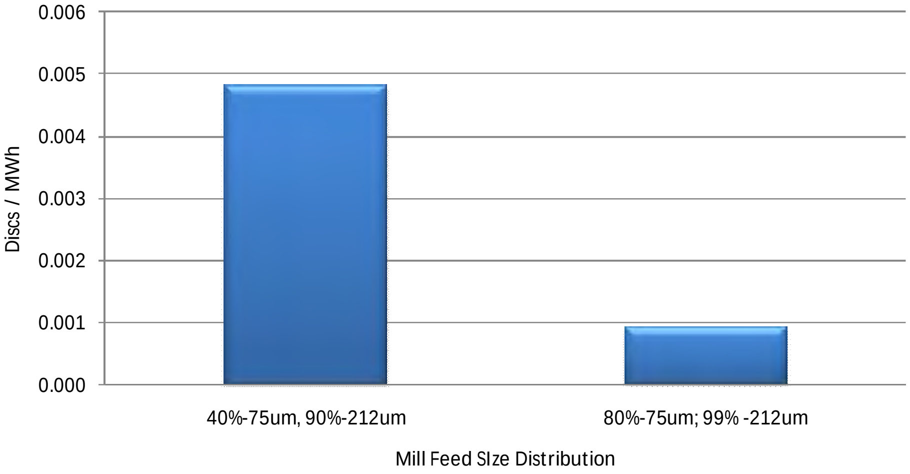


Figure 9 - Effect of Feed Size Distribution on Disc Consumption

In the initial primary grinding trials, the MRM M3000 IsaMill™ was serviced every 400 hours and usually required replacement of the shell liner as well as a complete set of discs. While it was still proving beneficial to the plant to operate the IsaMill™ in this duty there was a real need to reduce the maintenance cost and time.

Based on successful development work for the M10,000 IsaMills™ in South Africa (Rule & de Waal, 2011), a trial using reduced diameter discs (1,050 mm compared to standard 1,200 mm) was made on the M3000 IsaMill™ at McArthur River. The first 7 discs were replaced by the smaller discs, the final disc was left at 1,200 mm so as not to influence the product separator performance. The mill was able to run for 1,200 hours between inspections where a full disc and liner change was required. The drawback was a (not unexpected) reduction in power of about 35% as a result of the reduced diameter, limiting the power draw to about 650-700 kW maximum. The reduction in power draw limited the throughput by the same percentage in order to maintain the same grind performance. The disc consumption per MWh (Figure 10) reduced by more than half, however is still significantly higher than the data presented in Figure 9. Measures to address the power draw issue are discussed later in the paper.

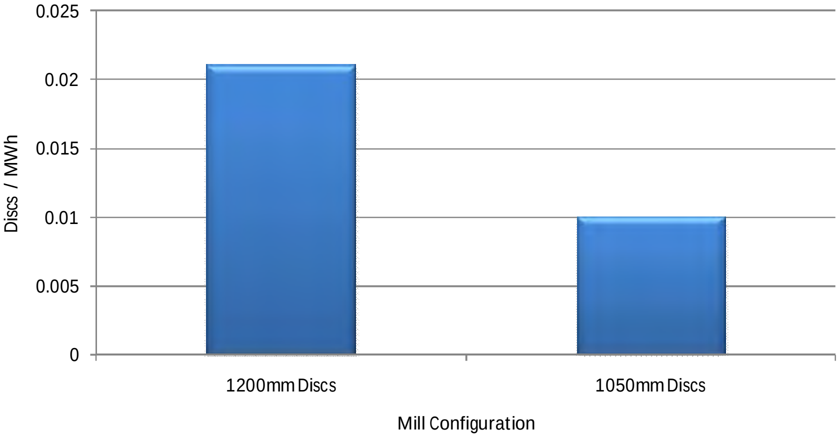


Figure 10 - MRM Disc Consumption Reduction per MWh after Mill Configuration Change

3.0 MW M10,000 IsaMills™

Based on the positive results from the M3000 IsaMill™ testwork and subsequent conversion to full time production, MRM made the decision to purchase and install two 3.0 MW M10,000 IsaMill™ units with the intention that they be configured in the primary grinding duty for the plant expansion. During the project construction phase, the M3000 development work continued. For a variety of reasons including that a lot of development work on the coarse grind had been done on the M3000 IsaMill™ and at that stage there wasn't scope to go through that learning experience again on the M10,000 IsaMill™, it was decided that the M10,000 IsaMills™ would be commissioned into a reconfigured UFG circuit (November 2008). This provided extra UFG capacity for the increased plant throughput and allowed free use of two of the M3000 IsaMills™ for the primary grinding duty.

Only one of the M10,000 IsaMills™ can be currently operated due to power supply limitations at site. The M10,000 currently treats all of the rougher concentrate (precyclone underflow) prior to grinding in the M3000 UFG IsaMills™ to produce the 7-8 µm cleaner feed.

Figure 11 - 3.0 MW M10,000 IsaMills™ at McArthur River

CURRENT PRIMARY GRINDING CIRCUIT PERFORMANCE

Currently, two 1.1 MW M3000 IsaMills™ are used in the primary grinding duties. Each of these mills process 40-60 tph at 80-95 m3/hr. The mills are operating with the reduced diameter discs to extend the service life intervals to 1,200 hours. The operation of both coarse grinding M3000 IsaMills™ allows plant throughput to be increased from 260 tph to 330 tph – a 27% increase in primary tonnage at reduced rougher flotation feed size distribution.

Summarised data from a recent survey across the entire primary grinding circuit is included in Table 2 where the circuit was configured as per Figure 2. The data shows the plant operating at 330 tph with both primary grinding IsaMills™ on line operating at 15 kWhr/t producing a product P_{80} sizing of sub 40 µm from a feed F_{80} of around 250 µm.

The primary grinding circuit is currently constrained by the dewatering capacity of the original double deck SAG discharge screen. The large recirculating load in the circuit is diverted away from the SAG mill and consequently the SAG discharge screen via the tower mill into the cyclone feed sump. The power to tonnage ratio of the tower mill is very low and the size reduction is minimal – however it plays a significant role in this circuit by effectively bypassing slurry around the SAG discharge screen. As a result, when the tower mill is offline for shoe replacements, there is a plant throughput decrease of 60 – 70 tph required to ensure that the screen deck does not become dewatering constrained. New trial pipe work is currently being installed to allow a bypass of the tower mill (during maintenance) – this will allow an evaluation of the real impact of the tower mill on the overall throughput and grind size.

Table 2 - Primary Grinding Circuit Survey Summary

	P_{80}	tph	kW	kWhr/t*
Fresh Feed to SAG Mill	6377	323	3800	11.76
SAG Cyclone Feed	586	1548		
SAG Cyclone OF	87	245		
SAG Cyclone UF	582	1303		
Tower Mill Feed	582	1187	950	0.80
Tower Mill Discharge	571	1187		
1st IsaMill™ Feed	253	39.6	600	15.15
1st IsaMill™ Discharge	43	39.6		
2nd IsaMill™ Feed	247	38.7	600	15.51
2nd IsaMill™ Discharge	36	38.7		
Total Rougher Feed	76	83.4		
Total		323	5950	18.41

*kWhr/t is based on the fresh feed tonnage actually processed through the individual unit.

FUTURE DEVELOPMENT OF THE PRIMARY GRINDING IsaMill™ AT MCARTHUR RIVER

Experience under the extreme grinding conditions of the McArthur River circuit has improved the understanding of wear in the IsaMill™ and the interactions of feed size distribution, feed density, media size, disc design, disc tip speed etc. These learnings have been adapted to other duties to improve wear life in all IsaMill™ applications – UFG, regrind and primary mainstream grinding. Work is currently underway on a number of initiatives to further improve the wear performance at McArthur River.

Laboratory testwork of a redesigned IsaMill™ disc has shown a 25% increase in power draw. Several of these discs will soon be trialled at McArthur River with the aim of restoring the mill to its full 1 MW power draw, thereby allowing further throughput increases using the M3000 IsaMills™.

Investigations into retrofitting the latest design M5000 (5,000 litre) IsaMill™ shell to the existing M3000 layout are underway. This would permit the mill to operate with the standard 1,200mm diameter discs at lower wear rates and regain the full power draw potential. The M5000 was developed as a result of lessons learned from the McArthur River experience. It is aimed at several market areas but its key advantage is in its geometry which allows it the ability to treat coarse feed distributions and draw up to 1.5MW without the high wear issues experienced in the M3000, which was specifically designed for UFG applications.

FUTURE DEVELOPMENT OF THE PRIMARY GRINDING CIRCUIT AT MCARTHUR RIVER

Xstrata Zinc has recently announced a proposed Phase 3 Development for McArthur River which includes increasing the processing capacity to 5 mtpa. The primary grinding circuit options for this increase are currently under investigation - with 3.0MW M10,000 IsaMills™ included in each of the cases under consideration. The M10,000 IsaMills™ will be used in either a secondary or tertiary duty to take F_{80} 150 µm feed to the rougher flotation feed size of P_{80} 45 µm. These mills will be configured in a similar manner to the two existing M10,000 mills, providing operational options for either primary or regrind duty. The project is scheduled to commission in 2014.

CONCLUSIONS

Successful laboratory, pilot scale and full scale testwork has seen the incorporation of IsaMill™ technology into the primary grinding circuit at McArthur River. The IsaMill™ has now been part of the primary grinding circuit for more than four years. The use of two IsaMills™ in the primary circuit has enabled a 27% increase in primary grinding throughput.

Significant improvements were made to the IsaMill™ wear performance albeit at some sacrifice of power draw and mill throughput. Initiatives are underway to recover the lost power draw without losing all of the gains made in wear performance. Learnings from this application have improved understanding of wear mechanisms and have led to improved designs for all IsaMill™ applications.

The success of the work is reflected in the fact that additional 3.0MW M10,000 IsaMills™ are under consideration for use in the primary grinding circuit of the upcoming expansion to 5mtpa processing capacity.

ACKNOWLEDGEMENTS

The authors wish to thank all those at McArthur River and Xstrata Technology involved in the original testwork and plant scale implementation associated with this project and for permission to publish the data.

NOMENCLATURE

kWhr/t = specific energy

REFERENCES

Anderson, G.S. & Burford, B.D. (2006). *IsaMill – The crossover from ultrafine to coarse grinding*, in Proceedings MetPlant 2006, Perth, Australia.

Anderson, G.S. (2006a). *SAG cyclone underflow grinding: M20 pilot plant – stage 1.* (Internal Technical Report). Brisbane, Australia.

Anderson, G.S. (2006b). *SAG cyclone underflow grinding: M20 pilot plant – stage 2.* (Internal Technical Report). Brisbane, Australia.

Buys, S., Rule, C. & Curry, D.C. (2005). *The application of large scale stirred milling to the retreatment of Merensky platinum tailings,* in Proceedings of the 37th Meeting of the Canadian Mineral Processors (CMP 2005) , Ottawa, Canada.

Curry, D.C., Clark L.W. & Rule C. (2005). *Collaborative technology development – Design and operation of the world's largest stirred mill,* in Proceedings of the Randol Innovative Metallurgy Conference 2005, Perth, Australia.

Curry, D.C. & Clermont, B (2005). *Improving the efficiency of fine grinding,* in Proceedings of the Randol Innovative Metallurgy Conference 2005, Perth, Australia.

Enderle, U., Woodall, P., Duffy, M., & Johnson, N.W. (1997). *Stirred mill technology for regrinding McArthur River and Mount Isa zinc/lead ores*, in Proceedings of the XX IMPC 1997, Aachen, Germany.

Logan, R.G., Leung, K & Karelse, G.J. (1993). *The McArthur River project*, in Proceedings of the International Symposium on Zinc (World Zinc 1993), Hobart, Australia.

Nihill, D.N., Stewart, C.M., Bowen, P. (1998). *The McArthur River mine – The first years of operation*, in Proceedings of AusIMM '98 – The Mining Cycle (1998), Mt Isa, Australia.

Pease, J.D., Anderson, G.S., Curry, D.C., Kazakoff, J., Musa, F., Shi, F. & Rule, C. (2006). *Autogenous and inert milling using the IsaMill*, in Proceedings SAG 2006, Vancouver, Canada.

Rule, C. (2010). *Stirred milling – New comminution technology in the PGM industry*, in Proceedings The 4th International Platinum Conference, Platinum in Transition 'Boom or Bust', SAIMM, Capetown, South Africa.

Rule, C. & deWaal, H (2011). *IsaMill design improvements and operational performance at Anglo Platinum*, in Proceedings MetPlant 2011, Perth, Australia

IsaMill™ Technology in the Primary Grinding Circuit

G Anderson (Xstrata Technology)
D Smith (McArthur River Mining)
S Strohmayr (McArthur River Mining)

IsaMill™ Development - UFG

- Developed to address ultra fine grinding requirements for McArthur River and Mt Isa Pb/Zn deposits
- McArthur River required 7µm grind
- Conventional technology was unsuitable due to poor efficiency and surface contamination
- Looked outside of the mining industry
- Collaboration with Netzsch of Germany – history of manufacturing stirred mills for paint, inks, pharmaceuticals
- 1.1MW M3000 (litre) IsaMill™ developed – became the enabling technology for McArthur River Mine 1995

IsaMill™ Development - UFG

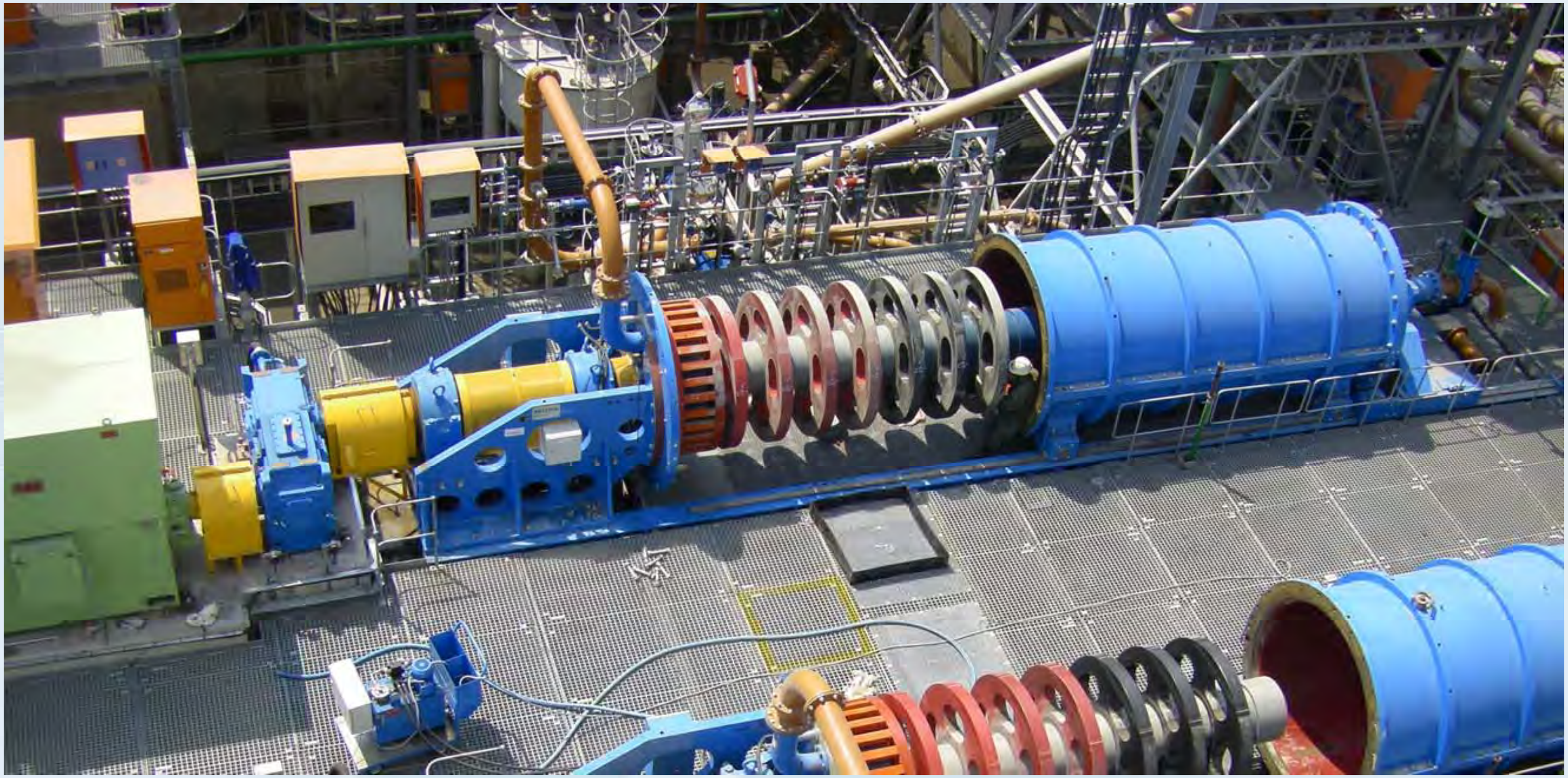
IsaMill™ Development – Coarse Grind

- Requirement for larger scale IsaMill™ identified by Anglo
  - Inert and efficient grinding to improve mineral liberation and recovery
- M10,000 IsaMill™ development with Anglo/Netzsch 2003
- 1st M10,000 in UFG at Western Limb Tailings Retreatment

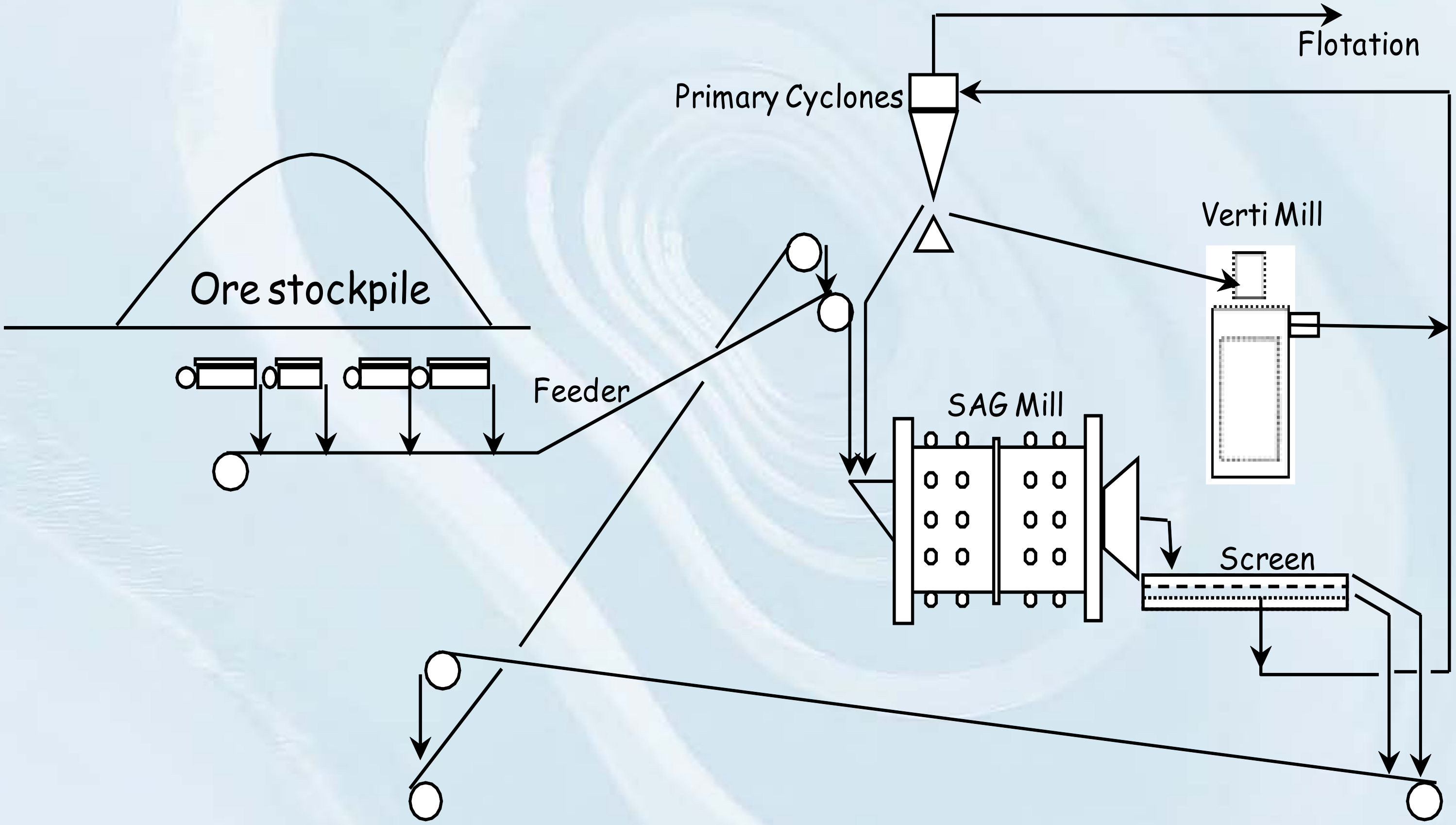
- Development of 3.5mm ceramic media with Magotteaux
- 18 x 3MW M10,000 IsaMills™ installed within Anglo
- F80 of 75-100µm; P80 of 53µm
- Further 4 x 3MW on order for Anglo
- Worldwide over 100 IsaMills™

IsaMill™ Development – Coarse Grind

IsaMill^{TM} Operation

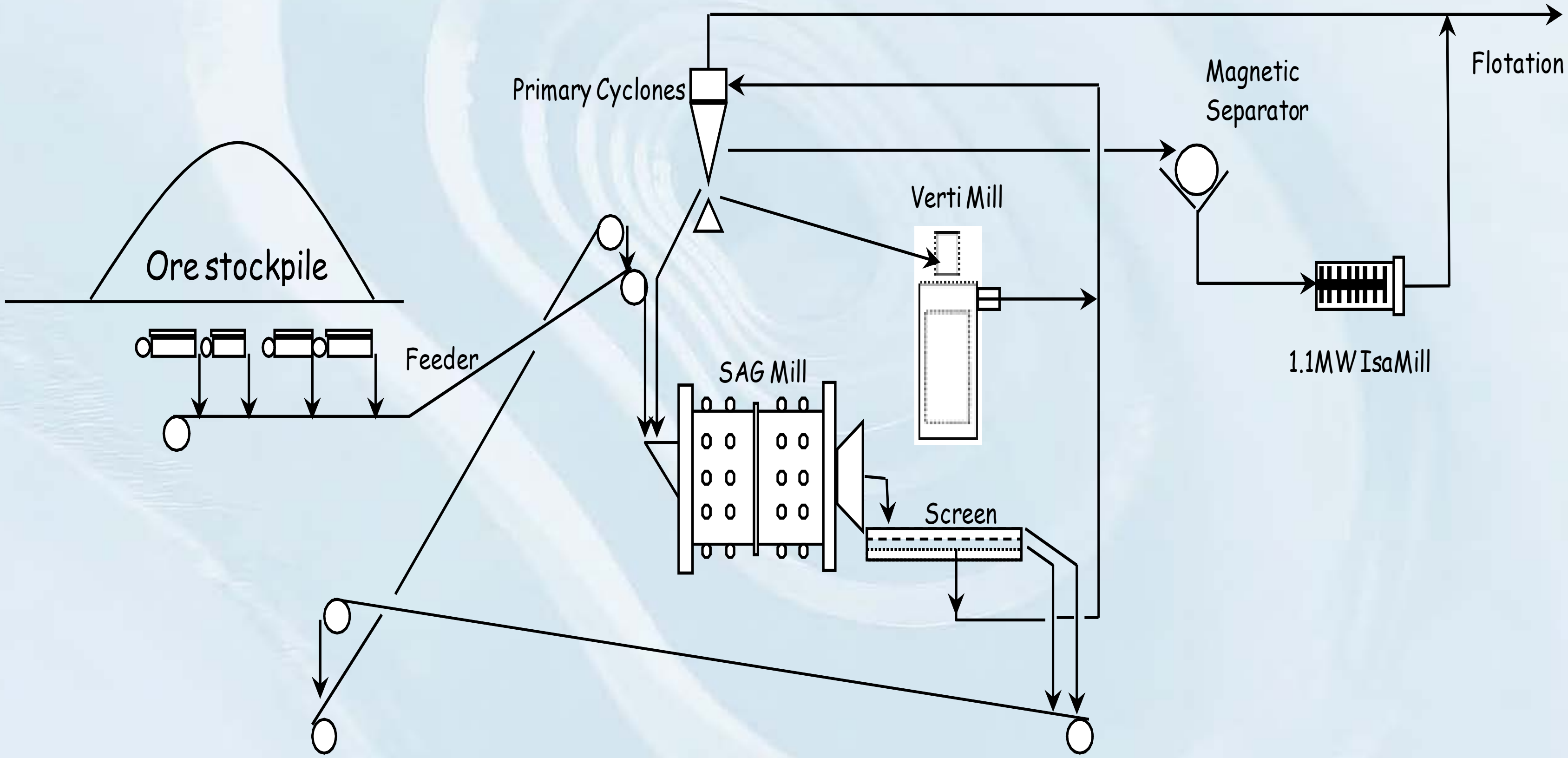


McArthur River - 2006

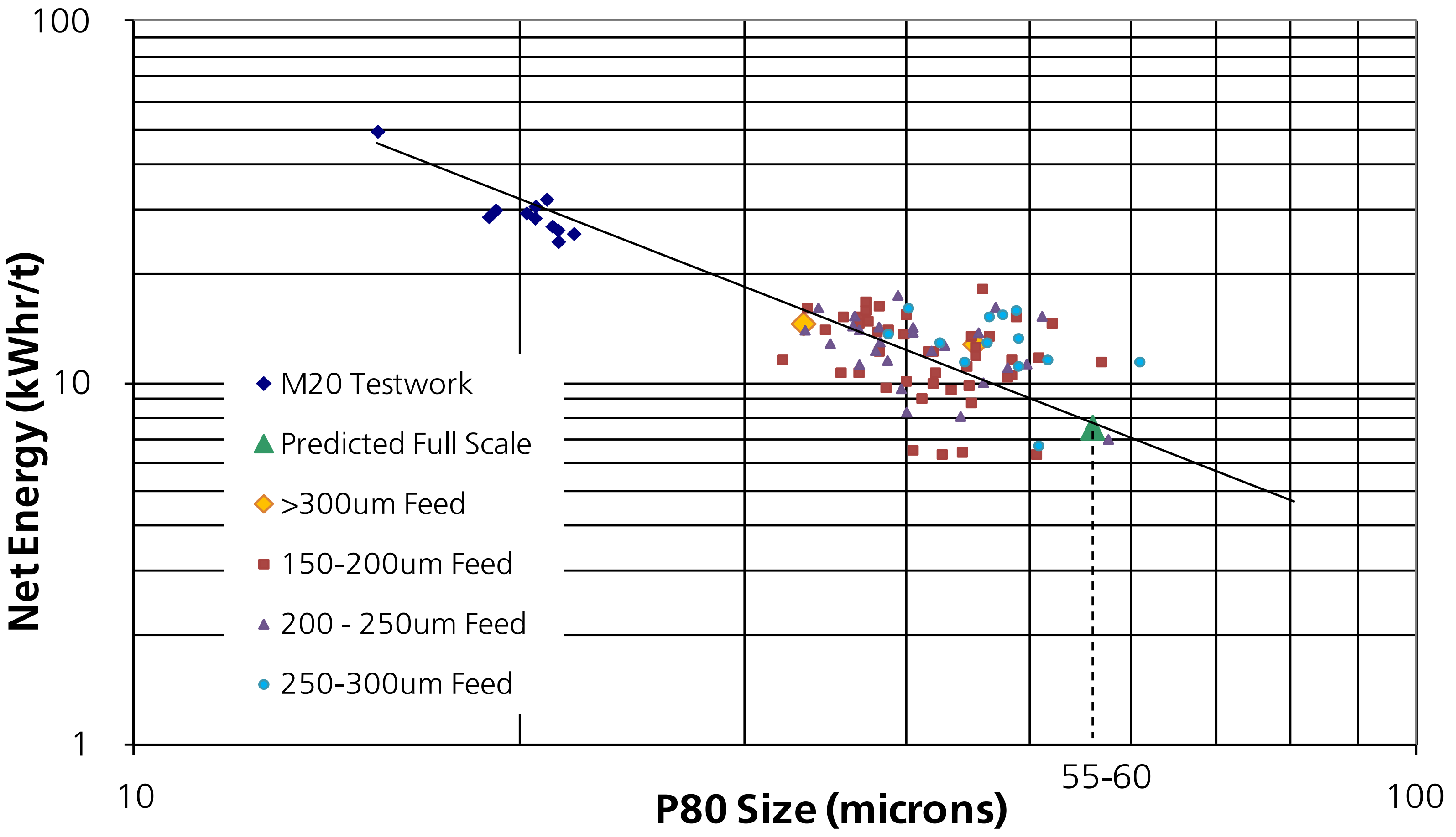


McArthur River – Plant Trial

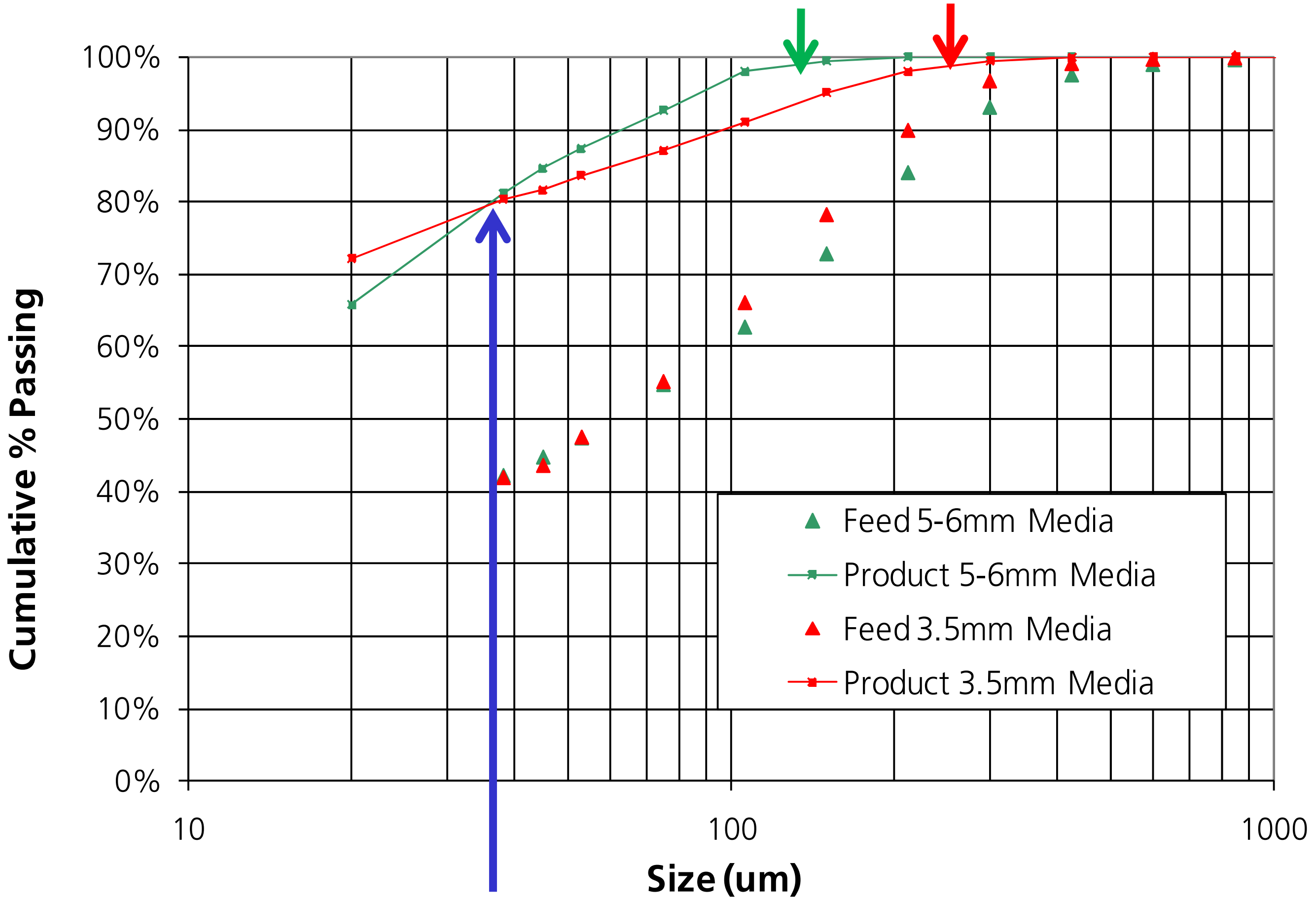
- Plant trial commenced in May 2007



IsaMill™ Survey Data



Effect of Media Size



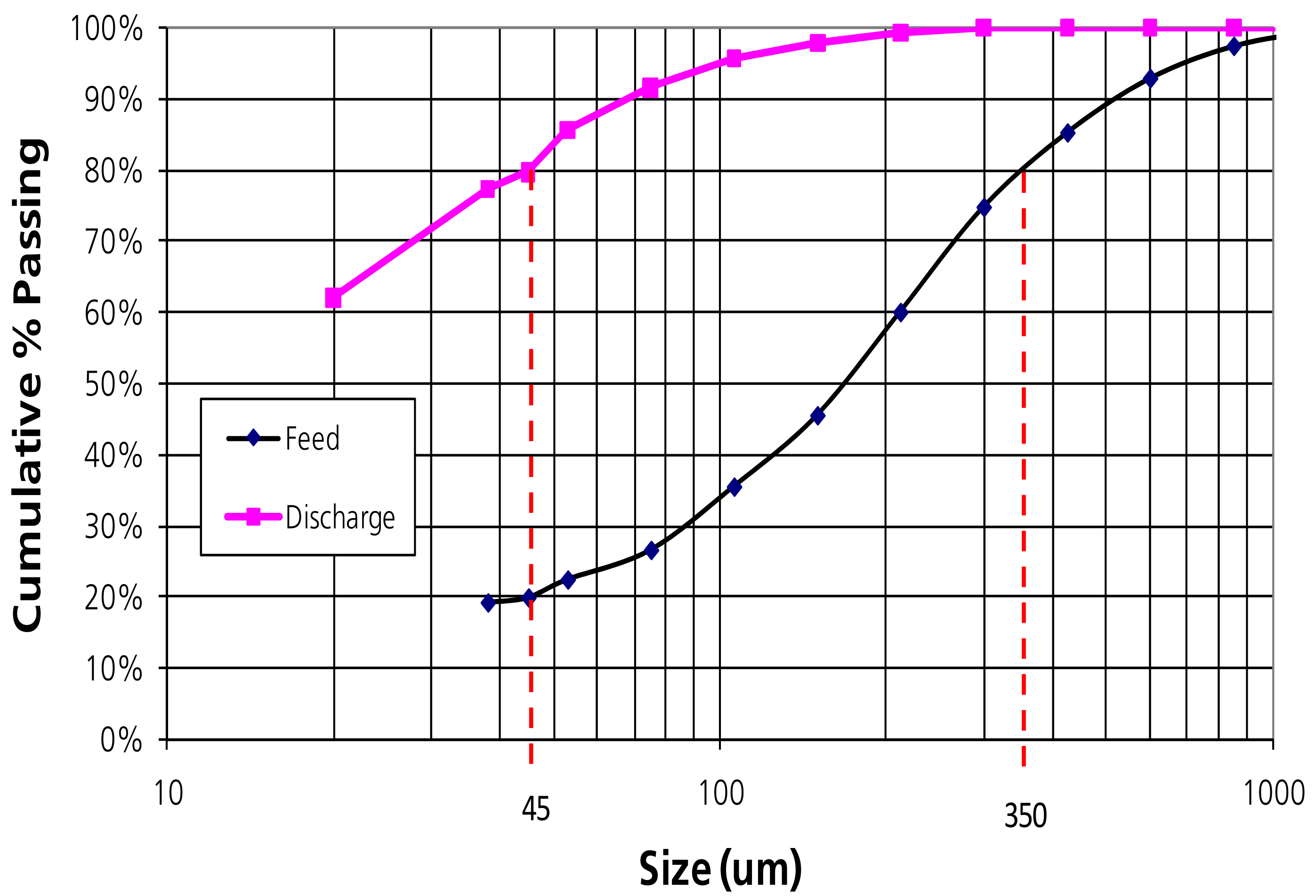
P98 110µm P98 220µm

P80 37µm

11kWhr/t

16kWhr/t

F80 350µm @ 12.8kWhr/t



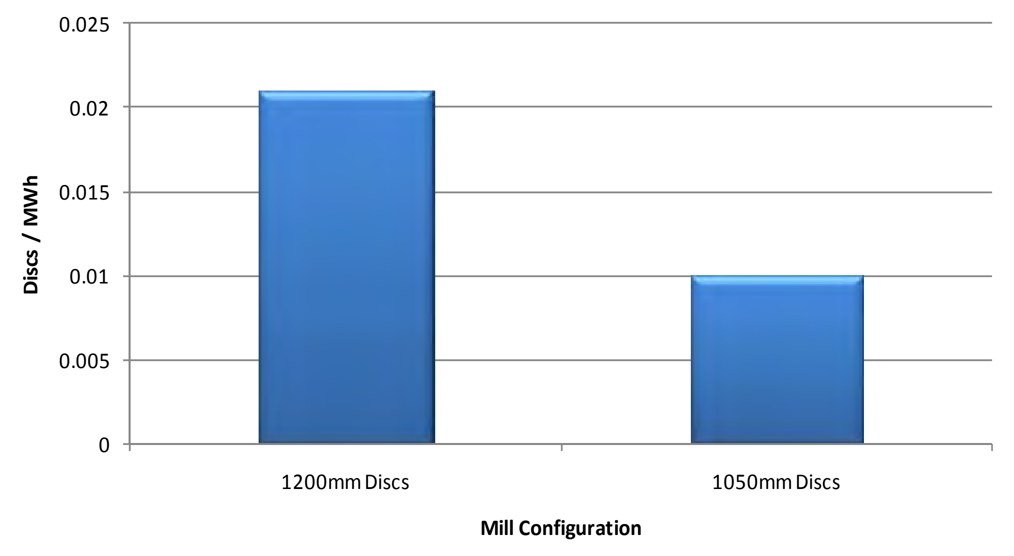
IsaMill™ Wear Performance

- Initial 400 hour service life
- Disc and shell replacement
- Still beneficial to operate - trial became permanent production mill

- UFG mills 2000 hour service intervals
- Higher wear a function of
  - coarser feed/media size
  - mineralogy
  - mill geometry designed for UFG

IsaMill™ Wear Performance

- Smaller Diameter Discs : 1200mm to 1050mm
- Service life increased 400 to 1200 hours



- Max power reduced from 1MW to 650-700kW
- Proportionally limited the throughput to maintain grind performance

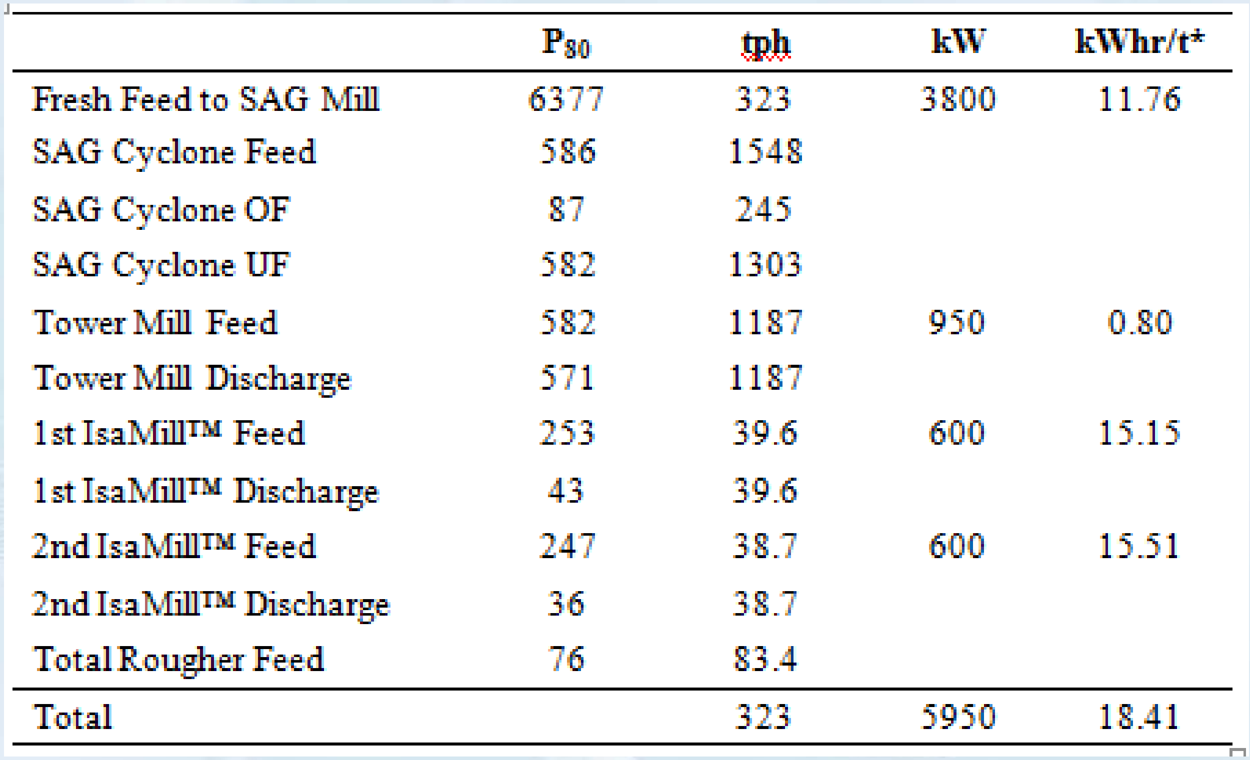
McArthur M10,000 IsaMills™

- Successful M3000 trial – permanent production mill
- 2 x M10,000 installed for plant expansion Nov 2008

- M3000 development work continued
- M10,000 commissioned in UFG circuit - allowed 2 x M3000 for primary grinding circuit

Primary Circuit Performance

- Plant throughput increased by 25% from 260 to 325 tph



	P_{80}	tph	kW	kWhr/t*
Fresh Feed to SAG Mill	6377	323	3800	11.76
SAG Cyclone Feed	586	1548		
SAG Cyclone OF	87	245		
SAG Cyclone UF	582	1303		
Tower Mill Feed	582	1187	950	0.80
Tower Mill Discharge	571	1187		
1st IsaMill™ Feed	253	39.6	600	15.15
1st IsaMill™ Discharge	43	39.6		
2nd IsaMill™ Feed	247	38.7	600	15.51
2nd IsaMill™ Discharge	36	38.7		
Total Rougher Feed	76	83.4		
Total		323	5950	18.41

- With no IsaMill™ case, P80 85µm, OWi = 19.3kWhr/t
- With 2 x IsaMill™ case, P80 76µm, OWi = 18.4kWhr/t

Development

- IsaMill™ Development
  - Alternative disc designs and configurations to recover power draw
  - Shell design – M5000 IsaMill™ - can be retrofit to M3000 IsaMill™
    - Allow geometrical ratios established with SDD in M3000 to be scaled up

- Circuit Development
  - McArthur River Phase 3 Development 2014
  - Increase capacity to 5 mtpa
  - M10,000 included in study for upgraded primary circuit
    - Secondary or tertiary duty ahead of flotation

Conclusions

- Successful lab, pilot and full scale testwork
- IsaMill™ now integral, permanent part of primary grinding circuit - operating for over 4 years
- Improvements made to wear life and performance with further development work ongoing
- Success of the program
  - IsaMill™ permanent part of grinding circuit
  - 2 x M10,000 installation
  - Further IsaMills™ under consideration for the Phase 3 expansion of the primary grinding circuit

Iron Ore 2011

Unlocking the value in waste and reducing tailings: Magnetite Production at Ernest Henry Mining

J Siliézar, D Stoll, J Twomey

Contact Author:
Full name: Jose Gerardo Siliezar
Position title: Magnetite Interface Metallurgist
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Phone: 07 4769 4500
Fax: 07 4769 4555
Email: jsiliezar@xstratacopper.com.au

Iron Ore 2011

Unlocking the value in waste and reducing tailings: Magnetite Production at Ernest Henry Mining

J. Siliezar (1), D Stoll (2), J. Twomey (3)

1.

Jose Siliezar (MAusIMM)
Position title: Magnetite Interface Metallurgist
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Email: jsiliezar@xstratacopper.com.au

2.

Dominic Stoll (MAusIMM)
Position title: Magnetite Project Metallurgist
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Email: dstoll@xstratacopper.com.au

3.

John Twomey (MAusIMM)
Position title: Manager Magnetite Project
Organisation Name: Ernest Henry Mining
Address: PO Box 527 Cloncurry Qld 4824
Email: jtwomey@xstratacopper.com.au

ABSTRACT

Ernest Henry Mining (EHM) is situated 38 km north-east of Cloncurry in the Mount Isa – Cloncurry mineral district of North-West Queensland. The EHM ore body is an iron oxide copper gold deposit with an average grade of 1 %copper, 0.5 g/tonne gold and 23% magnetite, with current reserves of approximately 88 million tonnes.

The copper concentrator is a single line plant with a nominal throughput rate of 1,300 tonnes per hour. An average of 350,000 tonnes of concentrate is produced each year containing around 100,000 tonnes of copper metal and 120,000 troy ounces of gold.

In December 2009 Xstrata Copper announced approval of a $589 million investment to extend the life of EHM to at least 2024, through the transformation of open pit mining operations to a major underground mine together with the construction of an associated magnetite extraction plant.

The EHM magnetite plant extracts the magnetite from the copper concentrator's tailings stream. It will produce approximately 1.2 million tonnes of magnetite concentrate per annum at full capacity for export to Asia, making EHM Queensland's first iron ore concentrate exporter. Construction of the magnetite extraction plant commenced in July 2010 and was completed in January 2011.

This paper gives an overview of the development of the EHM magnetite extraction plant flow sheet from the early mineralogy and laboratory test work through to the plant commissioning and early operation of the base plant. It has a major focus on the commissioning learnings from the plant start-up and includes a review of the operational performance of the plant versus original design and expectations.

INTRODUCTION

In December 2009 the Xstrata board announced the approval of $589 million to transform EHM into a major underground mine with an associated magnetite extraction plant which will extend the mine life to 2024. This will create 400 permanent jobs from 2013 in addition to the jobs created during the construction phase. The transformation from an open pit to a large scale underground operation will enable the extraction of 6M tpa of ore at full capacity to produce 1.2 Mtpa of tonnes of premium quality magnetite concentrate. The first exports of magnetite concentrate are planned for shipment from Townsville in the first half of this year.

The construction of the $79 million magnetite base plant was completed in January this year followed by a period of process commissioning. As part of the base plant Xstrata is developing an $8.6 million expansion of its 65 kt enclosed storage shed at the port of Townsville. While magnetite has traditionally been discarded as tailings at EHM, the magnetite extraction plant allows it to be captured as an important by-product of the copper-gold concentrating process, reducing the amount of tailings sent to the on-site storage facility. The production of magnetite concentrate enables EHM to maximise the value of its existing resources by creating an additional by-product, reduce the amount of tailings in the on-site storage facility and provide further employment opportunities in the region. The magnetite concentrate produced at EHM will be chiefly used to fuel Asia's steel industry, however, the commodity may also be utilised as a washing agent in coal operations.

The plant has 3 key circuits including an extraction plant which separates and upgrades the concentrator tailings to a marketable concentrate, a dewatering plant to remove the water from the magnetite concentrate for transport and a regrind circuit to maximise the recovery from concentrator tailings. The overview of the magnetite extraction process is shown in Figure 1.

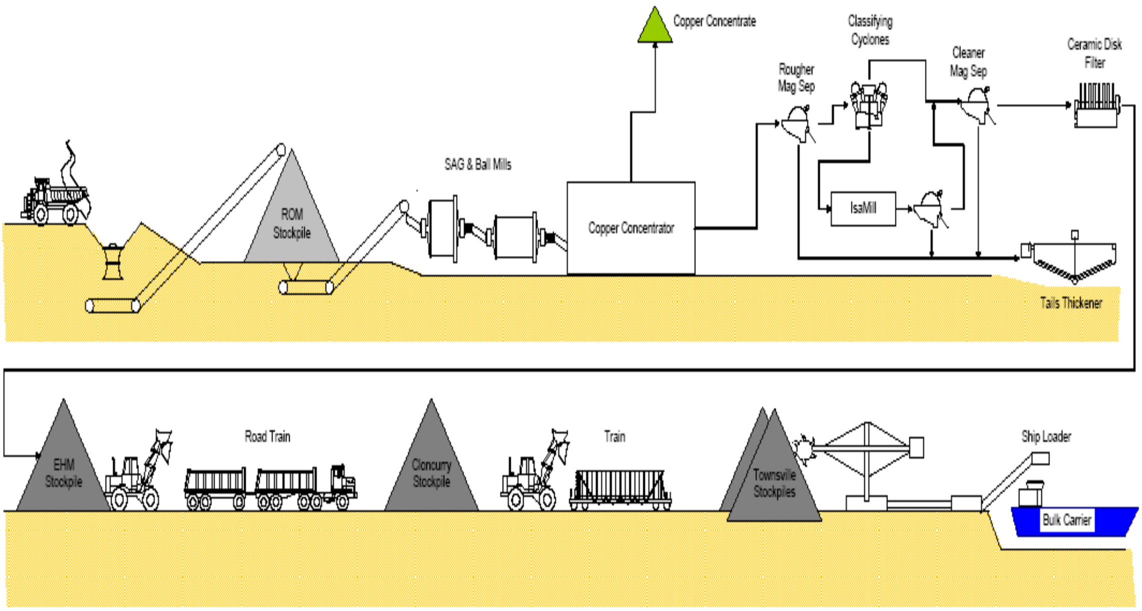


Figure 1: Overview of the EHM magnetite extraction process

MINERALISATION AND ORE RESERVES

The Ernest Henry ore body is an iron oxide copper gold deposit (IOCG) discovered in October 1991 by airborne geophysics. Similar examples are Olympic Dam and Prominent Hill in South Australia, and Osborne in Queensland. It is located in the Eastern Fold Belt of the Mount Isa Inlier under approximately 50m of sand and clay cover. The mineralisation at Ernest Henry has formed in a south-east plunging body of altered and variably brecciated felsic volcanic rock. The combined thickness of the mineralised sequence is approximately 250 m, width averages 300 m and the down dip length is approximately 1,000 m in a pipe like formation which dips 40 degrees to the SSE and is still open at depth (Ryan, 1998).

The primary ore mineralogy is dominated by chalcopyrite within a magnetite-carbonate gangue. The mean magnetite content of the primary ore is between 20 and 25 wt%. The copper grade increases with increasing magnetite grade (EHM Feasibility, 1995).

The Ernest Henry underground deposit is known to extend to at least 400 m below the open pit final stage 7 (575 m deep, 1.5 km x 1.3 km). In June 2010, lithology caving estimates were given at 76 million tonnes containing 1.3 %Cu, 0.7 g/t Au and 28% FexOy which is made up of measured and indicated resource, and 13 million tonnes containing 1.2 %Cu, 0.6 g/t Au and 26% FexOy of inferred resource. The width of the mineralised deposit is 200 m. The overview of the underground ore body is shown in Figure 2.

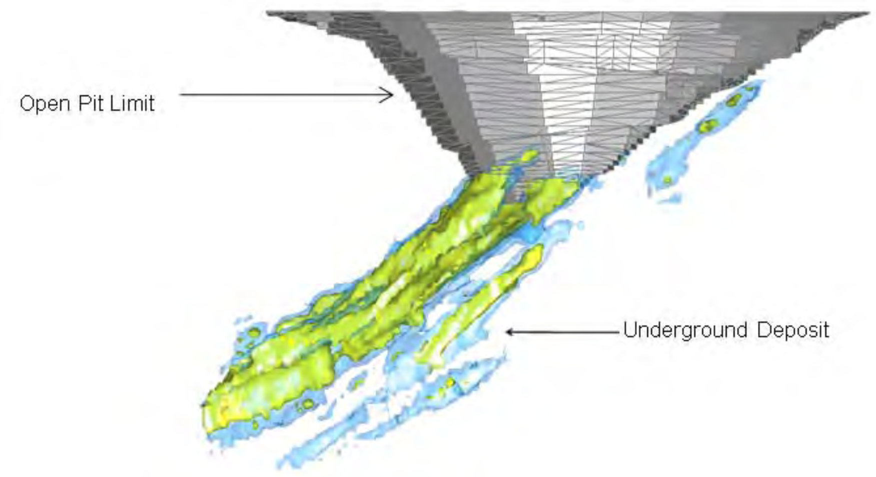


Figure 2: Overview of the EHM underground deposit

The dominant iron oxide species in the ore is magnetite; varying significantly from 4.6% to 46.7%. Historical plant averages from the open pit are 17.6 %. Monthly composite analyses of the final concentrator tail for the previous two years are in average 25% magnetite. Magnetite and hematite are well liberated (Middleditch, 2008) showing minimum association with chalcopyrite in the rougher tailings.

TESTWORK OVERVIEW

There have been numerous investigations into magnetite processing at EHM since operations started in 1997, with emphasis on how it can be removed, both for product development, and to assist in reducing the circulating load in the grinding circuit. Early studies using an onsite pilot plant in conjunction with large scale laboratory test work at CSIRO demonstrated that suitable product grades and production rates could be achieved These test programs demonstrated that magnetite production via magnetic separation was viable, but limited by impurity grades and liberation (Zhu et al 2007). Most of the programs contained either a comminution or impurity reduction stage. Where grinding was not employed, the yield achieved at acceptable product quality was very low. The results of the historical studies are shown in Table 1.

These early testwork campaigns showed that the magnetite in the EHM comminution circuit is preferentially milled due to the hydraulic rather than size-based nature of the magnetite in the primary classification circuit of the copper concentrator (Zhu et al 2004) . The consequence is that the classifiers treat fine magnetite in exactly the same way as coarser gangue, concentrating it to the cyclone underflow resulting in preferential liberation of magnetite in the ball mill. Practically, due to the relatively liberated nature of the feldspar and magnetite, the hydro- cyclones operate with two independent efficiency curves (one for magnetite and one for feldspar), which overlap as a function of co-liberation of the two minerals (Zhu et al 2006).

Table 1
Historical testwork results

Program	Tailings Sample		Product								
	Grind P80	Yield (%)	Fe (%)	Cu (%)	P (%)	S (%)	Al_2O_3 (%)	As (%)	SiO_2 (%)	K_2O (%)	Na_2O (%)
EHM 2001 Laboratory	No	0.60	66%	0.23	N/A	1.11	1.0	0.010	7.4	0.63	0.03
CSIRO August 2004 Pilot Run 1	No	47.2	34.7	0.23	N/A	1.11	6.9	0.010	30.4	N/A	N/A
CSIRO August 2004 Davis Tube	39µm	28.8	70.0	0.02	0.004	0.10	0.36	0.003	1.64	0.21	0.05
IMO Nov 2004 Pilot Phase 5	No	20.3	57.0	0.07	NA	0.20	2.6	NA	14.2	N/A	N/A
CSIRO Feb 2007 Laboratory	38µm	20.5	68.9	0.02	0.005	0.08	0.6	N/A	2.5	0.31	0.05
CSIRO June 2008 Laboratory	35µm	23.3	66.8	0.02	0.007	0.13	1.2	N/A	4.8	0.66	N/A

In 2008, after assessing the strengths and weaknesses of the previous programs, further metallurgical test work was conducted by Xstrata as part of a feasibility study to demonstrate if an iron ore concentrate meeting steel industry specifications could be produced at a reasonable yield by controlling particle size distribution. The work was completed at laboratory pilot scale, and used the laboratory flowsheet shown in Figure 3.

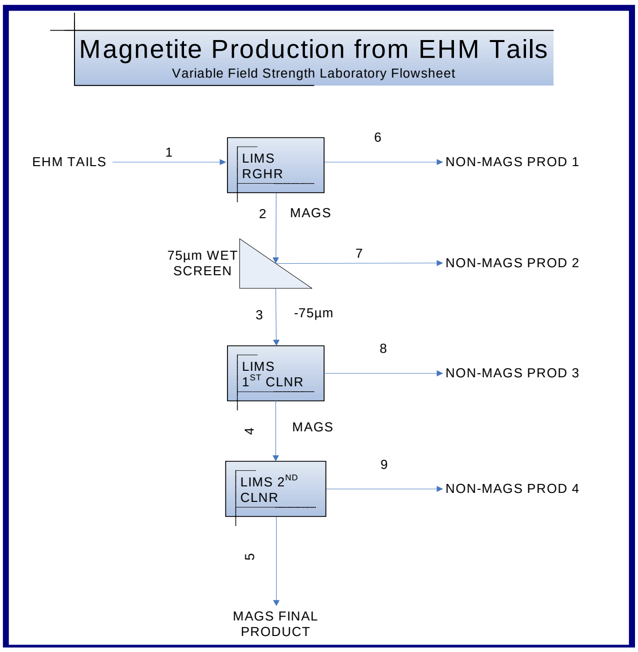


Figure 3 - Final design metallurgical testwork flowsheet

The metallurgical testwork completed confirmed that magnetite concentrates containing 69% iron with minimal impurities can be produced with suitable mass recoveries by controlling particle size distribution of the magnetite product (Magee, 2009).

In addition extensive mineralogy was completed on test products, and demonstrated that the majority of the rougher magnetic separator losses were in the <20wt% Fe-Oxides liberation class. The majority of the losses from the cleaner magnetic separators were in the form of hematite, which is co-reported with magnetite as Fe-Oxides, but is non-magnetic. Approximately 15-20% of the Fe-Oxides in the EHM final tailings are hematite, rather than magnetite. The hematite is present in rims or inclusions around the partially liberated magnetite particles as shown in Figure 4.

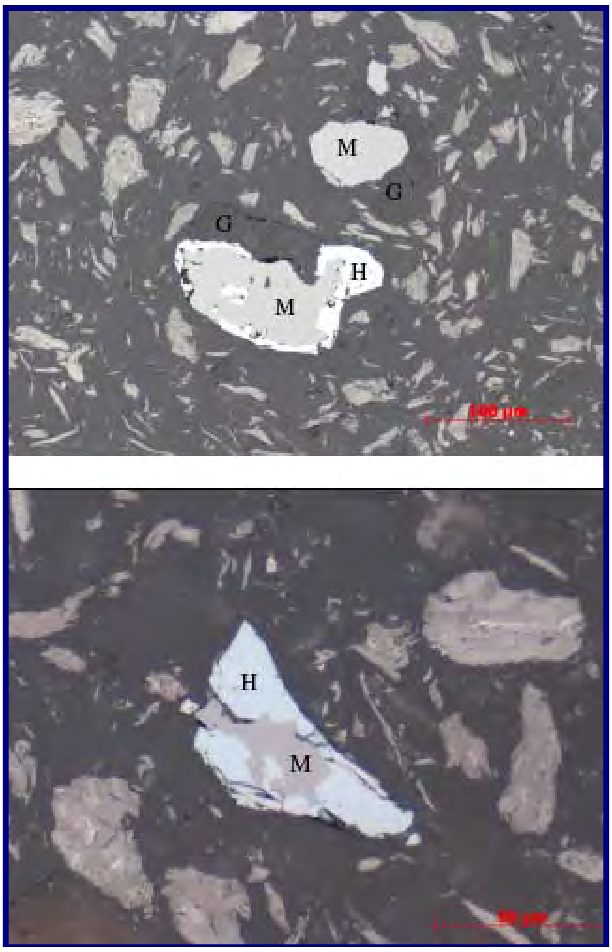


Figure 4 – Hematite rimming of magnetite particles

H = Hematite
M = Magnetite

In the 2009 testwork hematite represented the majority of the Fe-Oxide losses. Hematite recovery cannot be practically improved and hematite as it is not magnetic and will decrease with fine grinding due to derimming of the composite particles. Evidence of this exists throughout the circuit, but is clearest in the rougher non-magnetics stream, where 90% of the Fe Oxides that are lost are actually hematite rather than magnetite (Magee, 2009).

FLOW SHEET DEVELOPMENT AND EQUIPMENT

In order to benchmark the proposed flowsheet and to understand the process risks and plant operation site visits to La Candelaria and Los Pelambres in Chile were undertaken. These sites had similar magnetic separation and filtration technology to that being considered for the project. Ceramic disc filters were observed to be working effectively in concentrate filtration applications at these sites. The magnetic separation plant was observed to have poor process control but was achieving the desired product quality and demonstrated the robust nature of the magnetic separators to a range of operating conditions.

The magnetite plant is divided into 3 circuits. The extraction (magnetic separation) and dewatering (filters) form the base plant (which has been commissioned). The regrind circuit which is due for SMP construction in June is scheduled for process commissioning in August this year. A flow sheet of the base plant is shown in Figure 5.

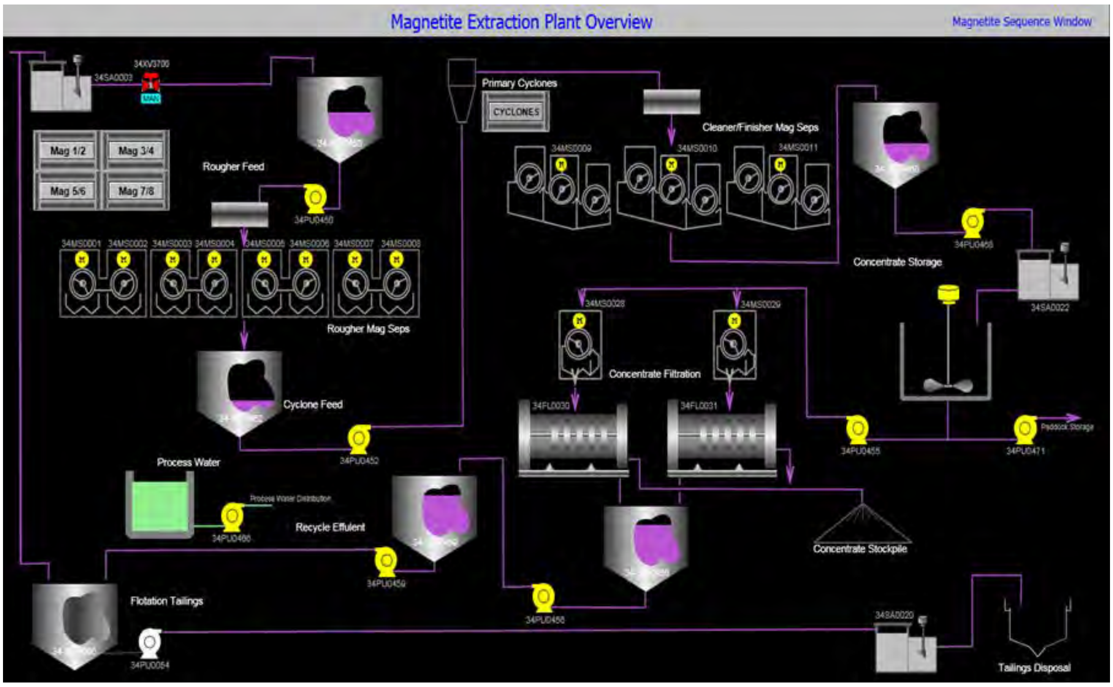


Figure 5 –Magnetite extraction plant flowsheet for Ernest Henry Mine in Australia

Base Plant

The base plant is a simple beneficiation process that uses magnetic separators to extract the magnetite form the copper concentrator tailings a detail of the process parameters is shown in Table 2. It is divided into the following circuits:

- Rougher magnetic separator circuit that processes raw flotation tailings
- Primary cyclone cluster that separates out particles less than 75 µm
- Cleaner/finisher magnetic separator circuit that upgrades the rougher magnetic concentrate
- Dewatering circuit that removes the excess water from the recovered magnetite

Table 2
EHM base plant process parameters

Circuit	Process Inputs	Process Outputs	Equipment Details
Rougher Circuit	Throughput: 721 t/h-1350 t/h Magnetite content: 14-24% Feed solids: 35-42%	Mass yield: 30% - 33%	Rougher units constrains (m drum h): Magnetic loading: 26 t Volumetric loading: 130 m^3
Classification Circuit	Throughput: 220 t/h- 525 t/h Feed solids: 33%	Product target: P_{99} 75 µm Mass yield to COF: 46% - 47% Product solids: 20%	Cyclone diameter: 400 mm No of cyclones: 8 Operating pressure: 80-120 KPA
Cleaner Circuit	Throughput: 95t/h- 230 t/h Feed solids: 20%	Mass yield: 97% - 99%	Cleaner units constrains (m drum h): Magnetic loading: 26 t Volumetric loading: 105 m^3
Dewatering Circuit	Throughput: 75/h- 207 t/h Feed solids: 60%	Moisture content: < 8%	Filtration rate: 1.187 t/m^3 Filtration area: 163 m^2

The purpose of the rougher magnetic separation circuit is to separate the coarse magnetite material from the copper flotation tail. The circuit treats the material using four (4) twin magnetic separation drum units. The rougher magnetic separators are permanent low intensity magnets with a magnetic strength of 1000+ Gauss. Each unit is composed of two counter-current wet drum separators running in parallel with one another and fitted with respective launders for concentrate and tailings discharge. The concentrate produced from the separator consists mostly of fully liberated magnetite with some unliberated magnetite and middlings. The rougher magnetic separator recovers 78% of the magnetite from the rougher feed. Figure 6 shows the design and operating conditions of the rougher circuit.

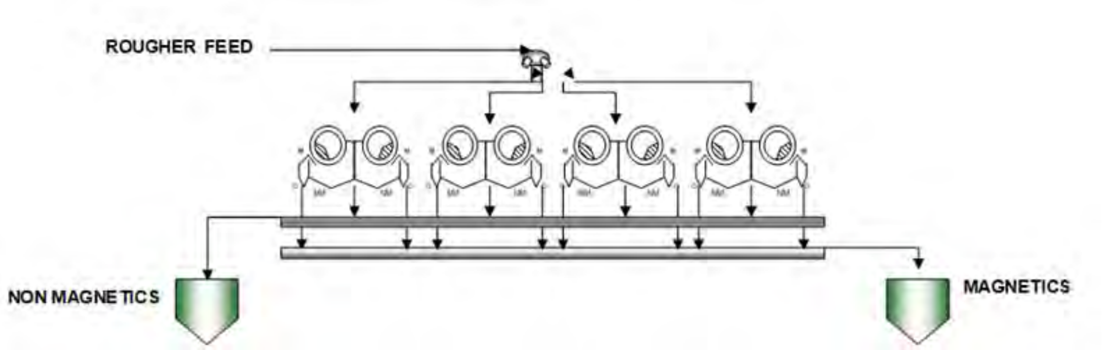


Figure 6 – EHM Magnetite rougher circuit

The concentrate produced from the rougher magnetic separation still contains considerable amounts of impurities. The test work results from the rougher feed material suggests that particle sizes of less than - 53 µm consist mostly of liberated magnetite while the upper size is composed of unliberated magnetite, which requires further size reduction by milling. A size separation through a cyclone classifier is applied to reject most of the tailings and upgrade the rougher concentrate.

The primary cyclone circuit is designed to separate the fine liberated magnetite from the coarse material. This process stage incorporates a cluster of 8 Cavex 400 cyclones. The feed density to the cluster is controlled to ensure the correct cut point is obtained while the number of cyclones operating can be altered to match the feed flow volume.

The cyclones produce an overflow product of 20% solids with a P80 of 53µm containing mostly liberated magnetite and a small amount of low-SG gangue material. The cyclone underflow which consists of mostly unliberated magnetite and low-SG gangue is sent to the existing flotation tailings sump. In the size classification circuit, 47% of the magnetite is recovered from the cyclone feed. A regrind circuit has been incorporated into the design to treat the cyclone underflow and recover the unliberated magnetite. Figure 7 shows the design and operating conditions of the classification circuit:

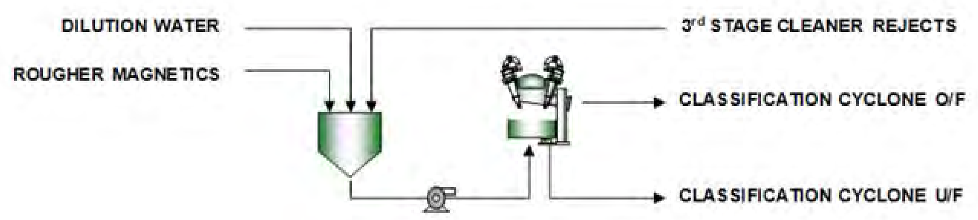


Figure 7 – EHM Magnetite classification circuit

The effective classification of the magnetite is paramount to the success of the separation process. This is complicated by the bimodal density properties of the high-SG magnetite and the low-SG gangue. This was evident throughout the metallurgical testwork campaigns and the JKTech was approached to model the suitability of 400CVX cyclones as a classification method.

A series of simulations were performed on the magnetite feed using the multicomponent Nageswararao cyclone model in JKSimFloat and independently verified using other tools, including JKSimMet. The plant feed was divided into three density categories and fewer than two feed conditions, underground and open cut mining, which produced different particle size distributions. The main objective was firstly to determine whether the proposed cyclone cluster could handle the feed flow rate and density in the magnetite recovery circuit. Secondly, the predicted flow splits, densities and size distributions of the overflow and underflow streams were also of importance. Finally, the recovery of magnetite to the overflow was a key variable. The outcome of the modelling was that the 400CVX cyclone was deemed to be suitable in the Magnetite recovery circuit at EHM.

The cyclone overflow is gravity fed to the cleaner/finisher circuit. The purpose of the cleaner/finisher circuit is to remove the final gangue that is entrained in the magnetite slurry. This stage incorporates three magnetic separator banks operating in parallel. Each bank comprises three magnetic separators operating in series.

The cleaner/finisher magnetic separator works under the same principle as the rougher magnetic separator but at a different magnetic strength. The slurry is introduced into the cleaner separator first, which is a permanent low intensity magnet and operates with a magnetic strength of 750 Gauss. The cleaner separator produces a concentrate which flows into the feed launder of the following finisher drum separator which also operates at 750 Gauss. The last finisher drum separator is operated at 550+ Gauss and serves as a polishing separator for magnetite and non-magnetic gangue materials. At the end of this stage, 99.8% of the magnetite is recovered from the feed and a clean concentrate containing up to 98% iron oxides is produced, which is suitable for sale. Figure 8 shows the design and operating conditions of the cleaner circuit.

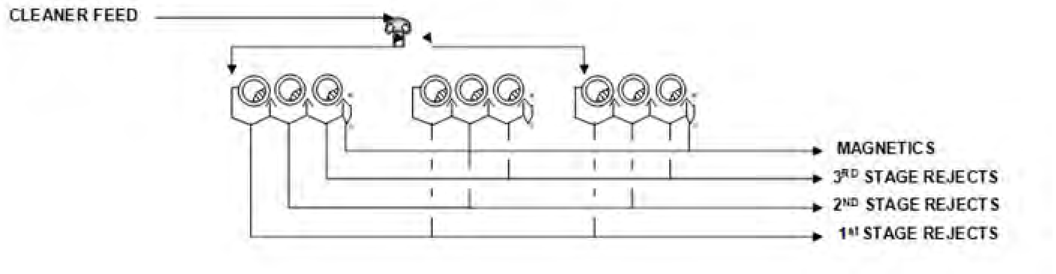


Figure 8 – EHM Magnetite cleaner circuit

The purpose of the dewatering circuit is to condition the extracted magnetite to make it more suitable for storage and transportation. The circuit consist of two CC60 Larox ceramec disc filters. Due to the physical properties of the magnetite and the location of the dewatering plant the clean magnetite is required to be maintained as slurry prior to the dewatering process. This is achieved by agitation in holding tanks and continuous circulation of the slurry prior to entering the dewatering filters. Once dried the product is transferred by gravity directly to the stockpile area located underneath the plant. Figure 9 shows the design and operating conditions of the dewatering circuit:

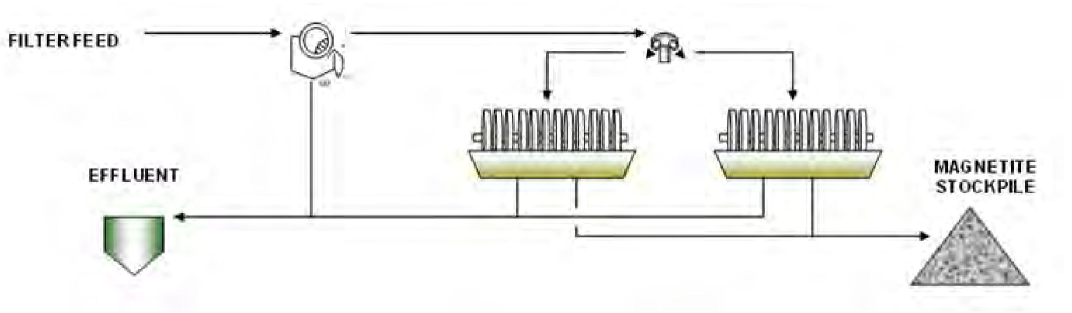


Figure 9 – EHM Magnetite dewatering circuit

Regrind Circuit

In May 2005, Xstrata Technology completed ISAMill characterisation testwork on a sample of EHM tailings. The testwork sample had a P80 of 130 µm, and was ground progressively to 8.4 µm. The testwork showed that the ISAMill technology was able to produce material down to 13 µm from a feed size of 113 µm. Alternative technology tested failed to produce material less than 31µm as shown in Figure 10 (Burford and Niva, 2008).

The reason for the difference between the mills was the smaller grinding media used in the ISAMill testwork. Media selection was based on what a full scale plant can realistically operate at. A full scale ISAMill can be supplied and operated with ceramic media from 1 to 3.5mm, however Tower Mills can only realistically operate with 12mm media and larger, which means that they cannot achieve the grind sizes that a full size ISAMill can.

ISAMills have been in fine grinding applications in base metal circuits since 1994. The ISAMill technology was selected due to its high energy efficiency and intense grinding action. Developments in this technology have allowed the mill to treat coarser feed sizes at high energy efficiency compared to traditional grinding technologies.

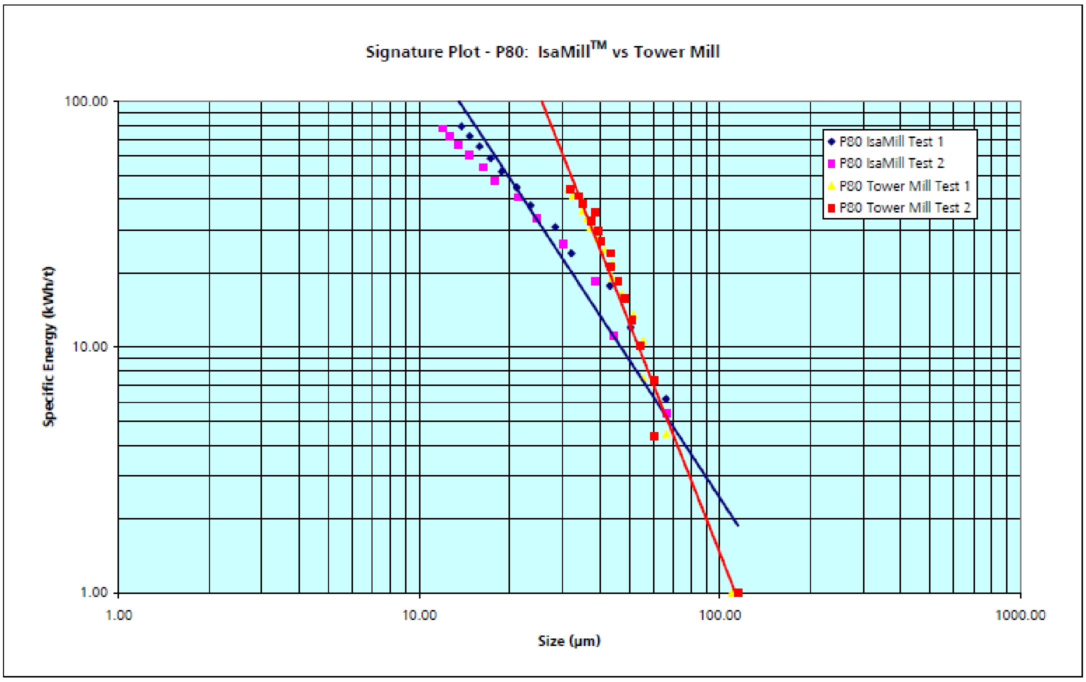


Figure 10 – Comparisons of IsaMill and Tower Mill

Further regrind studies before and after the magnetite project feasibility study included: Bond Index, Levin tests and ISAMill signature plots were carried out on EHM rougher magnetics. These additional studies identified that with rougher magnetic concentrate regrinding, Fe-Oxide recovery to final magnetite product could be increased from 48% to 80% (Magee, 2009). The regrind testwork showed that 1xM10000 (3MW) Isamill would be sufficient to process the cyclone underflow stream followed by rougher magnetic separation. The rougher magnetics will be pumped to the existing cleaner circuit in the base plant which has been designed to accommodate the extra flow.

COMMISIONING AND EARLY OPERATION

The EHM plant was designed to produce a premium product to supply steel industry requirements. The product specification will be published after the completion of the process commissioning. Process commissioning of the base plant commenced in January 2011 however it was delayed by the effects of the Brisbane Floods and Cyclone Yasi which affected a large area of South-East and North Queensland. Early results have achieved a product with the desired particle size distribute on as shown in Figure 13. In addition the magnetite concentrate grade is approaching the desired target. The current objective is to produce a consistent product specification followed a yield optimisation stage.

In addition the magnetite project team is also working to complete the operational readiness plan to hand over the operation of the magnetite plant to the copper concentrator team. The operational readiness tasks include asset management plans such as maintenance schedules and critical spares, training material and the development of operating parameters, targets and KPI's.

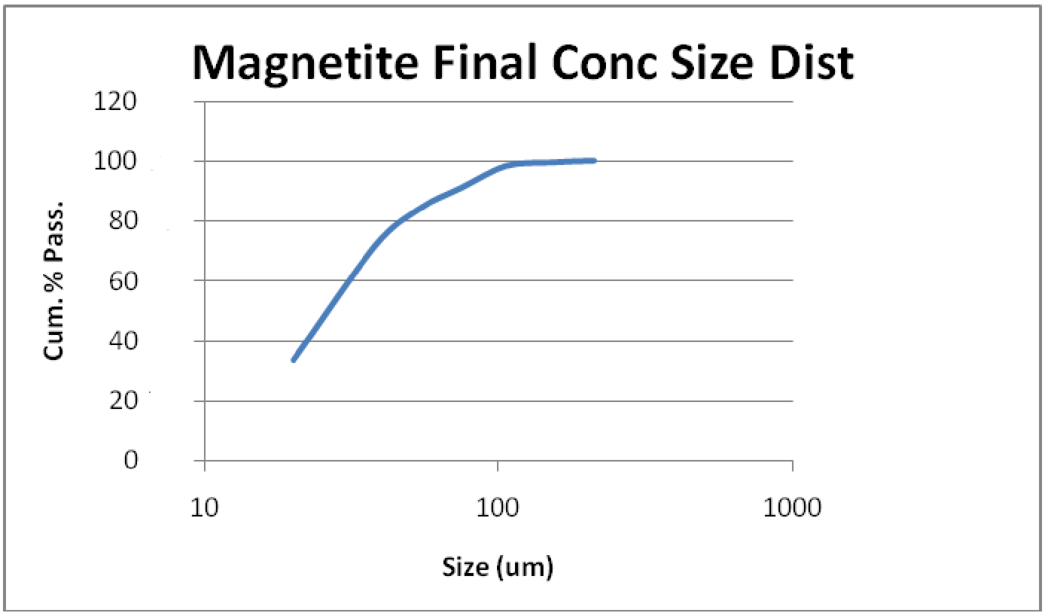


Figure 11– Final concentrate particle size distribution

The slurry commissioning plan includes a series of sampling campaigns to confirm the design parameters, optimise the operation of the plant, and develop guidelines for the daily operation. The sampling and testwork is focusing on collecting sufficient data points for statistical analysis.

- Confirm the specific gravities of all the process streams and compare with the design assumptions.
- Cyclone surveys to develop a size by size database to determine the optimum cut point for the cyclone overflow to achieve the desired magnetic separation.
- Determine the optimum cyclone operating parameters to achieve the target cut size.
- Determine cleaner efficiency and develop operational guidelines for water dilution to the cleaner circuit.
- Maximise ceramic filter throughput.
- Determine what impact if any water impurities could have on the final concentrate specification.

Future testwork is scheduled to understand and optimise the plant performance including full plant surveys, mineralogical analysis of all the process streams as well as IsaMill signature plots on the cyclone underflow.

CONCLUSIONS

The EHM magnetite extraction plant is part of the life of mine extension at Ernest Henry Mine and part of our strategic objective of maximising the value of our existing resources. It will provide an important source of growth and employment for the region. The EHM magnetite project has utilised a concise project management strategy and proven technology to maximise the value for shareholders in our pursuit for zero waste.

REFERENCES

Ryan, A, 1998. Ernest Henry copper-gold deposit, in Geology of Australian and Papua New Guinean Mineral Deposits (Eds: D A Berkman and D H Mackenzie), pp 759-768 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Ernest Henry Feasibility Study, January 1995.

Burford, B and Niva, E, 2008. Comparing Energy Efficiency in Grinding Mills, Metplant 2008 Perth WA

Middleditch, D, Langdon, A, Kormos, L, Stack, P, Chisholm, K, Oliveira, J and Whittaker, P, 2008. Ernest Henry Mine Metallurgical Report, Xstrata Process Support Final Report 4008206.00 28 October.

Magee, M, 2009. Magnetite Project Final Technical Report Internal Report.

Zhu, R, Nielsen, P, Suthers, S, Trudu, A, and Campbell, J, 2007. DMR3139 Ernest Henry Magnetite Pilot Plant Trials – Ball Milling and Magnetic Separations, CSIRO Minerals.

Zhu, R, Price, R and Campbell, J, 2004. DMR-2515 A Study of Magnetic Separation for the Extraction of Magnetite from Ernest Henry Mine Tailings, CSIRO Minerals.

Zhu, R, Nielsen, P, Price, R and Campbell, J, 2006. DMR-2975 Ernest Henry Magnetite Pilot Plant Trials – Stage 2: Hydrocyclone, CSIRO Minerals.

Development of an Innovative Copper Flowsheet at Phu Kham

M F Young and I Crnkovic

Phu Kham Location in Laos

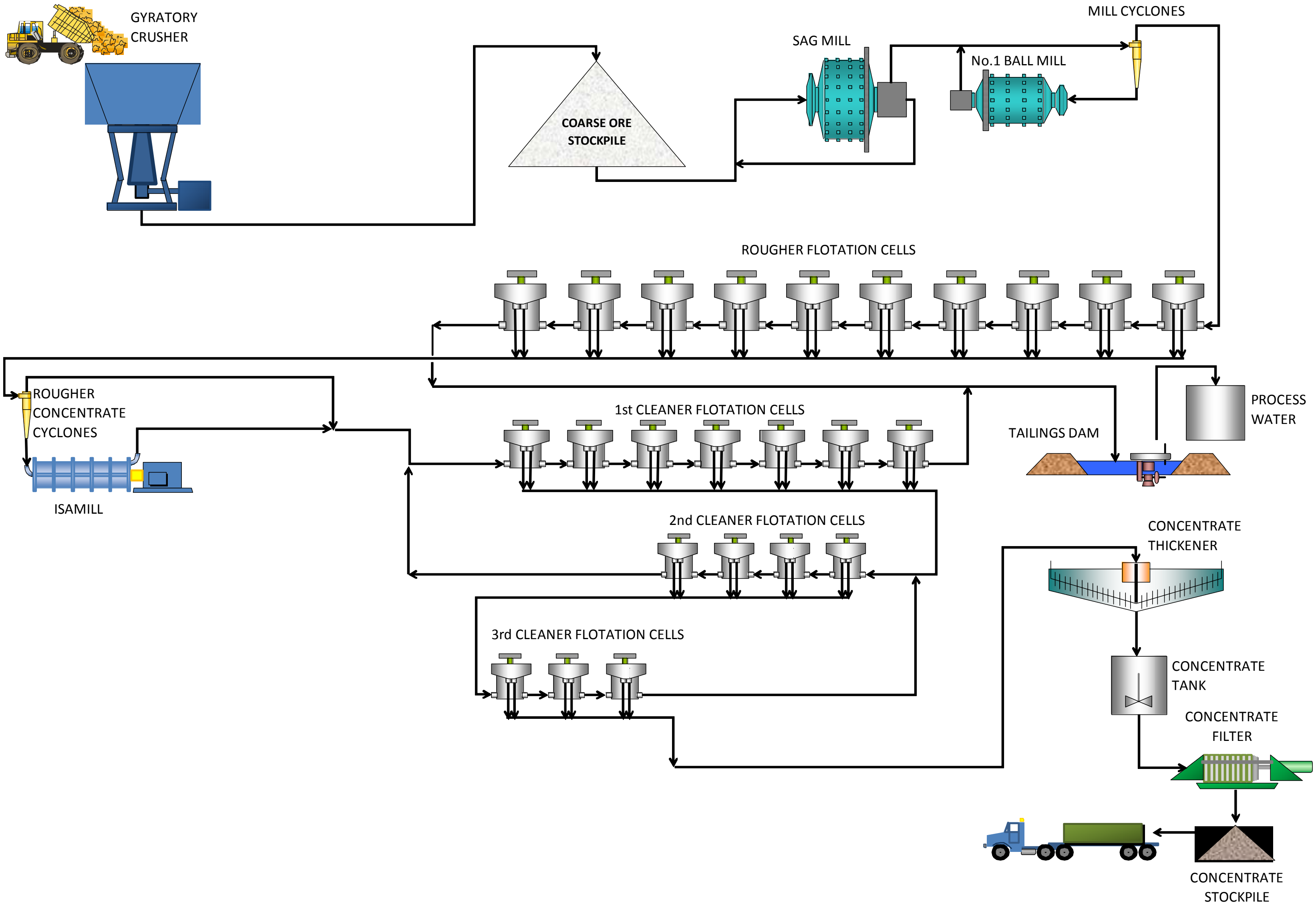
Phu Kham Plant Layout


Phu Kham process plant at night

Plant Layout showing

- Primary Grinding
- Rougher Flotation
- IsaMill Regrinding

Phu Kham Flowsheet



Phu Kham Flowsheet

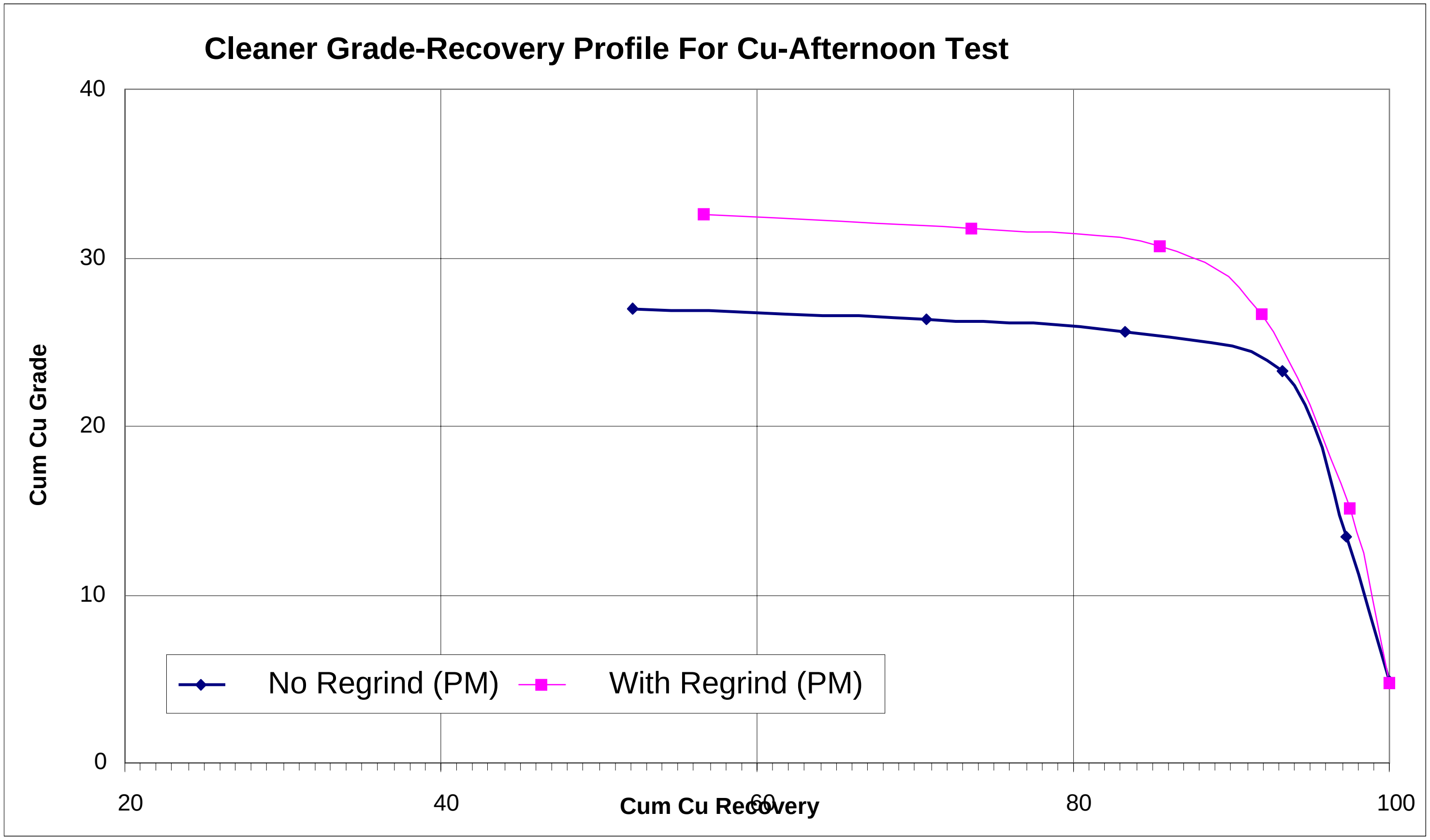
- 12.8 Mt/a of copper-gold bearing ore from the open pit
- Plant Feed Grades are 0.75% Cu and 0.33g/t of Au and 3.8g/t of Ag
- Concentrate quality is +25% Cu, 7 g/t of Au and 60g/t of Ag

- Primary Grinding Circuit contains
  - 34 ft × 18 ft (13MW) SAG mill
  - 24 ft × 39 ft (13 MW) ball mill
- Regrinding Circuit contains
  - M10,000 (2.6MW) IsaMill

- Ore has fine locking of copper and gangue minerals
- Ore requires regrinding rougher concentrate to 38 microns to make good quality final copper concentrate

Cleaner Feed Performance

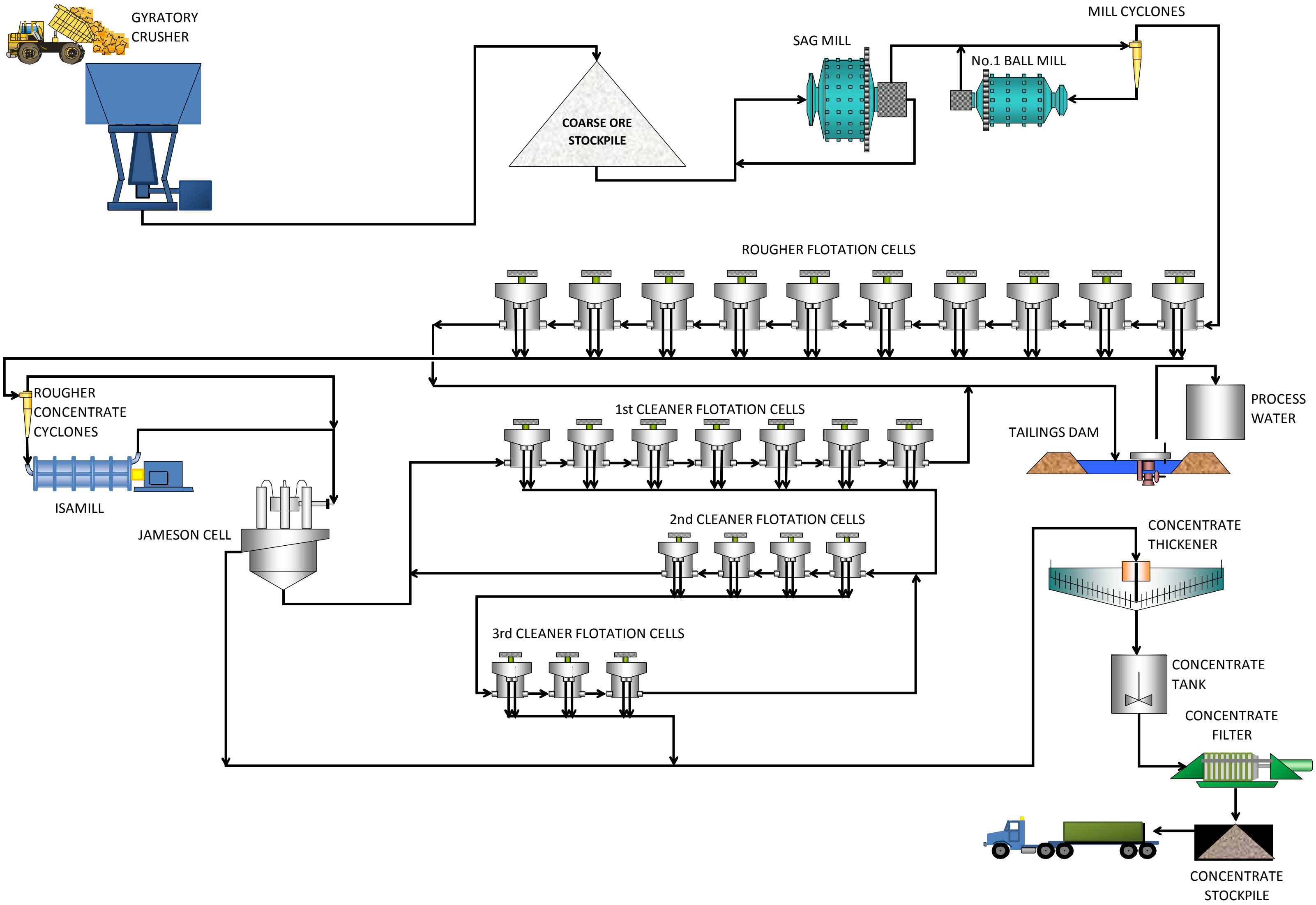
- IsaMill currently grinds from 90 microns to 38 microns
- Cleaner feed performance at 38 microns and 25 microns
- Laboratory Flotation Tests



New Phu Kham Flowsheet

- Cleaning Circuit was overloaded at times
- Fine grinding to improve concentrate quality, slowed flotation rate and increase frothing issues in the cleaner circuit, decreasing the performance of the cleaner circuit
- Jameson Cell with froth washing was installed at the head of the cleaner circuit to increase cleaning capacity
- This new circuit allowed final concentrate to be produced from the IsaMill regrind product in one flotation cell and remove more than half the load from the existing cleaning circuit

New Phu Kham Flowsheet

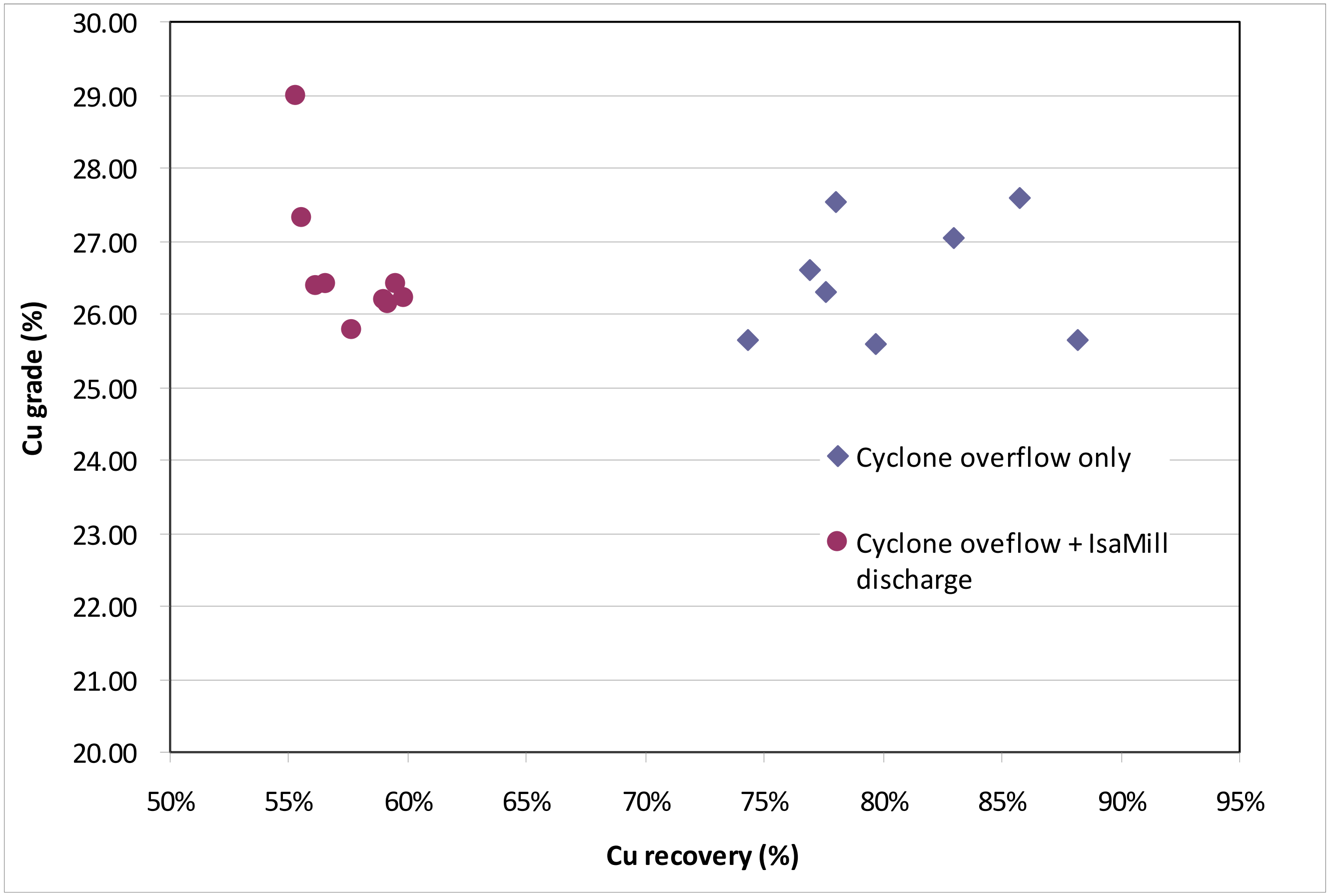


Layout of IsaMill and Jameson Cell

Jameson Cell Flotation at Phu Kham

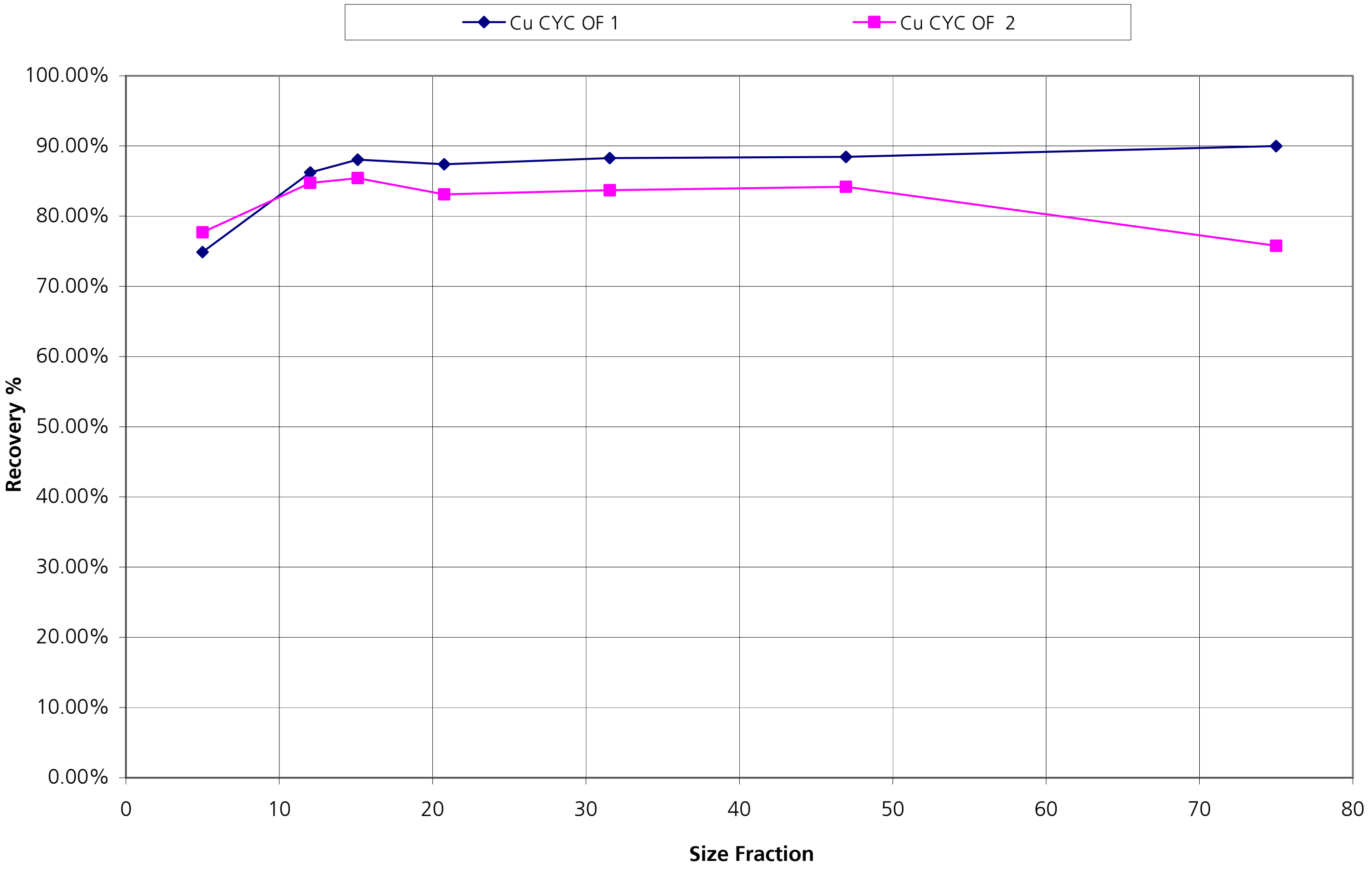
Phu Kham Jameson Cell
Grade-Recovery Curve Performance

- IsaMill Cyclone Overflow P80=20 microns
- IsaMill Discharge P80=38 microns
- Feed (Rougher Conc) Grade = 3-5% Cu



Jameson Cell Size by Size Recovery on IsaMill Cyclone Overflow Stream

Phu Kham Jameson Cell Size-By-Size Recovery

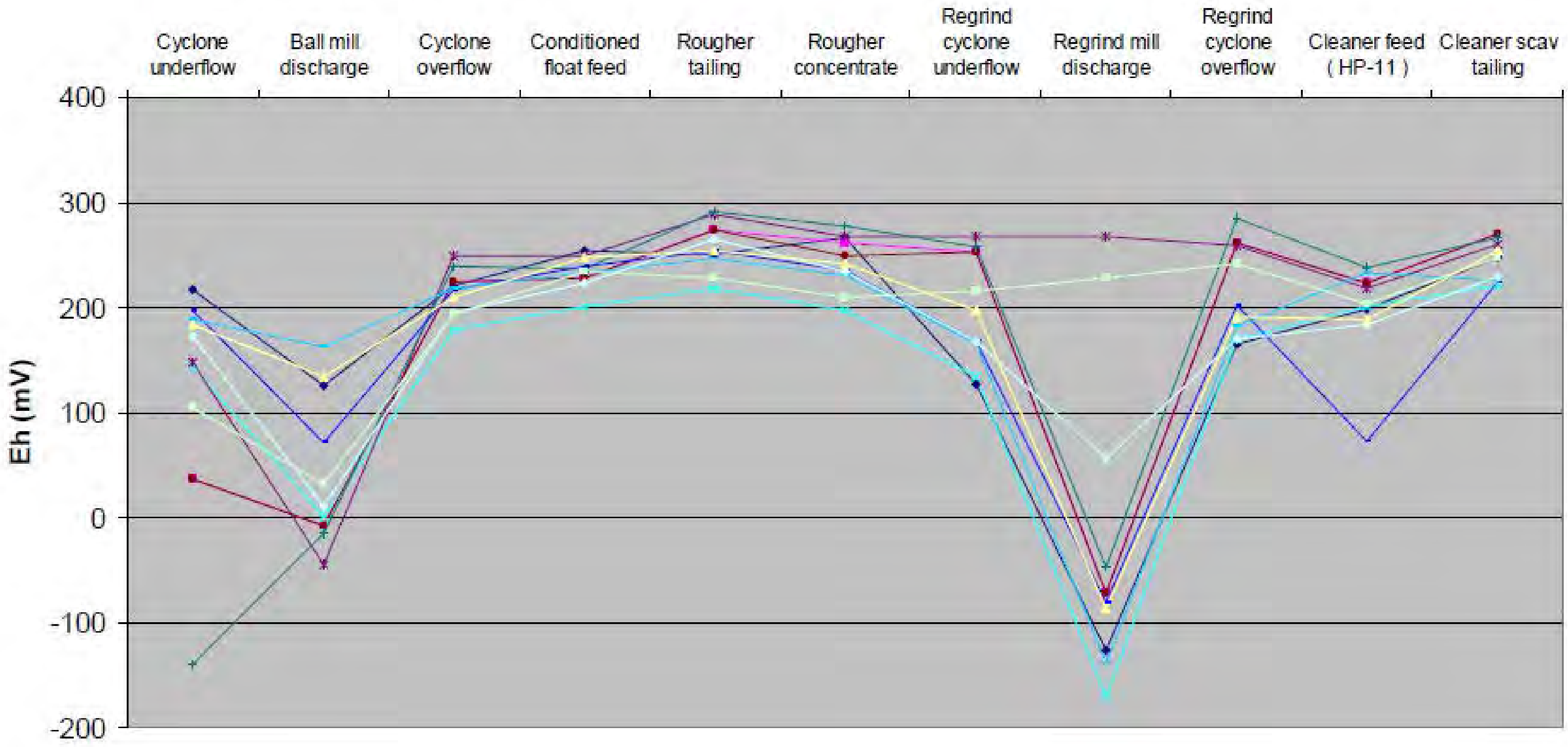


New Cleaning Circuit Results

- Jameson Cell was initially commissioned on the regrind cyclone overflow
- Cyclone Overflow was fine (P80=20 microns) and well liberated
- Jameson Cell produced 25.5% to 27.5% Cu grade at 75% to 90% Cu recovery on this stream
- Jameson Cell produced high recovery in all size fractions for this stream

- Jameson Cell then treated cyclone overflow plus IsaMill discharge
- Jameson Cell still produced high grade copper concentrate at 26% to 29% Cu grade at 55% to 60% copper recovery
- The recovery of the coarser size fraction was lower due to the locking and lower liberation in the size fractions

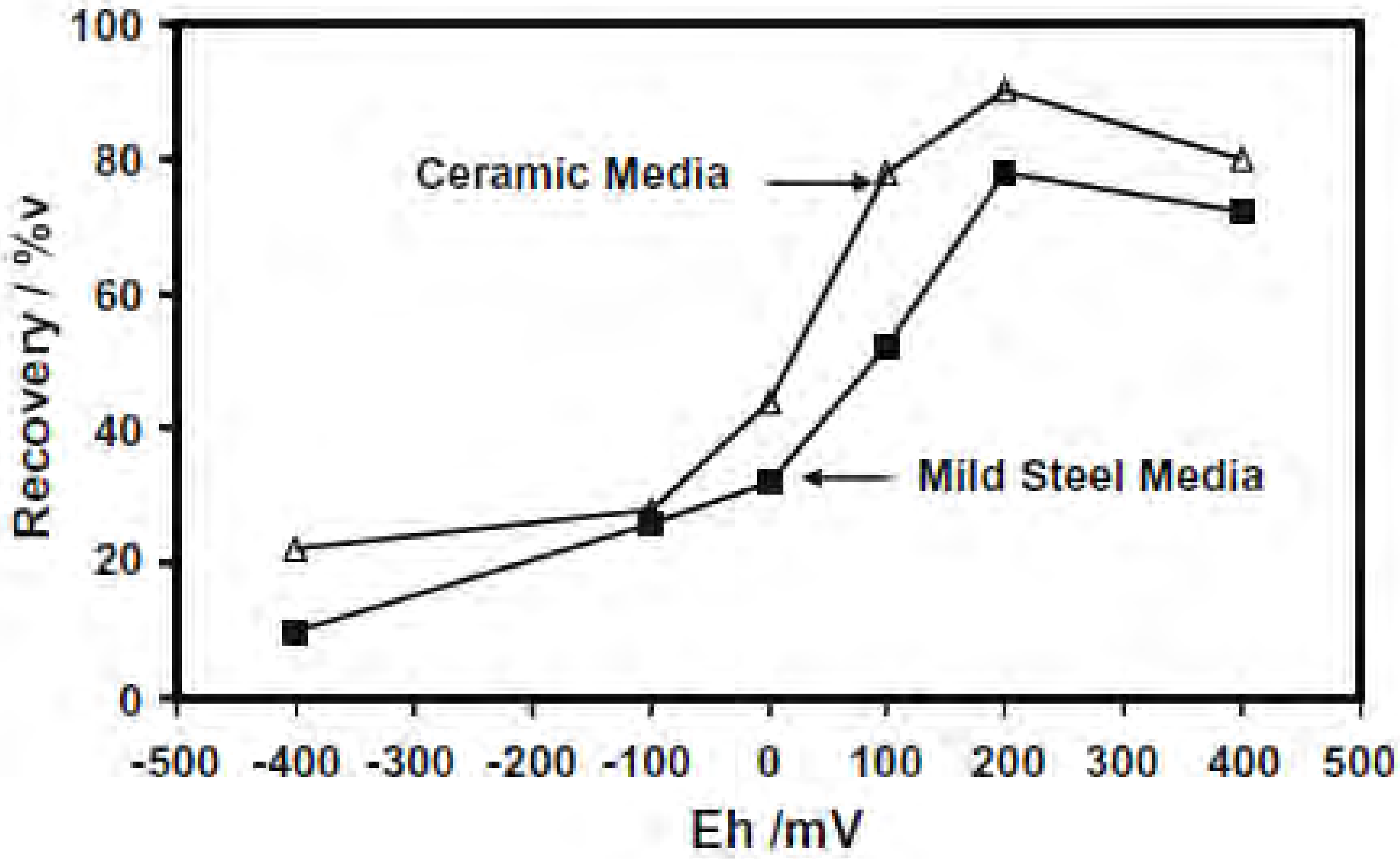
Redox Measurements in Plant



Low Eh reduces the copper mineral flotation performance

Impact of Inert Grinding - Chalcopyrite

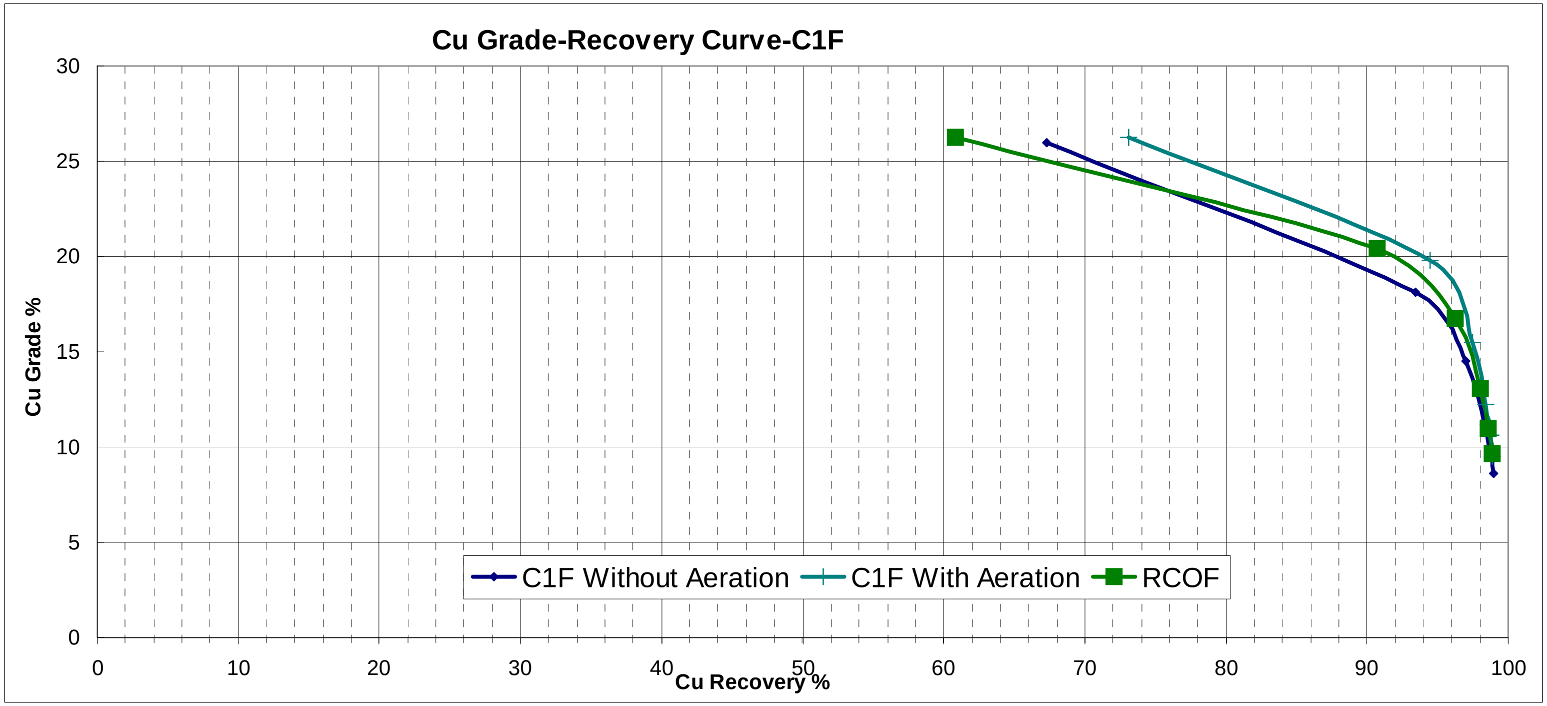
The Effect of Eh – Mt Isa Cu Ore



The critical importance of the grinding environment on fine particle recovery in flotation - Stephen Grano - 2009

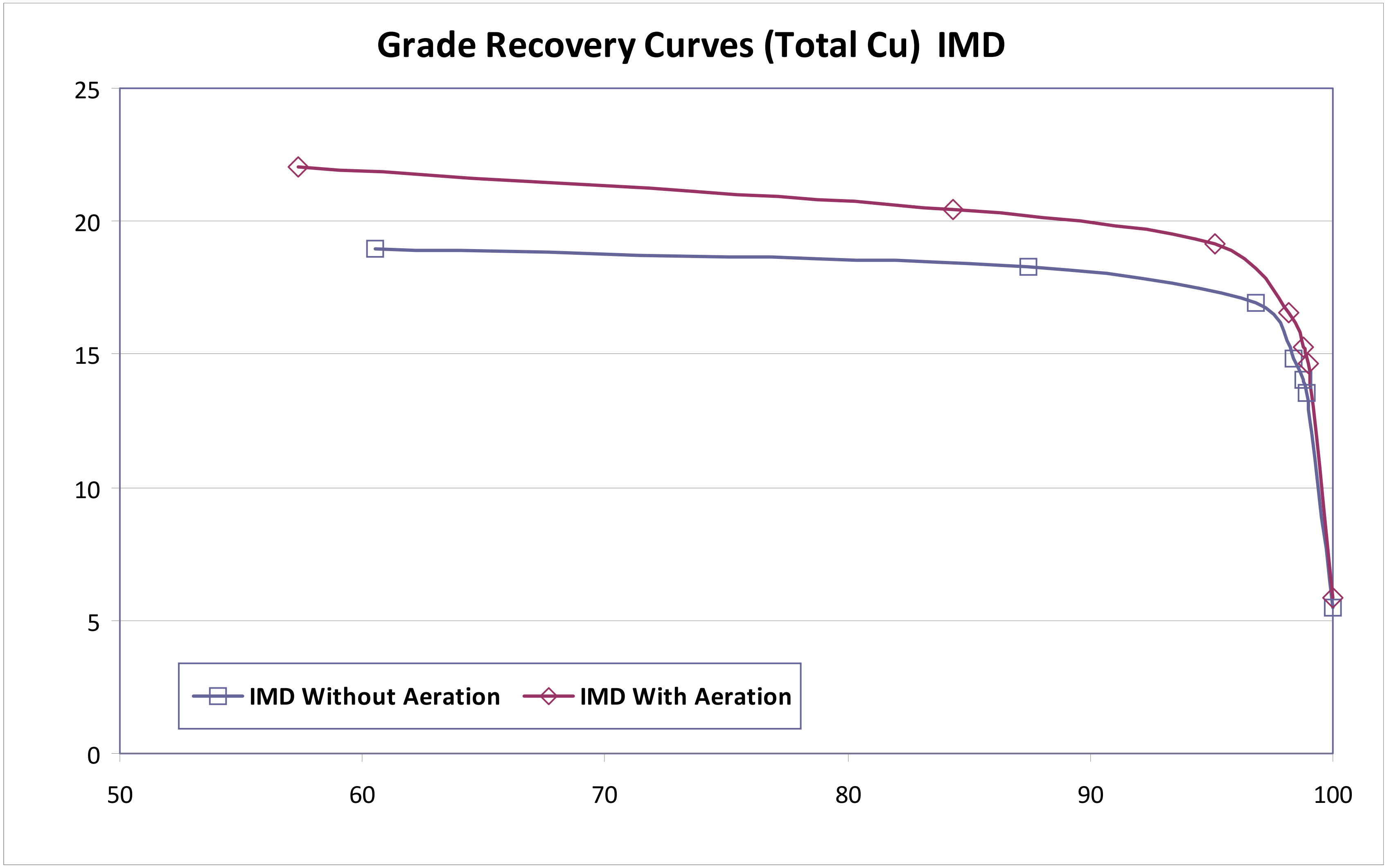
Cleaner Feed Performance

The Effect of Aeration in Laboratory Float



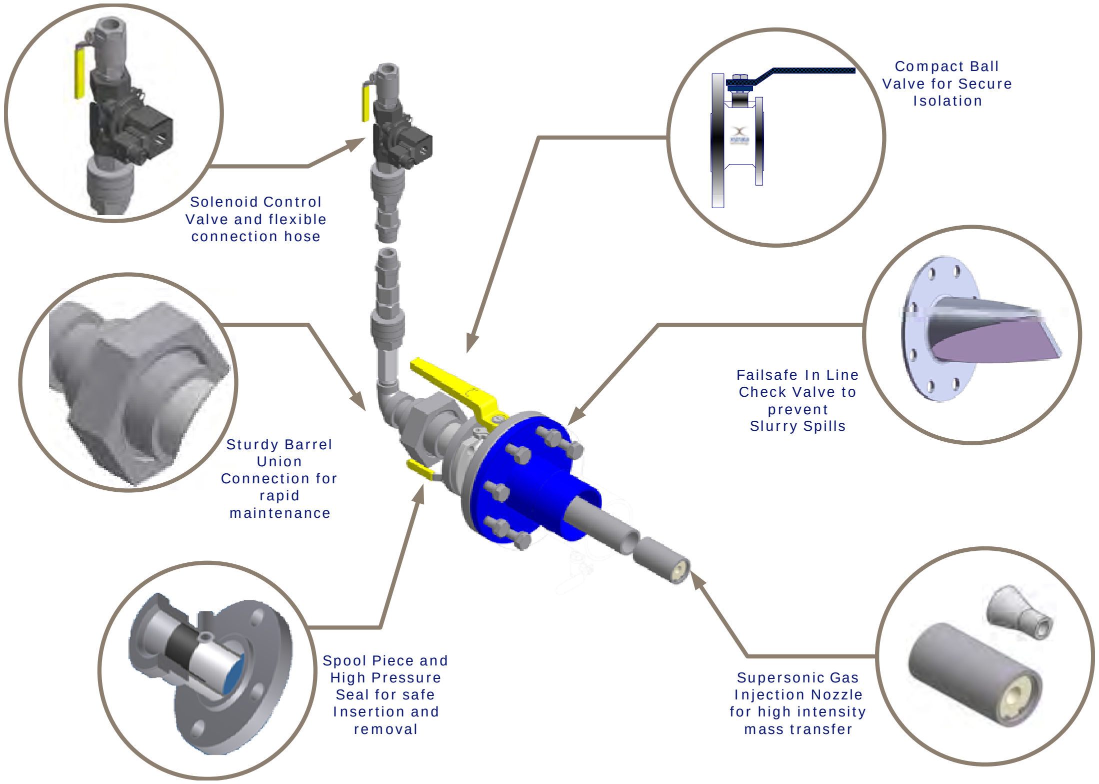
IsaMill Discharge Performance

The Effect of Aeration in Laboratory Float



HYPERSPARGE

Supersonic Gas Injection



Solenoid Control Valve and flexible connection hose

Compact Ball Valve for Secure Isolation

Sturdy Barrel Union Connection for rapid maintenance

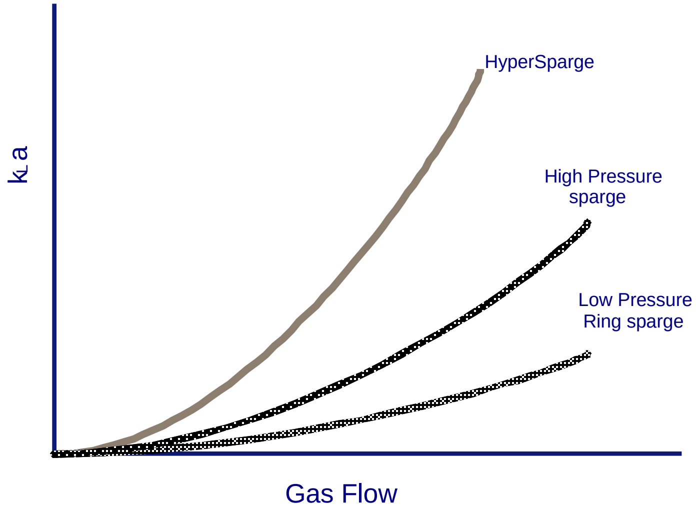
Failsafe In Line Check Valve to prevent Slurry Spills

Spool Piece and High Pressure Seal for safe Insertion and removal

Supersonic Gas Injection Nozzle for high intensity mass transfer

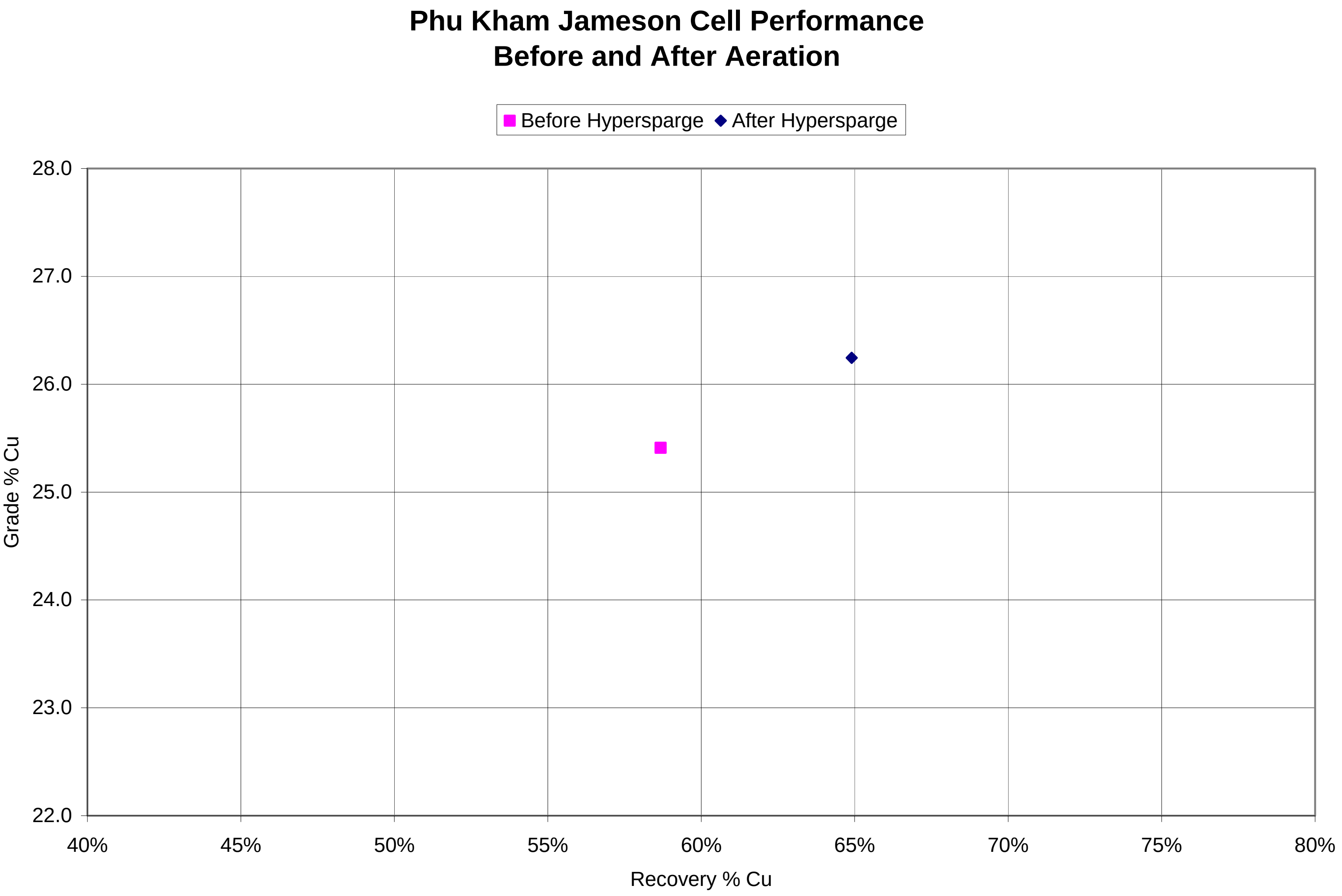
Fine Bubble Generation

Able to generate much finer bubbles for fast gas transfer rates

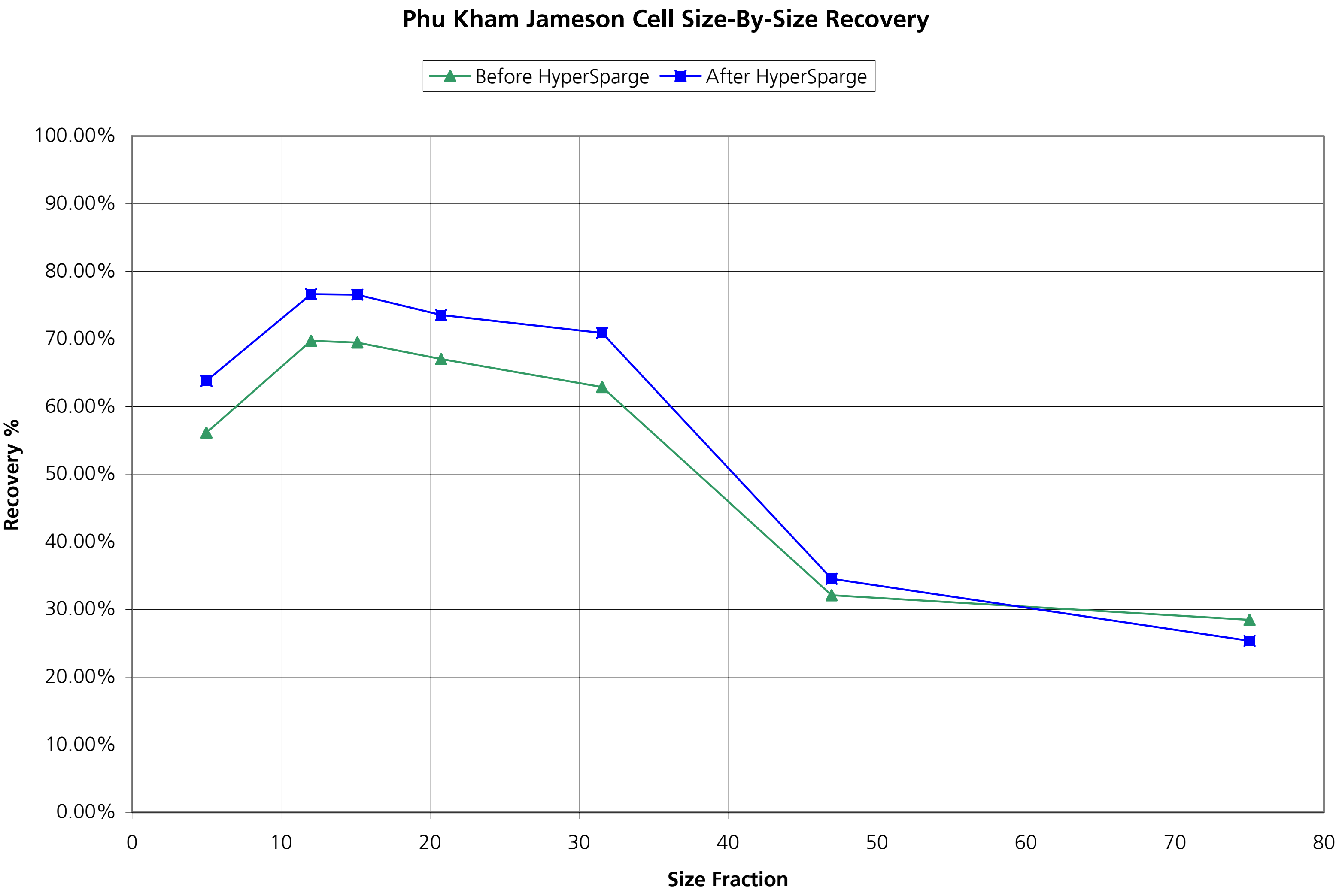


HyperSparge installed in IsaMill Discharge line

Phu Kham – Effect of Aeration Plant Surveys



Phu Kham – Effect of Aeration Plant Surveys



Summary

- IsaMill regrinds rougher concentrate to 38 micron to improve the final copper concentrate quality

- Cleaner feed performance at 38 microns is 25-28% Cu in final concentrate
- and regrinding 25 microns can produce +30% Cu in final concentrate

- Jameson Cell installation was an innovative flowsheet change to removed load from cleaning circuit and increased plant capacity and decrease cleaner circuit frothing issues

- Aeration of Cleaner feed has improve performance and fines recovery and selectivity

- Jameson Cell has allowed the IsaMill to operate at Increase power to improve the circuit performance.

Regrind Mills: Challenges of Scaleup

Michael Larson[1], Greg Anderson[1], Dr. Rob Morrison[2], Michael Young[1]
1- Xstrata Technology
2- Julius Kruttschnitt Mineral Research Centre

Abstract:

As ore deposits become finer grained the requirements for regrinding before cleaning or leaching have increased substantially. Despite this increasing need, there is no standard test to predict grinding energy requirement below 70 microns. The standard Bond Mill test applies for coarser ball milling, but is not appropriate for stirred milling with fine media grinding to 70 microns. With no industry -standard test, the energy requirement is often made on the basis of supplier estimates or benchmarking against similar applications. Yet suppliers use vastly different scale-up methods which result in widely different energy estimates. Estimates can differ by 100-300% even for similar mills. This must result in serious errors to install either too much or too little grinding power. This paper explores this by comparing the actual performance of full scale regrind mills against their original design estimate. It confirms that serious scale-up errors have been made, and have then been perpetuated by "benchmarking".

The test conditions to achieve accurate power estimation are discussed. The essential test conditions are: continuous (not batch) tests, ensure steady state (no coarse fraction retained in test mill), correctly account for classification, measure energy directly (not inferred), and using the same media size as the full scale installation. Failure to correctly address even one of these factors can bias results by 40%; failure to address several multiplies the error.

Introduction:

The requirement for regrinding mills has increased as ore deposits have become more complex and fine grained. This increased requirement has not yet been matched with accurate industry design tools for fine grinding. There is no industry-standard test for fine stirred milling. The Bond test and other scale-up tests designed for coarser grinding do not apply below for stirred milling with fine media to sizes below 70 microns. Using these coarse grinding techniques for finer regrinding underestimates energy requirement because of the different ball sizes, different trajectories and different ball–particle and ball-shell interactions in small mills. Modified Bond tests such as the Levin test are more accurate but are often regarded to be too time consuming and requiring too much sample. As a result, regrind mill suppliers have developed their own scale-up methodologies. Yet in recent experience these different scale-up methods result in 300-400% variations in estimated energy for the same duty. In several recent cases, estimates for the same mill from different manufacturers have varied by over 100%. Clearly they can't all be right. Plant designers need a standard technique to select the correct energy, and need rigorous data on *actual* operation performance (not *design* performance) for benchmarking. Without such a standard, the temptation is to choose the lowest energy estimate, perhaps supported by benchmarking against other *designs* rather than *performance*. This approach could perpetuate incorrect designs, and will lead to project failures.

This paper compares actual regrinding performance with design performance. The poor design record is explained by inadequate laboratory procedures. Different procedures and interpretations are reviewed, and the conditions required to achieve accurate scale-up for any regrinding technology are described.

Scale-up Procedures for IsaMill^{TM}

MIM developed the IsaMill^{TM} in the early 1990's in response to fine regrinding needs of its ore bodies. Because they were also the operators, the MIM engineers required accurate scale-up. Also, they were able to calibrate predictions against operating mills. As a result they developed procedures for accurate 1:1 scale-up from laboratory to plant, and achieved this with both the MIM George Fisher and the MRM 1.1 MW, 3,000 liter IsaMills (M3,000s) (Figure 1). In this scale-up procedure, an M4 (4 liter horizontal) IsaMill^{TM} is operated in multiple pass mode where the product from one continuous pass becomes feed for the next pass to develop a "signature plot" of energy versus different product sizes. This mill is filled 80% with media of the same size as final application (typically 1-3 mm), and agitated at 1500 RPM. The mill configuration and operation – discs, disc speed; media size, media type and media filling; and mill discharge classifier – are all very similar to the full scale operation. Therefore the feed particles "see" the same grinding and classification mechanisms as they do at full scale. Because the grinding media is small relative to the test unit, mill shell effects are minor, as demonstrated by the 1:1 scale-up performance.

MIM and MRM M3,000 Scale-up

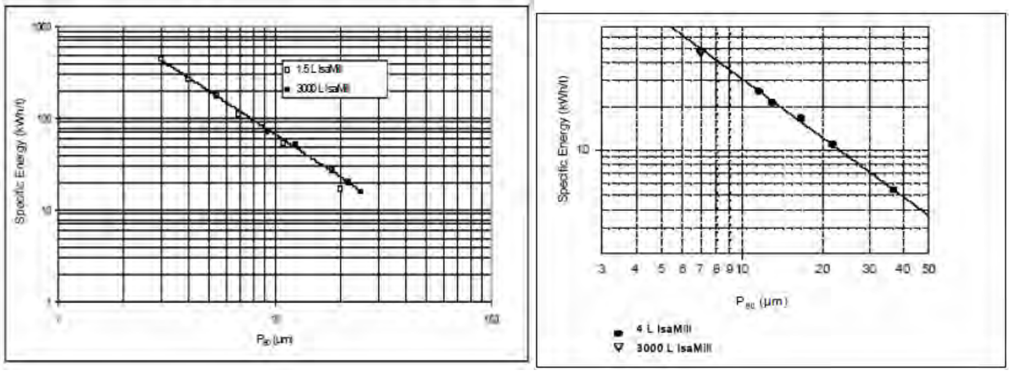


Figure 1. Scale-up of MIM(left) and MRM M3,000 IsaMills (Barns K, Curry D; 2006)

The reasons for the accurate direct scale-up from the laboratory to full size IsaMills are:

- Using the same size and type of grinding media as full scale - no correction factors are needed.
- The mill configuration and mechanism is the same – horizontal mill, similar grinding discs with similar tip-speeds, and closing the mill with an internal classifier similar to full scale. Therefore the mechanisms, velocities and physics within the laboratory mill are very similar to the full scale mills.
- A large test mill relative to particle and media sizing, so shell effects are minimal.
- Testing with the same slurry feed size and density.
- Ensuring enough feed material is processed to reach steady state. Each point on the "signature plot" results from a continuous pass through the mill, during which the mill reaches steady state. The mill contents must be displaced several times in each pass, as retention of even a small amount of coarse particles will seriously underestimate full scale steady-state power. This is a crucial point that simply cannot be achieved in a batch test, because it does do not continuously add new material and can't ensure steady state. This requirement to achieve steady state is crucial for fine regrinding – fortunately it is also possible with relatively small samples (15 kg).
- Direct measurement of energy consumption from the agitator shaft, in the same way that energy is measured for full scale mills. This also done to accurately subtract the no-load power ensuring the lab mill will scale-up to any size mill.

Having developed this procedure for internal MIM operations use, it has now been successfully applied in other installations. Figure 2 demonstrates the scale-up performance for the 2.6MW (3500 hp) M10,000 IsaMill for Anglo Platinum.

Western Limb Tailings Retreatment Scale-up

IsaMill Model	Installed Power (kW)	Chamber Volume (L)	Specific Energy (kWh/t)	Pulp % Solids	P_{98} (μm)	P_{80} (μm)
M4	4	3.5	37	39	47.5	16.0
M10,000	2,600	10,000	37	42	42.5	16.5

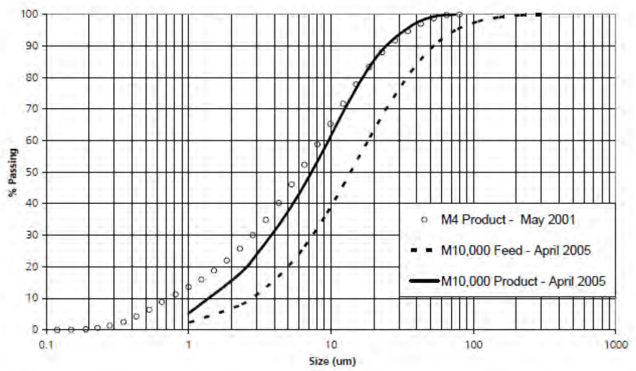


Figure 2. Western Limb M10,000 Scale-up (Curry D, Clark L, Rule C; 2005)

SMD Scale-up:

The scale-up performance of stirred mill detritors (SMD's) at the Century mine has been reported. (Gao M, Reemeyer L, Obeng D, Holmes R; 2007). The duty at Century is fine lead-zinc regrind, very similar to the George Fisher and MRM duties reported above. However for Century, an error of 20% was reported from the 0.55 kW lab mill to the full scale 355 kW mill- that is the full scale mill was less efficient than predicted from testwork. This may be attributed to measuring laboratory power draw by grinding chamber reaction torque rather than by direct measurement (Nesset J, Radziszewski P, Hardie C, Leroux D; 2006), but the power measurement had been corrected to be a direct power measurement and there was still a 20% error from the 0.55 kW lab mill to the full scale 355 kW mill, so this error may be due to not reaching a steady state operating condition in the mill.

SMD Scale-up

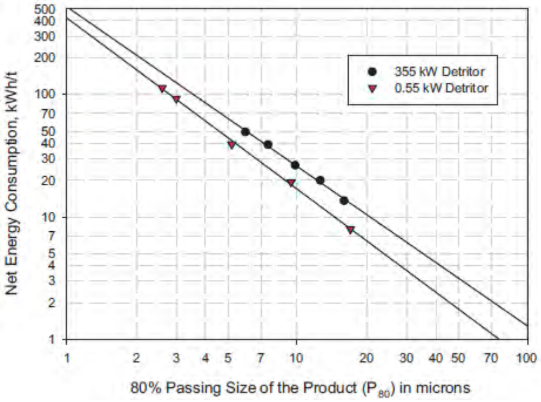
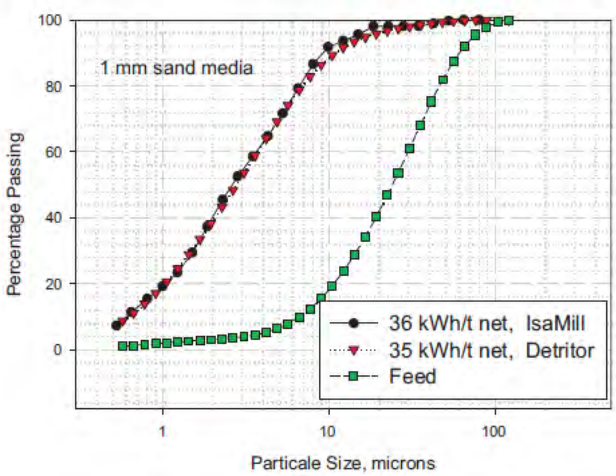


Figure 3. SMD Century scale-up (Gao et al, 2007)

The scale-up procedures developed for the IsaMill were then compared against the full scale Century SMD's. The IsaMill M4 procedure accurately predicted the full scale Century performance on a P_{80} basis. This is intuitively correct – though the mills have different designs (horizontal versus vertical), essentially they stir the same media at similar speeds, so should be expected to have similar energy requirements for gross size reduction. The only difference between the tests was the finer P_{98} (sharper size distribution) from the IsaMill, attributable to the internal product separator and the bypass losses that occur in the full scale SMDs.



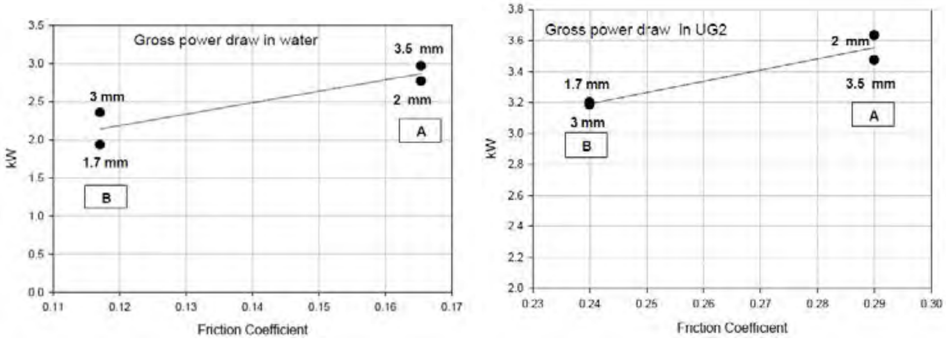
Industrial SMD vs. Lab Scale IsaMill

Figure 4. Full Scale SMD vs. 4 Liter IsaMill Comparison (Gao et al; 2007)

Errors Caused by Inadequate Top Size Breakage

A continuous test that simulates full scale grinding mechanism should give accurate scale up. However rigour is required to ensure coarse particles do not accumulate in the mill, and are adequately broken during the test. The importance of avoiding build up is discussed later – using too small a sample to reach steady state will *underestimate* energy. But failure to break top size particles can *overestimate* grinding power. If grinding media is too small to break the largest particles they will accumulate in the mill and displace grinding media. This reduces grinding efficiency while increasing power draw (analogous to "critical size" build up in a SAG mill). This reduction in grinding efficiency is not always evident if the particles are retained in the mill.

This was almost certainly the cause of the analogous data reported by Farber et al. This work compared two different media types, A and B. The smaller media appears to be more efficient but the results are misleading because of coarse material holdup in the mill. When this coarse material is taken into account the energy requirement is much higher. The first graph in Figure 5 demonstrates the lower friction of Type B media, leading to lower power draw on water. As expected, for any media type, smaller particles need less energy to mix than coarser ones. (it is easier to push your hand through a bucket of 3mm balls than a bucket of 12 mm balls). The same effect should be seen when operating with slurry, yet the second graph in Figure 5 shows the opposite effect – both media types are reported to consume more energy using fine media in ore grinding ("UG2" platinum ore). This almost certainly indicates the tests were in error – the fine media (1.7mm Type B and 2 mm Type A) was too small to break the top size, causing reduced grinding efficiency and holdup of coarse material in the mill. Therefore the resulting signature plots (Figure 6) are incorrect.



Media Power Draw

Figure 5. Power draw in water (left) and UG2 slurry (right) for first pass of Signature Plot test. (Farber et al; 2010)

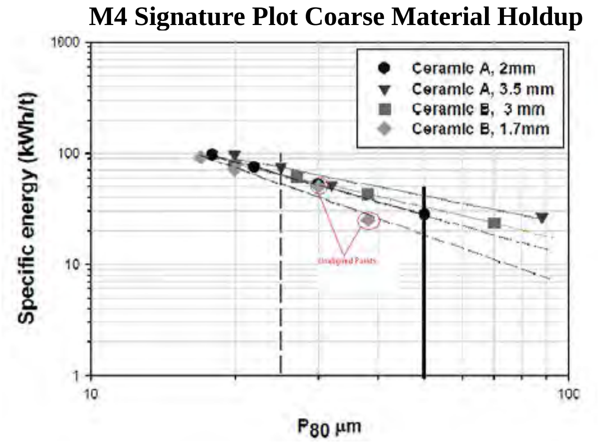


Figure 6. UG2 signature plot with different media size and media SG (Farber et al, 2010)

Indeed, there is a “fingerprint” of this error in Figure 6: the first point for 1.7mm Ceramic B does not fall on the same line as the other 3 points, rather it is finer than expected. This is a clear indication that coarse material was held in the mill in the first pass, that steady state had not been reached, and so the test is invalid. Note that this point’s deviation does not “look” very significant on the wide log-log scale used in Figure 6. However it is a large deviation that would be much more apparent if the graph was redrawn with a better scale (10 to 100 kwh/t rather than 1 to 1000 kwh/t). Beware of the dangers hidden by log-log scales in fine grinding!

Similar investigations which correctly ensured top size breakage and steady state were conducted by Larson. Different sizes and types of media were compared in water and grinding copper ore. This yielded the expected results – smaller media is more efficient, drawing less power in both water (Figure 7) and copper ore (Figure 8), ***so long as*** it breaks top size particles.

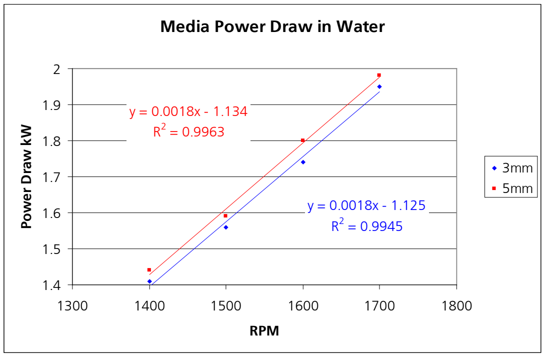


Figure 7. Hira media size vs gross power draw in water. (Larson M; 2010)

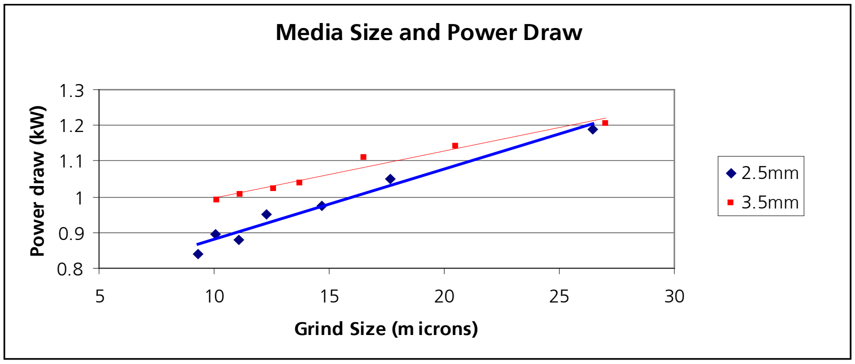


Figure 8. Media size power draw in copper concentrate slurry. (Larson M; 2010)

Figure 9 shows the difference in net power draw between the same two media charges but across different grind sizes. These were taken from individual signature plot tests on the same ore under the same operating conditions. Both media were capable of breaking the top size. The 5mm media resulted in a slightly smaller P_{98}/P_{80} ratio. It was 2.6 compared to 2.7 for the 3mm.

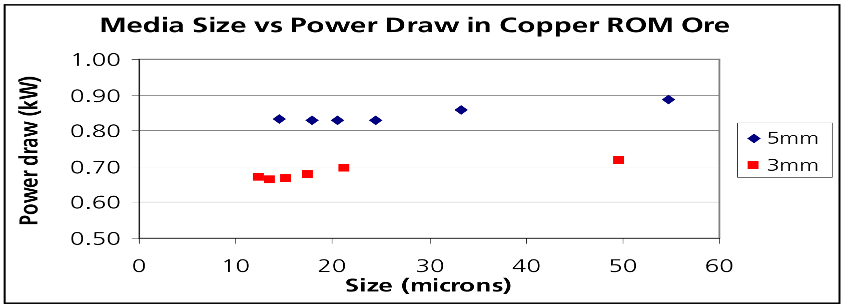


Figure 9. Hira media size vs. power draw in a copper ROM ore. (He M; 2010)

Again, as shown in Table 1 and Figures 8 & 9 the smaller MT1 media will also draw less power under the same slurry conditions. Further, as the slurry is ground finer the power draw will decrease. In Figure 8 the first points at 1.2 kW are almost the same power draw. At these points the 2.5mm media produced a P_{98} of 95µm and the 3.5mm media produced a P_{98} of 63µm. The finer points had closer ratios as the 2.5mm media was more capable of grinding the top size in these passes.

Net Mill Power at Different MT1 Media Sizes			
Media Size	1.5 mm	2.5 mm	3.5 mm
Net Mill Power	.80 kW	.98 kW	1.07 kW

Table 1. Average net mill power draw for copper concentrate varying only media size (Larson M; 2010)

Morenci M10,000 Scale-up:

The design of the Morenci M10,000 is described below and in Table 3. Though the M4 IsaMill test accurately predicted energy requirement to achieve P_{80}, it demonstrates other important factors required for accurate prediction of the P_{98}, as explained below (Cole J, Wilmot J; 2009)

.

> The design was for a P_{80} of 7 micron and a P_{98} of 15 micron. Actual feed sizes were P_{80} of 11 micron and a P_{98} of 34 micron with the Morenci concentrates. The Morenci concentrates contained variable amounts of pyrite which is more difficult to grind than a relatively pure chalcopyrite. Due to the mineralogy of Sierrita concentrates, containing approximately 85% chalcopyrite, a switch was made to process Sierrita concentrates. Sierrita concentrates were less variable in composition and softer than Morenci concentrates. Sierrita concentrates are also similar to the Bagdad concentrate which was used to size the IsaMill. The P_{80} and P_{98} from the Sierrita concentrates were finer than the Morenci concentrates, at 7.4 micron and 25 micron respectively, but were still not at the design criteria.
>
> The design pyrite level in the concentrates was 22.5%, the actual varied from the low teens to as high as 55% on Morenci concentrates.

Morenci concentrate leach plant design criteria.

Morenci MT-DEW-SX	
Design Criteria	
Feed Rate, mtph	26.4
Super Fine Grinding	
Feed Rate, mtph	32.0
P80, µm	7.0
P98, µm	15.0
kWh/t	68.0

Table 2. Morenci M10,000 design criteria (Cole J, Wilmot J; 2009)

Even with the difference in ore mineralogy, this inability to achieve the P_{98}/P_{80} ratio as designed from the original testwork suggests that a small fraction of coarse, hard material was not fully ground and remained in the test mill. Though this is the only case so far when P_{80} scaled but the P_{98} did not, it suggests that if the objective of the testwork is highly accurate prediction of P_{98}, then the standard 15 kg of sample for an M4 may be insufficient.

Centerra Gold’s Kumtor Mine has a similar duty with M10,000 IsaMill grinding gold/pyrite feed from 20µm to a P_{80} of 10µm prior to cyanide leach. Figure 10 shows this scale-up was successful, including the predicted P_{98} (in fact, the full sized mill produced a slightly better size distribution than predicted).

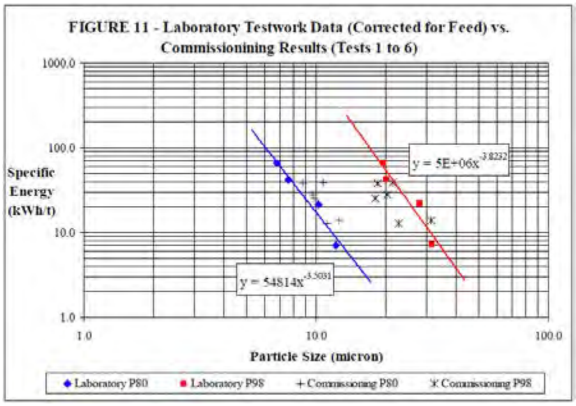


Figure 10. Kumtor M10,000 scale-up (Kazakoff J, Mortimore A, Smith S, Curry D; 06)

Tower Mill Procedures:

The test conditions for the Nippon Eirich Tower Mill are shown in Figure 11. These procedures raise several questions when compared with the requirements described above. Firstly, recirculating mill discharge directly to the mill feed pump means that steady state can never be reached – mill feed is almost instantly contaminated with fines from the “first pass” (in contrast, the M4 IsaMill tests passes the sample continuously from the feed tank through the mill into a separate product tank, reproducing full scale grinding).

Grindability test 易磨性实验

Model型号 : NE008 Tower Mill
Capacity容积 : 8 liter
Sample样品 : ~10 kg / test
Test period实验时间 : 1 week
Purpose : Preliminary grindability check
目的：初步的易磨性检测

Model 型号 : KM-5 Tower Mill
Capacity容积 : 120 liter
Sample样品 : ~150 kg / test
Test period 实验时间 : 2 weeks
Purpose目的 : Mill size selection for the performance warranty 设备选型实验

March, 2009
Page 25
NIPPON EIRICH CO., LTD.
EIRICH

Figure 11. Nippon Eirich Brochure; March 2009

Secondly, these tests appear to use a sample size far too small relative to the size of the test mill. Mill void volume must be replaced at least 3-4 times to achieve steady state and avoid hold-up of coarse particles (which causes serious underestimation of energy). Table 3 indicates this Tower Mill testwork uses about half the volume required to reach steady state.

Mill Type Test Feed vs. Void Space Ratio

Mill Type	Mill Open Volume(L)	Solid Volume(L)	Ratio (recommended above 3)
M4 IsaMill	1.35	5.0	3.70
NE008 Tower Mill	2.35	3.33	1.42
KM-5 Tower Mill	35.2	50.0	1.42
3.0 SG Solids, .66 Media Bulk Fill Ratio and .88 Mill Operating Volume Based on Geometry			

Table 3. Grinding test solids/open volume ratio comparison

The impact of too small a feed volume is demonstrated in Figure 12 from tests on a pilot Tower Mill. The signature plot shifts significantly from 50 liters of feed (0.7 times mill open volume) to 150 liters of feed (2 times open volume). Clearly the mill is not at steady state with the smaller volume. In this case the energy estimate to grind to a P_{80} of 70µm increases by 30% at the higher feed volume. In fact, the feed volume needs to be even higher than this, so the true error is even higher.

Combining too small a sample size with the error of diluting new feed with mill discharge compounds the error further, and is likely to lead to serious underestimation of true power requirement in a plant situation.

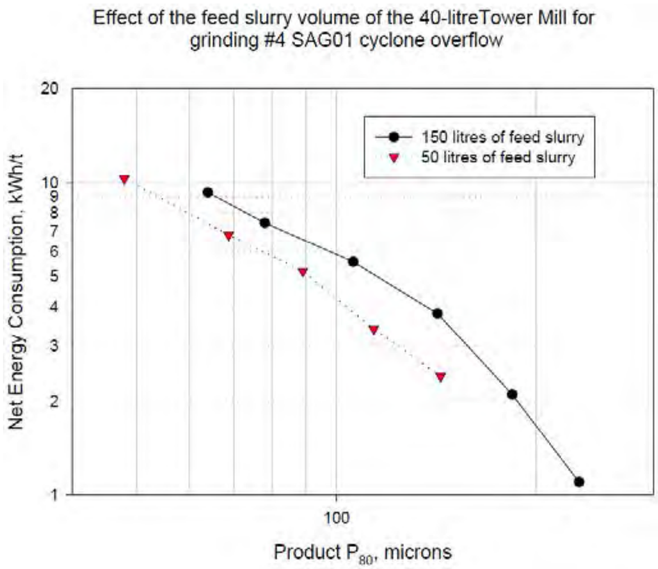


Figure 12. Effect of varying feed solids mass in a pilot tower mill (Gao M)

Extensive testwork has been done to ensure this is not a problem with the standard 15 kg M4 IsaMill testwork. One example is shown below comparing the signature plots generated with masses of 15 and 35 kg of common iron ore concentrate. In this case both tests consisted of a magnetite concentrate with a F_{80} of 65 microns being ground with 3.5mm media. The 3.5mm media was selected for its energy efficiency and ability to better break the top size than would happen with a smaller media. The resulting steep size distribution would show an improvement in final concentrate grade of 2-3% depending on the target grind. Both the P_{80} and P_{98} plots fall within the margin of error for the M4 test. The iron ore had a solid SG of 4.2. This is on the high side of what is normally tested. Anything lighter will take up more volume so the conclusions from this testwork will apply to those samples as well.

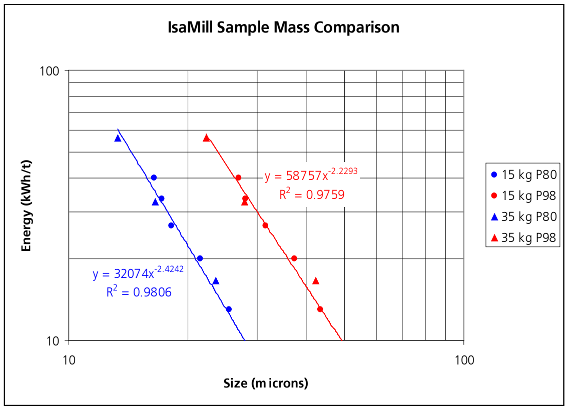


Figure 13. Effect on signature plot of varying feed mass in an M4 IsaMill (Larson M, Villadolid V; 2010)

The two tests resulted in respective points at 16.5 and 16.7 µm P_{80}'s at just over 40 kWh/t. A comparison of those two size distribution curves is given below. The shape is nearly identical. This would also indicate that at both sample masses the mill is discharging the same steady state sample with no segregation effects.

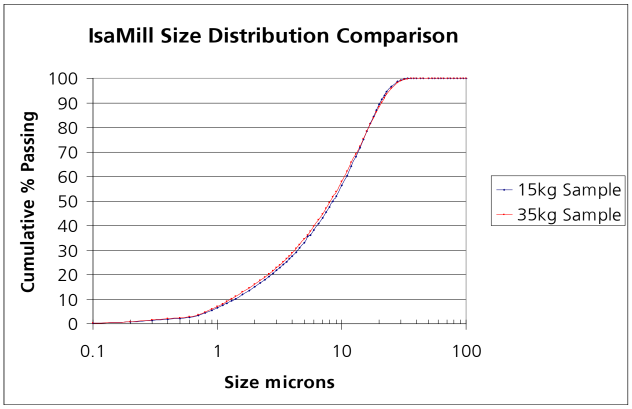


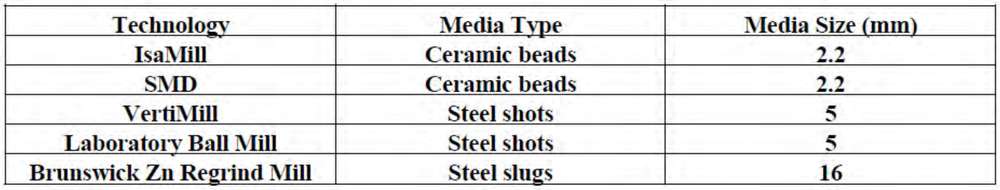
Figure 14. Effect of varying feed mass on M4 IsaMill product size distribution (Larson M, Villadolid V; 2010)

Nesset Regrind Comparison:

The 2006 paper by Nesset et al compared the grinding efficiency of a ball mill, IsaMill, SMD and Vertimill. It concluded that the SMD procedure of measuring power by reaction plate torque would underestimate energy requirements by 32% and 39% for the two media types that were tested. Once the SMD power was measured from the motor there was little difference in energy requirement between the IsaMill and SMD on a P_{80} basis, though there were differences at the P_{98} level. This supports the findings reported above. However, Nesset's paper also concluded that the laboratory Tower Mill was 43% more energy efficient than either the IsaMill or SMD, for the finest media tested in the Tower Mill. Since both the IsaMill and SMD were specifically developed and determined to be more efficient than Tower Mills for fine grinding, and this has been demonstrated consistently at plant scale, this conclusion needs further examination. It demonstrates another potential source of error in laboratory testing – using different size media than full scale.

Media size is crucial for fine grinding efficiency – the need to use finer media to increase energy efficiency drove the development of both the IsaMill and SMD. Nesset's Tower Mill tests used 5mm media, a size which is never used in plant installations, and the coarser media that were tested in the Tower Mill were much less energy efficient. The finest media practically used in operating Tower Mills is 12 mm (otherwise media floats from the mill). Per cubic metre, 5mm media has 2.4 times the surface area of 12 mm media – a huge difference for attrition grinding in a stirred mill. Nesset et al report the most efficient media sizes for different devices in Table 4. Though 5 mm media is most efficient in the ball and Tower Mill, the key point is that it cannot be practically used at large scale, both in terms of media retention and energy intensity (size of installation). This is precisely why the high speed stirred mill technologies (IsaMill and SMD) were developed – to practically achieve the energy efficiency benefits of fine media.

Efficient Media Size



Technology	Media Type	Media Size (mm)
IsaMill	Ceramic beads	2.2
SMD	Ceramic beads	2.2
VertiMill	Steel shots	5
Laboratory Ball Mill	Steel shots	5
Brunswick Zn Regrind Mill	Steel slugs	16

Table 4. Most efficient media sizes (Nesset et al; 2006)

Even the inappropriate media size does not fully explain the unusual Tower Mill efficiency reported by Nesset. It is likely that the Tower Mill testwork also failed to reach steady state and retained coarse particles in the mill. Not enough experimental information is available to confirm this, but based on reported sample size and number of tests, it appears that the Tower Mill feed may only have been 1.5 times mill void volume, not enough to reach steady state. This suspicion is supported by a graph (Figure 15 below) comparing product size distribution for three different media sizes. This reports that the 5mm media improved both P_{80} and P_{98} compared with coarser media. This is unexpected – usually coarse media breaks coarse particles more effectively (especially in stirred milling). This counter-intuitive result strongly suggests that coarse particles were retained in the Tower Mill in the test procedure, which would cause serious underestimation of energy requirements as described in earlier sections.

Vertimill Media Effect on Size Distribution

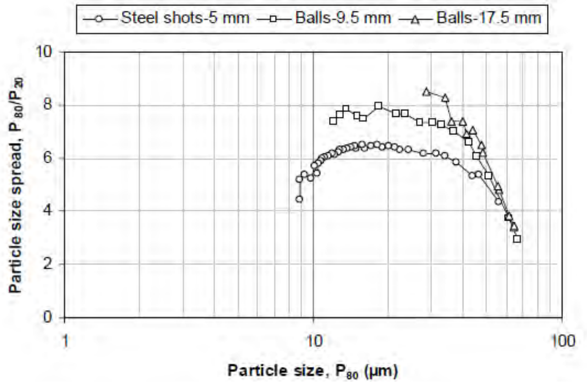


Figure 15. Vertimill product sharpness and media size (Nesset et al; 2006)

Comparison of Tower Mill Design and Operating Data

A short list of Tower Mill design and actual data is included in Table 5. This was compiled by publicly available data (Vendor installation list), AMIRA P336/P9 projects and Xstrata's own operating data.

Tower Mill Design vs. Operating Data

Application	Energy requirements for a target product size		Design/ Actual
	Vendor Design	Actual*	
Cu regrind	6 kWh/t for 45µm	13 kWh/t for 45µm	48%
Pb secondary grind	4.4 kWh/t for 63µm	7.9 kWh/t for 63µm	57%
Pb tertiary grind	7.1 kWh/t for 45µm	12 kWh/t for 45µm	63%
Zn regrind	5.7 kWh/t for 30µm	9.4 kWh/t for 30µm	61%
Zn regrind	19.4 kWh/t for 20µm	25.5 kWh/t for 20µm	76%
Pb regrind	16.7 kWh/t for 20µm	31 kWh/t for 20µm	54%
Ni regrind	11.7 kWh/t for 60µm	13.5 kWh/t for 60µm	84%
Fe regrind	9.4 kWh/t for 30µm	13.8 kWh/t for 30µm	68%

*Actual energy requirements were calculated using operating work index from plant surveys

Table 5. Tower Mill design and actual comparison

The operating work index was also calculated for all available full scale mill data and is shown in Figure 16. There is no design data included in this but it should serve as a guide as to what the mills are actually capable of. Recent regrinding designs falling in the lower area of the graph have significantly under quoted actual operating mills.

Tower Mill Operating Points

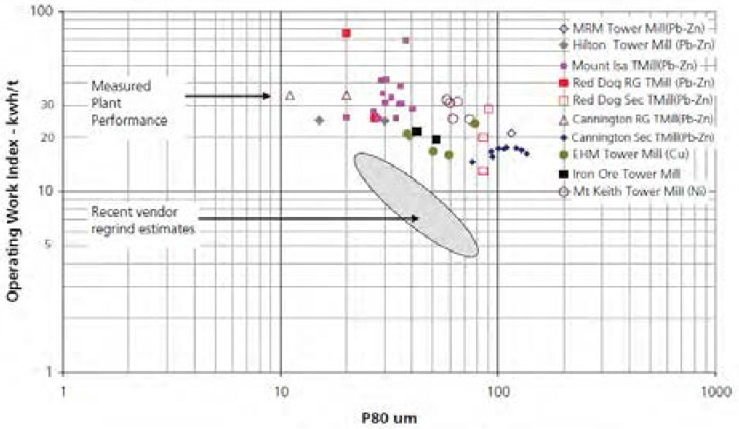


Figure 16. Tower mill operating work index summary (Pease J; 2010)

Modified Bond and Levin Tests:

The Levin test appears to be underused for sizing regrind ball mills. Examples of the results are shown in Table 6. The requirement of 20-30 kg may be a prohibitive factor. The Levin test is a modified Bond Ball test, but makes use of finer screen sizes, and finer feed sizing. In the Levin test the lab ball mill is run at varying lengths of time (energy). At the end of each interval the entire mill contents are emptied and screened at that intervals size. Any undersize is replaced with new top size to maintain a constant volume. This is completed from 75, 53, 45 and 38 microns. With 4 tests each requiring about 2 liters of material this could result in a requirement for more material than is possible to produce in a small pilot plant run given that it will likely be a rougher concentrate. However, if the material mass is possible it would appear that the test will give more accurate results than many of the alternatives.

COMPARISON OF PLANT DATA WITH RESULTS OF THE GRINDABILITY TEST

	kWh/t	
Materials and Origin	Plant	Grindability test for fine material
GOLD ore, East Driefontein	14.4	14.8
Gold ore, Libanon	10.8	11.3
Gold ore, Western Deep Levels, VCR Reef	14.7	13.8
Gold ore, Western Deep Levels, Carbon Leader	17.6	16.4
Fluorspar, Chemspar	4.3	3.3
Copper-lead-zinc ore, Black Mountain	12.8	11.1
Copper-zinc ore, Prieska	13.6	12.8
Sand tailing, Crown Mines	9.3	9.3

Table 6. Comparison of Levin test results with full scale ball mills (Levin J; 1989)

Operating conditions:

When downstream testwork has been done with product produced from one technology it should not be expected to transfer to a different grinding technology. Each will result in a different size distribution curve for a given P_{80}. The ratio between the P_{98} and the P_{80} should be considered for each.

This is important for the recovery in leaching as shown below. The more top size material present the less liberation there will be. Material ground to the same target P_{80} can have three times the top size present. This will also affect the final con grade during flotation as unliberated coarse particles will still float. Potentially these will contribute excess attached gangue minerals to the concentrate.

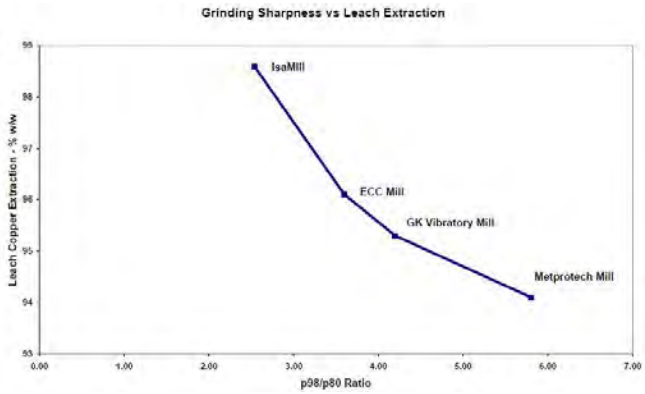


Figure 17. Effect of size distribution sharpness on leaching recovery (Pease J, Young M, Curry D; 2007)

Care should also be taken to do the scale-up testwork at the correct density. The effects of viscosity become more pronounced as the fine sizes common to regrinding are met. At some point depending on the surface area created the slurry will begin to carry the media rather than mixing it to grind. Xstrata Technology currently recommends the use of a Marsh funnel during testwork to control the density to optimize the effects of viscosity on energy efficiency. The Marsh funnel is simply a cone through which the time for the flow of one quart of slurry is measured. The Marsh funnel is not a comprehensive measurement of rheology but does serve to give a quick easy point of measurement ideal for use in multiple lab tests or in the field for site surveys.

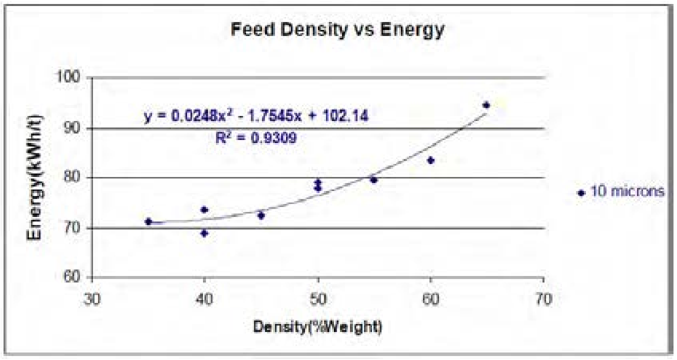


Figure 18. Effect of feed density on fine grinding efficiency (Larson M, Morrison R, Young M; 2008)

The Future:

Xstrata Technology is also striving to improve the scale-up ability of the IsaMill through sponsored and internal research. One of the most recent developments was the development of a JKSimMet model of the IsaMill. The basis of this model is a relatively simple function for predicting the creation of fines. The normal signature plot has one limitation in that the line created does not usually pass through the feed at 0 energy. This limits the signature plot to predicting energy requirements coarser than the band of sizes produced on the plot. It is possible to extrapolate finer however it is always best to actually cover those sizes in the testwork to be sure of the results.

The Squared Function for Fines Production also creates a linear relationship but also consistently passes through the feed at 0 energy. The model itself was inspired by McIvor's work (McIvor R and Finch J; 2007) demonstrating that new production finer than a certain target size is approximately linear for rod and ball mills. Plotting the percent passing a size vs. energy did not work for the IsaMill. It was found though that by squaring that % passing value a straight line through each point including the feed was developed as shown in Figure 19 for a variety of cooper ores. The only constraints are that the size shown has to be measureable in all of the samples. This will likely require a size between the F_{10} of the feed and the P_{90} of the product.

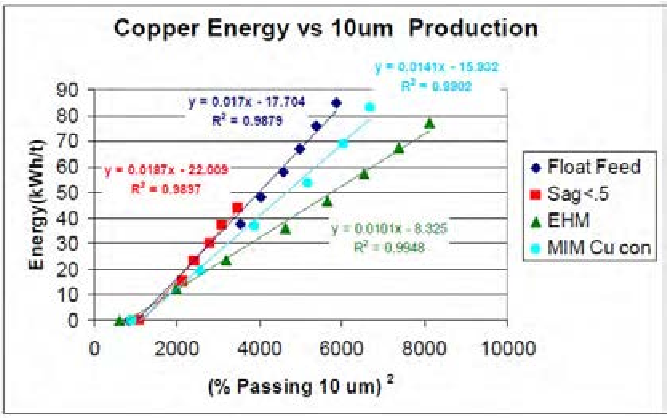


Figure 19. The squared function for fines production applied to different copper ores (Larson et al; 2008)

This method also enhances the ability to properly size a mill if the design feed size changes after testwork. By keeping the slope of the line parallel and changing the feed size a new energy can be calculated. In Figure 19 a signature plot line is plotted as the squared function. A new coarser feed can be plotted by calculating the new squared value and starting the new energy vs. size line at that point to the left of the original feed line. By keeping the new squared function line parallel to the original one a new energy can be calculated to the desired product.

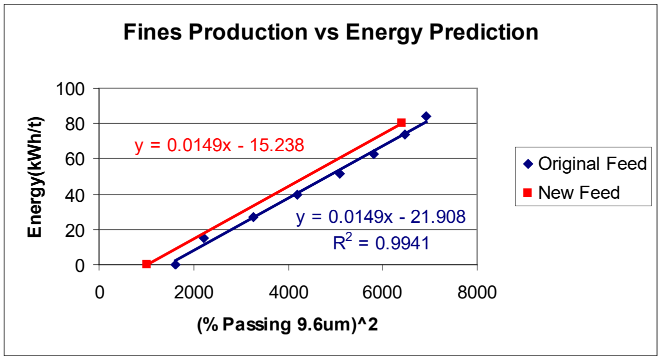


Figure 20. Squared function for fines production energy prediction for coarser feed (Larson; 2010)

The likely reason the squared function works is that the attrition grinding process is creating new surface area in a consistent manner. The IsaMill also results in a rounding of particles. As this should be the same for all attrition grinding processes there is a good probability that the squared function or something similar will apply to those as well.

Conclusions:

Currently a wide range of tests are used to predict regrinding requirements, and they result in enormous variation in predicted energy requirement. Clearly serious mistakes are being made, which will either lead to installing too much power, or too little power and resultant plant underperformance. The industry needs an accepted standard test to reliably predict regrinding energy. An essential part of this is an understanding of actual plant regrinding performance compared with design. The JKMRC Fine Grinding review aims to address these issues. Operators, designers and engineering companies should support a rigorous, independent review to establish industry performance and an industry standard.

Until this happens current scale-up methods have to be rigorously examined to ensure the regrind mills will perform as claimed. Designers should perform reality checks on all vendor estimates – not against other *designs*, but against actual *performance* of other installations.

Acknowledgements:

The authors wish to thank Xstrata Technology and the JKMRC for the research that went into this paper and for the helpful suggestions in writing it.

References:

1. Barns K, Curry D; 2006; Stirring the Pot: A New Direction for IsaMilling; Ultrafine Grinding 06
2. Cole J, Wilmot J; 2009; MORENCI CONCENTRATE LEACH PLANT FIRST YEAR REVIEW; SME 2009 Annual Meeting; Denver USA
3. Curry D, Clark L, Rule C; 2005; Collaborative Technology Development-Design and Operation of the World's Largest Stirred Mill; Proceedings 2005 Randol Conference; Perth Australia
4. Farber B, Durant B, Bedesi N; 2010; Effect of Media Size and Mechanical Properties on Milling Efficiency and Media Consumption; Comminution 10; Cape Town, South Africa
5. Gao M, Reemeyer L, Obeng D, Holmes R; 2007; Efficiency of the Detritor Mills at Zinifex Century Mine; SAIMM
6. Gao M; CSIRO report
7. He M; 2010; JKMRC private report to Xstrata Technology
8. Kazakoff J, Mortimore A, Smith S, Curry D; 2006; Introduction of IsaMill Technology into a Major Gold Flotation and Leaching Operation in Central Asia; Comminution 06; Perth Australia.
9. Larson M; 2010; Development of a Simulation Model of an IsaMill; MPhil Thesis, University of Queensland JKMRC
10. Larson M, Morrison R, Young M; 2008; Improving Grinding Efficiency With The IsaMill; Comminution 08, Falmouth England
11. Larson M, Villadolid V, Xstrata Technology Internal Report-Iron Ore, May 2010
12. Levin J; 1989; Observations on the Bond standard grindability test, and a proposal for a standard grindability test for fine materials; SAIMM Journal, January 1989
13. McIvor R and Finch J; 2007; The Finch-McIvor Functional Performance Based Grinding Circuit Modeling System; SME Annual Meeting
14. Nesset J, Radziszewski P, Hardie C, Leroux D; 2006; ASSESSING the PERFORMANCE and EFFICIENCY of FINE GRINDING TECHNOLOGIES; CMP; Ottawa Canada
15. Nippon Eirich Brochure; 2009
16. Pease J; Elephant in the Mill; 2010; XXV IMPC Proceedings; Brisbane Australia
17. Pease J, Young M, Curry D; 2007; Fine Grinding as Enabling Technology- The IsaMill

Comparing Energy Efficiency in Grinding Mills

B D Burford[1] and E Niva[2]

1. Senior Process Engineer, Xstrata Technology, Level 4, 307 Queen Street, Brisbane Qld 4000. Email: BBurford@xstratatech.com.au
2. Undergraduate Student, Luleå University of Technology, Sweden, working for Xstrata Technology, Level 4, 307 Queen Street, Brisbane Qld 4000.

ABSTRACT

The IsaMill™ is challenging the way that plants are designed and operated. This paper challenges existing designs of concentrator flow sheets, particularly focusing on magnetite circuits.

From what has started out as a small scale ultra fine grinding mill in the pharmaceutical and pigment industries, it has been redesigned and improved upon for mineral processing, and has been the mainstay of fine grinding applications for over 10 years. These applications have required energy efficient grinding to succeed, and have been predominately in the base metals industry.

Further development of the IsaMill™ has now resulted in the machines being able to treat larger tonnages, with higher capacity motors. This development, along with the introduction of purpose designed ceramic media, has allowed the mill to treat courser feed sizes. At the same time, the mill still offers highly efficient grinding, and has enabled it to be operated in coarser tertiary and secondary grinding applications.

The acceptance of the mill in coarser applications, predominately in base metals and PGM applications, has enabled the IsaMill™ to be a serious contender for beneficiation in other minerals. One such application is the potential for IsaMills™ to be part of magnetite flowsheets, which are being considered in Australia to meet the growing iron demand of China. The high energy efficiency of the IsaMill™ compared to conventional technologies, as well as the smaller infrastructure requirements, provides a great opportunity to reduce the power intensity of magnetite circuits.

This paper will examine the use of IsaMill™ technology in conventional grinding applications, including recent testwork in a primary grinding base metal circuit, as well as testwork on magnetite ore comparing a lab scale IsaMill™ with a lab scale Tower Mill.

The growing demand for minerals over the next decade, coupled with higher energy cost, will result in energy efficient technology, such as the IsaMill™, being included in standard circuit design.

INTRODUCTION and BACKGROUND

The development of IsaMill™ technology was driven by the metallurgical requirements of fine grained Lead/Zinc deposits at Mount Isa in Queensland and McArthur River in the Northern Territory, both of which were controlled by Mount Isa Mines Limited (now Xstrata).

The complex nature of both deposits required ultra fine grinding to sizes that were not possible to do economically with the technology that was available in the early 90's.

McArthur River orebodies were mineralogically complex, and required regrinding down to 7 µm to achieve sufficient liberation to allow the production of a bulk concentrate (Enderle et al, 1997; Pease et al, 2006). In the case of the Mount Isa orebodies, there was a gradual decrease in plant metallurgical performance from the mid 1980's as a result of decreasing liberation size and increased amounts of refractory pyrite in the ore that saw recovery decrease from 70% to 50% (Young et al, 1997; Pease et al, 2005; Pease et al, 2006). However using conventional ball and tower mill technology to achieve finer grinding for mineral liberation was uneconomic, as well as resulting in a high rate of steel media consumption which contaminated the mineral surfaces with iron deposits, resulting in poor flotation response post regrinding.

In both of these orebodies, a need had arisen for a technology that could grind to ultra fine sizes in metallurgical operations economically without serious contamination of mineral surfaces and pulp chemistry. Testwork was undertaken in the early 90's at Mount Isa Mines into high speed horizontal stirred mill technology, which was used in pigment and other industries. It was shown at pilot scale that such mills could grind down to the ultra fine sizes required for mineral liberation.

Arising from these findings, a program of major mechanical modification of horizontal stirred mill technology was undertaken between Mount Isa Mines Limited and Netzsch-Feinmahltechnik GmbH (Enderle et al, 1997), the manufacturer of the stirred milling technology, to make the technology more applicable for the mining industry.

After many prototypes, the first full scale model was developed and installed at the Mount Isa Mines' Lead Zinc Concentrator in 1994. The mill, the M3000 IsaMill™, was quickly installed in other circuits at this concentrator, and was installed in the McArthur River Concentrator in 1995 (Johnson et al, 1998). Later, in 1999, it was commercialised and sold outside of the Xstrata group.

Since commercialisation of the IsaMill™, there is now over 70 MW of installed IsaMill™ power operating around the world, treating materials such as copper/gold, lead/zinc and platinum. While the early installations were applied to ultra fine duties, the IsaMill™ today is being applied to coarser grinding applications, once the domain of tower and ball mills. The application to coarser duties, means all the advantages of the IsaMill™ that was developed for ultra fine grinding can now be applied to the coarser applications.

IsaMill™ OPERATION

The IsaMill™ is a horizontally stirred mill consisting of a series of discs rotating around a shaft driven through a motor and gearbox, at speeds ranging from 21-23m/s, with energy intensities up to 300kW/m^3. The general layout of the IsaMill™ is displayed in Figure 1, while the grinding mechanism is displayed in Figure 2.

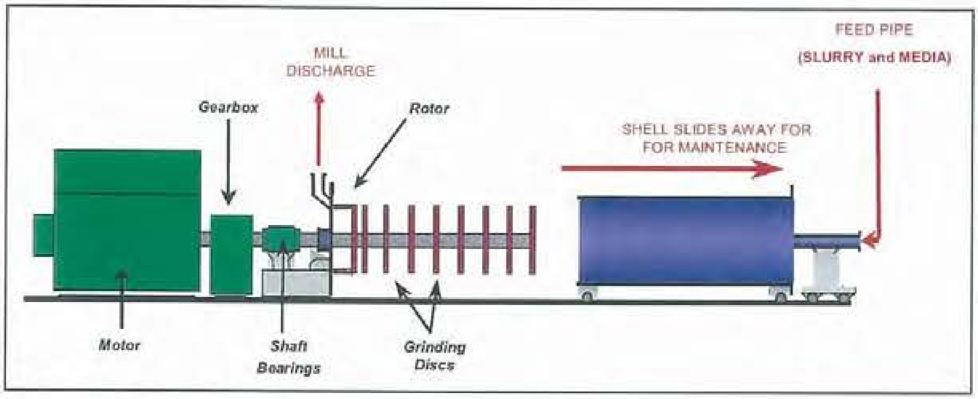


Figure 1: IsaMill^{TM} Layout

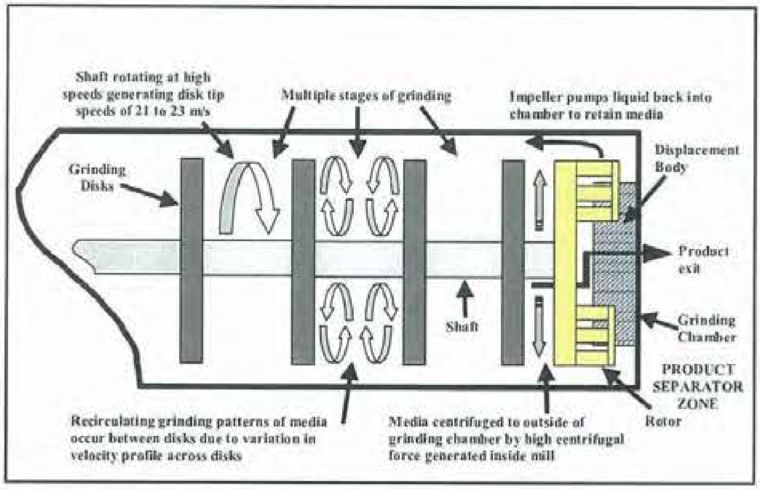


Figure 2: IsaMill^{TM} Grinding Mechanism

In operation, the mill is filled with grinding media between each disc, each one of these segments acting as an individual grinding chamber. When 8 discs are used in the mill, it effectively acts as 8 grinding chambers in series, which also minimises any short circuiting of the mill feed to the discharge. The action of the grinding discs when rotating, radially accelerates the media towards the shell. Between the discs, where the media is not subject to the high outwards acceleration of the disc face, the media is forced back in towards the shaft – creating a circulation of media between each set of discs. Minerals are ground by an attrition action, as a result of the agitated media, with the resulting high energy efficiency being achieved due to the high probability of media-particle collision.

Energy Intensity

The high tip speed of the IsaMill^{TM} results in a high energy intensive environment. Energy intensity of the IsaMill^{TM} is significantly higher than any other commercially available grinding equipment, as illustrated in Table 1. Combining the energy intensity and the high grinding efficiency leads to a compact mill, able to be fitted into existing plants where floor space is limited.

	Installed Power (kW)	**Mill Volume (m^3)**	**Power Intensity (kW/m^3)**
Autogenous Mill	6400	353	18
Ball Mill	2600	126	21
Regrind Mill	740	39	19
Tower Mill	1000	12	42
IsaMill™ - M10,000	3000	10	300

Table 1: Comparative Energy Intensity of Grinding Technologies

Media

The key to the energy efficiency of the IsaMill™ is the ability to use fine media. While tower mills are typically limited to 10-12mm fresh media sizing, the IsaMill™ can use media as small as 1mm. This results in significantly more surface area per unit volume of media in the IsaMill™ compared to a tower mill.

The IsaMill™ is versatile and able to use a range of media types, including low cost, locally available media such as sand or smelter slag, to provide good grinding performance at acceptable energy efficiency. However, the need for improved energy efficiency at many installations has resulted in the use of high quality, high density ceramics, designed specifically for stirred milling applications, such as Magotteaux Keramax ® MT1™.

Media Retention

Grinding media is retained in the mill without the need for screens, which is why IsaMills™ can use fine media. At the end of the mill is a patented product separator consisting of a rotor and displacement body. The distance between the last disc and the rotor disc creates a centrifuge, so that coarse particles and media move to the outside of the mill, which are pumped back towards the feed end of the mill from the action of the rotor. Meanwhile the fine particles flow through the rotor and discharge from the mill, which means no screens or cyclones are required, and allows the mill to be operated in open circuit without cyclones, reducing capital and simplifying circuit configuration.

Product Size Distribution

In open circuit operation, the IsaMill™ is able to produce a sharp product size distribution, which reduces overgrinding and the creation of ultra fines. Typically the ratio of the P98 to the P80 is around 2.5. This is a direct result of the individual grinding chambers acting in series, preventing short circuiting, as well as the classification action of the product separator. The ability to operate the mill in open circuit greatly simplifies the operating and maintenance strategies of the circuit. Also the sharp product size is beneficial in pipeline design and filtration, due to the reduction of ultra fines and oversize particles.

Inert Grinding

The operation of the IsaMill™ using sand or slag, or more often ceramic grinding media, has a big advantage over steel media in conventional grinding systems, as it greatly reduces the generation of ferric ions. These are detrimental to flotation and leaching circuits, as the ions form a coating on the mineral surface, which hinders the action of the flotation or leaching reagents, resulting in more reagents being used, and may also result in poor metallurgical performance.

Maintenance

The IsaMill™ has been designed to keep maintenance simple. The shell of the mill is simply rolled away from the mill on a set of rails, enabling the disc and internal wear surfaces to be examined and changed if required (Figure 3). The shell liner of the mill is easily replaced as the shell is designed in two pieces.

Wear within the mill is determined by the specific size reduction of the mill, as well as wear characteristics of the minerals, and it is common for IsaMills™ to be operating with availabilities 96% and higher.

Figure 3 – IsaMill™ Maintenance

COARSE GRINDING TRANSFORMATION

The IsaMill™ was developed to enable the fine grained orebodies of McArthur River and Mt Isa to be developed (Enderle et al. 1997; Pease et al, 2006). Grinding down to a P80 of 7um at high energy efficiency was a big step forward in mineral comminution, however only a small number of mine sites needed grinding down to this size.

However, the development of ceramic media and M10,000 IsaMills™ in recent years has enabled the IsaMill™ to treat coarser feed materials in tertiary and even secondary grinding duties, which has resulted in greater application of IsaMills™ in most concentrators, (Burford, 2007).

While the use of low cost natural media and slag was used in initial IsaMill™ applications, the quality variability and the certainty of supply had a big impact on IsaMill™ operation. In particular the variability of the media shape, SG and size constrained the energy efficiency of the mill when operated with sand or slag, (Curry et al 2005b)

The development of ceramic media designed for use in IsaMills™ by Magotteaux International, was a major step forward for application of IsaMills™ in coarse grinding. This was due to the media having good structural integrity, tough, high SG as well as being designed up to 3.5mm in diameter. (Anderson et al 2006)

In terms of the energy that can be provided by the media particle, the development of larger diameter media made from ceramic increases the energy available for grinding due to the increased diameter of the media, and the increased SG of the ceramic. In terms of the Keramax® MT1™, the SG of the ceramic is 3.7, over 40% higher than that of sand, (SG = 2.6). This is described by the Stress Intensity Relationship in Table 2 (Pease 2007).

$$\mathbf{E \ \alpha \ d^3.v^2.SG}$$

E = Energy per Media Particle
d = media diameter
v = media velocity
SG = media density

Table 2 – Stress Intensity Relationship

The other development in the transformation of the IsaMill™ from fine to coarse grinding applications, was the development of the larger M10,000 IsaMill™ (Curry et al 2005a). As previously described, the WLTRP project by Anglo Platinum in South Africa required large scale grinding mills to treat 53 tph, up to a maximum of 65tph, from a P80 of 75µm down to a P90 of 25µm. This duty required 35 kWh/t, and would have involved multiple numbers of the M3000 IsaMills™. However a joint development between Anglo Platinum, and Xstrata Technology, enabled the much larger M10,000 to be developed. Not only was this mill nearly 3 times bigger in volume than the M3000, it was powered by a 2.6MW, and provided considerable more energy available for grinding, (Figure 4). Later versions were supplied with 3MW motors (Anderson 2006), such versions offering 300kW/m^3. Larger flow rates could now be treated by IsaMill™ technology.

With the developments of ceramic media and M10,000 IsaMills™, energy efficiency and other benefits that were common in fine grinding circuits, were now transferred to coarse grinding applications.

Figure 4 – M10,000 IsaMill™

IsaMills™ AND THE POTENTIAL OF MAGNETITE GRINDING

The increasing worldwide demand for iron ore has triggered the development of Australia's magnetite resources. While once regarded as uneconomic to process, they are now being seriously considered as a potential iron source at the current iron ore prices, driven by high demand from the China.

Magnetite has been regarded as uneconomic to process in Australia due to the infrastructure requirements to produce an iron concentrate, as well as the high energy requirements to grind the ore. Haematite deposits in the Pilbara region of Western Australia have always been the preferred source of iron ore in Australia, due to its relatively low cost mining and processing methods, and its high quality and abundance. However, the surging iron ore price and the high demand for iron, has resulted in many magnetite projects, that were once regarded as marginal, being regarded as commercial propositions.

To date, there are approximately 5 projects planning to treat magnetite ore, scheduled to be started up over the next 5 years, with many others being considered. By 2014 this would mean approximately 40MT of ore will need to be processed yearly (Gardner-Bond, 2008; Australian Resources, 2008).

Magnetite is also a key source of iron in many countries where haematite resources are not present, and regions such as Northern Africa, Central Europe, China, and North America have operations mining and treating magnetite. In these operations, traditional comminution technology such as ball mills and tower mills are common practice, and the large tonnages treated in these plants results in large amounts of power being used for grinding the ore.

However, with the development of IsaMill™ technology for coarse grinding applications in base metals, there is also significant potential for the IsaMill™ to be used in magnetite operations. More energy efficient grinding in these plants with IsaMill™ technology, could result in less installed power being required, compared to traditional plant design utilizing ball mills and tower mills, with less infrastructure required such as cyclones and pumps.

MAGNETITE TESTWORK

The objective of the test was to do a side by side comparison for a magnetite sample being ground by an IsaMill™ and a tower mill to determine the signature plot of each mill. Both mills were to be run in open circuit. Davis tube testing was also undertaken on the samples after grinding.

The testwork was undertaken at the CSIRO facilities at Pullenvale, Queensland. At this site there is a M4 IsaMill™ lab scale unit, as well as a small scale tower mill.

Magnetite ore for the testwork was sourced from Ernest Henry Mine tailings (EHM). This mine is a copper mine in North West Queensland, with the host rock being a breccia, which is comprised of strongly altered and replaced felsic volcanic fragments in a matrix assemblage of predominantly magnetite, chalcopyrite and carbonate minerals. Post flotation, the majority of the chalcopyrite and some other sulfides have been separated from the gangue stream, leaving it rich in magnetite.

The M4 IsaMill™ is a 4 litre mill, containing 7 discs for this testwork, operating with 3.5mm ceramic Keramax® - MT1™ media. The tower mill is 40L capacity and operates using 12mm steel media. These are displayed in Figure 5.

Figure 5: M4 IsaMill and small scale Tower Mill

For the tests, the as received sample under went sample preparation and preliminary grinding to reduce the feed size from 163um to 113um to eliminate oversize particles blocking the test rigs.

In each test, 20 to 21kg of sample was made into slurry of 50% solids and pumped through each mill for a number of runs. The power used for each run is recorded, and a small sample of the discharge is taken for laser sizing. This procedure is carried out through each mill for a minimum of 12 times, or until there is no significant reduction in the sizing, i.e. the mill cannot reduce the sample any further. The data is then used to draw a signature plot (a log-log graph of P80 size versus the specific energy to obtain that size), as displayed in Figure 6.

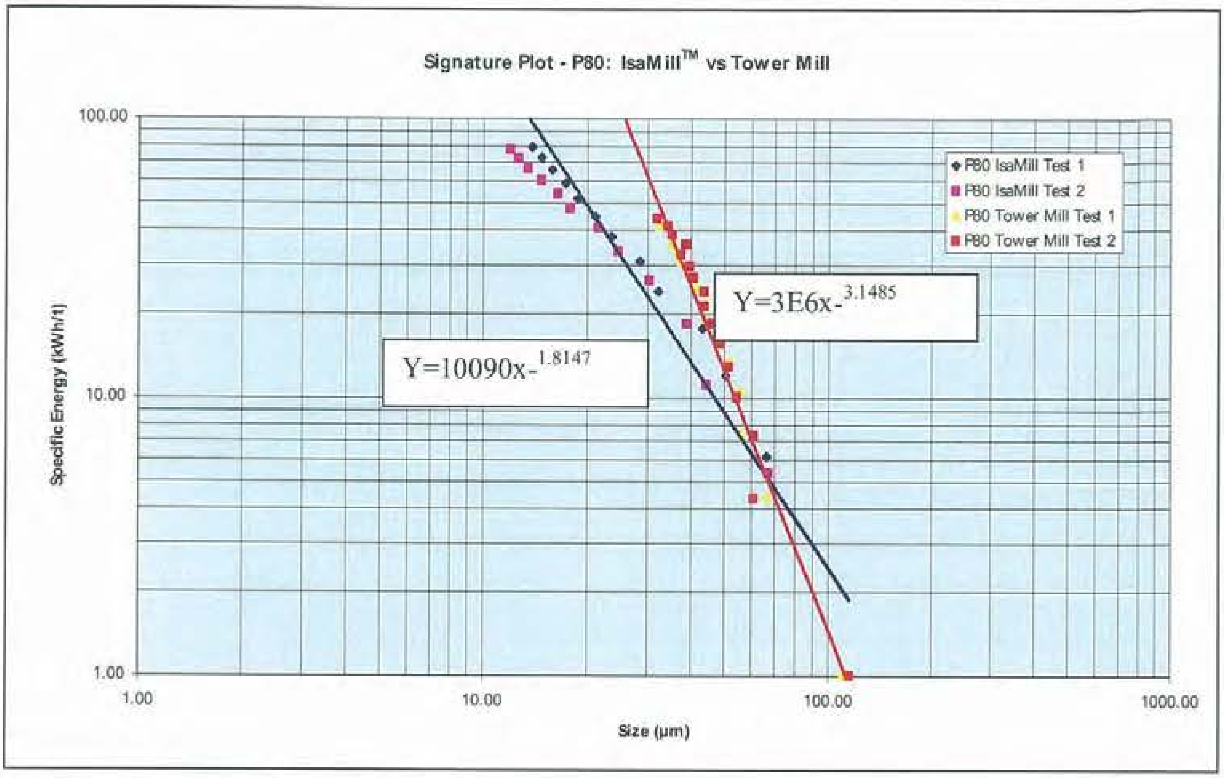


Figure 6 – Size versus Specific Energy - M4 IsaMill™ and Tower Mill

Grinding Results

The signature plot for the two tests, for each grinding technology, shows a good level of reproducibility.

The IsaMill™ was able to reduce the top size of the feed, at a F80 of 113µm, down to a P80 of 13µm. The tower mill also treated the same feed size, however couldn't produce grind sizes down any further than 31µm. For the testwork, a charge similar to what a full scale grinding unit would use was used. In the case of the IsaMill™, 3.5mm ceramic media (Magotteaux Keramax® MT1™) was used, while in the tower mill, 12mm steel media was used. As expected, the smaller media in the IsaMill™ enables finer grind sizes to be achieved, while the 12mm balls in the tower mill limits how fine the tower mill can grind.

The flatter curve for the IsaMill™ signature plot indicates less energy required to achieve grind size, than the steeper curve that was obtained with the tower mill. This difference has a big impact on the energy needed to grind down to the required product sizes. For instance, to reduce a theoretical sample with a P80 from 100 µm to 30µm, using the tower mill will take approximately 39 kWh/t. However to do the same size reduction with an IsaMill™ will take only 13 kWh/t, only a third of the tower mill power. The key to the signature plots is that for the coarser sizes, there is only a small power requirement to grind the coarse sizes. However as the size required becomes smaller, there is an exponential increase in the power required. Therefore while the tower mill may be more power efficient at sizes greater than 65µm, a reduction in particle size less than 65µm for this sample will result in the IsaMill™ being more efficient.

One scenario that was not tested was using the tower mill in a closed circuit, as is often the case in practice. However setting up such an experiment is difficult due to recirculating loads and ensuring the cyclones cut efficiently. This is one of the practical drawbacks of closed circuits, in that cyclones never operate efficiently, and are often poorly maintained, and small diameter cyclones required for fine cuts, are prone to blockages. Also the associated power of pumping at a reasonable pressure for the cyclone to operate needs to be taken into account in the energy use in these circuits.

The signature plot was restrictive in the scale that could be achieved with the IsaMill™, as the feed for each test was maintained at a P80 of 113µm. In practice, coarser grinding can take place at increased sizes between P80's of 250 to 300µm, and larger media is being developed to handle even coarser sizes.

Davis Tube Results

The samples from the grinding testwork underwent Davis Tube testing, which involved the separation of the magnetics from non magnetics using a small scale magnetic separator. The iron grade versus iron recovery obtained from this testwork is shown in Figure 7. The grade recovery relationship indicates the maximum grade for the ore type was 71% iron, at a maximum iron recovery of 90%. The assay from the magnetic separation from the Davis Tube testwork gives an indication of how the liberation of the minerals occurs as the particle size reduces.

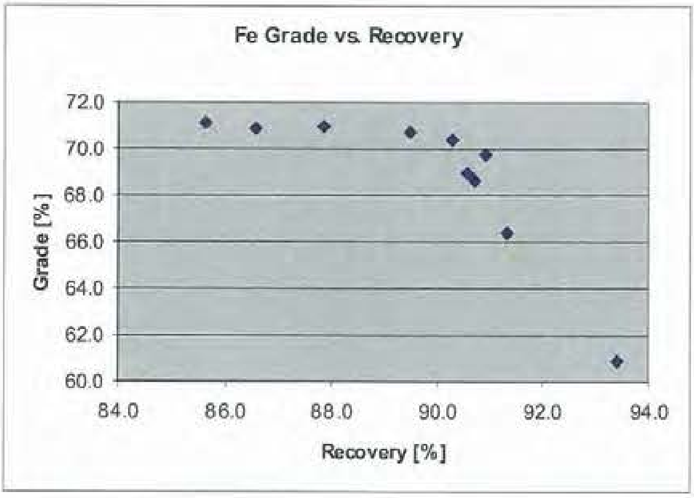


Figure 7 – Iron Grade vs. Recovery

The EHM tailing that was received and used in the test work contains 43% iron, with a P80 of 163µm. When this material under went magnetic separation without regrinding a concentrate containing 61% iron was achieved at an iron recovery of 93%. At this grade, the silica, sulfur, alumina and phosphorus levels are 9.3%, 0.34%, 2.01 and .02% respectively.

The data was also plotted to produce a grade versus size curve as shown in Figure 8.

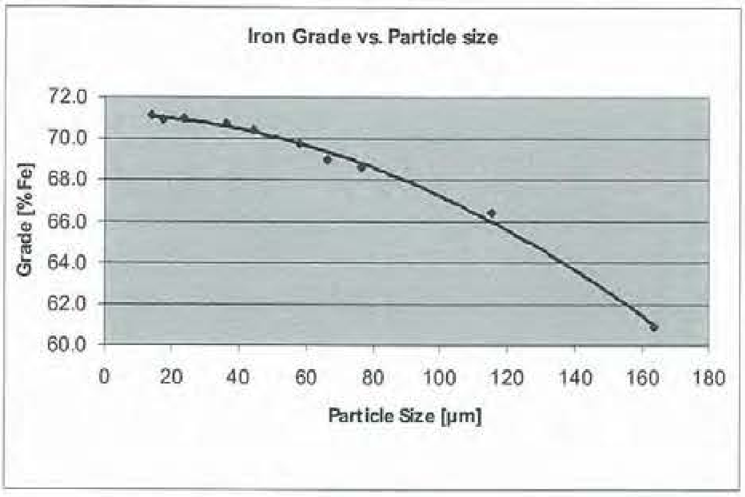


Figure 8 - Grade vs. Size Curve

Figure 8 shows that a 70% iron grade is achievable with a grind size P80 of 50 µm. Further grinding will improve the grade marginally to a 71% iron grade. As in any grinding circuit, the benefits of increased concentrate grade needs to be weighed up against the extra grinding power that is required. In cases where the grind size is quite fine, the increase in grade requires an exponential increase in grinding energy and could well require another grinding unit to achieve.

Figures 9, 10, 11 and 12 also show the particle size grade relationships of the impurity elements, silica, sulfur, alumina and phosphorus. As expected, grinding finer and separating the magentics from the non magnetics, will result in less of the non ferrous impurities reporting to the magnetics as they are liberated by finer grinding. At a grind size P80 of 50µm, silica, sulfur, alumina and phosphorus levels have dropped to 1.6%, 0.045%0.035 and .003% respectively.

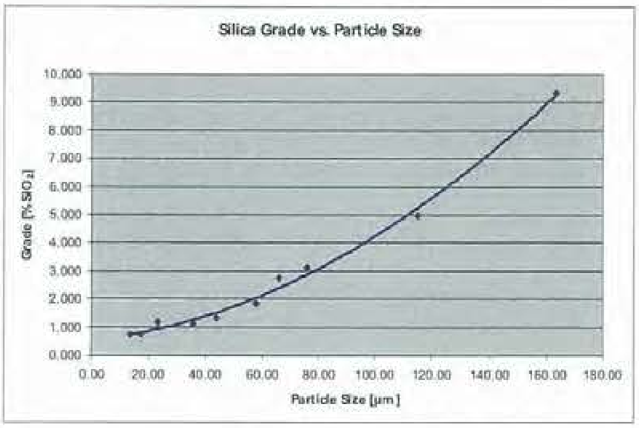


Figure 9 – Size vs. Silica Grade

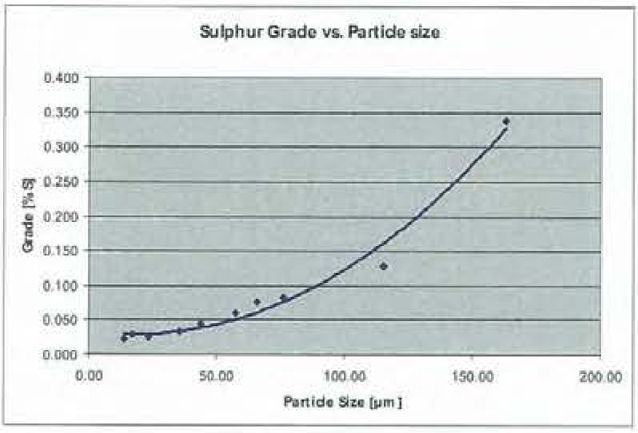


Figure 10 – Size vs. Sulfur Grade

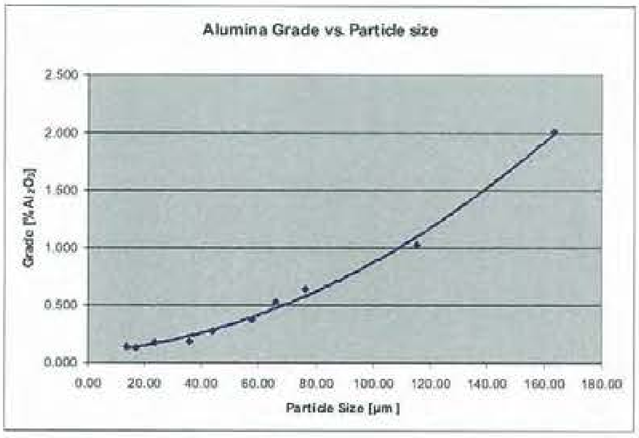


Figure 11 – Size vs. Alumina Grade

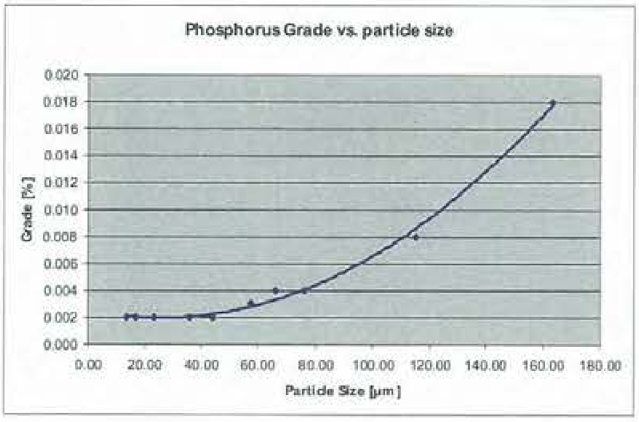


Figure 12 – Size vs. Phosphorus Grade

	Size (µm)	%Fe	%SiO$_2$	%S	%Al$_2$O$_3$	%P	%Cu	%As	%Co
As Received	163.30	60.890	9.340	0.339	2.010	0.018	0.083	0.007	0.009
Ground Feed	115.35	66.440	4.980	0.128	1.030	0.008	0.037	0.004	0.006
Ground Sample 1	76.20	68.620	3.090	0.082	0.640	0.004	0.022	0.003	0.005
Ground Sample 2	66.20	69.000	2.730	0.076	0.530	0.004	0.020	0.003	0.005
Ground Sample 3	57.83	69.780	1.840	0.060	0.370	0.003	0.016	0.002	0.005
Ground Sample 4	44.14	70.420	1.330	0.045	0.270	0.002	0.015	0.002	0.005
Ground Sample 5	36.04	70.690	1.080	0.033	0.180	0.002	0.012	0.002	0.005
Ground Sample 6	23.46	70.980	1.150	0.026	0.170	0.002	0.010	0.000	0.004
Ground Sample 7	17.27	70.900	0.720	0.028	0.130	0.002	0.011	0.000	0.004
Ground Sample 8	13.82	71.120	0.730	0.022	0.140	0.002	0.010	0.002	0.004

Table 2 – Assay per Size Fraction – Magnetic Concentrate

In relation to other magnetite ores, there have been several M4 IsaMill™ tests conducted on other deposits, although not as many as undertaken on base metal deposits. The magnetite samples come from deposits in the Yilgarn Craton in Western Australia, and Central Europe

The M4 IsaMill™ on the magnetite material to date, indicate that it is in the middle of the range in terms of the power required to reduce the size of the sample. Magnetite, copper and PGM (Platinum Group Minerals) signature plots using an M4 IsaMill™ have been plotted on Figure 13, as well as in Table 3, for the power required to grind samples from an F80 of 45 µm, to a product size P80 of 25µm. This range was chosen as the majority of magnetite samples that have been tested have been in this range. However, as observed in Table 2, the IsaMill™ could treat much coarser magnetite feed sizes than these.

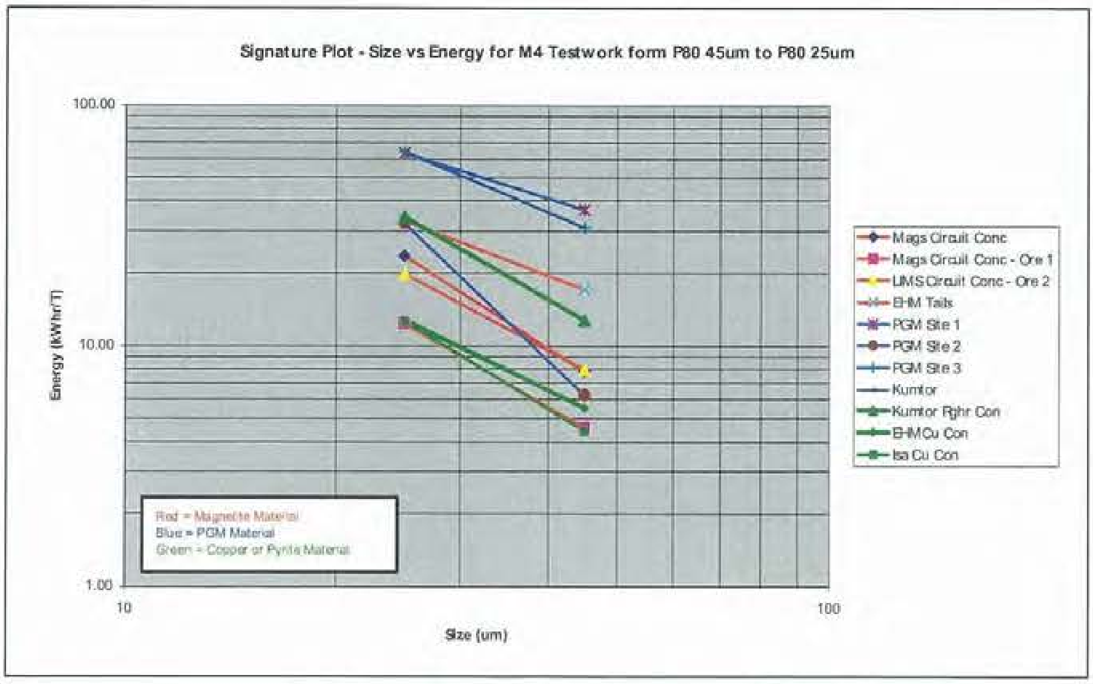


Figure 13 – Signature Plots of Different Materials

Material	Circuit	kWh/t
Magnetitie Material	Mags Circuit Conc	15.8
Magnetitie Material	Mags Circuit Conc - Ore 1	7.8
Magnetitie Material	LIMS Circuit Conc - Ore 2	12.0
Magnetitie Material	EHM Tails	16.0
PGM Material	PGM Site 1	26.2
PGM Material	PGM Site 2	26.1
PGM Material	PGM Site 3	33.1
Pyrite Material	Kumtor Rghr Con	21.5
Copper Material	EHM Cu Con	7.3
Copper Material	Isa Cu Con	8.2

Table 3 – Power Requirement for Grinding from 45µm to 25µm

The information from Table 3 and Figure 13 indicates that the 4 magnetite samples required 8 to 16 kWh/t to achieve the grind required. The hardest of the magnetite material was the EHM tail sample. In comparison to the EHM copper concentrate sample, EHM Cu Con, which is floated off before the tailing, and contains mainly chalcopyrite and other sulfides with low levels of magnetite, required less than half the energy compared to energy required to do a similar size reduction for the EHM Tails sample.

The PGM Material came from several sites across the Bushveld in South Africa. The power required for the grinding duty ranged from 26 to 33 kWh/T, and was significantly higher than the energy to grind the magnetite sample, which ranged in energy from 8 to 16 kWh/t. Both PGM Site 2 and PGM Site 3 have M10,000 IsaMills operating, treating the PGM minerals which are associated with the hard chromite host rock. The other site in the list where a M10,000 IsaMill™ operates is the Kumtor mine in Kyrgyzstan, which treats a pyrite concentrate at 21.5 kWh/t.

NEW INSTALLATIONS USING IsaMill™ in COARSE DUTIES

Phu Kham Project

The Phu Kham deposit is located approximately 100km north of the Laos capital Vientiane. It is owned by Pan Australian, an Australian listed mining company. The Phu Kham deposit hosts two distinct styles of mineralization: an oxide gold cap and beneath this transitional/primary copper-gold. The Phu Kham oxide gold cap is the principal deposit for the Phu Bia heap leach gold mine, the first phase of the development of the Phu Kham deposit, which entered into production in 2005. Feed to the concentrator will consist of 12MT on average, with planned annual output from this mine being over 200,000 tonnes of concentrate (grading 25% copper), containing 50,000 tonnes copper, 40,000 ounces gold and 400,000 ounces silver, (on average). The concentrate will be exported for further treatment and refining by custom smelters in the Asia Pacific region.

Process technology employed for Phu Kham Copper-Gold is conventional comminution at the head of the circuit, followed by flotation to produce a copper-precious metal concentrate, (Pan Australian, 2006)

Rougher concentrate will be treated through a M10,000 IsaMill™, powered by a 2.6MW motor, treating approximately 168 tph and reducing the feed size from a F80 of 106µm to a P80 of 38µm, before further flotation. The grinding media for the operation will be MT1. The Phu Kham Copper-Gold operation is planned for start-up in mid 2008.

Prominent Hill Project

Oxiana Limited owns 100% of the Prominent Hill copper-gold project located 650 kilometers north west of Adelaide, and 130 kilometers north west of BHP Billiton's Olympic Dam in South Australia.

The ore body consisting of a copper gold breccia, will be mined via an open pit. The ore will be treated through a conventional grinding and flotation processing plant, with a designed capacity of 8MTPa. The initial planned concentrate production will be on average 187,000 tpa, peaking at 230,000 tpa in 2009, with average concentrate grades of 45% copper, 19g/t gold, 57g/t silver. The high grade concentrate will be sold to smelters in Australia and Asia. (Oxiana, 2007)

One M10,000 IsaMill™, powered by a 3.0MW motor, has been selected to treat the rougher concentrate. It will treat approximately 138 tph, reducing the feed size from a F80 of 125μm to a P80 of 24μm for further flotation. The planned commissioning of the mill will be mid to late 2008. The grinding media for the operation will be MT1.

Anglo Platinum Installations

In 2007, Anglo Platinum had ordered five, M10,000 IsaMills™, following the successful installation and operation of the first M10,000 IsaMill™ at their Western Limb Tailings Retreatment Project in 2003. The typical duty of these installations is from 75-100 μm feed size down to 53 μm product size.

Installations have been successfully commissioned at Potgietersrust Platinum mine (C-Section), Potgietersrust A and B Sections (2 mills), and two more at the Rustenburg Watervaal UG2 operation.

The Potgietersrust Platinum mine (C-Section) mill is designed to operate with a 3MW motor and use MT1 media, treating scats from A and B section primary milling circuits, with the ore having a Bond Work Index (BWi) over 30 kWh/t. Figure 14 illustrates the simplified C section flowsheet using an IsaMill™.

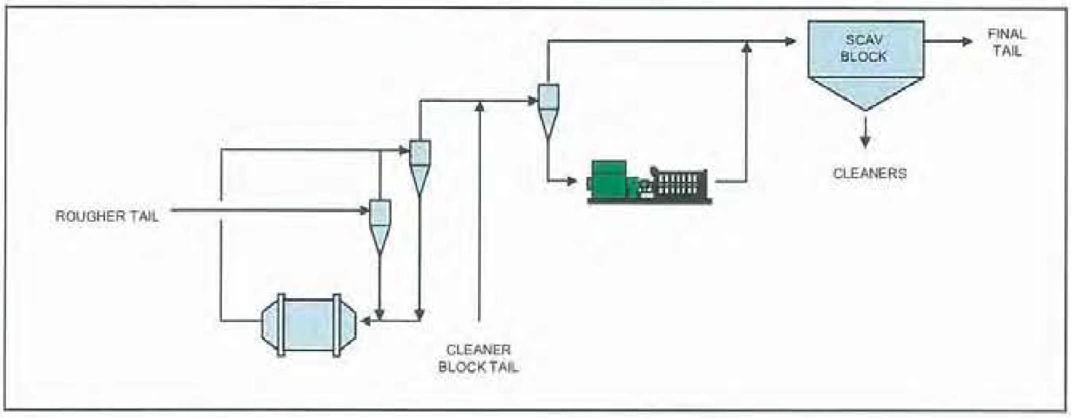


Figure 14: Simplified PPL C Section Flowsheet with a M10,000 IsaMill™

McArthur River Mines (MRM)

McArthur River Mine (MRM) is a zinc/lead mine in the Northern Territory, Australia, and is operated by Xstrata Zinc. It was commissioned in 1995, and was where the IsaMill™ was developed to regrind streams down to a P80 of 7 µm, which was the enabling technology that allowed the mine to be developed. Initially there were 4 x M3000, 1.1MW IsaMills™ in the regrind duty. This has since been expanded to 6 IsaMills™ with a combined installed power of 6.7MW. The current plant flowsheet is shown in Figure 15.

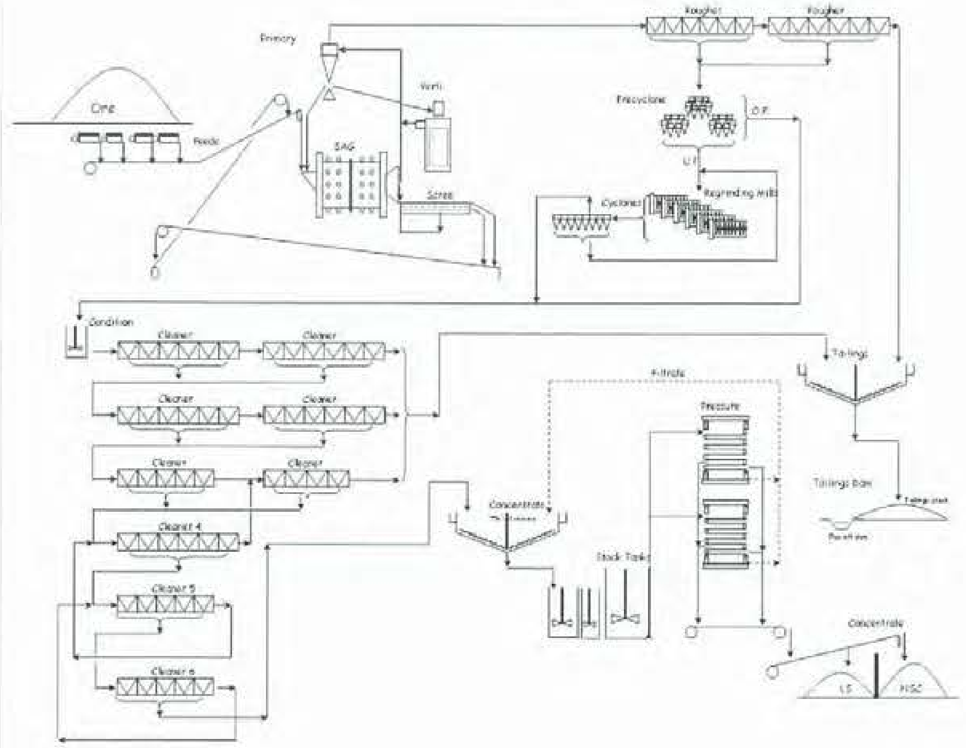


Figure 15: McArthur River Flowsheet

MRM have a need to increase milling capacity from 230tph to 305tph to account for decreased head grades as the operation shifts from underground to open cut. Flotation feed size is also to be reduced from the current P80 of 75µm, back to the original size of a P80 of 45 µm. At the same time there was a desire to reduce downtime and reduce operating cost by eliminating the Tower Mill from the circuit, hence MRM have been keen to explore the effectiveness of a M10,000 IsaMill™ in the primary grinding circuit.

Testwork has been undertaken using a M4 and M20 IsaMills™ treating SAG mill cyclone underflow (Anderson 2006, Burford 2006), with further testwork using a M20 IsaMill™ in late 2006 designed to overcome the presence of scats which caused problems in the earlier testwork. Figure 12 displays the flowsheet that was used for this testwork that incorporated a magnetic separator to remove steel scats in the cyclone underflow. Feed to the mill was also screened at 1mm.

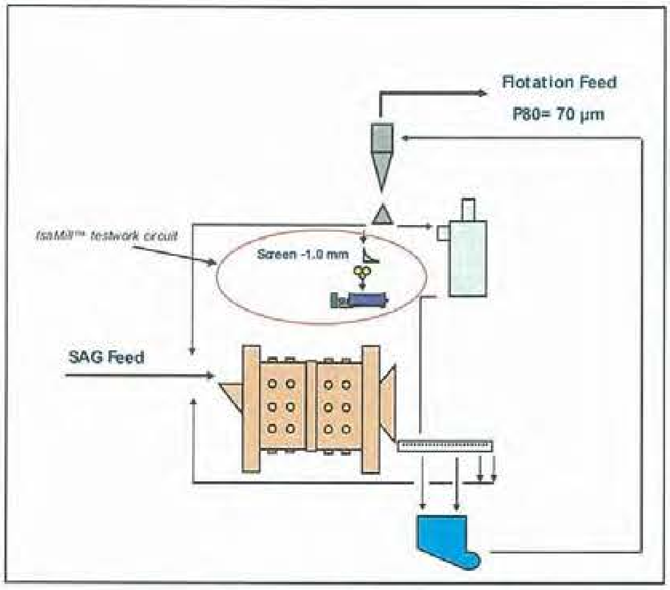


Figure 16: Site Testwork at MRM Using M20 IsaMill™ with a Magnetic Separator on Feed

The M20 IsaMill™ was able to treat material from a feed sizing of 300um, down to a product sizing of 20 to 25 um, (finer than the 40 µm target), in a single pass The data was able to permit a size energy relationship to be established, as shown in Figure 13, compared with the current Tower Mill operation in that circuit.

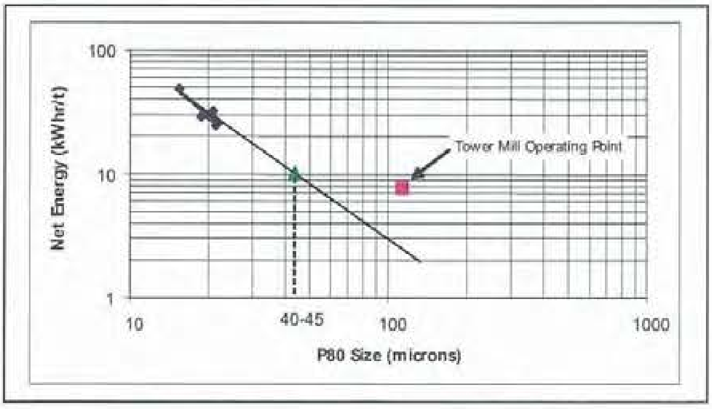


Figure 17: Size versus Net Energy Comparison for IsaMill™ and Tower Mill

Using the energy data from the M20 IsaMill™ testwork, and the current energy use for the tower mill in the primary circuit, it has been conservatively estimated that the IsaMill™ could produce a P80 size of 45 µm to 50 µm, while the tower mill could produce a P80 of 100 µm using a similar amount of energy.

However the flowsheet was not the most efficient use of both of the technologies, especially the IsaMill™, as IsaMill™ operation was hampered by the need to operate the mill to control the coarse particles, rather than achieve target grind size. Simulations followed the testwork with different circuit configurations, which lead to a much better circuit design based on the main advantage of the IsaMill™, open circuit operation.

MRM is planning to use IsaMills™ to treat the flotation feed instead of cyclone underflow, with the eventual circuit designed to eliminate cyclones and the tower mill. The eventual circuit configuration is displayed in Figure 18.

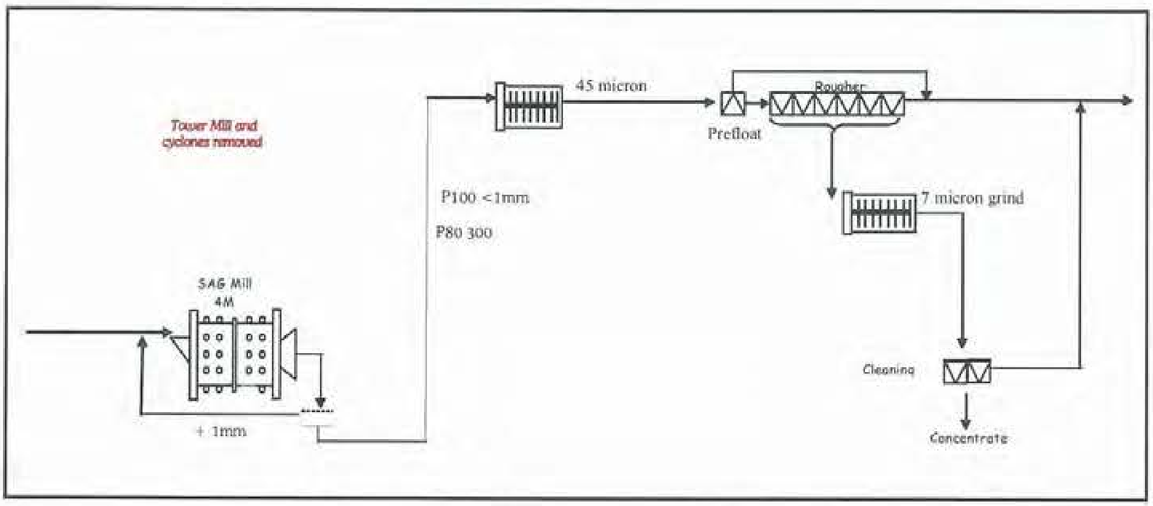


Figure 18 – IsaMill™ in MRM Grinding Circuit

This benefit of this circuit is that it allows the IsaMills™ to concentrate on the particles that it is ideally suited to, with an estimated F80 of 300μm, to produce the P80 of 45μm.

To date, testwork is being undertaken with one M3000, 1.1MW IsaMill™ continuously operating in this circuit with different grinding media, wear materials and feed sizing being trialed by the mill. Later in 2008, two M10,000 IsaMills™ powered with 3MW motors will be installed in the primary grinding circuit, with the rated capacity of the primary grinding circuit to be increased from 1.9 Mtpa to 2.5 Mtpa.

It is expected that the introduction of open circuit IsaMilling™ on the flotation feed using inert media will improve metallurgy as has been observed at other lead/zinc circuits (Pease *et al* 2005, Young *et al* 1997), such as improving the selectivity of fines, improving flotation rates, and reducing circulating loads and flotation reagents.

CONCLUSION

IsaMill™ technology is becoming a realistic alternative in coarse grinding applications since the development of ceramic media as well as the large scale M10,000 IsaMill™. These developments have enabled coarse grinding to be undertaken at a number of mineral processing sites, where energy efficient grinding is required.

The application of IsaMill™ technology to new applications, such as grinding in magnetite concentrators, offers magnetite operators an exciting alternative to conventional ball mill and tower mill technologies. Magnetite ore was found to be amenable to grinding with IsaMills™, in much the same manner as base metal ore is, with several of the magnetite samples being softer than base metal oretypes that are treated using IsaMill™ technology.

With the pressure being applied to all industries today for improved sustainability and the potential cost implications of carbon emission in the future, the need for increased energy efficiency in grinding is as important as ever.

ACKNOWLEDGEMENT

The authors wish to thank the management of Xstrata Technology for their support in the preparation and publication of this paper.

REFERENCES

1. Curry, D.C., Clark, L.W., Rule, C., 2005 - Collaborative Technology Development – Design and Operation of the World's Largest Stirred Mill. Randol Innovative Metallurgy Conference, Perth, Australia, August.

2. Curry, D.C. and Clermont, B., 2005 - Improving the Efficiency of Fine Grinding – Developments in Ceramic Media Technology. Randol Innovative Metallurgy Conference, Perth, Australia, August.

3. Enderle, U; Woodall, P; Duffy, M and Johnson, N W, 1997 - Stirred Mill Technology for Regrinding McArthur River and Mount Isa Lead/Zinc Ores. XX International Mineral Processing Conference, Aachen, Germany, September 21-26, Vol 2, pp 71-78.

4. Pease, J D; Curry, D C; Barns K E; Young M F and Rule, C, 2006 - Transforming Flowsheet Design with Inert Grinding – The IsaMill, 38th Annual Canadian Mineral Processors Conference, Ottawa Canada.

5. Pease, J D; Curry, D C and Young, M F, 2005 - Designing Flotation Circuits for High Fines Recovery, AusIMM Centenary of Flotation Conference, Brisbane, Australia, 6-9 June.

6. Young M F; Pease, J D; Johnson, N W and Munro, P D, 1997 - Developments in Milling Practice at the Lead/Zinc Concentrator of Mount Isa Mines Limited from 1990, Sixth Mill Operators Conference, Madang PNG, 6-8 October.

7. Burford B.D; Clark L.W., 2007 - IsaMill™ Technology Used in Efficient Grinding Circuits, V111 International Conference on Non-Ferrous Ore Processing, KGHM Cuprum, Wroclaw, Poland,May 21-23

8. Pease, JD – Crushing and Grinding, IIR, 2007, Brisbane, Australia, August 21-23

9. Gardner-Bond S; Spring.S. March, 2008 - Iron Ore Sector Review, Ocean Equities

10. Anderson, G S and Burford, B D. IsaMill™, 2006 -The Crossover from Ultrafine to Coarse Grinding, Metallurgical Plant Design and Operating Strategies, Metplant 2006, Perth, Western Australia, September 18 to 19

11. Johnson N.W; Gao, M; Young M.F and Cronin B, 1998 – Application of the IsaMill™ (A Horizontal Stirred Mill) to the Lead-Zinc Concentrator (Mount Isa Mines Ltd) and the Mining Cycle, Ausimm Annual Conference, Mt Isa, April 19-23

11. Pan Australian webpage; www.panaustralian.com.au, 2008 – Operation, Projects and Exploration, Phu Kham Copper-Gold Mine, March

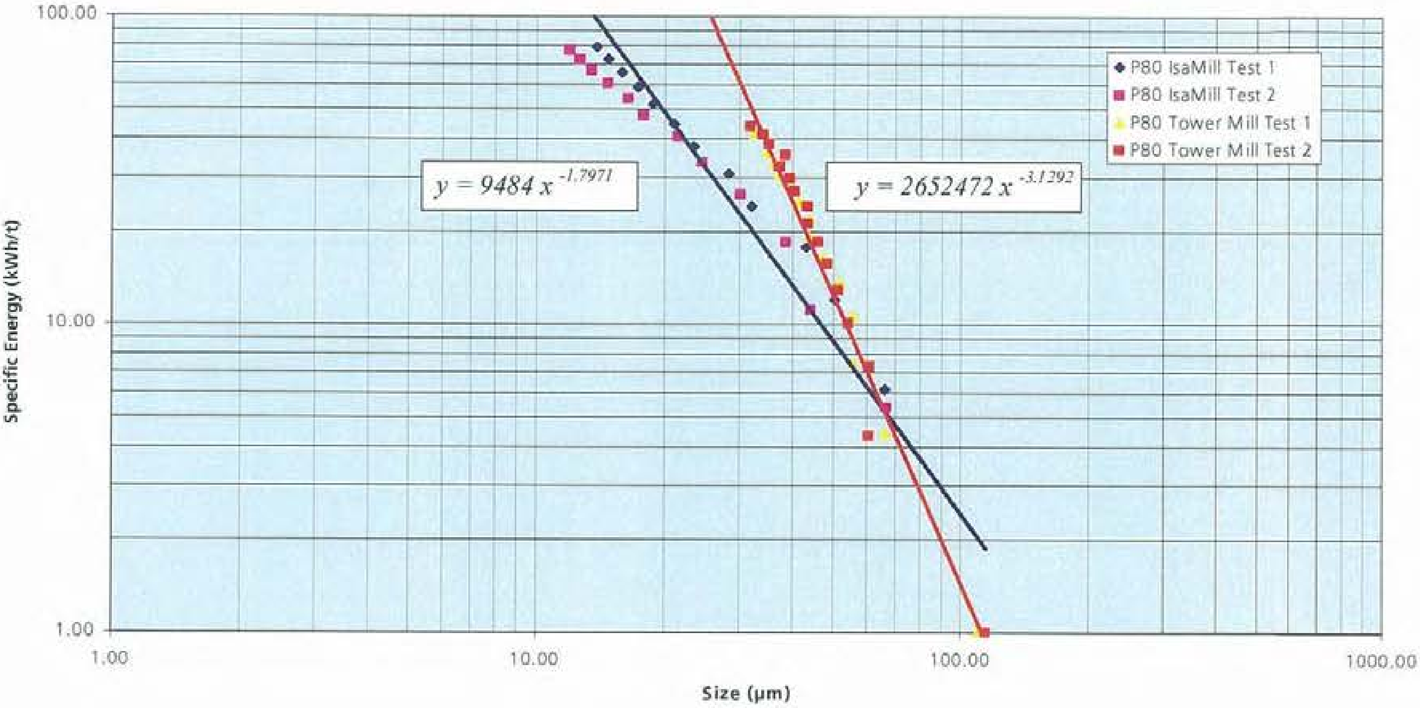
12. Oxiana webpage; www.oxiana.com.au, 2008 - Our Operations, Prominent Hill, March

13. Australasian Resources webpage; wwwww.austresources.com.au, 2008 – Investor Relations – Global Iron Ore and Steel Presentation (Amended Presentation) , March

AMMENDMENT

Figure 6 - Size versus Specific Energy – M4 IsaMill™ and Tower Mill

Equations updated with more data post publishing of original paper.



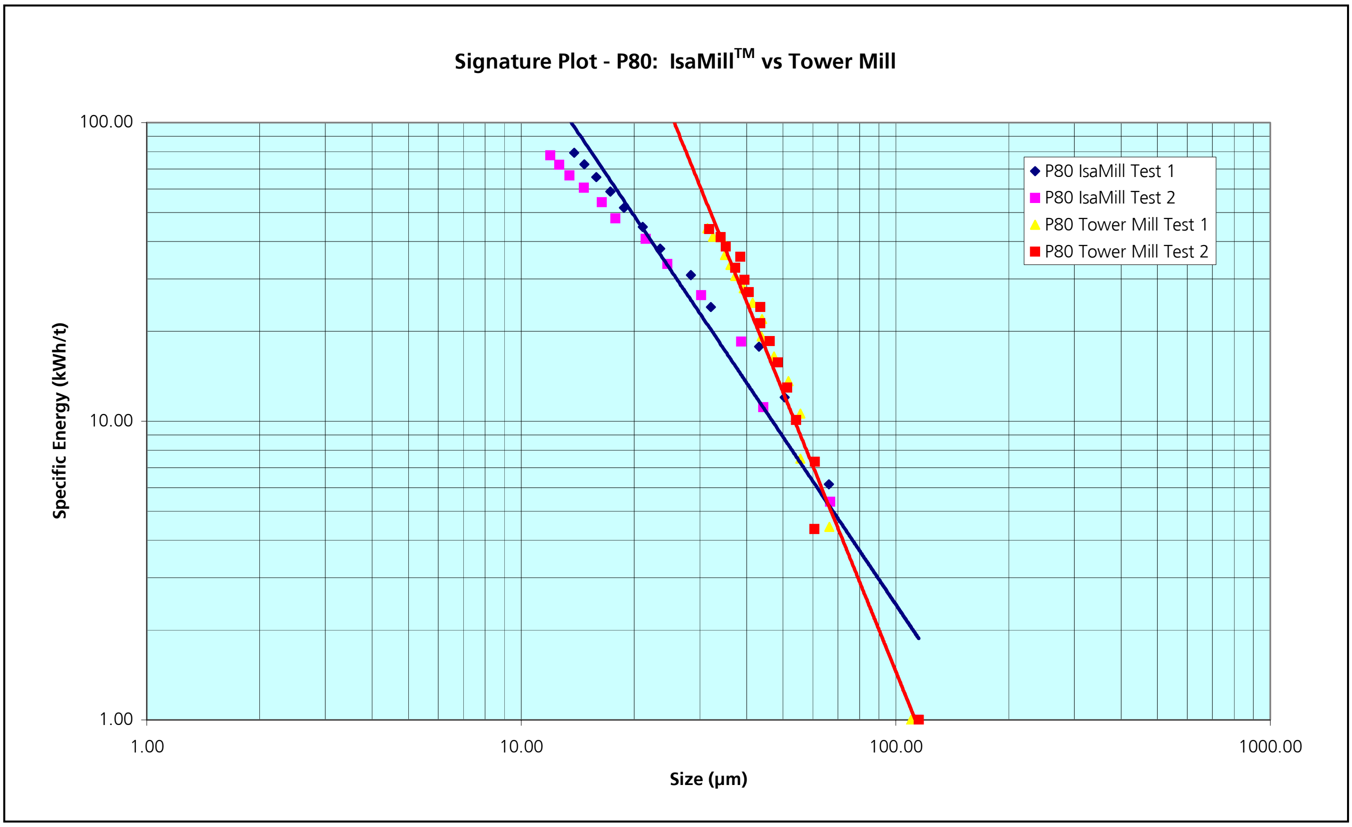
GRINDING COMPARISON TEST OF IsaMills™ WITH TOWER MILLS USING MAGNETITE

INTRODUCTION

IsaMills™ have been in fine grinding applications in base metal circuits since 1994. The IsaMill™ has been enabling technology, that has allowed these applications to be developed, due to its high energy efficiency and intense grinding action. Over the years the IsaMill™ has evolved from its initial design as an “ultra fine grinding mill” to its use in more traditional regrind duties following the increase in capacity of the units and the development of a ceramic media by Magotteaux – Keramax® MT1™. These developments have allowed the mill to treat coarser feed sizes at high energy efficiency compared to traditional grinding technologies. There is a range of models now available from the 500kW M1000, to the largest stirred mill currently available, a 3MW M10,000.

Today the mills undertake coarse grinding duties in PGM, copper/pyrite, and zinc/lead applications. While most of the work to date has been with base metals, there is no reason why the benefits of base metal operation cannot be transferred to other minerals.

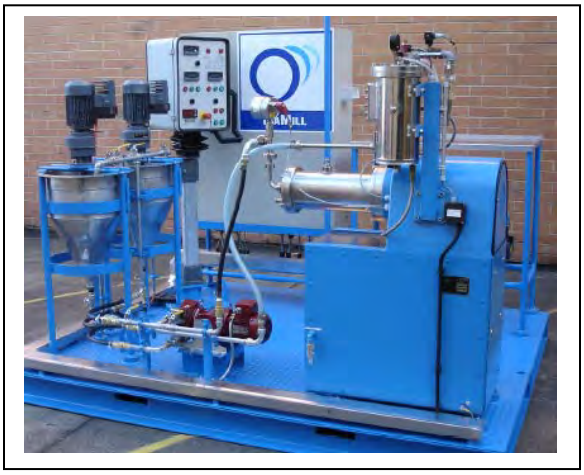
RESULTS OF TESTWORK ON MAGNETITE RECENTLY CONDUCTED (SEE OVERLEAF)



TESTWORK

The current testwork compares the operation of a lab scale M4 IsaMill™, with a lab scale Tower Mill using Ernest Henry Mine tailings. The tailings are the result of the flotation process, where chalcopyrite has been recovered from the breccia host rock, which is comprised of strongly altered and replaced felsic volcanic fragments in a matrix assemblage of predominantly magnetite, chalcopyrite and carbonate minerals

The M4 IsaMill™ is a 4 litre mill, containing 6 to 9 discs, operating with 3.5mm ceramic Keramax® - MT1™ media. The Tower Mill is 40L capacity and operates using 12mm steel media.



Top: M4 IsaMill™ 4L unit, similar to the unit at CSIRO where testwork was conducted.

Right: Tower Mill – 40L capacity, operated at CSIRO for comparison testwork.

In each test, 20 to 21kg of sample is made into a slurry of 50% solids and pumped through each mill for a number of runs. The power used for each run is recorded, and a small sample of the discharge is taken to permit laser sizing. This procedure is carried out through each mill for a minimum of 12 times, or until there is no significant reduction in the sizing, i.e. the mill cannot reduce the sample any further. A log-log, P80 size versus the specific energy to obtain that size graph, is then drawn (signature plot).

The test was repeated for each mill twice.

DISCUSSION

The results of the two test for each grinding technology shows a good level of reproducibility.

Key findings were:

- The IsaMill™ was able to produce material down to 13µm from a feed size of 113µm. However the Tower Mill failed to produce material less than 31µm. The reason for the difference between the mills was the grinding media used in the IsaMill™ testwork was a lot smaller. However media selection was based upon what realistically a full scale plant can operate with. A full scale IsaMill™ can be supplied and operated with ceramic media from 1 to 3.5mm, however Tower Mills can only realistically operate with media 12mm and upwards, which means that they cannot achieve the grind sizes that a full size IsaMill™ can achieve.

- The IsaMill™ used less power to produce a product 65µm or smaller compared to the Tower Mill. This means for this magnetite feed, an operator targeting a product of 40µm with an IsaMill™ will need approximately 14 to 18Kwh/T, while a Tower Mill will need a lot more energy between 26 to 27Kwh/T. The implications for this is smaller installed power for an IsaMill™ circuit and the resultant lower installation cost due to it operating in open circuit without the need for cyclones, as well as having a smaller footprint. It may also result in less grinding units compared with a Tower Mill circuit when large power requirements are needed.

Testwork is ongoing to determine the energy efficiency of the IsaMill™ on other magnetite deposits.

For further information please contact:
Lindsay Clark, Mineral Processing Business Manager – Xstrata Technology
Telephone +61 7 3833 8500; Email LClark@xstratatech.com.au; Web www.isamill.com

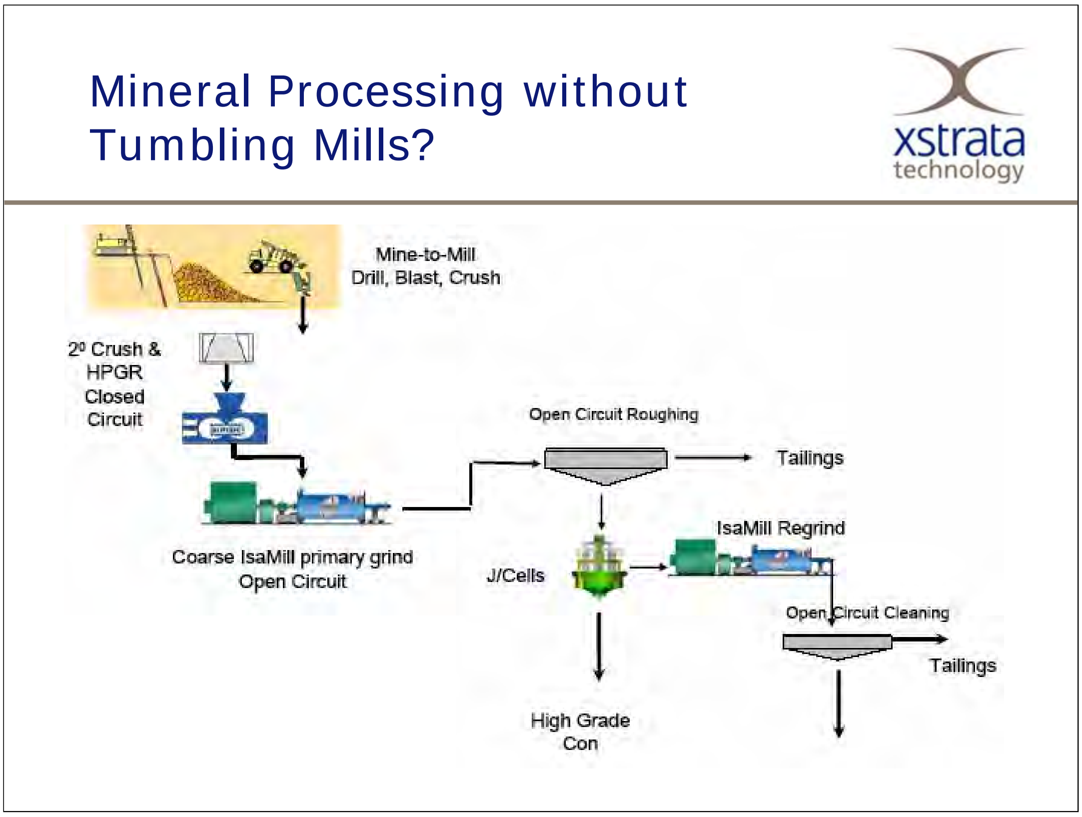
Case Study

Coarse IsaMilling at McArthur River

Joe Pease

General Manager Xstrata Technology

September 2007



Mineral Processing without Tumbling Mills?

xstrata technology

Mine-to-Mill
Drill, Blast, Crush

2^0 Crush &
HPGR
Closed
Circuit

Coarse IsaMill primary grind
Open Circuit

Open Circuit Roughing

Tailings

J/Cells

IsaMill Regrind

Open Circuit Cleaning

Tailings

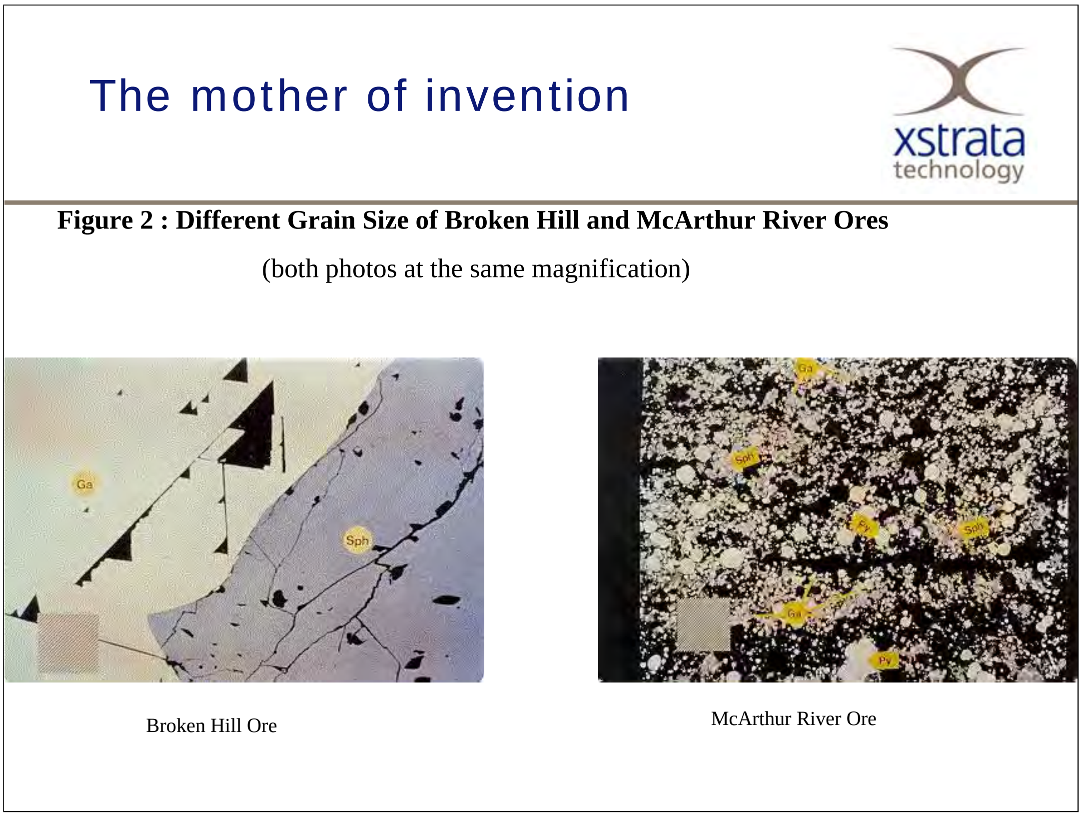
High Grade
Con

At this conference in 2006, I concluded with this slide, a picture of a possible energy efficient circuit of the future. A circuit without any tumbling mills at all. I said that perhaps in a decade we would see the first plant built without any tumbling mills. Or with only a small conventional grinding stage between HPGR and IsaMills.

Well, we aren't there yet, but we are a lot closer than we thought we were even a year ago.

This is a story about how you can find answers in unexpected places.

And how solving one problem may end up solving other, even bigger, problems.



The mother of invention

Figure 2 : Different Grain Size of Broken Hill and McArthur River Ores

(both photos at the same magnification)

Broken Hill Ore

McArthur River Ore

The story starts at McArthur River. This huge lead zinc deposit was discovered in 1955, but remained undeveloped for 40 years. There was simply no technology that could economically treat the ultra fine grained minerals.

As always, necessity was the mother of invention. Keeping the orebody, and survival in our other fine grained orebodies, simply required that we find a more efficient way to grind fine, or else go out of business. We had to make a step change in fine grinding. Conventional grinding, in ball or tower mills, was uneconomic for 3 reasons :

- it used too much energy
- the media cost was too high
- the large amount of steel media consumed harmed subsequent flotation.

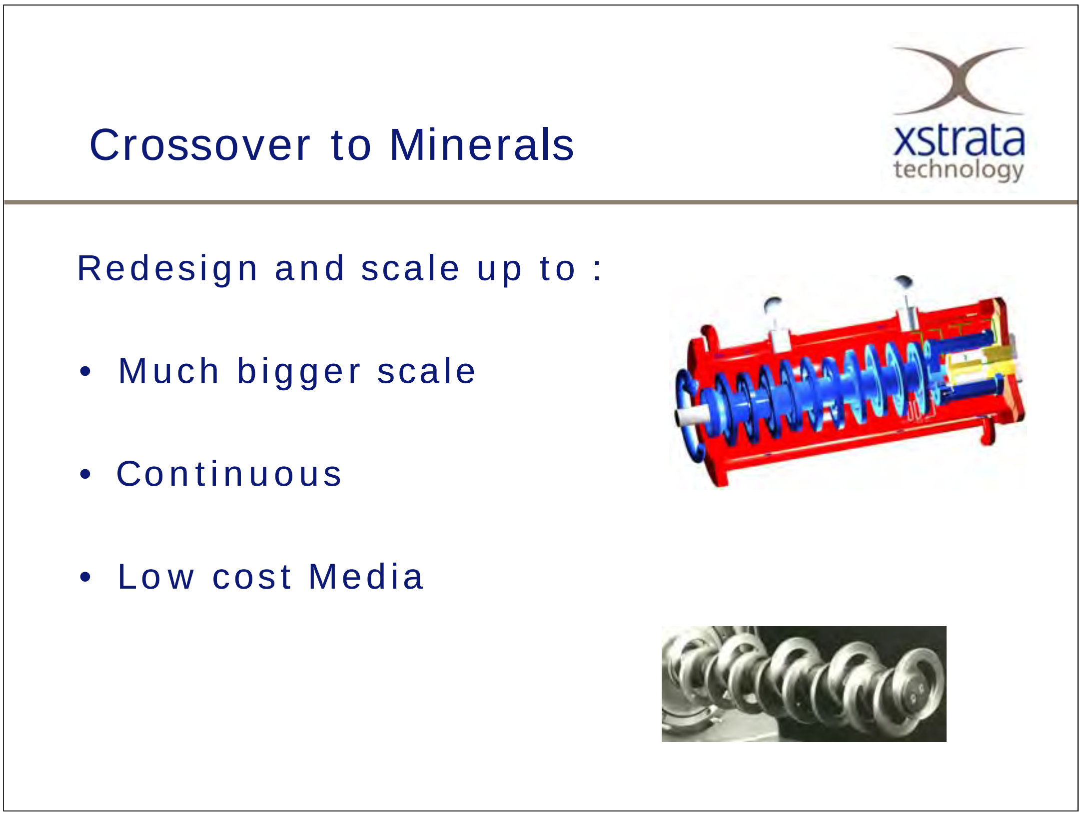
Fine Grinding in Manufacturing

Mt Isa’s head of research, Bill Johnson, knew that the answer to the problem didn’t lie in the minerals industry – we had looked there for decades. So he asked the question, who else has to grind fine ?

In fact, there was good established technology for fine grinding of high value manufactured products – like printer inks, pharmaceuticals, paint pigments, chocolate.

The pioneers in the field, and still the leaders, were Netzsch of Germany. We chose to work with them

As an aside, my wife says that if Bill had had only listened to her, he would have looked to chocolate to find answers long before he actually did.



Crossover to Minerals

Redesign and scale up to :

- Much bigger scale
- Continuous
- Low cost Media

So the concept was in manufacturing, but it had to be modified to minerals.

The manufacturing applications were for very high value products – like printer ink and chocolate. They are much higher value than zinc, which at the time was trading at about 40 c/lb. So these applications were small mills, often batch, and used very high cost media – ensuring no contamination of the product was much more important than the cost of grinding media.

To be economic to treat large tonnages of low value streams, we had to make the mills much bigger, operate continuously and robustly, and be able to use low cost media.

McArthur River

The end result was the IsaMill. This was the enabling technology for McArthur River, and then for the George Fisher and Black Star orebodies at Mt Isa.

This shows the 6 mills at McArthur River. They are 1 MW drive – 6 times bigger than the previous biggest Netzsch mills.

They operate continuously, and for the first 7 years their grinding media was ore gravel screened from the SAG mill discharge. That is, this was fully autogenous grinding to 7 microns !

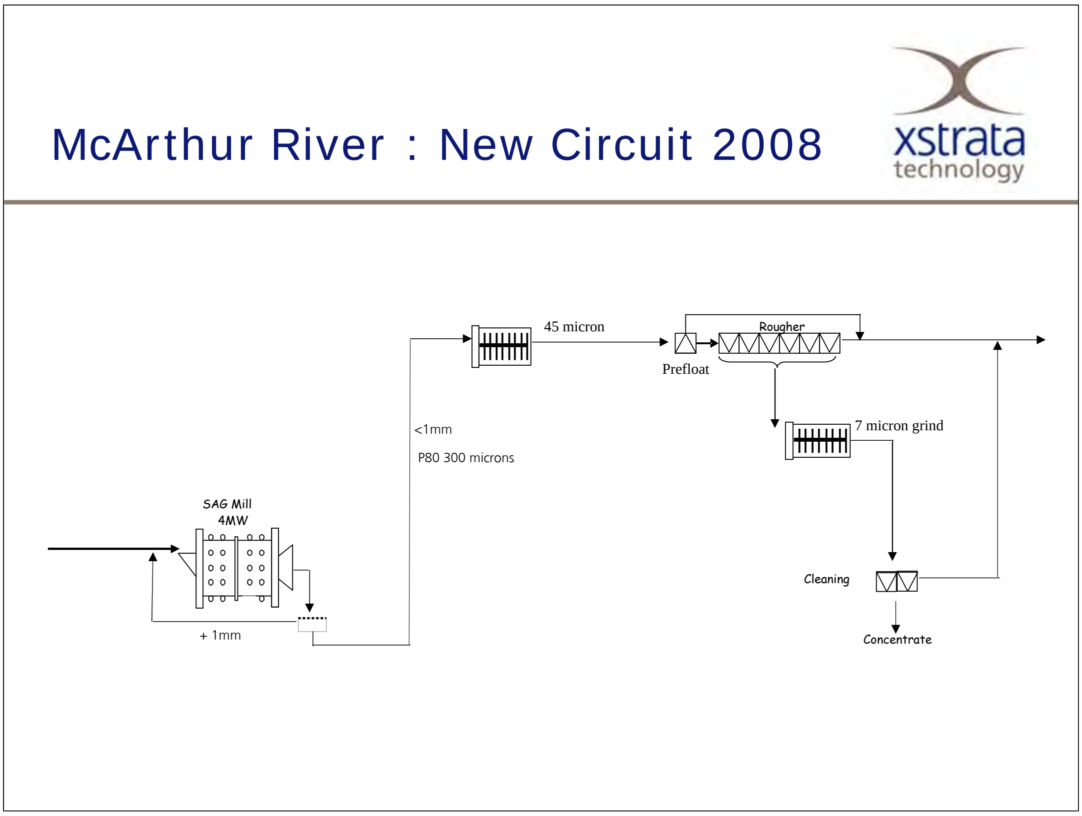
The previously untreatable orebody became economic, and achieves over 80% recovery into a 55% Zn+ Pb bulk concentrate.

3 MW, M10,000 IsaMill

So by looking in an unexpected area, a problem was solved, and a new technology was developed.

For many in the audience there is nothing new in this story, it has been told before. But now there is a a new, unexpected twist to the story, and again McArthur River is at the forefront.

They are currently installing IsaMills to grind SAG mill discharge.



McArthur River : New Circuit 2008

In early 2008, two 3 MW M10,000 IsaMills will be installed to treat SAG mill product before flotation.

So we are a lot closer to a circuit without tumbling mills than most of us expected.

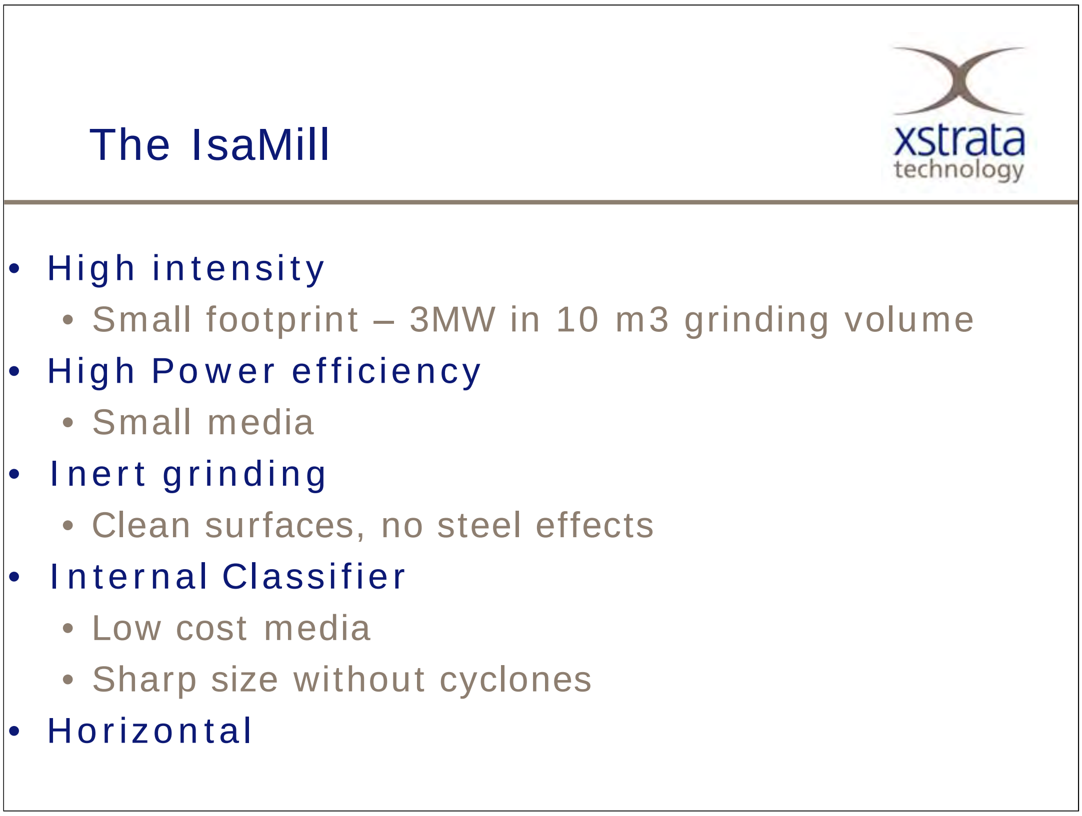
What is the IsaMill?

Firstly, what is the IsaMill?

The IsaMill was developed at Mt Isa in the early 1990’s as an economical grinding solution for fine grained ore bodies.

Pictured here is a 3.0 MW M10,000 (litre) IsaMill. This is the largest IsaMill currently available with up to a 3.0MW motor. Motor, gearbox, bearings, mill

The motor turns a horizontal shaft within the mill, the shell remains stationary.



The IsaMill

xstrata technology

- High intensity
    - Small footprint – 3MW in 10 m3 grinding volume
- High Power efficiency
    - Small media
- Inert grinding
    - Clean surfaces, no steel effects
- Internal Classifier
    - Low cost media
    - Sharp size without cyclones
- Horizontal

These are the characteristics of the IsaMill that make it quite different from conventional grinding.

The high intensity means a small mill footprint and installation size. The IsaMill has a power intensity of 300 kW per cubic metre, versus about 20 for a ball mill or Tower Mill - that is about 10 times higher. This means a significantly different installation, even for things like media – the entire first charge of media for a 3 MW mill is only 7 cubic metres.

The high power efficiency simply comes from the small media, as discussed in my previous presentation. In fact, the high power intensity and high power efficiency are linked in practice. A slow stirred ball or Tower Mill using 2 mm media would also be efficient for fine grinding. But the low power intensity would make the installation uneconomic – a huge installation, with prohibitively high consumption of prohibitively high cost media. The high power intensity in the IsaMill comes from the high stirring speed – about 20 m/second. This means that the fine media can do a lot of work in a small volume.

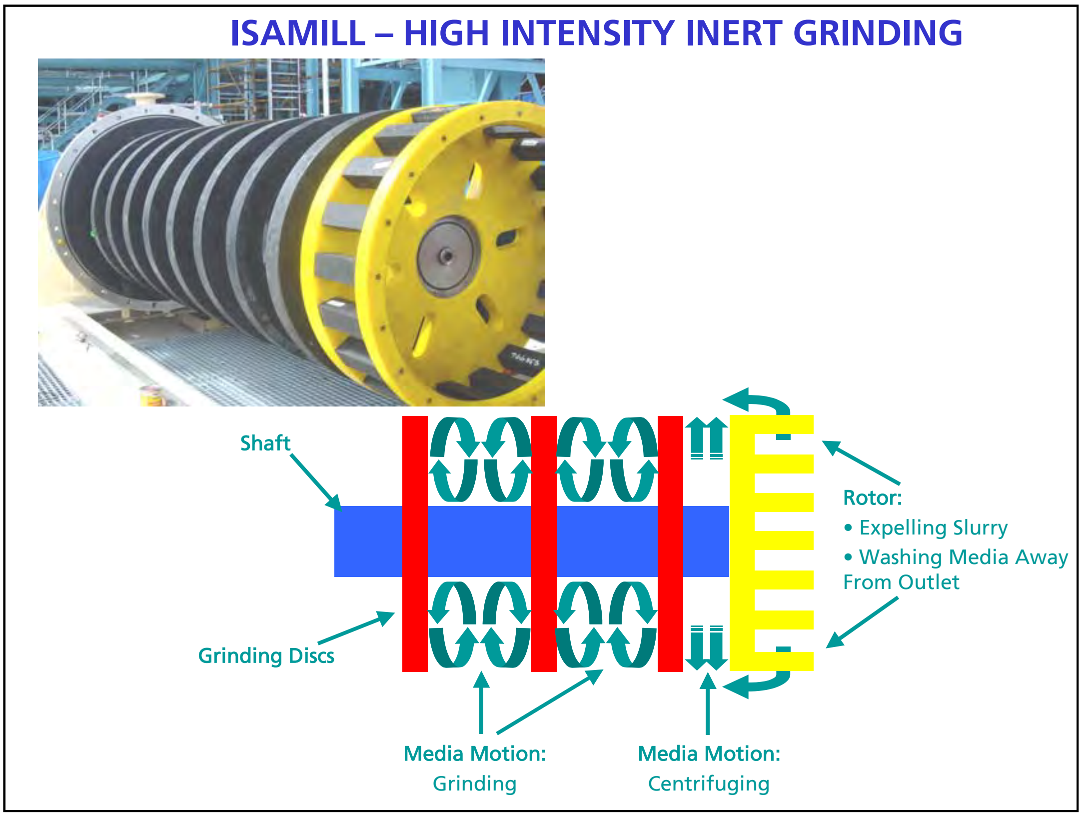
The internal classifier really is the great innovation of the IsaMill. How do you keep fine media, eg 1-2 mm, inside a mill, while allowing product to exit, and without using screens. We knew any solution that used screens just wasn't going to work at a large scale – the screens would block and peg, and would be an operating and maintenance nightmare. Further, having screens would dictate that you could only use high qualtiy, high cost media, because low quality media would break and block the screen. The answer was the product separator, basically a centrifuge to keep media in the mill without screens. It was the breakthrough that allowed us to use virtually free media – ore gravel at McArthur River, discard smelter slag at Mt Isa.

A happy consequence of this is that the mill doesn't need to be closed circuited with cyclones. This is hard to get used to – we have always known that grinding mills need cyclones and a circulating load. But the IsaMill doesn't.

The final point is obvious – this is a horizontal mill. But this simple point has big implications.

Our partners, Netzsch, make both vertical and horizontal stirred mills for manufacturing. They told us that both configurations have advantages. But there was one overwhelming factor in favour of the horizontal configuration – scale up to larger sizes. This is because of start-up torque. Netzsch advised that they simply wouldn't build a stirred mill over 400 kW, because the mechanical design would be dominated by the start-up torque on the bottom stirrer after a shutdown – the bottom stirrer has to be able to remobilise the settled load. For big mills this would dominate the design of stirrer, shaft, crane etc. In contrast, in a horizontal mill, all 8 stirrers are available. This helps explain why the IsaMill has scaled up 10 fold from 300 kw to 3 MW in a decade, already almost 3 times bigger than the biggest installation Tower Mills have achieved in over 50 years.

The picture shows the 8 grinding discs (black) and the product separator (yellow) at the mill discharge.



ISAMILL – HIGH INTENSITY INERT GRINDING

Shaft

Grinding Discs

Rotor:
- Expelling Slurry
- Washing Media Away From Outlet

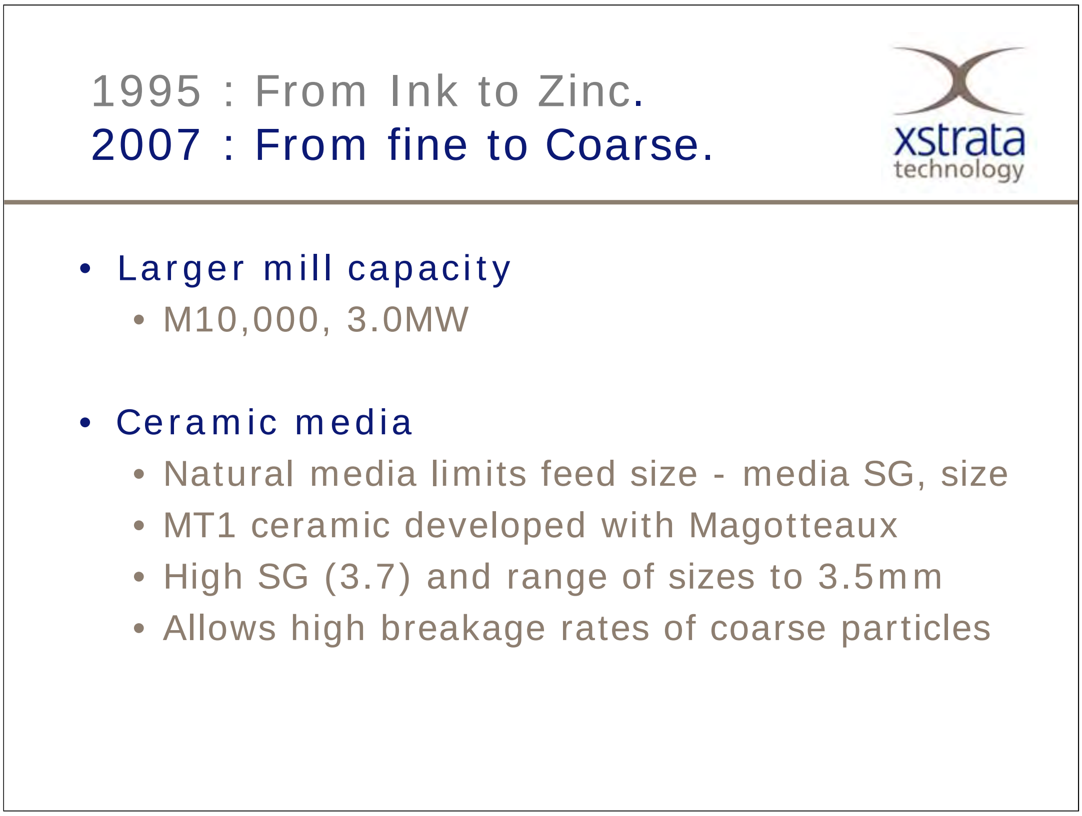
Media Motion: Grinding

Media Motion: Centrifuging

In operation the mill is 70-80% filled with media, which is stirred at high speed – up to the stirrer tip speed of about 20 m/s. New feed has to pass through 8 different grinding chambers before it gets to see the product separator, or centrifuge, at the end of the mill. At this point media is returned to the grinding discs, and fine solids and water is discharged. The mill operates full and pressurised, with average residence time in the mill is 30-60 seconds. Once again, this is fundamentally different from conventional grinding. It also explains why the mill doesn’t need cyclones to make a sharp product size – there is little time for overgrinding, but particles have to pass through 8 grinding chambers in series before leaving the mill.

The best way to describe this is to have a look at a small mill in operation.

This video shows why the IsaMill gets a sharp size distribution. The action of the product separator compresses the media between the 8 grinding discs. Feed has to pass through effectively 8 grinding chambers in series before it reaches the exit. When dye is introduced to the mill, its slow movement down the mill demonstrates the almost plug-flow path for new feed.



1995 : From Ink to Zinc.
2007 : From fine to Coarse.

- Larger mill capacity
  - M10,000, 3.0MW
- Ceramic media
  - Natural media limits feed size - media SG, size
  - MT1 ceramic developed with Magotteaux
  - High SG (3.7) and range of sizes to 3.5mm
  - Allows high breakage rates of coarse particles

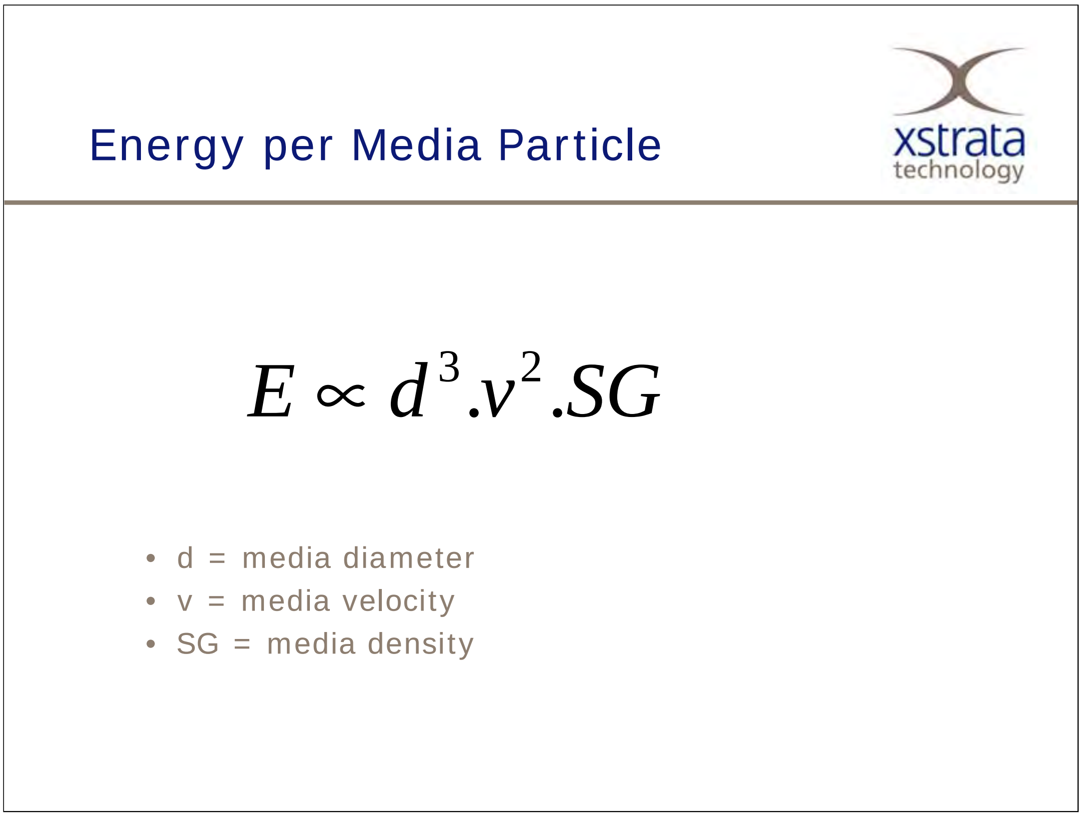
Developing the mill from grinding printer ink and chocolate to grinding zinc was the big step.

Compared with that, the move from fine grinding minerals to coarser grinds was relatively simple.

We were always fascinated by the potential to take the energy efficiency, small footprint, and inert grinding to mainstream applications. But there were two things we needed to do. We needed a bigger mill. We now have this with the 3 MW M10,000 mill.

Secondly, we needed better media. The ability to use low quality, free media like slag, ore gravel, or local sand, is remarkable, but it restricts the use of the mill to fine grinding. These grinding medias have relatively low density, and small natural grain size. This restricts the size of the ore particle they can break, even at the high stirring speed in an IsaMill. To be practical for mainstream coarse grinding, the media needed to be consistent quality, and high enough density and size to break the biggest particles in mill feed. And still be lower operating cost than using steel balls in conventional mills.

This was achieved by the development of MT1 Keramax ceramic by Maggoteaux.



Energy per Media Particle

xstrata technology

$$E \propto d^{3}.v^{2}.SG$$

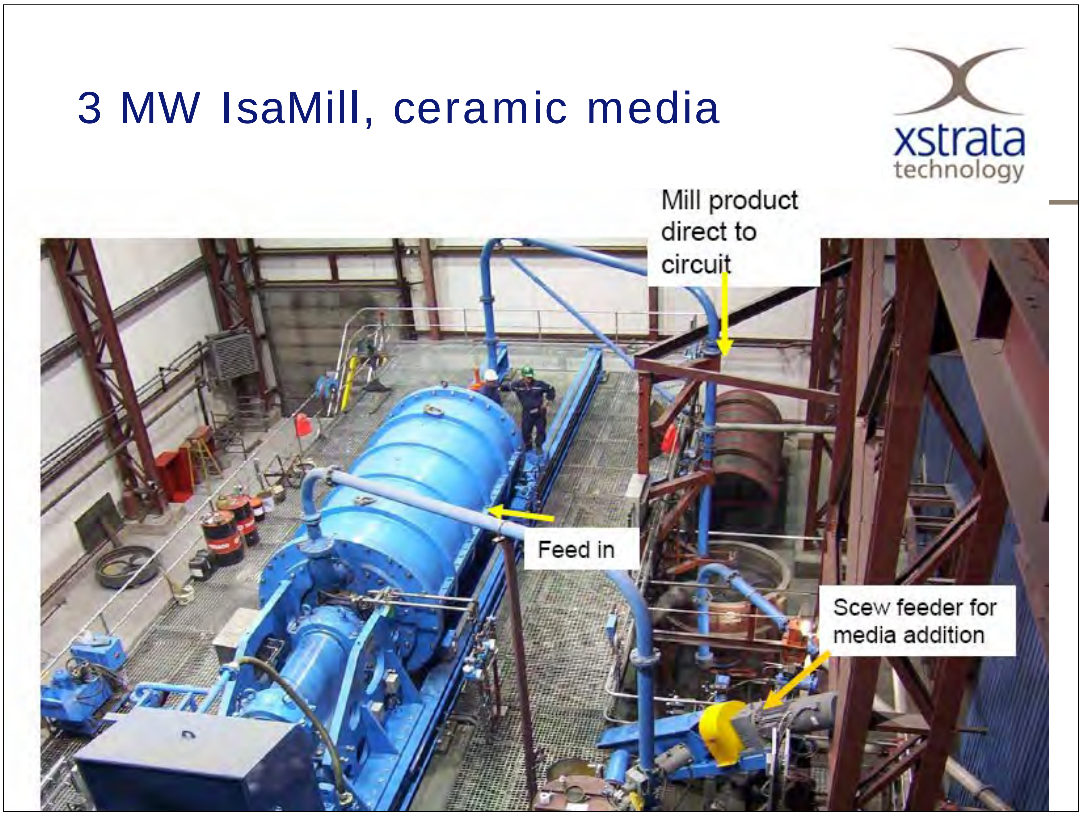
- d = media diameter
- v = media velocity
- SG = media density

I promise that this will be the only formula in this presentation. But it is important to the McArthur River coarse grinding story.

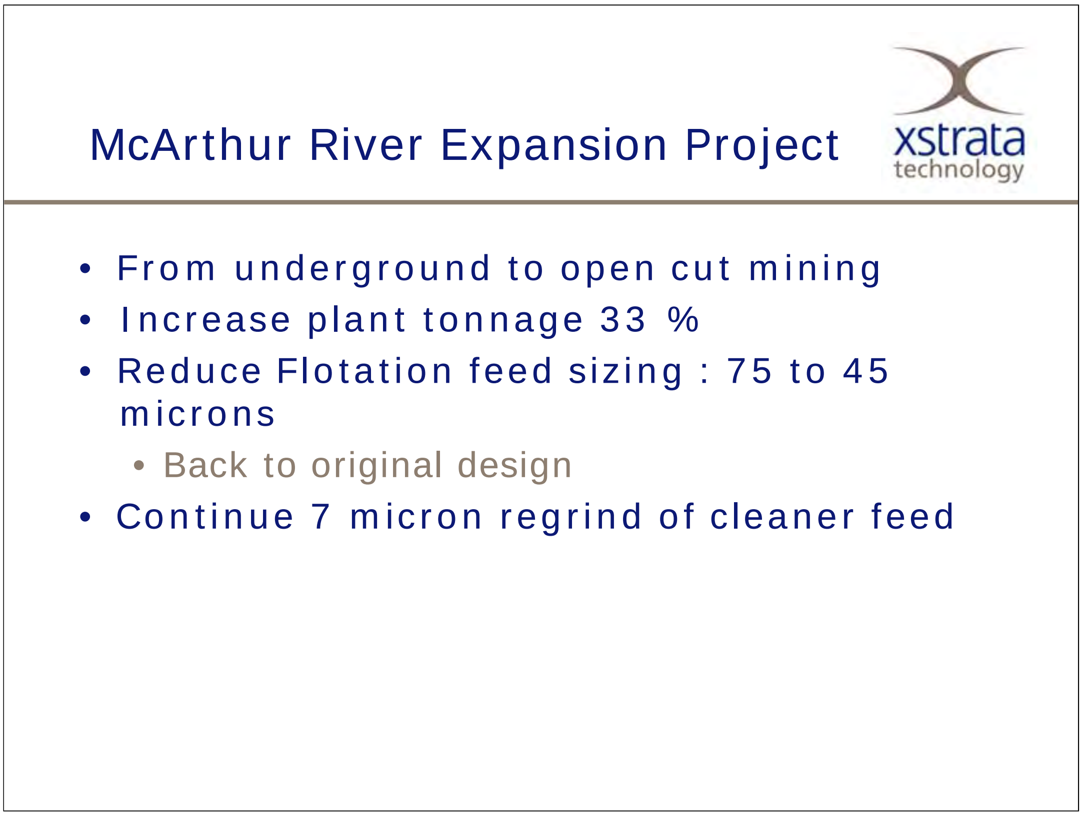
Media particles need enough energy to break the largest particles in the mill feed, as quickly as they are entering the mill. If they don’t, the coarse particles will build up in the mill and reduce the grinding efficiency – we are all familiar with critical size fractions from Autogenous and SAG milling.

In a ball mill, some steel balls are picked up and dropped the diameter of the mill – plenty of energy. In fact, probably too much energy, so some is wasted, but the coarsest particles will be broken.

Similarly in the IsaMill, we need enough energy in enough collisions to break the feed. We can either increase the SG of the media, increase its diameter, or increase its speed. The much high speeds in the IsaMill, , 20 m/sec, explains why small media can grind coarse particles. We don’t want to increase speed any more, so we increased the media diameter and density. The MT1 ceramic has density of 3.7, compared with 2.4 for sand. Maggoteaux currently produce it up to 3.5 mm diameter, enough for grinding duties up to 300 micron P80 feed.



3 MW IsaMill, ceramic media



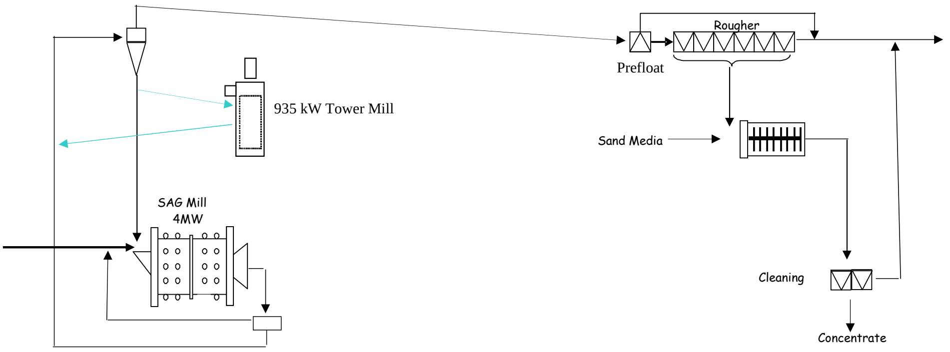
McArthur River Expansion Project

- From underground to open cut mining
- Increase plant tonnage 33 %
- Reduce Flotation feed sizing : 75 to 45 microns
  - Back to original design
- Continue 7 micron regrind of cleaner feed

McArthur River is now going through its next phase of development. It is moving from underground mining to Open Cut mining. Feed tonnage will be increased by 33%, from 230 t/h to 305 t/h (2.4 Mt/y). The project will also reduce flotation feed sizing from 75 microns back to the original design of 45 microns (increases in feed tonnage over the years caused the coarsening flotation feed)..

Of course, to make concentrate grade and recovery, all minerals have to be ground to 7 microns before cleaning.

McArthur River Existing Circuit 2007

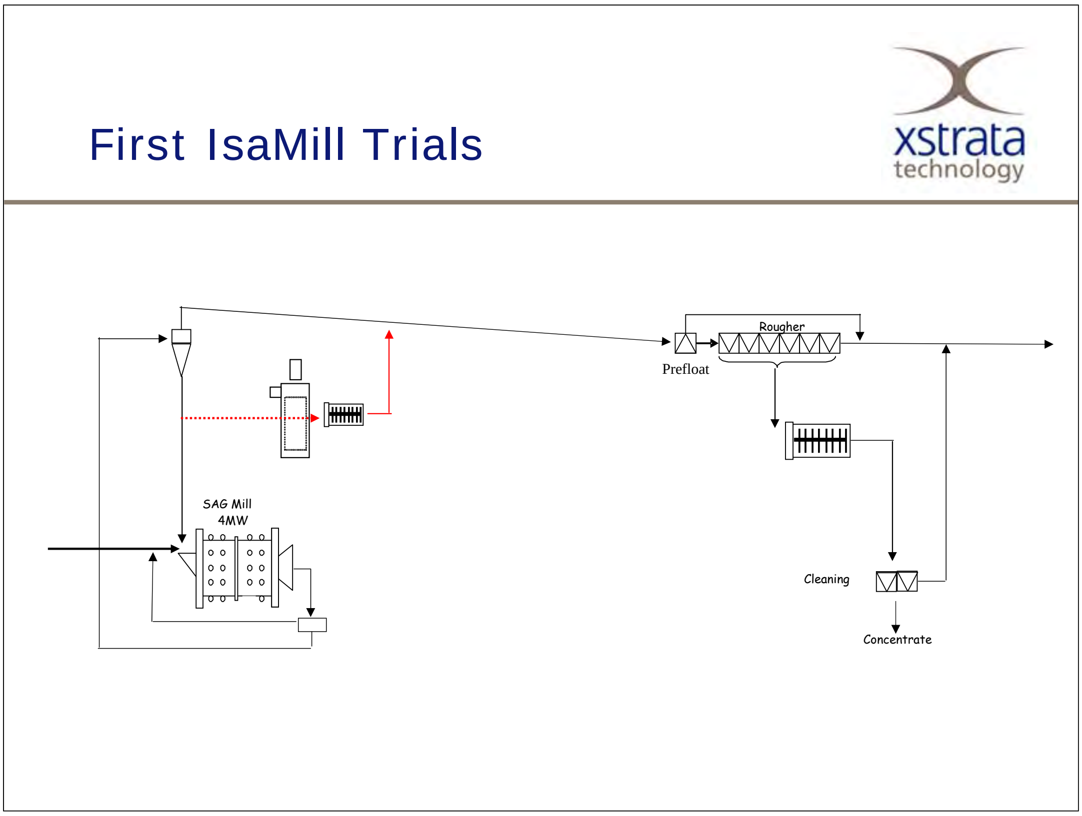


The original circuit was a SAG mill in closed circuit with cyclone and double deck screen to produce a P80 45 microns product for flotation. Media for the IsaMills was from the second screen deck (0.8-1.8mm). The IsaMills grind rougher con from 50 to 7 microns. Later, the IsaMill media was changed to sand. This decoupled SAG mill operation from IsaMill operation, allowing both to improve efficiency and throughput.

Feed tonnage has increased since start up. A 935 kW Tower Mill was installed to grind part of the cyclone underflow to partly compensate the tonnage increase. This is not the ideal duty for a Tower Mill, but it was chosen simply because a second hand unit was available – these were the days of 37 cent zinc ! Even so, flotation feed has coarsened to about 70 microns at the higher tonnage. Increasing the efficiency of the IsaMills helped compensate for the coarser feed, by maintaining a 7 micron grind of cleaner feed.

The ore from the open cut is lower grade and a bit more complex than the original orebodies treated. So McArthur needed to both increase tonnage, and return to design flotation feed size, while maintaining regrind size at 7 microns. IsaMills offered an efficient way to do this, while also bringing some of the advantages of inert grinding to rougher flotation.

McArthur River undertook a test program to evaluate the IsaMill for the coarse grinding duty.

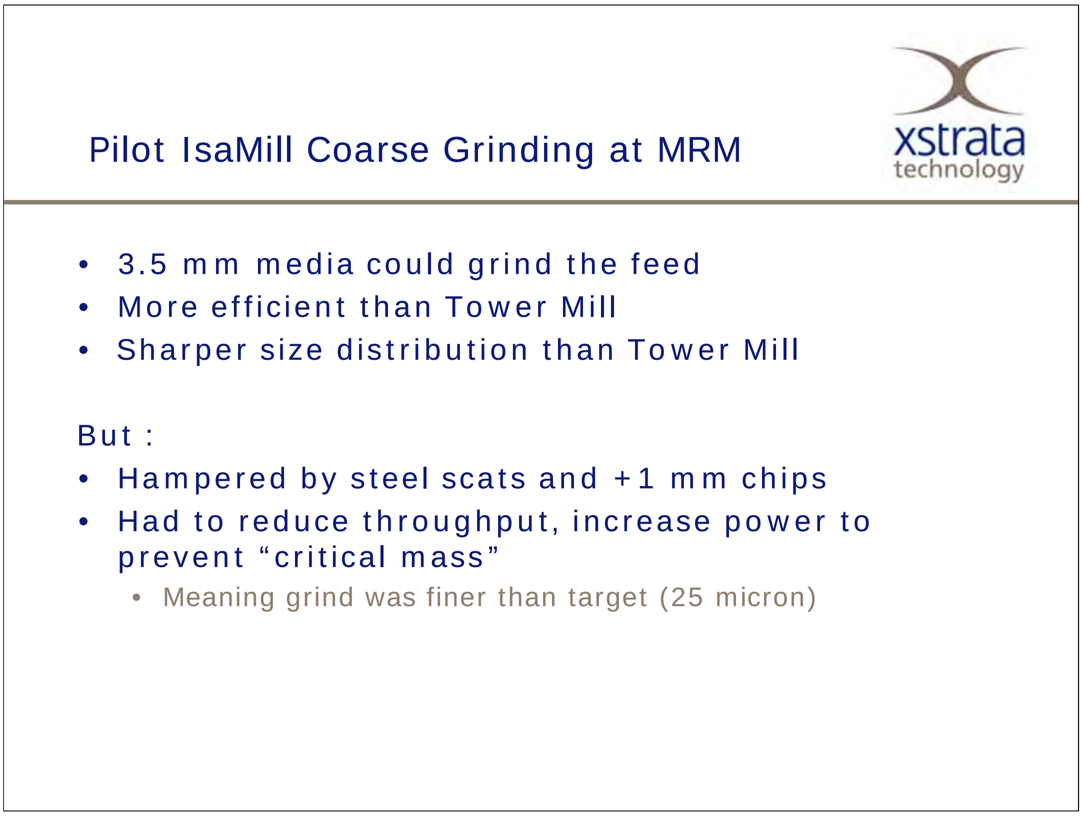


First IsaMill Trials

The first trial was to test the mill on a similar stream to the Tower Mill, ie treat a portion of SAG mill underflow.

Initial testwork was done initially in the 4 litre M4 laboratory mill. This indicated that 3.5 mm ceramic media would grind the stream at higher efficiency and with a sharper size distribution than the Tower Mill. As a result, the product could be sent straight to flotation rather than back to the SAG mill cylcone.

We then did in-plant pilot work using the M20 (20 litre) mill, treating SAG cyclone underflow. The target was to produce approximately 45 micron product from the IsaMill.



Pilot IsaMill Coarse Grinding at MRM

xstrata technology

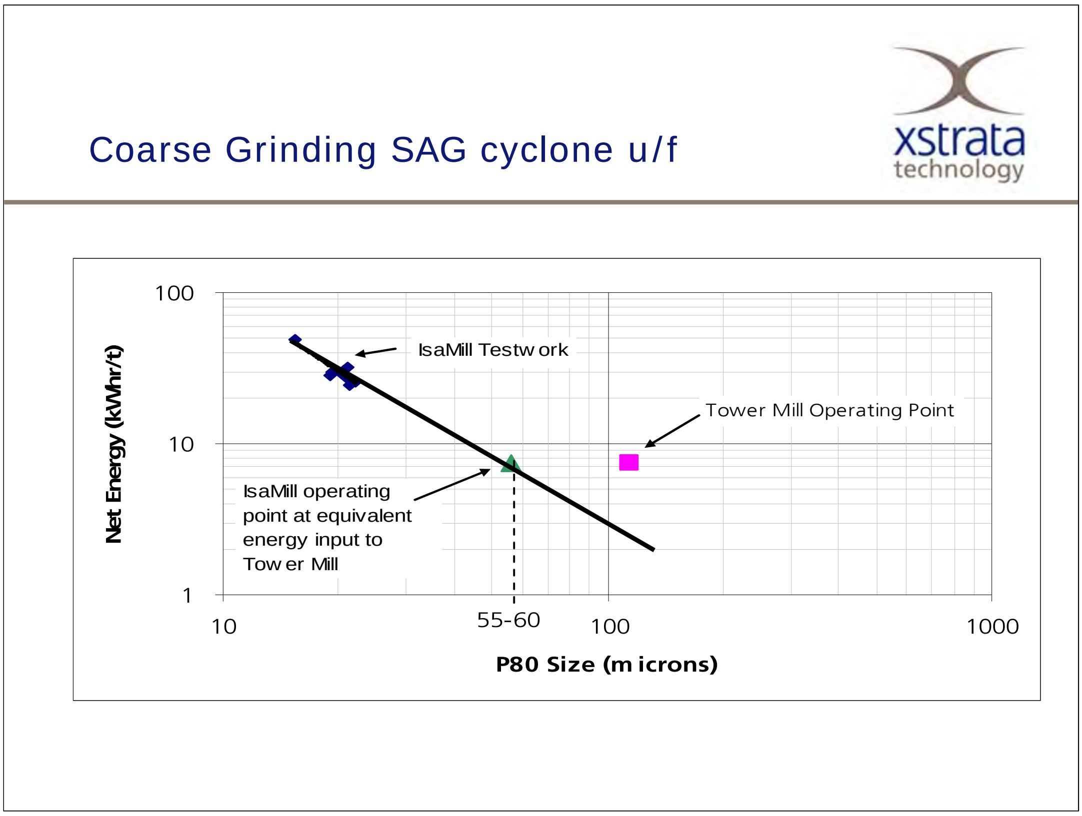
- 3.5 mm media could grind the feed
- More efficient than Tower Mill
- Sharper size distribution than Tower Mill

But :

- Hampered by steel scats and +1 mm chips
- Had to reduce throughput, increase power to prevent “critical mass”
  - Meaning grind was finer than target (25 micron)

The site pilot testwork confirmed the laboratory work that the IsaMill with 3.5 mm ceramic media could grind the cyclone underflow with the +1mm screened out. Efficiency and size distribution was better than the Tower Mill But the operation was hampered by build up of steel scats and coarse (up to 1mm) particles in this stream.

The steel scats could be removed with a magnetic separator. However the breakage rate of coarse particles was still too low. To prevent build up of coarse particles, the mill power had to be kept high, but throughput reduced. This meant an increase in kwh/t, and the feed was ground finer than necessary, to 25 microns rather than 45 micron target. Theoretically the grind was efficient, but we didn’t need to grind that fine.



Coarse Grinding SAG cyclone u/f

As expected, the IsaMill was more efficient than the Tower Mill – for the same energy input (7kwh/t) it would produce a 55-60 micron product versus a 100 micron product from the Tower Mill. However this comparison is not particularly valid, as neither mill is suited to this grinding application. The IsaMill had to be operated to control coarse particles rather than achieve target grind size. So it became clear that this was not the right location for the IsaMills.

So our first attempt told us that the IsaMill would certainly grind, but it needed to be in a better position in the circuit. We were trying to do too many stages of grinding in a single mill.

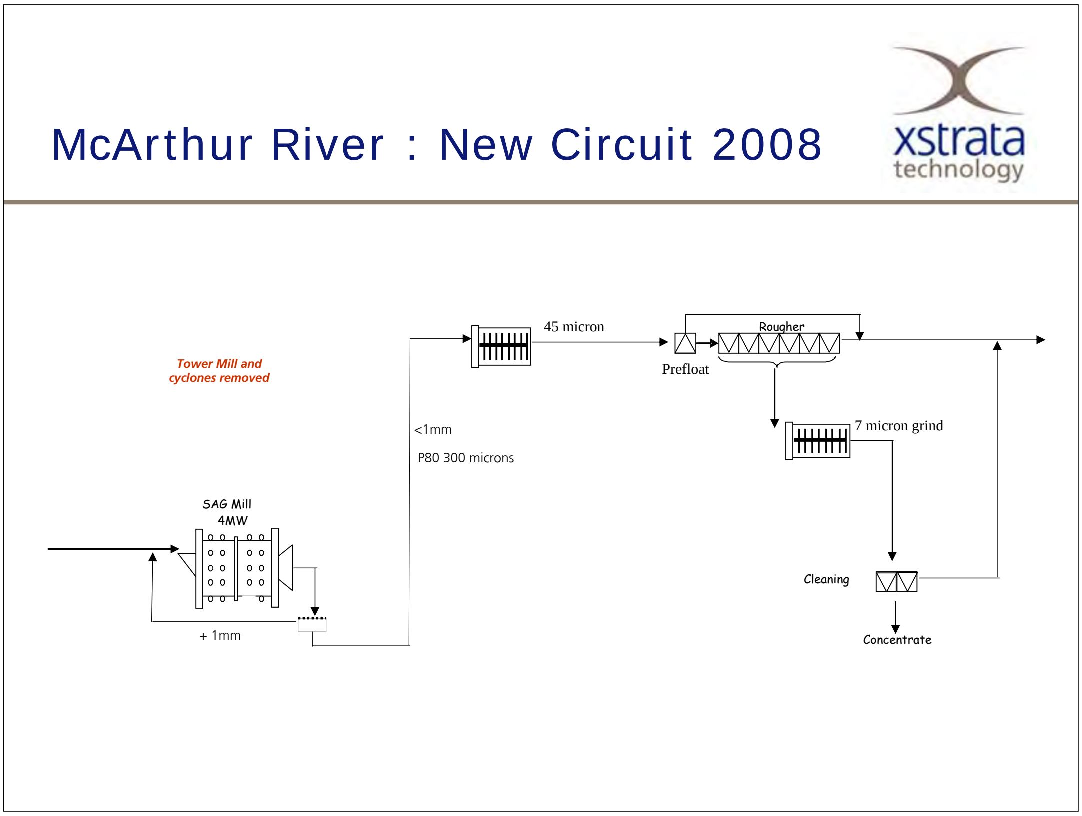
Dealing with +1mm fraction

- Increase breakage energy
  - $E \approx d^3 . v^2 . SG$
  - 7mm media has 8 times the breakage energy of 3.5mm
- Find a better location
  - Screened SAG discharge, not cyclone underflow
- 2006-7 : SAG simulation, pilot testing, full scale testing in existing 1 MW IsaMill
- Project approval early 2007

The breakage rate of coarse particles could be improved by returning to our formula – by increasing speed, media SG, or media size. In fact, the full scale mills operate with higher tip speed than the M20 unit, so should have performed better. But the biggest impact would be from bigger media – doubling the top size of media will mean an 8 times increase in maximum breakage energy available.

Currently the coarsest size of MT1 commercially available is 3.5 mm. This will increase in future. But the main point is that it became obvious we were trying to grind the wrong stream. A much better solution was to take the IsaMill out of the SAG mill circulating load. Let the SAG mill grind SAG mill feed, the IsaMill can grind the final product of that circuit.

During 2006 and 2007, McArthur River surveyed and simulated a different SAG mill configuration. This included closing the SAG mill with a 1 mm screen. Screen undersize is nominally -1mm, with a P80 of 300 micron feed. This stream would be ground in open circuit IsaMills, then sent to flotation. One of the existing 1MW has been used for full scale evaluation of IsaMilling and different medias, including ceramic and coarse sand. As a result, McArthur River are proceeding with a full scale project to install IsaMills to grind flotation feed.



McArthur River : New Circuit 2008

This is the new circuit being installed at MacArthur River. Two 3 MW M10,000 IsaMills will be installed in early 2008. Screen undersize will be sent direct to open circuit IsaMills, which will produce a P80 45 micron flotation feed. The Tower Mill and cyclones will be removed from the circuit.

MRM Expansion Project

xstrata technology

- Add 2 IsaMills to treat 305 t/h at 40% solids ***(open circuit)***
- Remove Tower Mill and cyclones
- Increase feed 1.9 Mt/y to 2.5 Mt/y (33%)
- Reduce flotation feed P80 75 to 45 microns
- Potential downstream benefits
  - Inert grinding before roughing
  - Reduced regrinding energy (finer float feed)

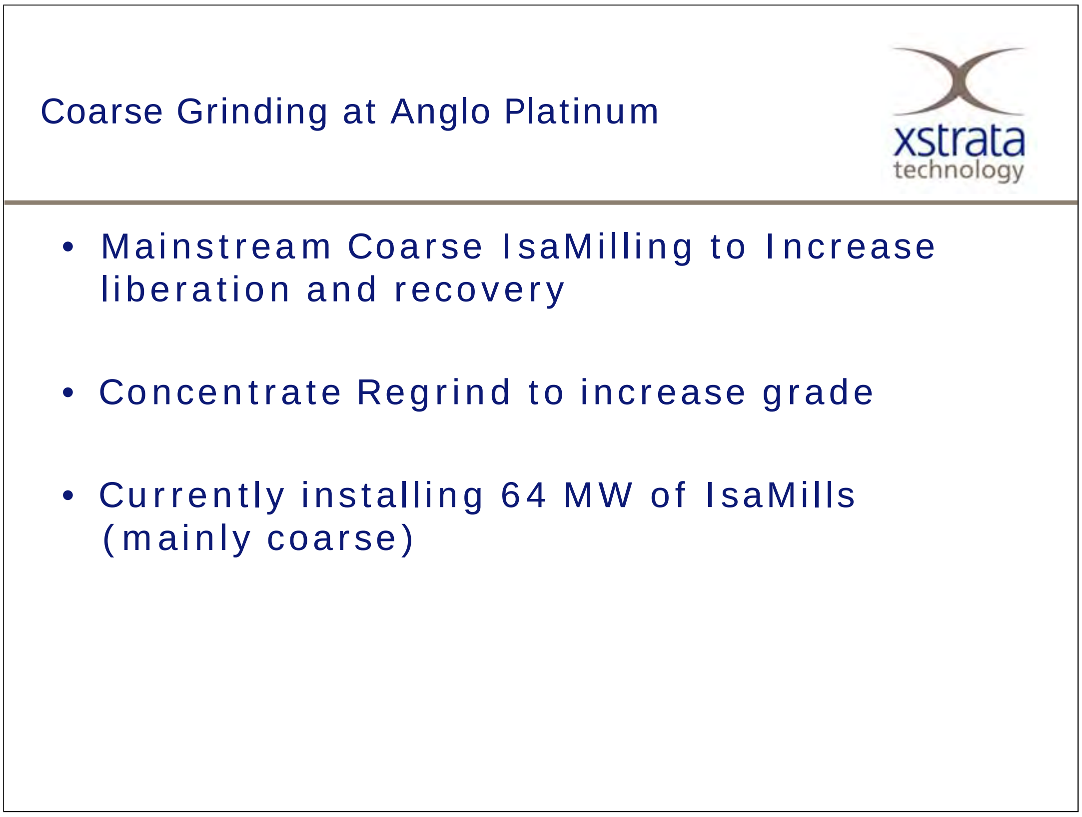
The mills will be fed at 40% solids.

The finer flotation feed sizing should improve roughing performance, and will slightly reduce regrinding energy need. Primary roughing may also benefit from the inert grinding – experience at other installations is improved flotation rates and selectivity for fines, and reduced reagent needs. These gains can only be evaluated at full scale, where the full impact on water chemistry and circulating loads is evident. So any gains from inert grinding are an “upside”, not part of the justification.

Coarse Grinding at Anglo Platinum

PPL 100 to 50 micron grind. Commissioned Dec 06

McArthur River have taken the coarse grinding applications of IsaMilling to the next level. Anglo Platinum have already committed to a major program of IsaMilling. This is the first installation at their PPL concentrator, which grinds scavenger feed in open circuit from around 100 microns to around 50 microns. This mill was commissioned in December 2006.

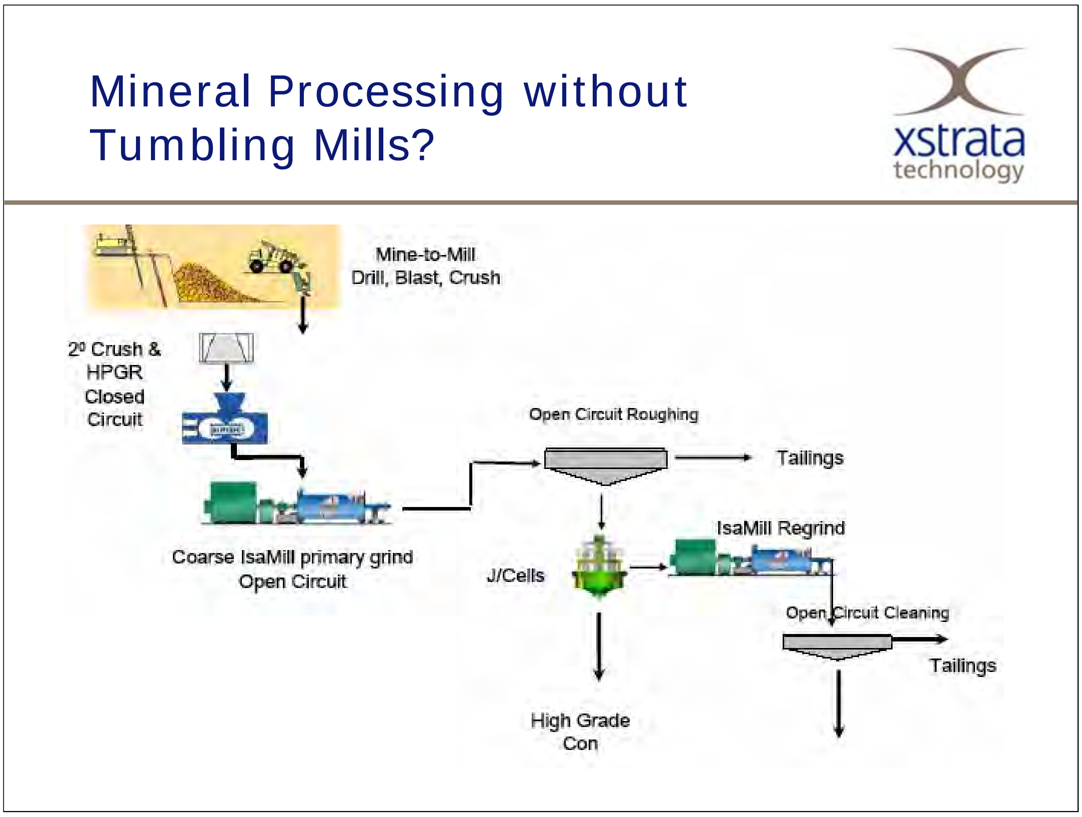


Coarse Grinding at Anglo Platinum

- Mainstream Coarse IsaMilling to Increase liberation and recovery
- Concentrate Regrind to increase grade
- Currently installing 64 MW of IsaMills (mainly coarse)

The success of the Anglo Platinum installations has led them to undertake a major installation program for IsaMills. Over the next 18 months they will be installing 64 MW of IsaMilling capacity. Much of it will be grinding mainstream feeds like the PPL installation – to increase liberation and recovery in roughing and/or scavenging. Other mills will be installed to increase concentrate grade, to increase smelter throughput, and reduce smelting energy and cost.

By the end of 2008, there will be 100 MW of operating IsaMill capacity. The technology has quickly moved from fine grinding to coarse – over 70% of the installations by the end of 2008 will be in conventional regrinding or mainstream grinding. (product sizes 25 to 60 microns).



Mineral Processing without Tumbling Mills?

So I return to my opening slide. Most people think of the IsaMill as a fine grinding mill. No more. A little over a decade ago it moved from grinding ink to grinding zinc. It has quickly made the adjustment to coarser grinding.

With the work at McArthur River, we are much closer to the hypothetical circuit without tumbling mills than any of us imagined even a year ago.

We aren't there yet, but we are a lot closer than we were even in 2006. The work at McArthur River for the first time will take a SAG mill discharge directly to IsaMills rather than to conventional ball mills. So further advances in increasing the feed size of IsaMills (eg with coarser or higher density media), and/or reducing the product size of HPGRs could enable this circuit. Reducing the product size of HPGRs could be achieved in several ways, including by using two or more HPGR stages in series, with each stage in either closed circuit, or in a simpler open circuit configuration.

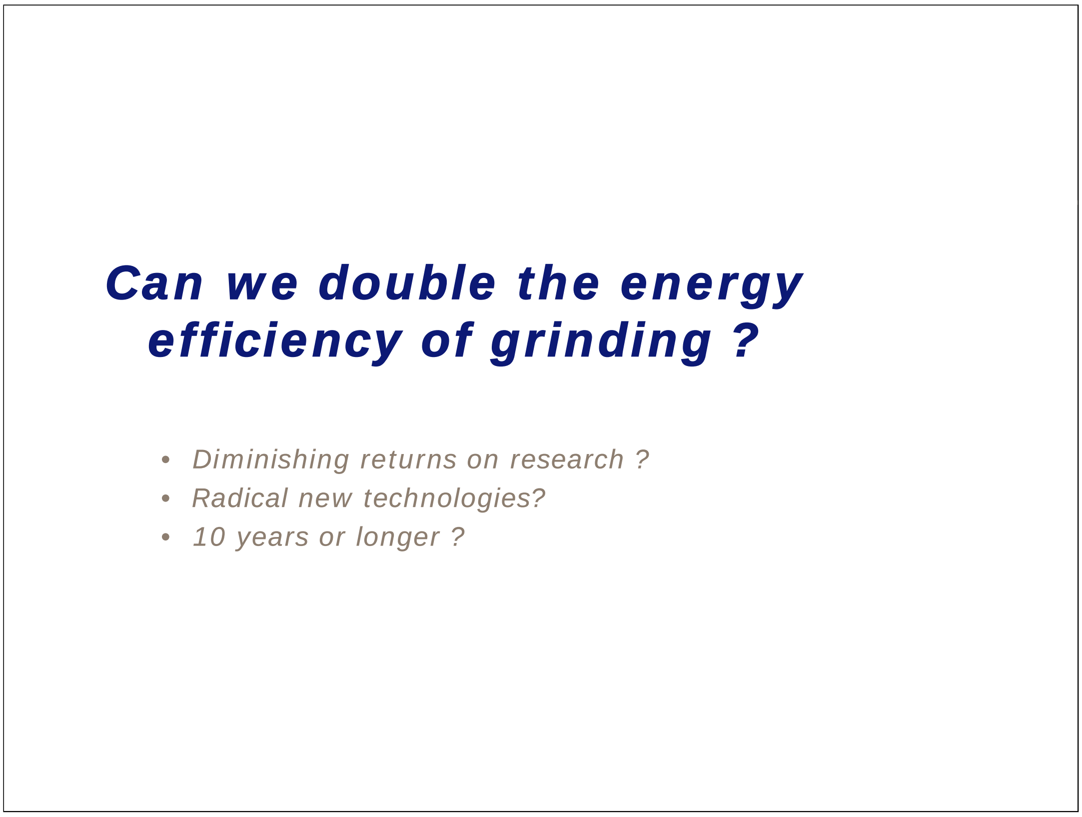
Perhaps the HPGR discharge sizing and IsaMill feed sizing will not converge quite close enough for them to meet as we have shown in the above simplified diagram. Perhaps the P80's will be mismatched, or maybe just the coarse fraction from HPGR will be too coarse for the IsaMill. In this case, the simplified circuit above could be modified to include a conventional grinding and/or classification stage between HPGR and the IsaMill. In this way, the benefits of energy efficiency and low steel consumption of HPGR and IsaMills would be maintained, with the addition of a relatively small conventional grinding and classification step (eg ball mill or Tower Mill with or without a classification stage).

Increasing the Energy Efficiency of Processing

Joe Pease

General Manager Xstrata Technology

August 2007



Can we double the energy efficiency of grinding ?

- *Diminishing returns on research ?*
- *Radical new technologies?*
- *10 years or longer ?*

Mineral comminution consumes about 3% of the world’s energy.

So improving the efficiency of grinding has long been an important objective for operators and researchers. Much progress has been made. Entire research institutions like the JKMRC have dedicated decades of study with almost the sole objective of improving comminution efficiency.

Does all this work mean that the big gains have been made, and that we are on diminishing returns from our research ?

Has conventional technology reached its limit ? Is equipment just getting bigger, but not more efficient ?

Do we need radical new technologies like microwave grinding to make a step change in grinding efficiency ?

A new collaborative research program, AMSRI has a $20 M budget and the ambitious target to double grinding efficiency in 10 years. Is this a realistic target, given the enormous of amount of work already done?

No, we don’t think 10 years is a reasonable target. We think it can be done today !

The tools are already available

- ***Laboratory grinding and flotation characterisation***
- ***Quantitative mineralogy***
- ***Basic thermodynamics***

In fact, processing efficiency can be significantly increased by using tools already developed by our previous research.

We don’t need radical new technologies, we need to combine the tools we already have in the right way.

These tools are :

•Laboratory grinding and laboratory flotation tests

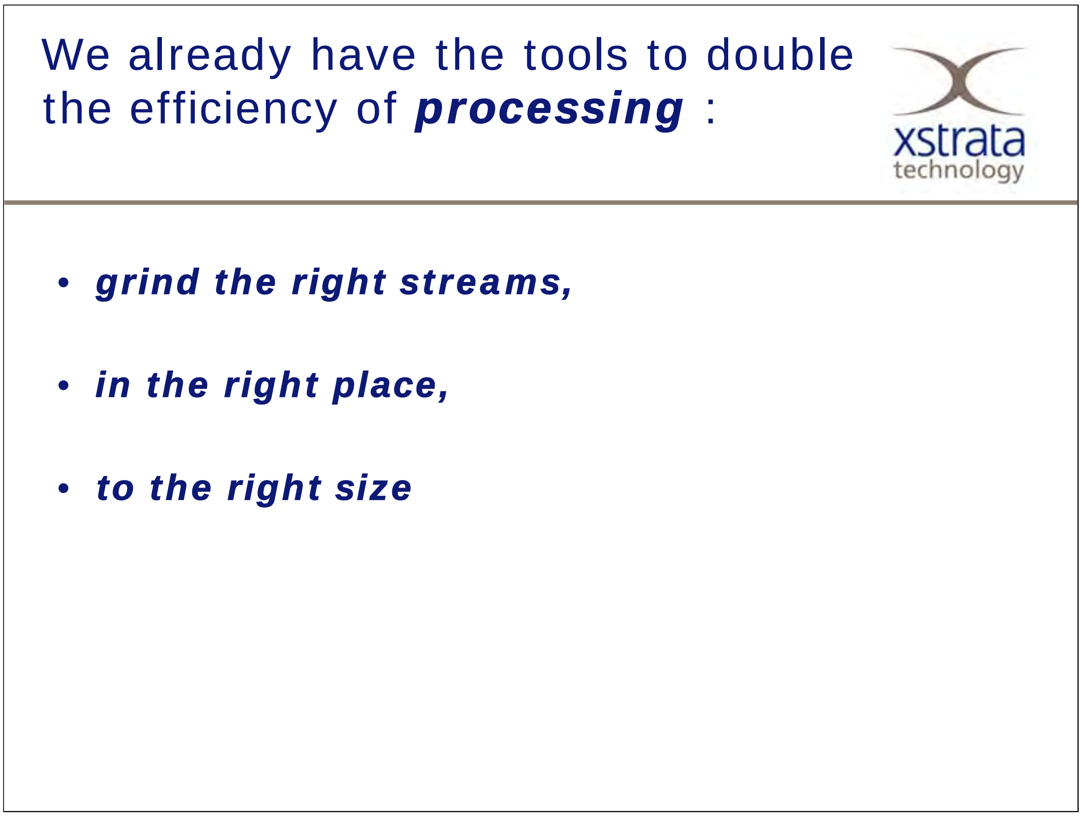
•Quantitative mineralogy

• basic thermodynamics of smelting and refining

To grind the right mineral in the right place to the right size.

In many cases this approach can double the energy efficiency of ***processing.***

Note that I refer to processing efficiency, not grinding efficiency. Grinding is not an end in itself. It is just one step in the processing chain. It cannot be considered in isolation of the downstream process. This is why I have included smelting thermodynamics in the list of tools. It is no point reducing grinding energy if this increases smelting energy.



We already have the tools to double the efficiency of ***processing*** :

xstrata technology

- ***grind the right streams,***
- ***in the right place,***
- ***to the right size***

In fact, processing efficiency can be significantly increased by using tools already developed by our previous research.

We don’t need radical new technologies like microwave grinding, we need to combine the tools we already have in the right way.

These tools are :

•Laboratory grinding and laboratory flotation tests

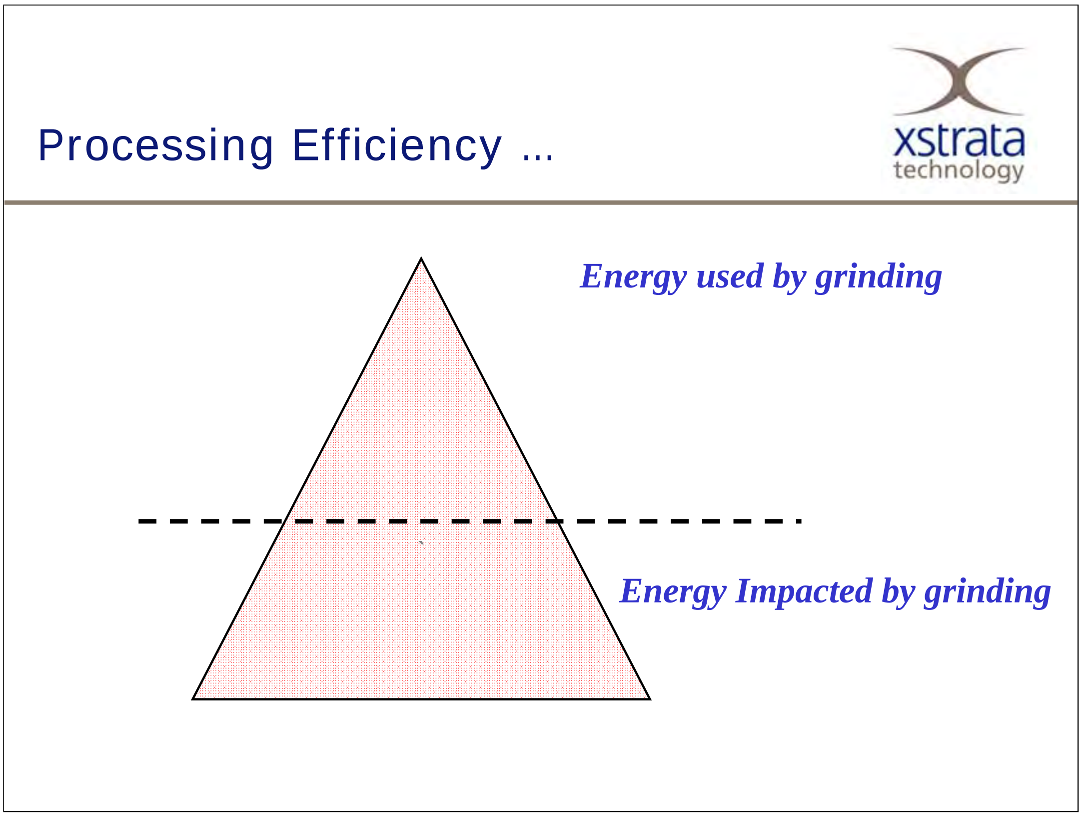
•Quantitative mineralogy

• basic thermodynamics of smelting and refining

To grind the right mineral in the right place to the right size.

In many cases this approach can double the energy efficiency of ***processing.***

Note that I refer to processing efficiency, not grinding efficiency. Grinding is not an end in itself. It is just one step in the processing chain. It cannot be considered in isolation of the downstream process. This is why I have included smelting thermodynamics in the list of tools. Sometimes we should increase grinding energy, if we can reduce smelting energy by more.



Processing Efficiency …

Energy used by grinding

Energy Impacted by grinding

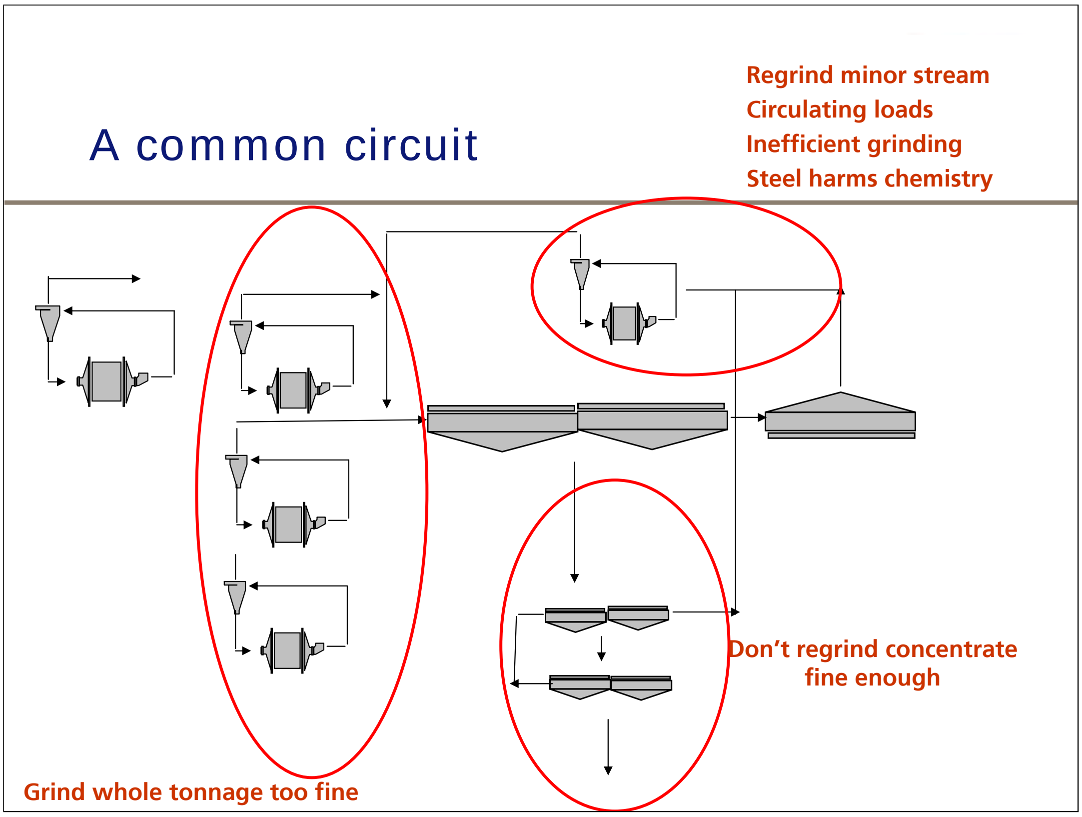
So of course, the real question is about the energy efficiency of the whole process, not just the grinding step.

It is no point reducing grinding energy if it increases downstream energy by more.

This is the same logic we all understand about mine-to-mill. One of the advances in the last decade has been to recognise the importance of the mine-to-mill interaction. Most of us will be familiar with the classic example of the mining engineer who reduced his blasting cost by 20 c/tonne, only to reduce SAG mill throughput by 10% and increase milling cost by $1/tonne. Sometimes the right answer is to spend a bit more and blast finer in the mine, and save much more in the mill.

Well of course, the same applies between the mill and the smelter. Yet how many of us check the thermodynamics of smelting, and see what we can do in the mill to reduce smelting energy ? Maybe that’s not a fair question, because we don’t have simple tools to help us do it. Have you ever seen a grinding model that includes some simple smelting thermodynamics to help you minimise processing energy ? Which research group is working on it ? We have research proposals to develop more and more sophisticated grinding simulations, to seek the last little bit of grinding efficiecny. But where do they consider smelting energy ?

I think this is a huge oversight, and a huge opportunity – we are just like the mining engineer who tried to reduce his blasting cost with no consideration of what it would do to grinding efficiency.



A common circuit

Regrind minor stream
Circulating loads
Inefficient grinding
Steel harms chemistry

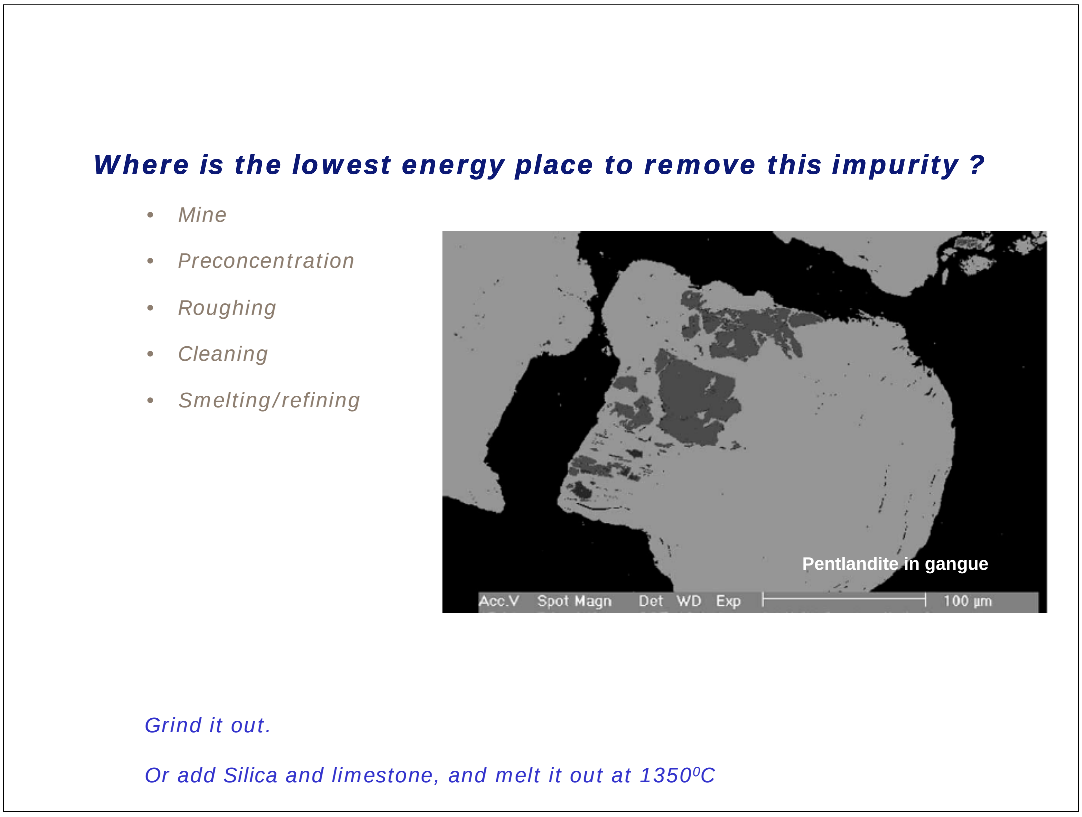
Don’t regrind concentrate fine enough

Grind whole tonnage too fine

This is a flowsheet of a common circuit. How can we use existing tools to reduce the energy need.

Often the primary grind can be coarsened. We don’t need complete liberation for roughing, we just need to liberate enough gangue away from the valuable mineral to get high rougher recovery. We can then achieve the liberation we want for cleaning by regrinding a smaller tonnage of rougher concentrate (or a part of it).

This is the biggest single impact on grinding energy – why grind big tonnages of silica finer than you need to? Grinding too fine up front also causes us to put in more roughing flotation capacity than we really need. Coarsening primary grind also has other advantages for tailings storage and reactivity.



Where is the lowest energy place to remove this impurity ?

- *Mine*
- *Preconcentration*
- *Roughing*
- *Cleaning*
- *Smelting/refining*

Grind it out.

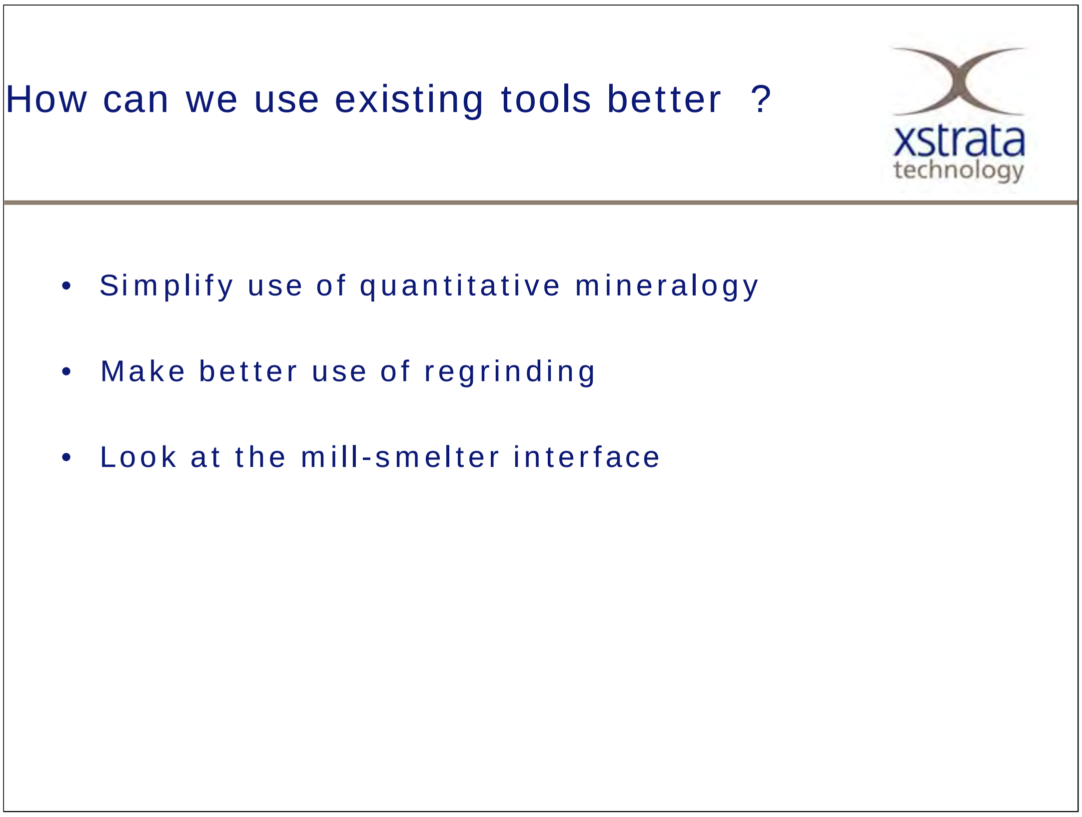
Or add Silica and limestone, and melt it out at 1350^0C

Improving energy efficiency means asking a simple question : where is the cheapest, lowest energy place to remove this impurity ?

- in the mine (by dilution control) – requiring no energy
- in a preconcentration stage (removing some gangue while it is still coarse)
- in roughing – ie by grinding the whole mill feed
- in cleaning – ie by grinding a much smaller tonnage of rougher concentrate
- in the smelter – by melting it out ?

Of course, cleaner feed tonnage is much lower than rougher feed tonnage. So ideally, so it is obviously better to grind the lower tonnage of rougher concentrate rather than rougher feed. So long as you grind rougher feed just enough to get recovery into the rougher concentrate.

The photo of the composite shown is taken from a paper on Nickel metallurgy. The gray is pentlandite, the white is MgO. Excellent work was done in the flotation circuit to allow recovery of this low grade composite. Even so, this particle causes a problem for the smelter – MgO causes highly viscous slag. The smelter has to add Silica and lime fluxes, then heat the whole lot up to 1350^0C to remove the MgO. This consumes a lot of energy – over 10 times the energy it would take to grind the MgO out. The high temperatures also make smelting difficult and reduce refractory campaign life. So to improve energy efficiency, the concentrator should regrind this particle – so long as the mill can still recover the pentlandite.



How can we use existing tools better ?

- Simplify use of quantitative mineralogy
- Make better use of regrinding
- Look at the mill-smelter interface

I think there three areas where we can improve our use of the existing tools to improve energy efficiency, which I will discuss in the next few slides.

.

Quantitative Mineralogy …

…. See the forest for the trees

What do we want to achieve in each part of the circuit, eg

- Primary grind and rough for recovery
  - Look for gangue liberation, not mineral liberation
- Regrind and clean for grade
  - Look for mineral liberation and entrainment

Distinguish symptoms from causes

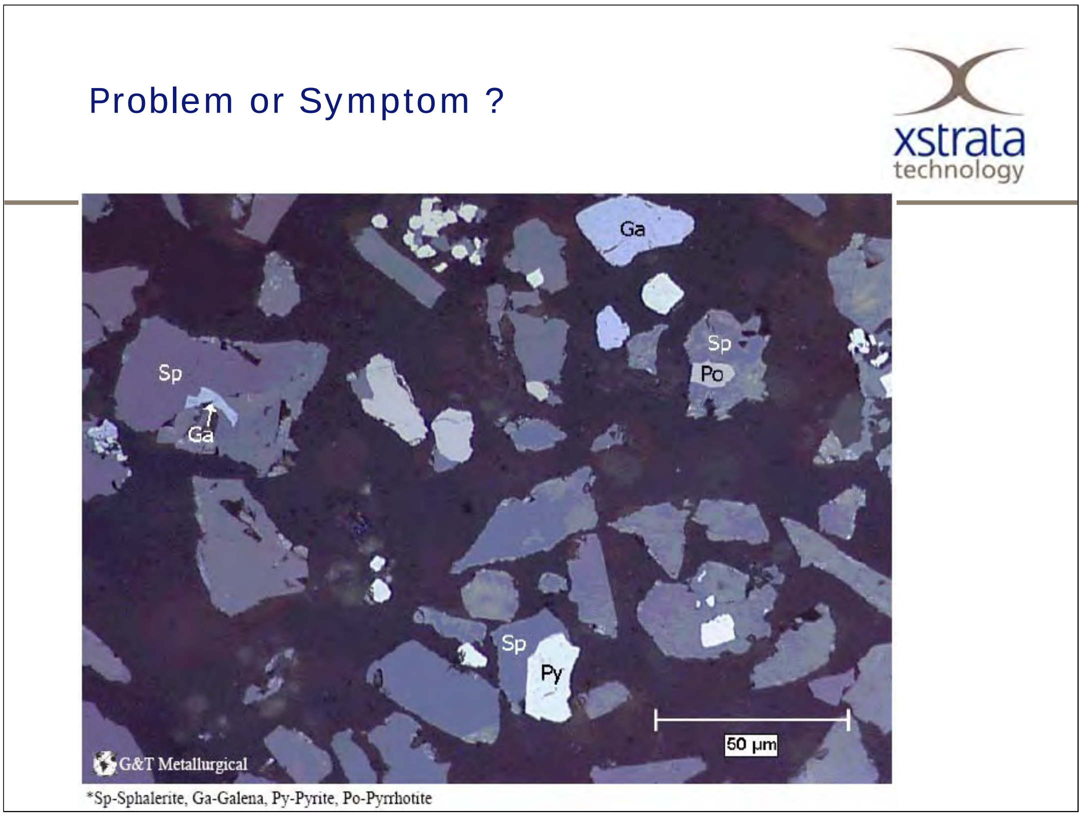
- Eg fines in tails may mean …
  - … you are grinding too fine
  - OR you are grinding too coarse !

Because quantitative mineralogy is so powerful, sometimes we overcomplicate it. It can generate so much data that we get lost in detail, and can't see the forest for the trees. We need to keep it simple, and apply the 80/20 rule. Avoid the temptation to find out everything about every mineral in every liberation class in every stream. Ask a few simple questions about what you are trying to achieve in each stage of the process. And look for root causes rather than symptoms. The mineralogy can only be interpreted with a good understanding of how the circuit is operated.

Two simple questions are :

1. in roughing, how coarse can I grind and still be able to get good recovery to rougher concentrate. You don't need to worry about grade yet, that is, you don't need to worry about the mineral liberation, but rather the gangue liberation – is enough gangue liberated from the mineral to allow it to float into rougher concentrate.
2. In cleaning, the question changes to how fine you need to grind to make a high grade concentrate. Now mineral liberation is important. So is entrainment of liberated gangue. If there are coarse composites in concentrate that could be easily liberated, these are "low hanging fruit". Even if you are already at concentrate grade, by removing the easy impurities you make room to "pull" the rest of the circuit harder.

This point is an example of distinguishing root causes from symptoms. You will always find fine liberated values in cleaner tailings and rougher tailings. Perhaps this means you are grinding too fine – "overgrinding" your valuable minerals. But, paradoxically, more often it means you aren't grinding fine enough ! The next slide shows why.



Problem or Symptom ?

*Sp-Sphalerite, Ga-Galena, Py-Pyrite, Po-Pyrrhotite

This is a photomicrograph of zinc concentrate. Look at the composite particle of sphalerite and pyrite at the bottom. It has a fair amount of exposed sphalerite surface, so it floats well, and is too valuable to throw out. Yet it is also too low grade to accept into concentrate - it probably assays about 20% Zn, it needs to be about 50% to be payable.

Because it floats relatively easily, the operator only has two choices :

•If he accepts it in concentrate, it lowers con grade. He has strict limits on the concentrate grade and Fe content, so he now has to pull the roughers and cleaners more gently to keep the concentrate grade. He can’t chase coarse composites, or slow floating fines in tailings, because he doesn’t have the processing power to recover them at a high grade.

•Alternatively, he tries to reject this composite, in this case by starving collector and running high pH. But these are the very conditions that will cause low recovery of the fines. Look at the fine liberated sphalerite particles near the composite – they have less sphalerite surface than the composite, so depressing the composite will probably depress them as well.

So no matter what the operator does, if he has to make a concentrate grade, then dealing with this composite will cause him to lose fines (and coarse composites). Instead, regrinds the composite removes the pyrite which increases concentrate grade without depressant. The higher concentrate grade leaves him room to pull his roughers and cleaners harder. Regrinding will inmprove his fines recovery.

This shows the importance of using quantitative mineralogy like forensics, not like a computer sledgehammer. Don’t get hung up on there being fines in tails. Look at the composites in concentrates, if you fix this root cause many of the symptoms disappear.

Before you can ask the right questions of your mineralogy, you **have** to understand how the circuit is operated.

Grind the right minerals in the right place

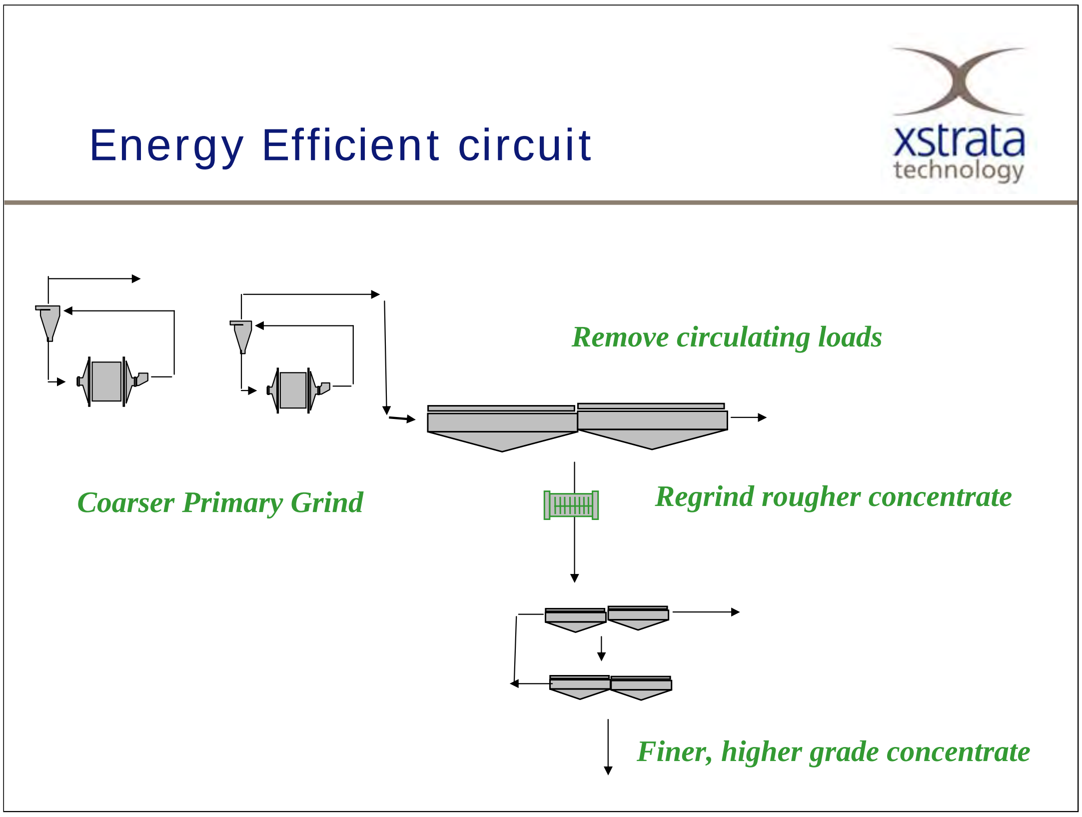
- ***Primary grind as coarse as you can***
- ***Regrind as fine as you need***
- ***Regrind cleaner feed, not cleaner tail***
  - *No step change from grinding a minor stream*

In summary, use quantitative mineralogy to ask these simple questions. In roughing, how coarse can I grind and still get recovery to rougher concentrate.

In cleaning, how many composites can I remove from the concentrate by regrinding finer.

We think the result is that many circuits could grind the main stream coarser, but need to regrind the cleaner feed finer than they do. They can make a higher grade at the same recovery, with less energy.

So many of our circuits regrind cleaner tailings, not cleaner feed. But if regrinding is a powerful tool, why save it until after you have made most of your concentrate? If composites like the one in the previous photo go straight to concentrate, the damage is already done. Cleaner tailing has a small portion of the metal, so you can't get a step change by regrinding it.



Energy Efficient circuit

Remove circulating loads

Coarser Primary Grind

Regrind rougher concentrate

Finer, higher grade concentrate

Compare this flowsheet to the one on page 6. Often the primary grind can be coarsened. We don’t need complete liberation for roughing, we just need enough liberation of valuable from gangue to allow high recovery in roughing. We can then achieve the liberation we want for cleaning by regrinding a smaller tonnage of rougher concentrate (or a part of it).

This is the biggest single impact on grinding energy – why grind big tonnages of silica fine when you don’t need to. Grinding too fine up front also causes us to put in more roughing flotation capacity than we really need.

The regrind mill shown here is an IsaMill. These mills grind in open circuit, producing a sharp sizer distribution than conventional grinding and hydrocylones.

Coarsening primary grind also has other advantages for tailings storage and reactivity.

Do we do enough Regrinding ?

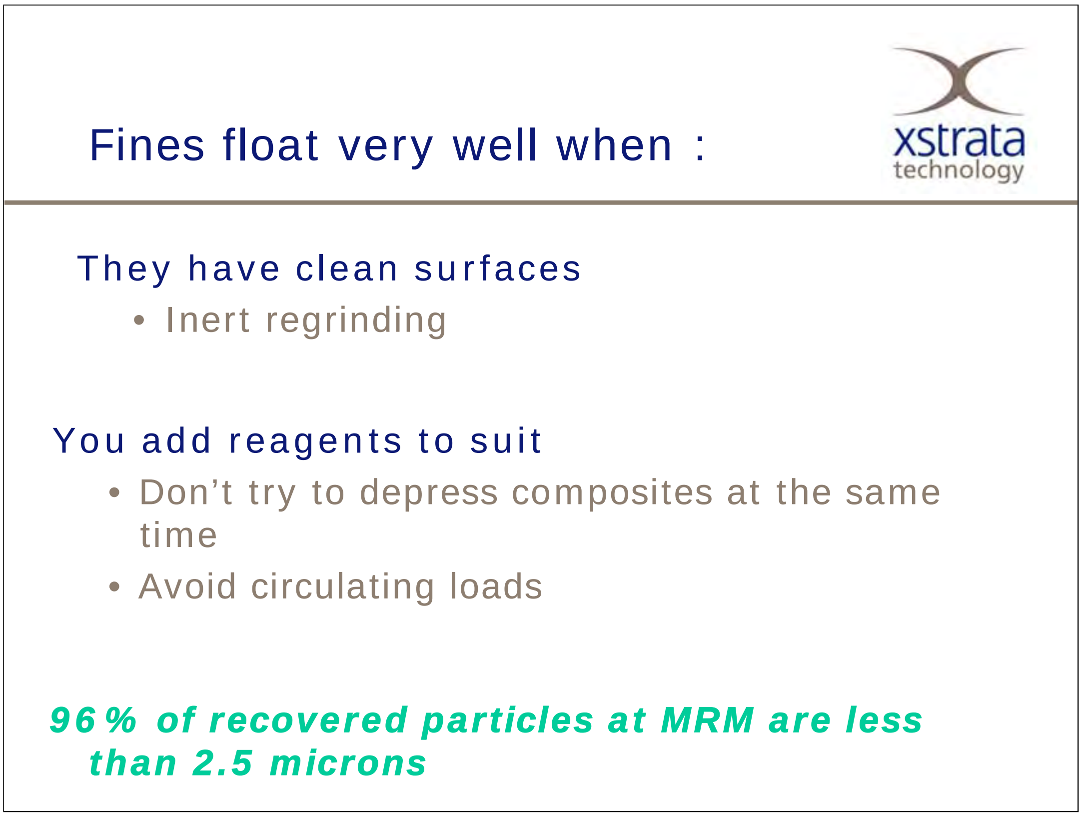
- Fines are hard to float ?
  - Not with clean surfaces and good size distributions
- Regrinding hurts chemistry ?
  - Not with inert grinding
- Regrinding is inefficient ?
  - Not any more

We think many concentrators don't regrind fine enough. When they do regrind, they regrind small difficult streams like cleaner tails or scavenger concentrate, not the main cleaner feed.

This probably reflects some perceptions about regrinding and subsequent flotation. Indeed, these concerns are valid, and most plant operators have experienced them.   Conventional regrinding to fine sizes is inefficient, it uses lots of media, and the steel media hurts flotation. Therefore we avoid regrinding cleaner feed so we don't risk harming our main stream; we save regrinding to cleaner tailing, a relatively small and difficult stream.

But this has all changed with the advent of high intensity inert grinding. The IsaMill is much more efficient than ball or Tower Mills for fine grinding, it uses inert media that produces clean surfaces for flotation, and produces sharp size distributions.

Regrinding used to be one step forward for liberation, one step back from chemistry. Not any more.



Fines float very well when :

They have clean surfaces

- Inert regrinding

You add reagents to suit

- Don't try to depress composites at the same time
- Avoid circulating loads

96% of recovered particles at MRM are less than 2.5 microns

Traditionally we thought that fines don't float well, that we need special reagents and flotation cells, and still get poor performance.

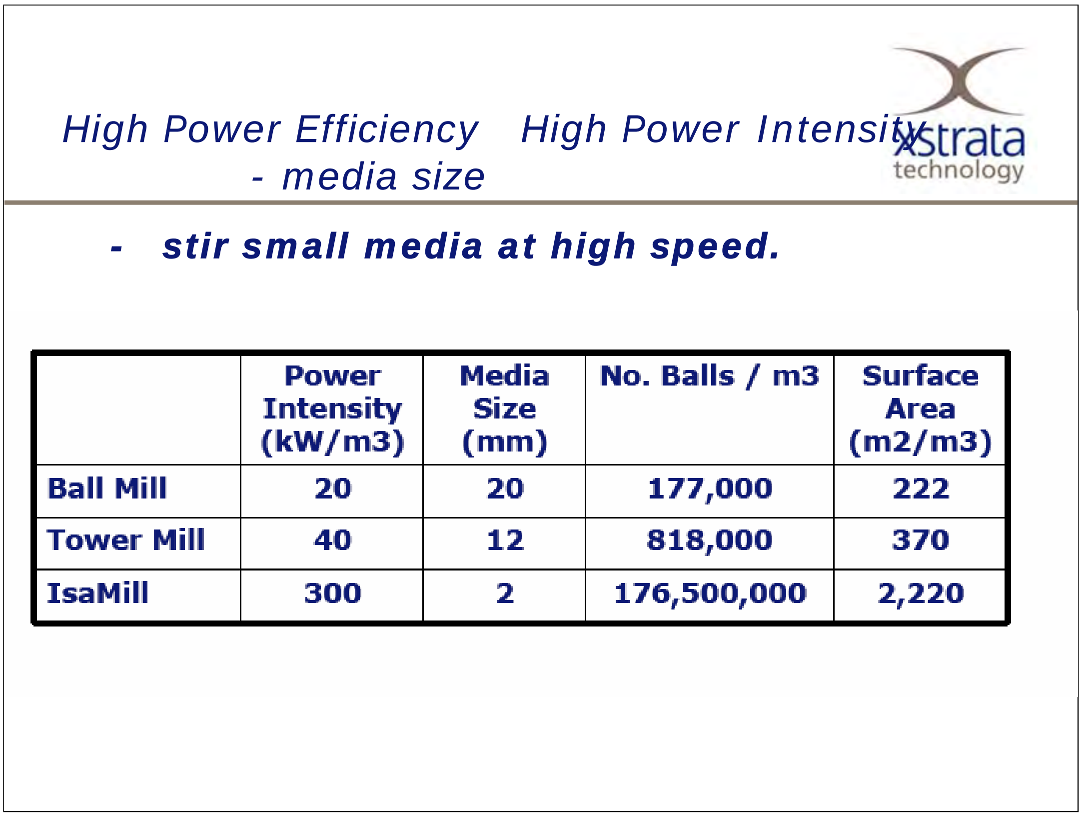
We now know this isn't true. Numerous operations get excellent fines flotation, with conventional flotation cells and standard reagents. We just have to give them clean surfaces, and make sure they aren't competing with composites – that is, we need to be able to add enough collector to float the fines, without having to worry about floating unwanted composites at the same time (as we discussed before).

When we do this, we find that fines down to 1 micron float just as well as intermediate size particles, and almost as quickly. In the Mt Isa lead and zinc circuit, cleaning recovery after IsaMilling of cleaner feed is above 95% for all size fractions above 1 micron, until dropping above 37 microns due to composites.

For another example, consider Macarthur River, an operation whose entire concentrate is finer than 7 microns, floated in conventional cells. Lets look at what this means in terms of individual particles. I know we only get paid for weight, not number of particles. But flotation is about individual particles connecting with bubbles.

So 96 percent of the successful particle-bubble intereactions at McArthur River occur for particles less than 2.5 microns.

So fines do float,, and very well indeed.



High Power Efficiency   High Power Intensity
- media size

Xstrata technology

- stir small media at high speed.

	Power Intensity (kW/m3)	Media Size (mm)	No. Balls / m3	Surface Area (m2/m3)
Ball Mill	20	20	177,000	222
Tower Mill	40	12	818,000	370
IsaMill	300	2	176,500,000	2,220

A lot has been written about grinding efficiency, and the relative efficiencies of ball mills and Tower Mills and high speed stirred mills. In fact, it is fairly simply. Overwhelmingly the biggest effect is media size. If you run a ball mill slowly with 12 mm media, it will be as efficient as a Tower Mill. Its just that we don't normally run ball mills with 12 mm media.   Similarly, a Tower Mill with 5 mm or smaller media will be more efficient again. But practically, no one ever runs a Tower Mill with 5 mm media – the media cost and wear would be prohibitive. As an aside, if you get testwork done, make sure it is done with the same media size you will use in the plant. We have heard of occasions when a Tower Mill energy need has been predicted from a test with 6 mm media, whereas the plant will use 12 mm or coarser, which will require significantly more energy.

The other important parameter in this slide is power ***intensity*** – it is no point being power efficient (like a Tower Mill with 5 mm media) if the power intensity is so low that you need a huge installation to get the necessary installed power.

So let me summarise – a Tower Mill with small media will be just as efficient as a high speed stirred mill with small media (the optimum media size for each device will be different). But the low power intensity of Tower Mills means you need a much bigger installation. These points are made by Nesset et al in their paper *"Assessing the Performance and Efficiency of Fine Grinding Technology", CIM 38th Annual Canadian Mineral Processing Operators Conference, Ottawa 2006.*

3 MW IsaMill Installation

Mill product direct to circuit

Feed in

Scew feeder for media addition

This is a picture of a 3 MW IsaMill installation, showing the features that make it such a different technology.

Firstly, note the scale from the people at the discharge – this is a 3 MW grinding mill. Consider the size of a 3 MW ball mill or three, 1 MW Tower Mills.

Note the simple installation. The mill is pressured, with feed in, outlet pipe straight to process (no closed circuit cyclones) and simple media system under the floor. Media is added to the mill feed pump by the screw feeder. Note the low footprint, low head height, and simple crane needs.

A Step Change in Technology

1.1 MW Tower Mill

- plus media handling
- plus close circuit cyclones

3 MW Isa Mill

Media system under floor

Open circuit operation

To give an idea of difference in scale – the IsaMill is not an incremental change, it is a step change – an order of magnitude smaller than conventional technology. Until now, the Tower Mill was considered the most efficient, modern technology for regrinding. After 55 years, the biggest model is still only 1.1 MW (shown).

The IsaMill next to it has been scaled to keep the people about the same size. This is a 3 MW IsaMill – three times the power of the much bigger Tower Mill. This is what we mean by power intensity. Think of the difference in installation cost if you need 3 MW of Tower Mills (with associated media handling, cranes, and closed circuit cylcones).

This is why we say that the IsaMill changes the way we can think about circuit design. It is much easier to imagine putting 3 MW of open circuit IsaMill in the middle of a flotation circuit.

A new tool for plant design

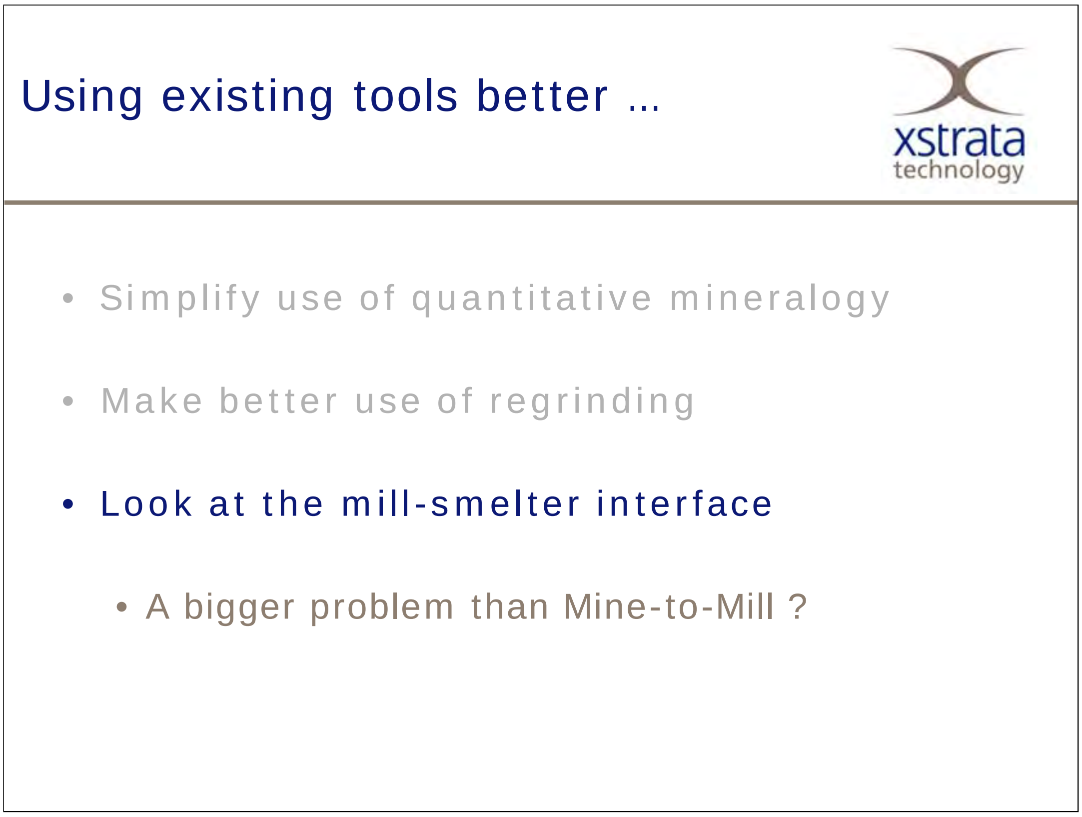
Regrinding that

- is energy efficient
- improves mineral surfaces and chemistry
- has a small footprint and head height
- doesn't need cyclones
- can be installed throughout the circuit

The tools to grind the right minerals in the right place

The IsaMill is a step change in regrinding technology. It was developed to change the economics and efficiency of ultrafine grinding. But it has now brought the same advantages to conventional grinding and regrinding, where it will have its biggest effect on the industry.

This fundamentally changes the way we can think about the design of circuits. A lot of regrinding power can be installed in a small space and in a simple installation that can be located where it is needed throughout the plant.



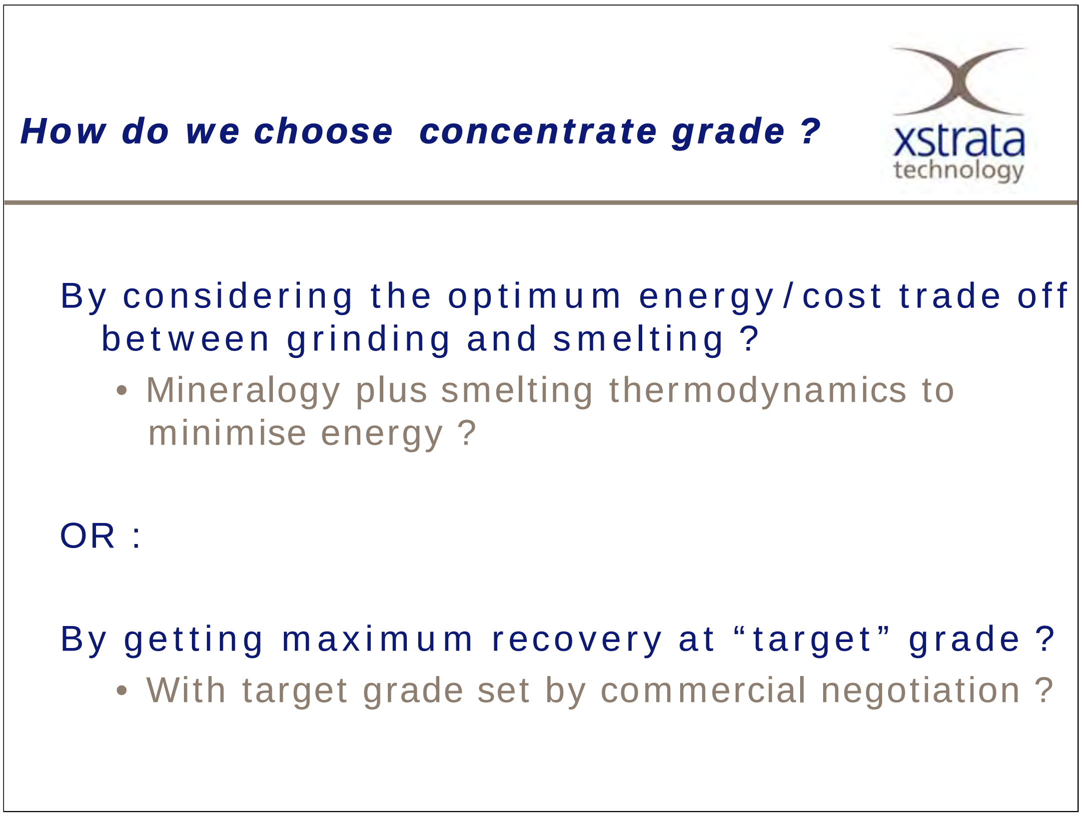
Using existing tools better ...

- Simplify use of quantitative mineralogy
- Make better use of regrinding
- Look at the mill-smelter interface
  - A bigger problem than Mine-to-Mill ?

The third reason we don’t take achieve the energy efficiency we could is because of the mill to smelter interface.

Some time ago we recognised the mine to mill interface as a problem. We knew it was a problem because we could hear the miners and metallurgists squabbling on sites. Since then a lot of work has been done to better co-ordinate mining and milling to improve overall efficiency – eg is it better to spend a bit more on mine blasting (drilling and chemical energy) to save more energy in the SAG mill.

But how often do we hear the same questions being asked about concentrating and smelting ? Or do we just accept that the target grade is 25% Cu because that suits the smelter contract. I think the mill-to-smelter interface is a more serious problem than mine to mill, because there is virtually no dialogue between us. You can’t even hear us fighting, because we don’t talk. The smelter may be owned by another company, and based overseas. My commercial people talk to your commercial people, and they argue about supply/demand balance. Can anyone imagine that they are negotiating about the most energy efficient way to remove impurities.



How do we choose concentrate grade ?

By considering the optimum energy/cost trade off between grinding and smelting ?

- Mineralogy plus smelting thermodynamics to minimise energy ?

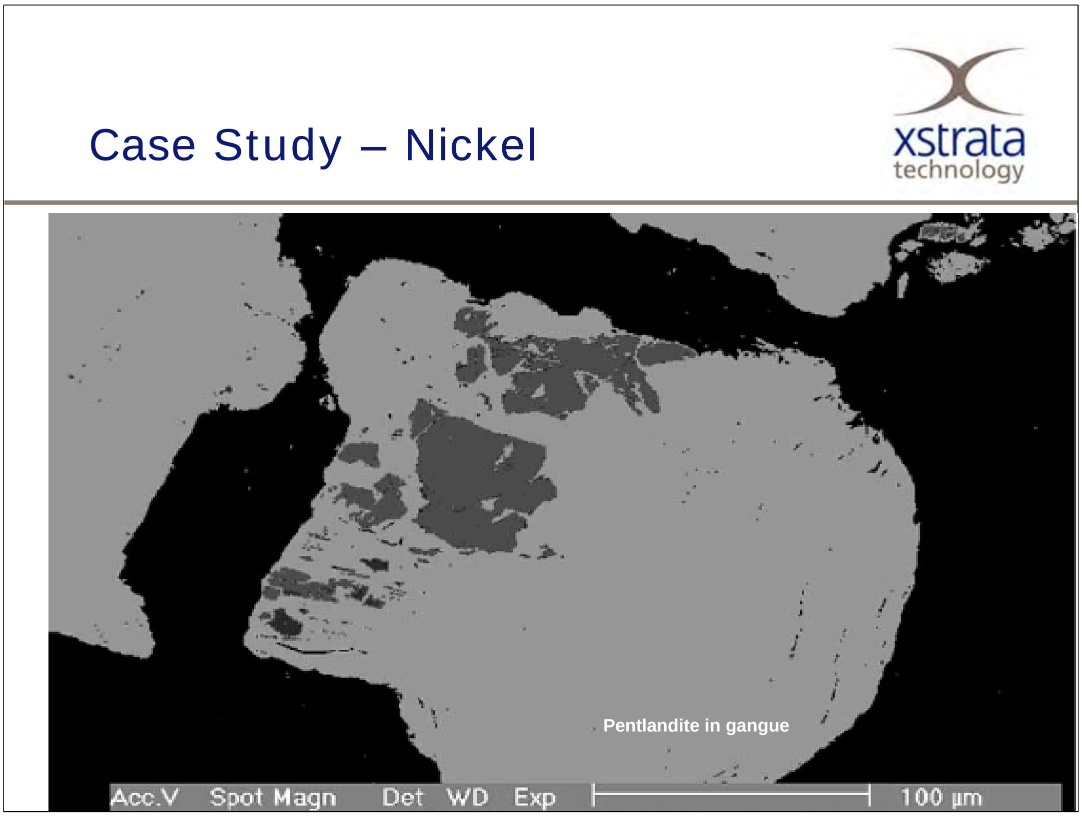
OR :

By getting maximum recovery at “target” grade ?

- With target grade set by commercial negotiation ?

It is logical for the concentrator to maximise return based on the smelter contract. They will work out what grade concentrate gives them the best return given contract terms and transport costs. Then they will maximise recovery and minimise costs at this grade. They will set KPI’s, and be rewarded to minimise cost and maximise recovery – which will “lock in” that target grade. What incentive do they have to question the grade ? Why would they volunteer to increase costs, maybe even drop some recovery, in order to make a higher concentrate grade, if this just helps another company’s smelter. Of course, they won’t.
The concentrator can only respond to the contract. And can we expect the contract negotiators to be optimising the trade off between mineralogy, smelting thermodynamics and energy efficiency ? Of course we can’t.

This is why I think mill to smelter is a bigger problem than mine to mill. There is simply no dialogue. The metallurgy and calculations are fairly easy, but we don’t do it. We don’t even recognise it as a problem. Of all the research projects you have heard of to improve grinding efficiency, how many include the smelter ?



Case Study – Nickel

xstrata technology

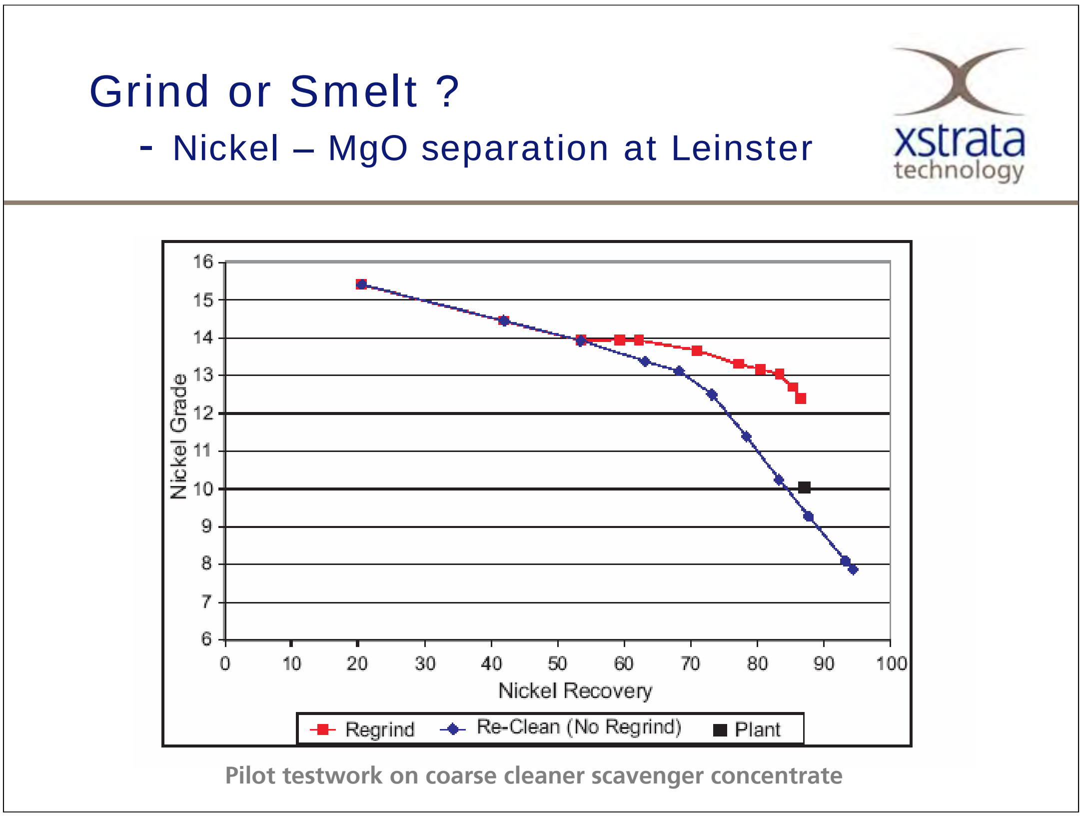
Let me give an example. I return to the Nickel composite we looked at earlier. This photo was taken from a paper by G. Senior and S.Thomas, *“Development of a New Flowsheet for the Flotation of Low Grade Nickel Ore”, International Journal of Mineral Processing 78(2005), Elsevier.*

It is a photo of a low grade composite floated into scavenger concentrate at Mt Keith circuit. Excellent work was done in the concentrator to be able to recover this low grade composite, without requiring a finer primary grind.

So the nickel recovery has been increased, and the process is more efficient. It could be even more efficient if we could then eliminate some of the additional MgO from the final concentrate, while still recovering the pentlandite. This is a liberation issue.

We were originally asked by what was then WMC to look at the potential for IsaSmelt for the Kalgoolie Nickel smelter, to handle the high MgO content. But the more we looked at smelting, the more we suspected that it would be easier to deal with the MgO in the concentrator.

So we worked with the Leinster operation to see whether we could improve the Nickel-MgO separation with inert regrinding in an IsaMill.



Grind or Smelt ?

- Nickel – MgO separation at Leinster

Pilot testwork on coarse cleaner scavenger concentrate

The answer is yes. This graph is from D.Seaman et al, *“Process Design of a Regrind Facility at the Leinster Nickel Operation to Improve Concentrator Recovery”, AUSIMM, 9th Mill Operators Conference, Freemantle 2007.*

The graph shows the results of pilot work conducted at Leinster. The coarse cleaner scavenger concentrate is shown with and without IsaMilling. The regrinding is effective in significantly increasing the Nickel concentrate grade at the same recovery. Most of the increase comes from removing MgO, as shown by the next graph.

Big impact MgO - and smelting energy

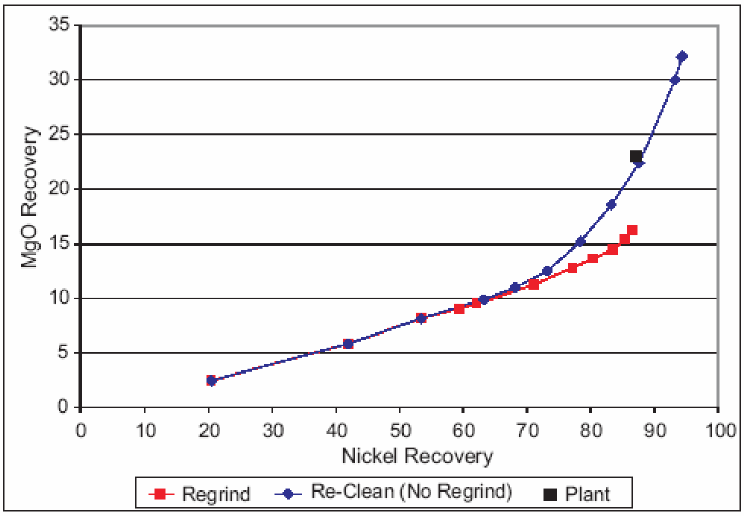
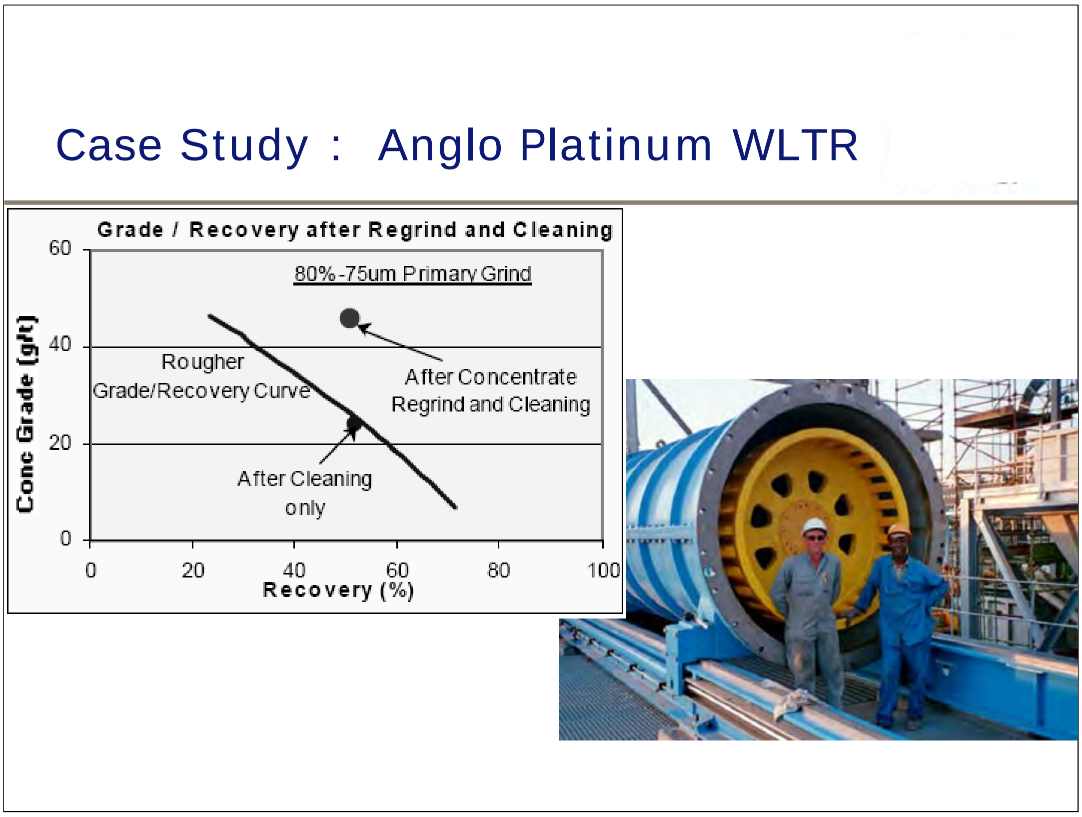


FIG 9 - MgO to nickel selectivity with and without regrind.

In this case, MgO can be removed from the concentrate without any loss of recovery with relatively little regrinding energy. This requires much less energy than smelting it out from concentrate.

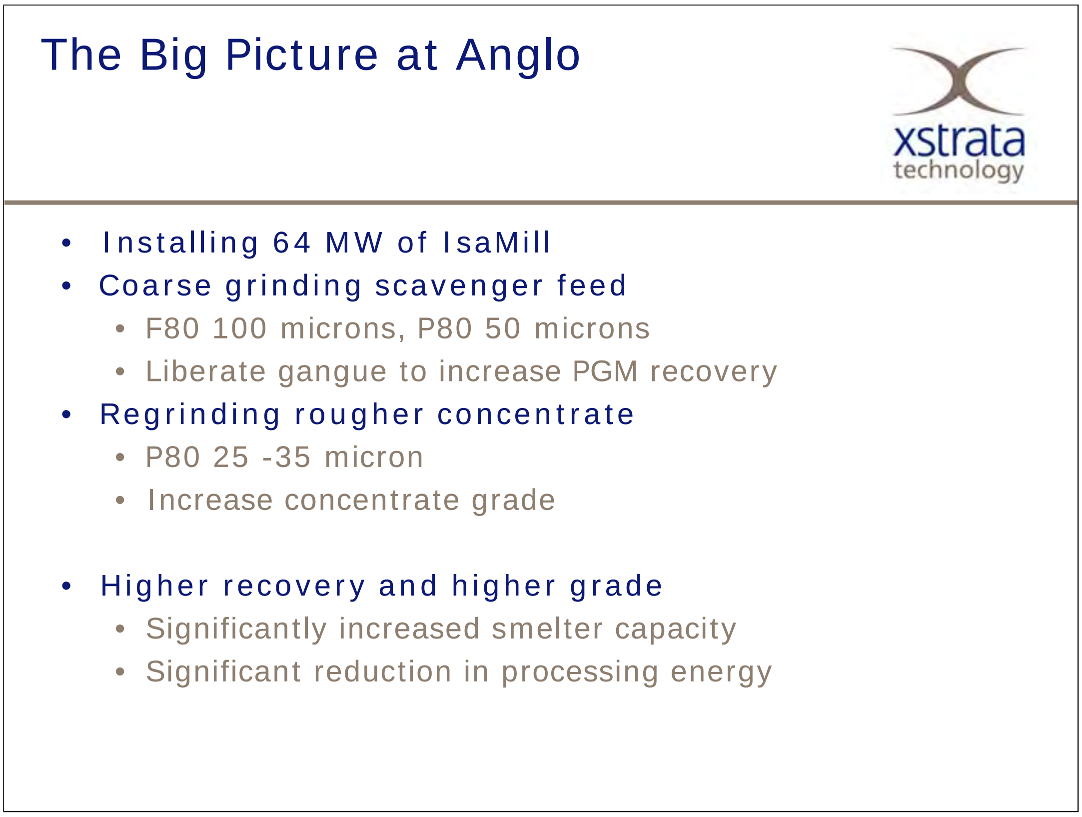


Case Study : Anglo Platinum WLTR

The next example is from Anglo Platinum. This mill is their first installation, at the Western Limb Tailings Retreatment Plant. In many ways this was typical of operations that needed a step change in technology – old mill tailings that had been in dams for up to 100 years, fine grained and with altered surfaces. The ability of the IsaMill to efficiently liberate, and provide clean new surfaces to flotation, was enabling technology for this operation.

This was the development site for the big M10,000 IsaMill – this unit is 2.6 MW and operates with sand media, the same mill model is now rated at 3 MW with ceramic media.

This is a fine grinding mill, operating on a difficult stream. But Anglo Platinum realised that the advantages of IsaMilling didn't stop with these difficult fine grained applications. They saw that the ability to improve liberation, and at the same time improve flotation selectivity, had much wider application to their mainstream applications.



The Big Picture at Anglo

xstrata technology

- Installing 64 MW of IsaMill
- Coarse grinding scavenger feed
  - F80 100 microns, P80 50 microns
  - Liberate gangue to increase PGM recovery
- Regrinding rougher concentrate
  - P80 25 -35 micron
  - Increase concentrate grade

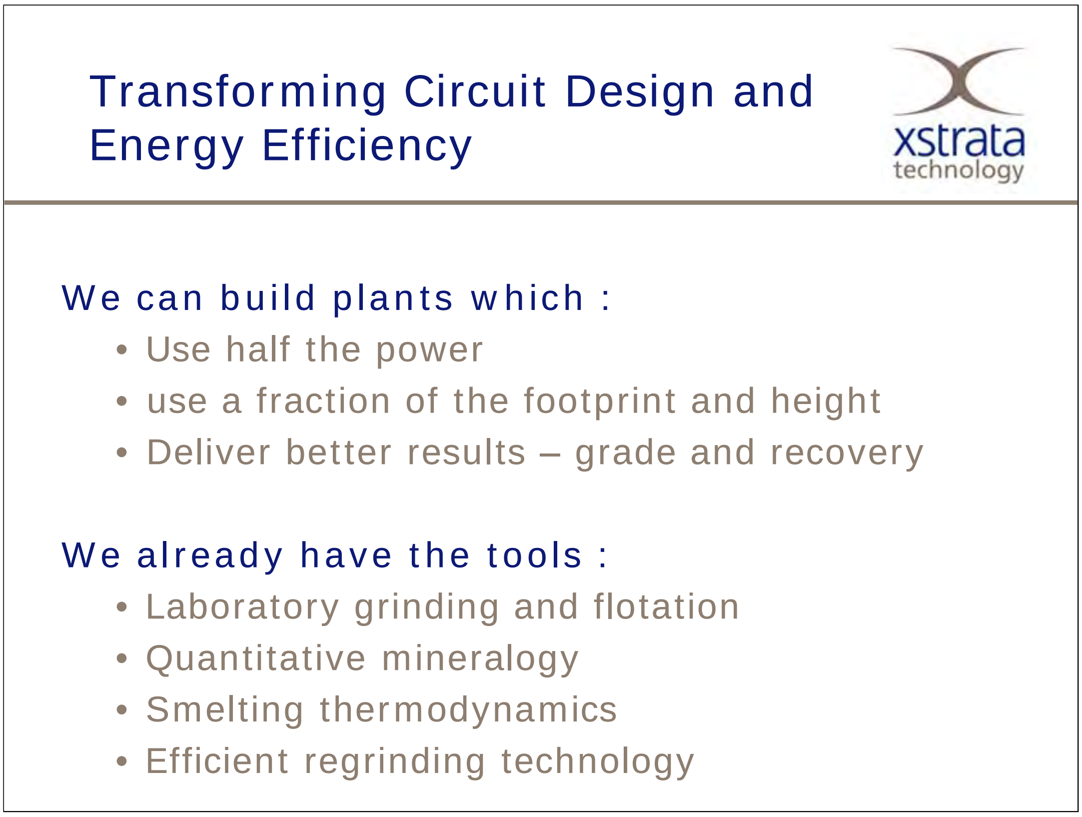
- Higher recovery and higher grade
  - Significantly increased smelter capacity
  - Significant reduction in processing energy

The end result is that Anglo Platinum have embarked on a major program to fit IsaMills to their concentrators, in a combination of mainstream grinding and regrinding duties. This program will significantly improve both recovery and concentrate grade. The increase in concentrate grade means a significant increase in capacity through existing smelting, and significant reduction in overall processing energy.

The mainstream mills will grind from typically F80 100 microns to P80 45 to 60 microns, and will treat rougher tail/scavenger feed. Each mill will process around 300 t/h of solids.

The regrind mills will treat rougher concentrate, and will regrind to P80 15-20 microns before cleaning.

It is clear that the IsaMill is now a mainstream grinding and regrinding mill, not a niche fine grinding mill.



Transforming Circuit Design and Energy Efficiency

xstrata technology

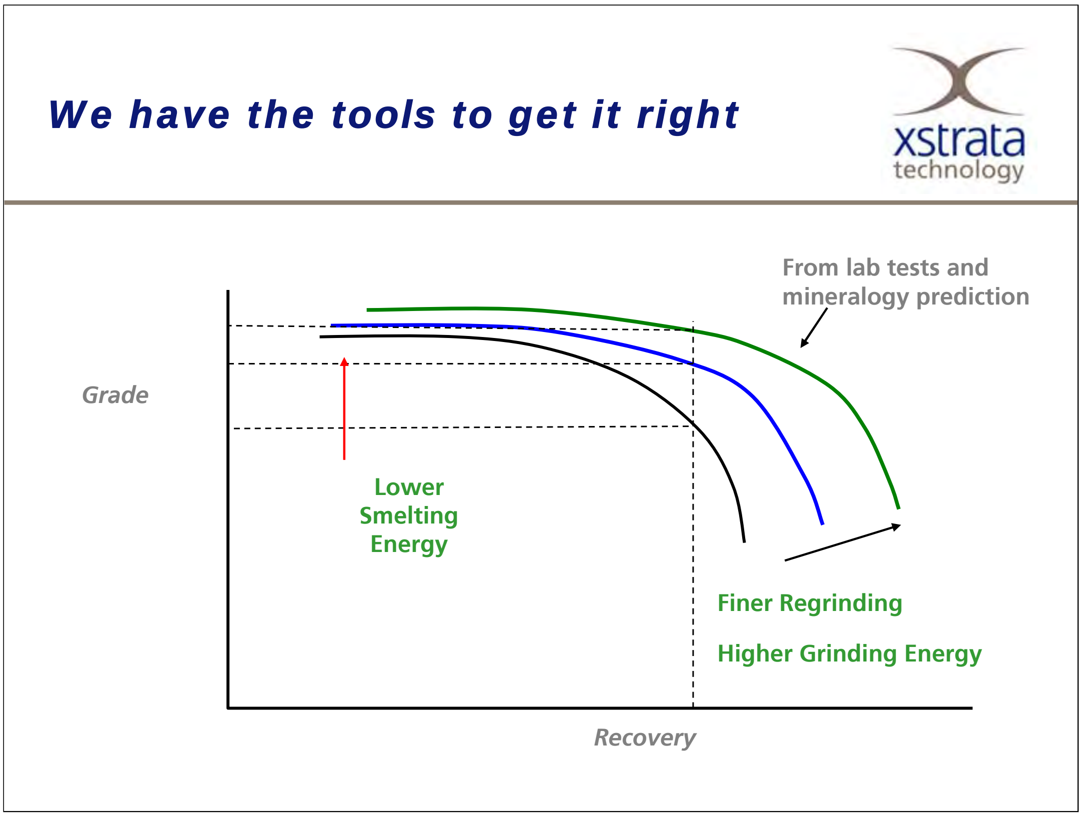
We can build plants which :

- Use half the power
- use a fraction of the footprint and height
- Deliver better results – grade and recovery

We already have the tools :

- Laboratory grinding and flotation
- Quantitative mineralogy
- Smelting thermodynamics
- Efficient regrinding technology

To summarise, we think we already have all the tools we need to double the energy efficiency of processing. We just need to use them together in the right way.



We have the tools to get it right

This is an illustration of what we can currently do. We can use laboratory tests to develop a range of grade/recovery curves at different grinding and regrinding sizes.

We can use quantitative mineralogy to help us design and interpret these tests, and interpolate between them.

We can predict the grinding energy needed to create the different grade/recovery curves.

We can use simple smelting thermodynamics to calculate the different smelting energy needed for the different concentrates.

This gives us enough information to find the most efficient operating point; the lowest energy trade off between grinding and smelting for that ore.

Challenge to Researchers :

xstrata technology

Develop a standard "energy index" to rank ores
  - Develop grade/recovery curves at different grinds
  - Calculate grinding energy and smelting energy for different options
  - Use for new ores, and benchmark existing plants

Reduce the energy of ***processing***, not just energy of ***grinding***

So in closing, I issue a challenge to researchers. If we have the tools, why don't we use them in the right way to optimise energy efficiency.

I think the previous concepts can be captured in a relatively simple "energy index" to describe an ore. For a long time the industry has used the Bond Work Index to summarise grinding energy needs. Even though there are much more sophisticated grinding models, we still find a single index to be useful.

So why don't we develop a modern energy index, like the Bond Index, but to include grinding and regrinding energy, the grade recovery curve, and basis smelting thermodynamics. A simple measure that for each ore predicts how much energy will be needed to produce metal (or final product). And a simple technique to indicate the lowest energy route for that ore, the best trade off between grinding and smelting/refining.

The same index and technique could be used to benchmark our existing operations – how do we perform against the estimate of the most efficient way to treat the ore.

Our focus is to improve the energy efficiency of processing, not just the energy efficiency of grinding.

IsaMill™ Technology Used in Efficient Grinding Circuits

B.D. Burford[1] and L.W. Clark[2]

High intensity stirred milling is now an industry accepted method to efficiently grind fine and coarse particles. In particular, the IsaMill™, which was invented for, and transformed the fine grinding industry, is now being included in many new comminution circuits in coarser applications. While comminution has always been regarded as important from a processing perspective, the pressure being applied by environmental concerns on all large scale power users, now make highly energy efficient processes more important than ever.

The advantages that were developed in fine grinding in the early IsaMill™ installations have been carried over into coarse grinding applications. These advantages include a simple grinding circuit that operates in open circuit with a small footprint, the ability to offer sharp product size classification, as well as the use of inert media in a high energy intensive environment.

This paper will examine the use of IsaMill™ technology in fine grinding (P80 below 15 micron), and examine the use of the technology in conventional grinding applications (P80 20 - 150 µm). Recent installations will be examined, including fine and coarse grinding applications, as well as the recent test work that was undertaken using an IsaMill™ in a primary grinding circuit, and the resulting circuit proposal for this site.

While comminution has been relatively unchanged for the last century, the need to install energy efficient technology will promote further growth in IsaMill™ installations, and result in one of the biggest challenges to traditional comminution design.

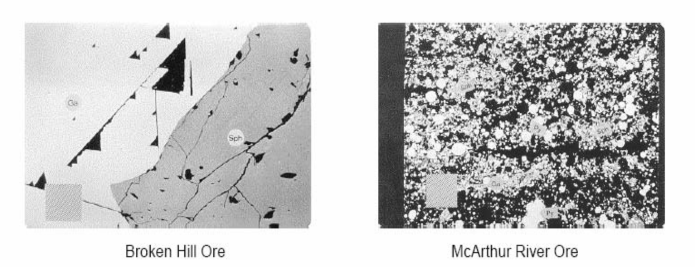
1. Senior Process Engineer, Xstrata Technology, L4, 307 Queen Street, Brisbane 4000, Qld, Australia
2. Business Manager – Mineral Processing, Xstrata Technology, L4, 307 Queen Street, Brisbane 4000, Qld, Australia

INTRODUCTION and BACKGROUND

The development of IsaMill™ technology was driven by the metallurgical requirements of Lead/Zinc deposits at Mount Isa in Queensland and McArthur River in the Northern Territory, both of which were controlled by Mount Isa Mines Limited (now Xstrata).

The McArthur River deposit was discovered in 1955 but, despite the efforts of numerous mining companies, an economic method for treatment of the fine grained deposit to produce saleable Pb/Zn concentrates was not achieved in 25 years of investigations (Enderle et al, 1997; Pease et al, 2006). In 1989 it was determined that regrinding down to 80% passing 7 µm was necessary to achieve sufficient non sulphide gangue liberation to allow the production of a bulk concentrate. Figure 1 is a comparison of the relative grain sizes of McArthur River and Broken Hill ore and illustrates the complexity of the mineralogy at McArthur River.

Figure 1: Comparison of McArthur River and Broken Hill Ore Grain Size (Grey square is 40 µm)



Broken Hill Ore

McArthur River Ore

In the case of Mount Isa, there was a gradual decrease in plant metallurgical performance from the mid 1980's as a result of decreasing liberation size and increased amounts of refractory pyrite in the ore. Concentrate grade targets were reduced to maintain zinc recovery, however plant performance continued to deteriorate to such an extent that by the early 1990's the zinc recovery had decreased from 70% to 50% (Young et al, 1997; Pease et al, 2005; Pease et al, 2006).

Significant work was conducted at Mt Isa on projects investigating finer regrinding using conventional ball and tower mill technology however the power consumptions necessary to achieve the required fine liberation sizes made them uneconomic. Further, the high rate of steel media consumption contaminated the mineral surfaces with iron, resulting in poor flotation response post regrinding.

A real need had arisen for a technology that could grind to ultrafine sizes in metallurgical operations economically and without serious contamination of mineral surfaces and pulp chemistry. However in 1990, there was no generally accepted technology for regrinding economically to such sizes in base metals. So testwork was undertaken at the time into high speed horizontal stirred mill technology, which was used in pigment and other industries. It was shown that such mills could grind down to the ultrafine sizes required for mineral liberation.

Arising from these findings, a program of major mechanical modification of horizontal stirred mill technology was undertaken between Mount Isa Mines Limited and NETZSCH-Feinmahltechnik GmbH (Enderle et al, 1997), the manufacturer of stirred milling technology used for other industries.

After many prototypes of increasing capacities, the first full scale model was developed and installed at the Mount Isa Mines' Lead Zinc Concentrator in 1994. The mill, the M3000 IsaMill™, was quickly installed in other circuits at this concentrator, and was installed in the McArthur River Concentrator in 1995 (Johnson et al, 1998). Later, in 1999, it was commercialised and sold outside of the Xstrata group.

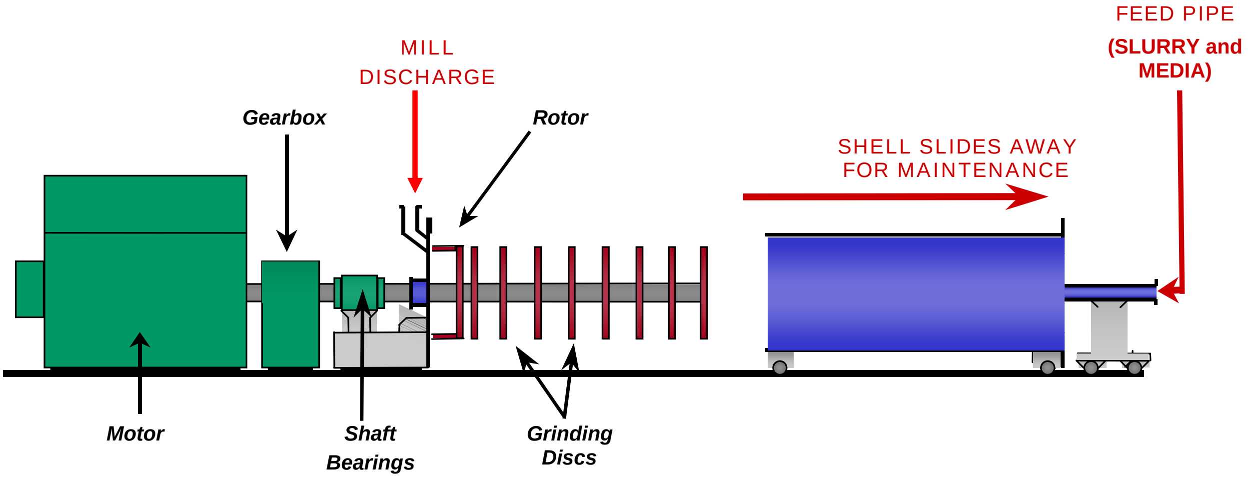
Since commercialisation of the IsaMill™, there are now over 50MW of installed IsaMills™ operating around the world, treating materials including copper/gold, lead/zinc and platinum. While the early installations treated only ultrafine sizes, the current mill installations are treating courser sized materials, once the domain of tower and ball mills. The need for energy efficient grinding circuits will only result in more IsaMill™ circuits being applied in the future.

IsaMill™ OPERATION

Grinding Mechanism

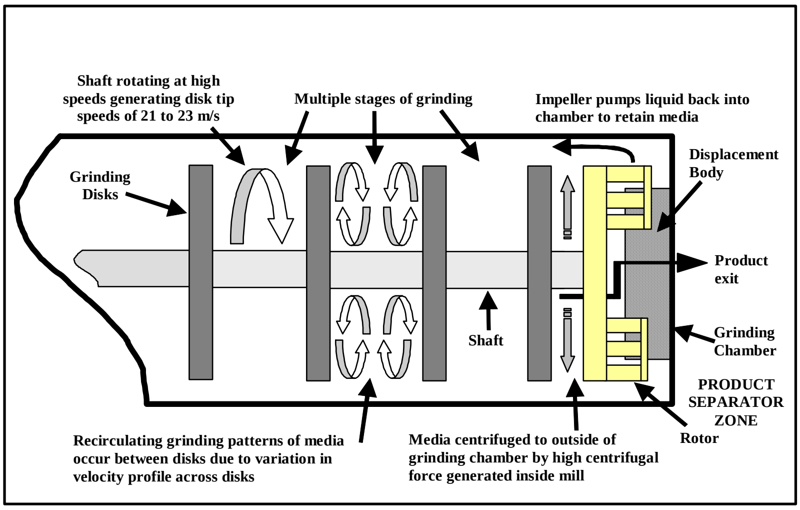
The IsaMill™ is a horizontally stirred mill consisting of a series of 8 discs rotating around a shaft driven through a motor and gearbox. The discs operate at tip speeds of 21-23m/s resulting in high energy intensities of up to 300kW/m3. Figure 2 illustrates the layout of the IsaMill™.

Figure 2: IsaMill™ Layout



The mill is filled with a suitable grinding media and the area between each disc is essentially an individual grinding chamber. As a result the mill is effectively 8 grinding chambers in series. The media is set in motion by the action of the grinding discs which radially accelerate the media towards the shell. Between the discs, where the media is not as subject to the high outwards acceleration of the disc face, the media is forced back in towards the shaft – creating a circulation of media between each set of discs. Minerals are ground as a result of the agitated media, the predominant mechanism being attrition grinding. The mechanism is best illustrated in Figure 3.

Figure 3: IsaMill Grinding Mechanism



As a result of having 8 chambers in series, short circuiting of mill feed to the discharge is virtually impossible. There is a very high probability of media-particle collision as a result of the high energy intensity and the 8 chambers in series.

Media

The key to the efficiency of the IsaMill™ is the ability to use fine media. While tower mills are typically limited to 10-12mm fresh media sizing, the IsaMill™ can use media as small as 1mm. This results in significantly more surface area per unit volume of media in the IsaMill™ than in a Tower Mill – a 2mm charge has 90 times more particles per unit volume compared to 12mm media. As a result, there is a significantly higher chance of media-particle collision, particularly at fine sizes.

The IsaMill™ is able to use a range of media types. Typically, low cost, locally available media such as sand or smelter slag have been used, which provide good grinding performance at acceptable energy efficiency. However, the need for improved energy efficiency at many installations has resulted in the use of high quality, high density ceramics, designed specifically for stirred milling applications, such as Maggotteaux's Keramax MT1.

Media Retention

Grinding media is retained in the mill without the need for screens, which is why IsaMills™ can use fine media. At the end of the mill is a patented product separator consisting of a rotor and displacement body (refer to Figure 3). The close distance between the last disc and the rotor disc centrifuges any coarse particles towards the outside of the mill. Ground product flows into the rotor area where it is essentially pumped back towards the feed end of the mill. This pumping action retains the media in the mill. The balance of the product (equivalent to the feed flowrate) exits the mill through the displacement body. This unique mechanism means that screens or cyclones are not required to retain media in the IsaMill™ which can therefore be operated in open circuit without cyclones, which simplifies the circuit, and reduces capital.

Energy Intensity

The high tip speed of the IsaMill™ results in a high energy intensive environment. Energy intensity of the IsaMill™ is significantly higher than any other commercially available grinding equipment as illustrated in Table 1. Combining the energy intensity and the high grinding efficiency leads to a compact mill, able to be fitted into existing plants where floor space is limited. On the other hand, Tower Mills require a settlement zone at the top to separate the media from the slurry – this limits the agitation speed to a tip speed of 3m/s (compared to IsaMill™ at 21-23m/s) and therefore limits the energy intensity, while ball mills can only have a relatively low amount of ball loading before media empties from the mill.

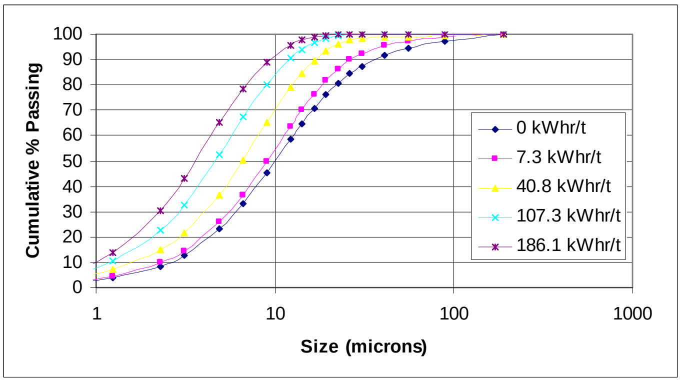
Table 1: Comparative Energy Intensity of Grinding Technologies

	Installed Power (kW)	Mill Volume (m^3)	Power Intensity (kW/m^3)
Autogenous Mill	6400	353	18
Ball Mill	2600	126	21
Regrind Mill	740	39	19
Tower Mill	1000	12	42
IsaMill™	3000	10	300

Product Size Distribution

In open circuit operation, the IsaMill™ is able to produce a sharp product size distribution. Typically the ratio of the P98 to the P80 is around 2.5. This is a direct result of the effect of 8 chambers in series preventing short circuiting and the classification action of the product separator. The ability to operate the mill in open circuit greatly simplifies the operating and maintenance strategies of the circuit. Figure 4 illustrates a typical IsaMill™ product distribution in open circuit at varying power inputs – note the steepness of the curve and the lack of ultra fines that would be expected from a Tower Mill distribution.

Figure 4: Typical IsaMill™ Product Distribution in Open Circuit (South American pyritic gold concentrate)



Inert Grinding

After the initial comminution stages of crushing and/or SAG/AG milling, the following grinding stages is usually carried out using steel charged ball or tower mills. The impact of grinding using steel media can offset any benefits gained by improved liberation, particularly as the target size decreases below 25 µm. Grinding in a steel environment results in the precipitation of metal and iron hydroxides on to the surface of ground particles. These conditions affect flotability, flotation selectivity and lead to higher reagent consumptions to overcome the surface coatings and regain recovery (Trahar, 1984; Pease et al, 2006). The benefits of inert grinding at several locations have been well reported (Pease et al, 2006, 2005, 2004; Young et al, 1997; Grano et al, 1994).

While the negative impacts of steel grinding will be greatest at fine sizings due to the large surface areas and high media consumptions involved, inert grinding has also been shown to produce benefits at coarser sizings (Grano et al, 1994; Greet et al, 2004; Pease et al, 2006). For a long time, chrome media has been offered to, and investigated by, ball and tower mill operators as a means of improving pulp flotation chemistry by reducing the amount of iron released into the grinding pulp and contaminating freshly ground surfaces. Greet and Steiner, 2004, analysed the surface of galena ground in three different environments for the presence of iron. It is clear from Table 2 that while grinding in a high chrome environment reduced the surface iron composition from 16.6% to 10.2%, grinding in a ceramic environment reduced the detectable surface iron to less than 0.1% - a significant improvement over both media types.

Table 2: Composition determined via XPS, of the unetched surfaces of Rapid Bay Galena ground with different media (Greet 2004)

Media Type	**Surface Atomic Composition (%)**			
	O	**Pb**	**Fe**	**S**
Mild Steel	53.1	15.6	16.6	14.7
High Chrome	50.0	20.6	10.2	19.2
Ceramic	33.6	32.0	<0.1	34.4

IsaMill™ RECENT DEVELOPMENTS

Until recently, inert grinding to coarser sizes (+25µm) was generally impractical in traditional milling equipment due to the high capital investment required to fill the mill. However two recent developments in IsaMill™ technology have allowed coarse grinding in an inert environment to become a reality, namely MT1 ceramic media and the scale up of the IsaMill™ to the M10,000 model.

MT1 Ceramic Media

The development of high intensity stirred mills, such as the IsaMill™, increased energy efficiency by the use of higher agitation speeds and finer media, compared to those associated with tower mills or ball mills. (Gao et al, 1999, 2002). The IsaMill™ has been able to use low cost, cheap media, such as sand, discarded slag, river pebbles, scats etc to obtain very fine grind sizes. However while the mills have been run with low cost media, the low quality media limits the energy efficiency of the mill. Quality issues such as particle shape, grain size, specific gravity etc all constrain the mills energy efficiency as well as the size of feed that can be milled. It is this latter limitation that has constrained the mill to grinding in fine size ranges. Previously available ceramics were of variable quality, high cost, high wearing and low SG which generally made their use uneconomic.

A recent initiative between Magotteaux International and Xstrata Technology brought about the development of a ceramic specifically designed for use in the IsaMill™ to address the limitations of current media. The product, Keramax MT1, was designed to address all the key characteristics for the ideal media, listed by Lichter and Davey, 2002, as:

- Hardness
- High sphericity
- High roundness
- Mechanical Integrity
- SG
- Definite initial charge PSD and top up size
- Chemical composition

Keramax MT1 is a high density, high quality alumina ceramic media. It has been shown to have a consistent hardness when sectioned with no air bubbles or structural defects (Curry et al, 2005). It has been specifically designed for grinding applications, with relatively high SG and low wear characteristics. Keramax MT1 properties are listed in Table 3.

Table 3: Keramax MT1 Properties

Keramax MT1	**Properties**
Composition	79% Al_2O_3 6.5% SiO_2 14.0% ZrO_2
Hardness	1300-1400 HV
Fracture Toughness	5-6
Specific Gravity	3.7
Bulk Density	2.3-2.4

The surface of the media is smooth and ‘pearl-like’ to touch. These surface properties mean that the energy loss in grinding due to friction is minimized, improving the efficiency of grinding. At the same time there is negligible contamination of ground material with deleterious ions.

The true benefits of this material can be seen when it is compared to other media in a lab environment. This was undertaken by Xstrata Technology and Magotteaux separately in 2004 on gold bearing concentrate from the Eastern Gold fields region of WA, using a M4 IsaMill™ (Curry et al, 2005). The results are summarized in Table 4.

Table 4: Media Type vs Energy, Consumption and Net Power for 250t/hr Treatment Rate

Media Type	Consumtion Rate (g/kWhr)	Specific Energy (kWhr/t)	Consumption Rate (Kg/t)	Net Power (MW) for 250 t/hr example
MT1 (-4 +3mm)	15	7.6	0.11	1.9
Alumina 1 (-4 +3mm)	128	13.1	1.68	3.3
Alumina 2 (-4 +3mm)	295	12.4	3.66	3.1
Australian River Pebble (-4 +3mm)	200	27.9	5.58	7.0
Australian Silica Sand (-6 +3mm)	781	11.2	8.77	2.8
Ni Slag (-4 +1mm)	1305	17.8	23.23	4.4

Keramax MT1 was found to be more efficient than the other ceramics or sands tested. This has big implications to grinding circuit design, as it requires less installed power for an application on MT1 compared a to much larger installed power that would be required if sand was used. This is highlighted in the final column of Table 4, where for a hypothetical throughput of 250tph, 1.9MW is required to do the grinding duty using MT1, compared to 7.0MW on River Pebble (-4mm to +3mm).

Keramax MT1 has all the attributes of inert media that has been in use in IsaMills™ over the last decade, but has the toughness and high SG to make it a more efficient and harder wearing media than slags and sand, while at the same time being able to handle coarser feed size distributions.

M10,000 IsaMill™

In late 2002, Xstrata Technology, in collaboration with Netzsch and Anglo Platinum, developed the M10,000 IsaMill™ ; the worlds largest stirred mill. The mill was developed for Anglo's Western Limb Tailings Retreatment Project near Rustenburg in South Africa. The project involved the retreatment of dormant platinum containing tailings, which represented a major economic resource. Fine grinding testwork had established that the PGMs could be recovered from the tailings. Pilot scale work resulted in the circuit needing to produce a primary grind, no less than 80% passing 75 µm, and a rougher concentrate regrind of no less than 85% passing 25 µm, which would enable good recovery of PGM's. The tonnage required for the regrind circuit was 53tph (max 65tph), to produce a P90 of 25 µm, requiring 35kWh/t. This duty could not be performed in a single M3000 IsaMill™. To maximize the economics of the project, a larger mill was required leading to the development of the 2.6MW, M10,000 IsaMill™, (Curry et al 2005).

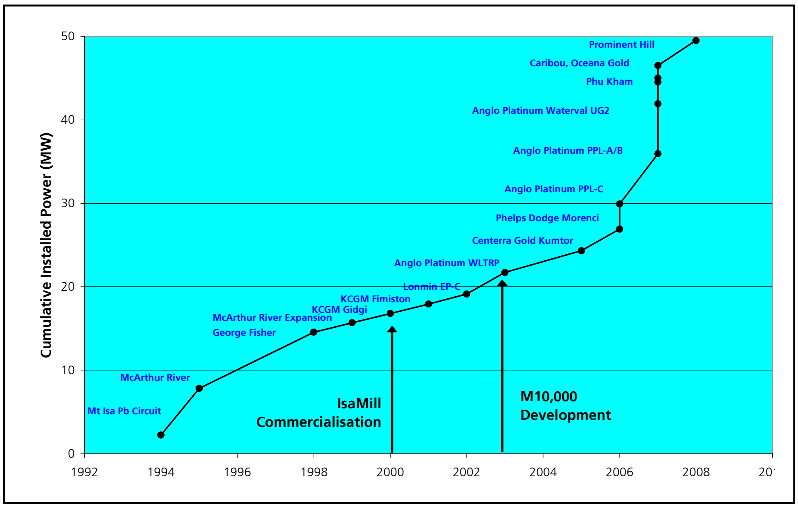
Results from the Western Limb Tailings Retreatment Project have confirmed the accuracy of the scale up from the lab scale M4 IsaMill™ and pilot plant IsaMill™ testwork to the full scale unit, Table 5.

Table 5 Laboratory versus Full Scale Grinding Efficiency

IsaMill™ Model	Installed Power (kW)	Chamber Volume	Specific Energy (kWh/t)	Pulp Solids	P98 (µm)	P98 (µm)
M4	4	3.5	37	39	47.5	16.0
M10,000	2,600	10,000	37	42	42.5	16.5

Since Anglo's Western Limb Tailings Retreatment Project, there have been 9 other M10,000 IsaMills ™ either installed or designed at the time of writing. They are either installed with 2.6MW or 3.0MW motors, and all use ceramic media. The rapid growth of IsaMill™ technology is shown in Figure 5.

Figure 5 – Time versus Cumulative Power for IsaMill™ Installations



Today, large scale coarse grinding using inert media is both practical and economic, due to the development of large scale IsaMill™ technology and MT1 ceramic media. The use of the IsaMill™ with inert media in coarse grinding applications not only has significantly improved energy efficiency, but also has improved downstream separation processes due to the production of minerals with clean, iron free surfaces.

The following Case Studies describe IsaMill™ installations being applied to coarse grinding duties.

CASE STUDY 1 - KUMTOR M10,000 INSTALLATION

The M10,000 IsaMill™ installation at the Kumtor gold mine in the Kyrgyz Republic, is one of the largest gold mines operating in Central Asia, producing over 500,000 ounces of gold per annum. It is based 4000m above sea level in the Tien Shan Mountains, and is owned by the Kumtor Operating Company, a fully-owned subsidiary of the Canadian company Centerra Gold Inc.

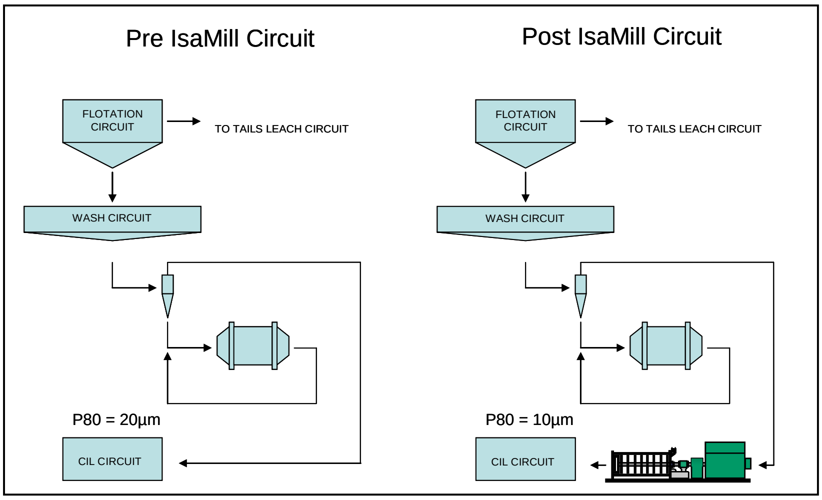
The Kumtor installation treats a hard refractory orebody of finely disseminated gold linked with significant quantities of pyrite. The flowsheet is conventional crushing, SAG and ball mill circuits, before the ball mill cyclone overflow is floated. The tailings are thickened and then undergo CIL before passing to the tails disposal circuit. The concentrate is also thickened, and is then reground in a conventional ball mill, prior to CIL, before stripping, electro winning and refining to doré. The regrind ball mill, operates in closed circuit with 25 mm diameter balls and very high recirculation loads (~600 %), to produce a product with a P80 of ~20 µm. Liberation studies undertaken by Kumtor established the gold recovery benefits of grinding finer, with an optimal P80 sizing of 10 µm selected for the design.

In late 2005, a 2.6MW M10,000 IsaMill™ was commissioned at Kumtor, treating the regrind ball mill discharge to a product sizing of 80% passing ~10 µm. Figure 6 illustrates the before and after flowsheets. This installation represented the first commercial application of the Magotteaux MT1 media. The decision to use the MT1 media was based on several key factors:

- Low wear rates resulted in less media required to be transported to the remote site.
- High SG ceramic media translates into more efficient power usage compared to sand, enabling 1 x M10,000 IsaMill™ operating in open circuit to do the work equivalent to 2 x M3000 IsaMills™ operating on sand, reducing capital cost and simplifying the circuit.
- Cyanide consumption would be reduced

Soon after commissioning the IsaMill™ was treating on average 72 tph, compared to the design tonnage of 65 tph. The current power draw was approximately 1950 KW, which equates to 23 KWhr/t specific grinding energy. Shortly after commissioning, metallurgical results had seen a drop in the gold tailings grade by 30%, (which represents an extra 20,000 ounces of gold per annum being recovered).

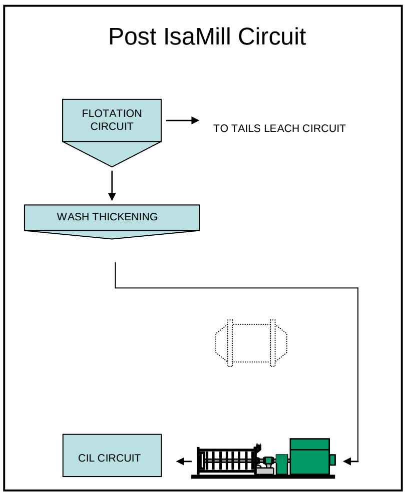
Figure 6: Pre and Post IsaMill Circuits at Kumtor



Replacement of Kumtor Ball Mill with IsaMill™ During Ball Mill Maintenance

While the Kumtor IsaMill™ is designed as an ultra fine grinding application, it was noted by Kumtor management that the IsaMill™ would be required from time to time to operate without the ball mill, when the ball mill required maintenance, Figure 7. This would allow the Kumtor operation to continue without any lost production from the maintenance. In this operation the mill would be expected to produce a coarser product, although it was hoped to match the ball mill discharge.

Figure 7: Circuit Configuration during Ball Mill Maintenance



Testwork was conducted on samples of rougher concentrate, which predicted that a coarse charge of ceramic media could produce a similar output to the ball mill, i.e. a P80 of 20 to 25um.

In March 2006, shortly after the IsaMill™ was installed for fine grind duty, the ball mill was relined for a 4 week campaign, and the IsaMill™ feed was coarsened from 20 um to 150um. However during the relining of the ball mill, a much finer ball charge was actually used in the IsaMill™ than used during the testwork, while the IsaMill™ was also operated with a non optimized feed density and rotor selection. The 4 week timeframe did not permit these key variables to be optimized, which, along with the fine media, impacted on the power draw, which was only 1885 KW (average) throughout the trial.

The key findings of the trial were

- IsaMill™ has reduced the feed from F80 of 130µm to 150µm to P80 of 60µm to 65µm
- Size reduction has been undertaken by finer media than used in the testwork
- Media consumption has been good at 17.5g/KWhr
- IsaMill™ operated continuously during the 4 week ball mill shutdown enabling full scale production to be achieved, with no breakdowns
- No significant wear issues were noticed during the trial, with the same mill liner and disc used throughout and after the trial
- Coarser media, optimal density and the use of a high flow rotor could have improved the power draw and further reduced the discharge sizings

CASE STUDY 2 - NEW INSTALLATIONS USING IsaMill™ in COARSE DUTIES

Phu Kham Project

The Phu Kham deposit is located approximately 100km north of the Laos capital Vientiane. It is owned by Pan Australian, an Australian listed mining company. The Phu Kham deposit hosts two distinct styles of mineralisation: an oxide gold cap and beneath this transitional/primary copper-gold. The Phu Kham oxide gold cap is the principal deposit for the Phu Bia heap leach gold mine, the first phase of the development of the Phu Kham deposit, which entered into production in 2005. The Phu Kham Copper-Gold operation is planned for start-up in mid 2008. Feed to the concentrator will consist of 12MT on average, with planned annual output from this mine being over 200,000 dry metric tonnes (dmt) of concentrate (grading 25% copper), containing 50,000 tonnes copper, 40,000 ounces gold and 400,000 ounces silver, (on average). The concentrate will be exported for further treatment and refining by custom smelters in the Asia Pacific region.

Process technology employed for Phu Kham Copper-Gold is conventional comminution at the head of the circuit, followed by flotation to produce a copper-precious metal concentrate, (Pan Australian, 2006)

Rougher concentrate will be treated through a M10,000 IsaMill™, powered by a 2.6MW motor, treating approximately 168 tph and reducing the feed size from a F80 of 106um to a P80um of 38um, before further flotation. The grinding media for the operation will be MT1.

Prominent Hill Project

Oxiana Limited owns 100% of the Prominent Hill copper-gold project located 650 kilometers north west of Adelaide, and130 kilometers north west of BHP Billiton's Olympic Dam in South Australia, Australia.

The ore body consist of copper gold breccia, and will be mined via an open pit. The ore will be treated through a conventional grinding and flotation processing plant, with a designed capacity of 8MTPa. The initial planned concentrate production will be on average 187,000 dry t/a peaking at 230,000t in 2009, with average concentrate grades of 45% copper, 19g/t gold, 57g/t silver. The high grade concentrate will be sold to smelters in Australia and in Asia. (Oxiana, 2007)

One M10,000 IsaMill™, powered by a 3.0MW motor, has been selected to treat the rougher concentrate. It will treat approximately 138 tph, reducing the feed size from a F80 of 125um to P80um of 24um for further flotation. The planned commissioning of the mill will be mid to late 2008. The grinding media for the operation will be MT1

Anglo Platinum Installations

Anglo Platinum installed the first M10,000 in 2003, in their Western Limb Tailings Retreatment Project, after working with Xstrata Technology to scale up the existing M3000 IsaMill™. The resulting mill, the M10000, was a powered by a 2.6MW motor, which had a variable speed drive (this was a precautionary measure, as it was the first M10,000 to be built).

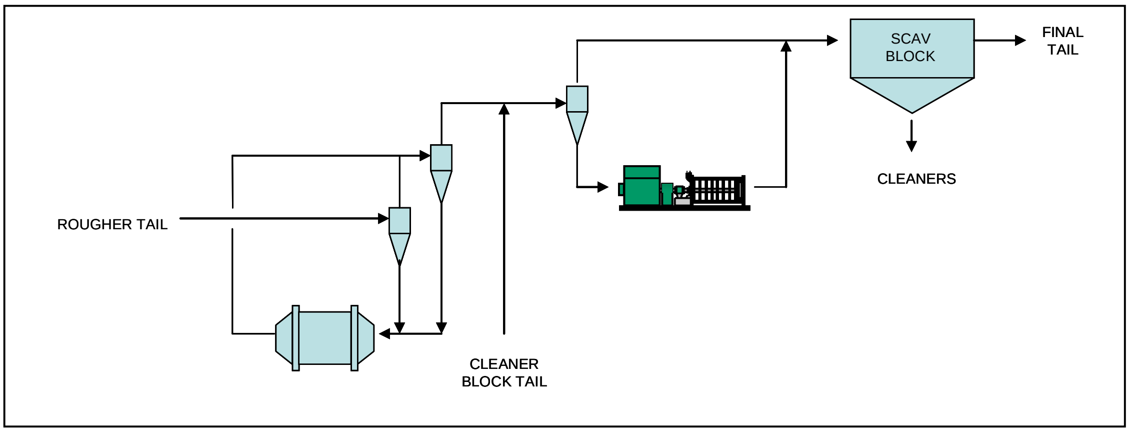
The mill treats oxidized PGM's from a tailings dam in a precleaner tail stream, and reduces feed from 75um to 25um.

Four years later, in 2007, Anglo Platinum has ordered another five, M10,000 IsaMills™. This time the mills come with fixed speed drives, and are powered by 3MW motors. The high energy efficiency demanded by the sites have called for ceramic media, and the duty of the mills will be similar, i.e. typical duty from 75-100 um feed size down to 53 um product size.

To date, the first IsaMill™ has been successfully commissioned at Potgietersrust Platinum mine (C-Section), and there are mills being manufactured/installed at Potgietersrust A and B Sections (2 mills), followed by two more at the Rustenburg Watervaal UG2 operation in late 2007.

The Potgietersrust Platinum mine (C-Section) mill is designed to operate with a 3MW motor and use MT1 media, treating scats from A and B section primary milling circuits, with the ore having a Bond Work Index, BWi, over 30 kWh/t. Figure 8 illustrates the simplified C section flowsheet with an IsaMill™.

Figure 8: Simplified PPL C Section Flowsheet with a M10,000 IsaMill™



The rougher tail is milled in closed circuit with a ball mill. Cyclone overflow, at a nominal P80 of 75 µm, proceeds to the IsaMill™ predensifying cyclone which increases the percent solids for efficient IsaMill™ operation, while sending the fines to the scavenger flotation circuit. The IsaMill™ then treats the predensifying cyclone underflow, before the product enters the scavenger flotation circuit.

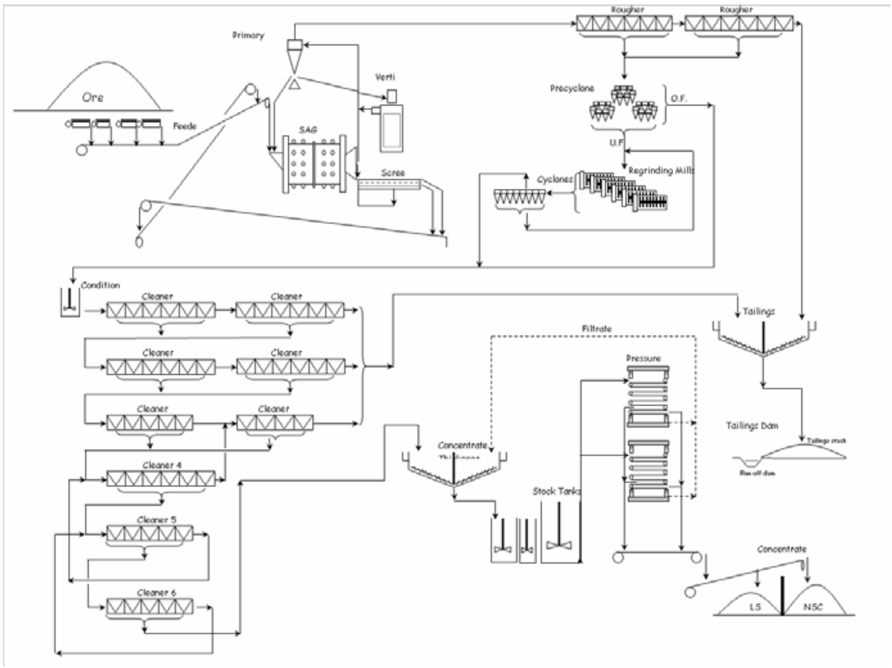
The IsaMill™ circuit is based on testwork which found very favourable flotation results after IsaMilling™ the cyclone overflow from an F80 of 75 µm down to a product sizing P80 of 53 µm. The circuit is designed to treat 162t/hr (nominally) through the IsaMill™, which is then rejoined with the predensifying cyclone overflow, before it passes to the scavenger flotation circuit. Design energy consumption grinding from F80 75µm to P80 53µm is 9 kWhr/t. The IsaMill will operate with top size 3.5mm MT1 ceramic media.

Before the implementation of the IsaMill™, it was calculated that a 8MW conventional ball mill operated in closed circuit, would be needed to do this duty. However with the M10,000 IsaMill™, less than half the power is required due to open circuit operation and the use of ceramic.

CASE STUDY 3 – IsaMills™ in McARTHUR RIVER SAG CIRCUIT

McArthur River Mine (MRM) is part of Xstrata Zinc, an operating subsidiary of Xstrata PLC. It is a zinc/lead mine, operating in Northern Territory, Australia, and was commissioned in 1995. The development of the IsaMill™ for regrinding down to a P80 of 7 µm was the enabling technology that allowed the mine to be developed. Initially there were 4 x M3000, 1.1MW IsaMills™ in the regrind duty. This has since been expanded to 6 with a combined installed power of 6.7MW. The current plant flowsheet is shown in Figure 9.

Figure 9: McArthur River Flowsheet

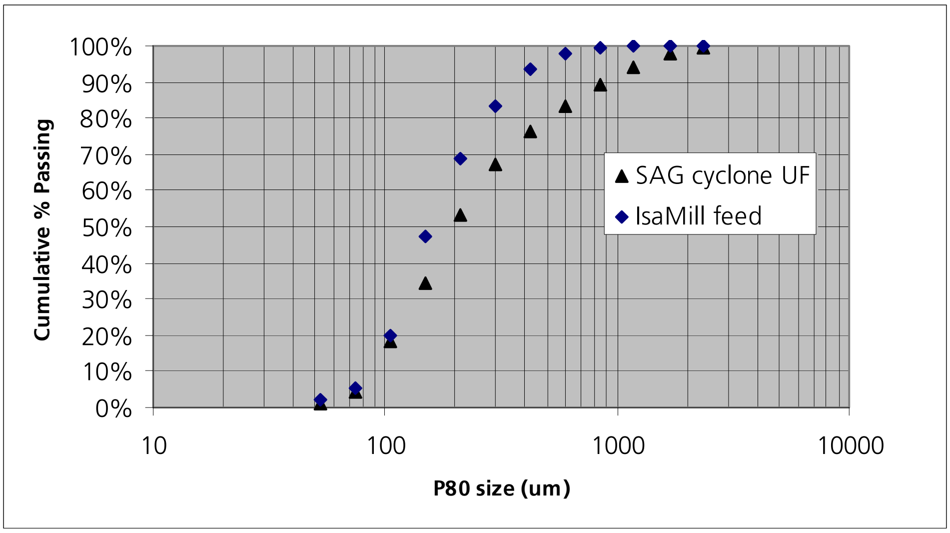


The grinding circuit consists of a primary SAG mill, closed by a double deck screen and cyclones. All screen oversize is returned to the SAG Mill. Cyclone overflow (currently at P80 of 70µm) feeds flotation, while cyclone underflow is split between a Tower Mill or returned to the SAG Mill. The Tower Mill product is returned to the SAG sump where it is pumped with the SAG screen underflow to the primary cyclones.

MRM have a need to increase milling capacity to account for decreased head grades as the operation shifts from underground to open cut. At the same time there was a desire to reduce downtime and reduce operating cost by eliminating the Tower Mill from the circuit, hence MRM have been keen to explore the effectiveness of a M10,000 IsaMill™ in the primary grinding circuit. In the primary grinding circuit, the IsaMill™ will be treating material of the order of 300 to 350um, the coarsest any IsaMill™ has been designed for. The IsaMill™ in the primary grinding circuit will be operating in open circuit, producing a P80 of 40µm product that could be pumped directly to flotation, with the SAG cyclone overflow. The Tower Mill will be decommissioned.

Testwork has been undertaken using a M4 and M20 IsaMills™ and reported by Anderson and Burford (2006). The findings from this work indicated that while the IsaMill™ could treat the screened cyclone underflow feed at 623um, the presence of SAG mill scats also in the underflow caused blockages of the small scale mill. The feed to the IsaMill™ can be summarised by the Figure 10.

Figure 10: McArthur River SAG Cyclone Underflow and IsaMill M20 Feed Distribution

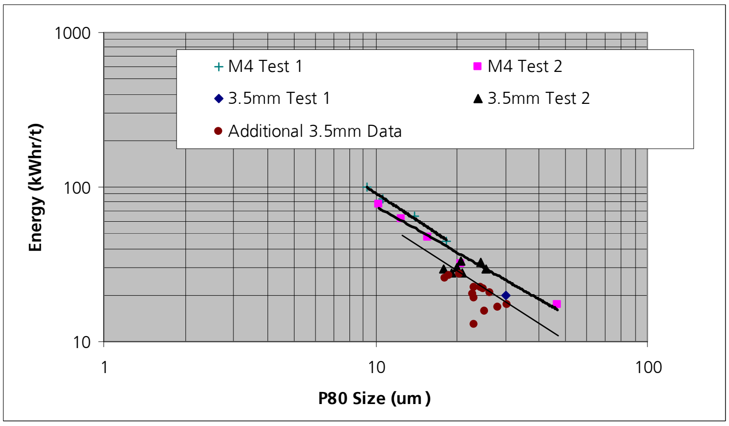


The average results from the two site test conducted with the M20 IsaMill™ indicted that the mill could reduce an average feed size of F80 of 350um, to a product with a P80 of 20 to 30 um, using MT1 as media (3.5mm in diameter). The testwork also indicated that an energy input of 10-15kWhr/t was required to reduce the feed, screened at a top size of 623µm, to a product P80 of 40µm. The data is shown in Figure 11, marked as "3.5mm Test 1" and "3.5mm Test 2".

Results were also presented on Figure 11 showing two trials that were also conducted using the 4 litre laboratory IsaMill™, ("M4 test 1" was conducted with unscreened SAG cyclone underflow; "M4 test 2" was conducted with SAG cyclone underflow top screened at 1.7mm).

Data from the M20 site tests shows higher efficiency for a given target size compared to the M4 IsaMill™. Some of the increased efficiency may have been due to the decreased feed size distribution of the site test, but it is unknown what is the impact of this difference, if any. Also the variability and accuracy of the data is affected by the fact that the mill was filling up with steel during the trials, which indicated further testwork was required to confirm the initial results...

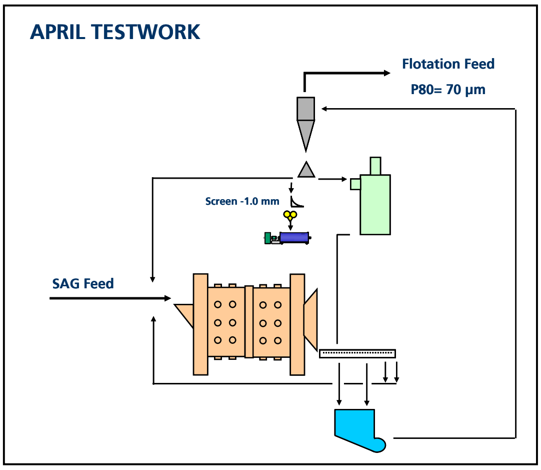
Figure 11: Product Size - Energy Relationship for 3.5mm MT1



Continuing Testwork

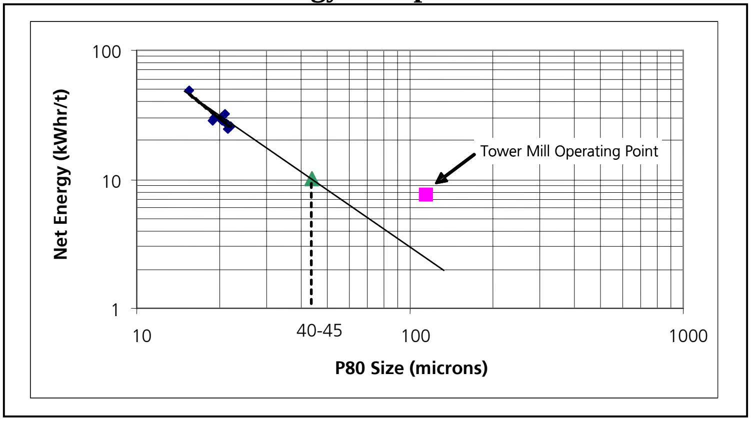
Further testwork was conducted later in 2006 to support the initial testwork, as well as overcome the presence of scats in the feed stream that lead to blockages in the M20 IsaMill™. Figure 12 displays the flowsheet that was used in this testwork that incorporated a magnetic separator to remove steel scats in the cyclone underflow. Aso the feed was screened this time at 1mm (previous screening was at 623um).

Figure 12: Site Testwork at MRM Using M20 IsaMill™ with a Magnetic Separator on Feed



The M20 IsaMill™ was able to treat material, that was slightly finer than the previous work, from a feed sizing of 300um, down to a product sizing of 20 to 25 um, (finer than the 40um target). The data was able to permit a size energy relationship to be established, as shown in Figure 13, compared with the current Tower Mill operation in that circuit.

Figure 13: Size versus Net Energy Comparison for IsaMill™ and Tower Mill



Using the energy data from the M20 IsaMill™ testwork, and the current energy use for the Tower Mill in the primary circuit, it has been conservatively estimated that the IsaMill™ could do the same duty as the Tower Mill for a third of the energy, a saving of 5KWhr/T less energy, i.e. Tower Mill uses 7 to 8 KWHr/T for a P80 of 100um, while the IsaMill™ achieves same sizing for 2 to 3 KWHr/T.

McArthur River Mine Future Circuit

McArthur River Mine has included two M10,000 IsaMills™ in the upgrade of the concentrator to cope with the increased tonnage as the mine goes from underground to open cut. As at March 2007, the concentrator has been planned to increase throughput from 1.8MTpa to 2.5MTPa, with a resulting increase in concentrate production from 320,000T of zinc lead concentrate to 430,000T of zinc lead concentrate. It is planned that the circuit will be ready for commissioning in mid 2008, (Xstrata Zinc, 2007). The IsaMills™ will take 1 year to be manufactured before they are delivered to site. They will take between 1 to 2 weeks to be installed.

McArthur River Mine managements' selection of IsaMills™ technology for their upgrade was due to the higher efficiency of the mills compared to existing technology. This overcome the need to increase the number of power generators at the remote mine site, while the small footprint of the mills enabled them to fit into the existing concentrator.

Final design parameters are still to be finalised at the time of writing, but the new circuit will have the cyclone underflow from the SAG mill reporting to magnetic separators and oversize screens to ensure all iron scats are taken from the stream before treatment by the two M10,000 IsaMills™.

Further testwork will also be undertaken during 2007 to trial one of the M3000 IsaMills™ in the fine grinding circuit to undertake the coarser duty. The first trial is planned for May, 2007.

CONCLUSION

IsaMill™ technology is becoming the preferred technology in efficient coarse grinding circuits. The development of reliable ceramic media, such as MT1, as well as the development of high capacity M10,000 IsaMills™, have lead to the IsaMill™ being a realistic alternative in coarse grinding applications.

IsaMills™ in coarse grinding applications have all the advantages of fine grinding applications, such as simple circuit design, small footprint and iron free contamination when used with ceramic media, which has profound beneficial impact on metallurgy, recirculating loads and reagent use. However the biggest advantage of IsaMills™ is its high energy efficiency, as a result of its high speed stirring action in a packed bed.

With the pressure being applied to all industries today for improved sustainability and the need for increased energy efficiency, the adoption of IsaMill technology in coarse applications is good news for operators, as the development of the IsaMills™ allows for significant energy and capital savings for the application.

IsaMill™ technology is a true alternative to ball mill and tower mill circuits.

REFERENCES

1. Curry, D.C., Clark, L.W., Rule, C., 2005
Collaborative Technology Development – Design and Operation of the World's Largest Stirred Mill. Randol Innovative Metallurgy Conference, Perth, Australia, August.

2. Curry, D.C. and Clermont, B., 2005
Improving The Efficiency Of Fine Grinding – Developments in Ceramic Media Technology. Randol Innovative Metallurgy Conference, Perth, Australia, August.

3. Enderle, U; Woodall, P; Duffy, M and Johnson, N W, 1997.
Stirred Mill Technology for Regrinding McArthur River and Mount Isa Lead/Zinc Ores. XX International Mineral Processing Conference, Aachen, Germany, September 21-26, Vol 2, pp 71-78.

4. Gao, M., Weller, K.R. and Allum, P., 1999. Scaling-Up Horizontal Stirred Mills from a 4-Litre Test Mill to a 4000-Litre "IsaMill". Powder Technology Symposium, Pennsylvania State University, Pennsylvania, USA, September.

5. Gao, M., Young, M.F. and Allum, P., 2002. IsaMill Fine Grinding Technology and its Industrial Applications at Mount Isa Mines. 34th Annual Meeting of the Canadian Mineral Processors, Ottawa, Canada, January.

6. Grano S.R., Wong, P., Skinner, W., Johnson, N.W. and Ralston, J., 1994. The effect of autogenous and ball mill grinding on the chemical environment and flotation of the copper ore of Mount Isa Mines Limited. In: III Latin-American Congress on Froth Flotation (University of Concepcion Publ.), Concepcion, Chile, pp. 351-388.

7. Greet, C J., Steiner, P., 2004. Grinding – The Primary Conditioner, Metallurgical Plant Design and Operating Strategies. AUSIMM, Perth Australia, pp 319-336.

8. Johnson, N W; Gao, M; Young, M F and Cronin, B, 1998. Application of the IsaMill (A Horizontal Stirred Mill) to the Lead-Zinc Concentrator (Mount Isa Mines Ltd) and the Mining Cycle. AusIMM Annual Conference, Mt Isa, Australia, 19-23 April.

9. Kazakoff, J; Mortimer, A; Smith, S and Curry, D, 2006. Introduction of IsaMill Grinding Technology into a Major Gold Flotation and Leaching Operation in Central Asia, Comminution 06, Perth, Western Australia.

10. Lichter, J and Davey, G, 2002. Selection and Sizing of Ultrafine and Stirred Grinding Mills, SME Mineral Processing Plant Design Symposium, Vancouver, Canada

11. Pease, J D; Curry, D C; Barns K E; Young M F and Rule, C, 2006. Transforming Flowsheet Design with Inert Grinding – The IsaMill, 38th Annual Canadian Mineral Processors Conference, Ottawa Canada.

12. Pease, J D; Curry, D C and Young, M F, 2005. Designing Flotation Circuits for High Fines Recovery, AusIMM Centenary of Flotation Conference, Brisbane, Australia, 6-9 June.

13. Young M F; Pease, J D; Johnson, N W and Munro, P D, 1997. Developments in Milling Practice at the Lead/Zinc Concentrator of Mount Isa Mines Limited from 1990, Sixth Mill Operators Conference, Madang PNG, 6-8 October.

14. Trahar, W.J, 1983. The Influence of Pulp Chemistry in Sulphide Flotation, Principles of Mineral Flotation – The Wark Sympoium, pp 117-135 (The Australian Institute of Mining and Metallurgy, Adelaide)
.

15. Casteel, K 2006 . HPGR: Making Its Case for Commercialisation, E&MJ, pp 61-64.

16. US NRC Report, 1981. Report of the US NRC Committee on Comminution and Energy Consumption.

17. Anderson, G S and Burford, B D. IsaMill-The Crossover from Ultrafine to Coarse Grinding, Metallurgical Plant Design and Operating Strategies (Metplant 2006), 18 to 19 September 2006, Perth, Western Australia.;

18 Pan Australian Quarterly Report to ASX, 31/12/2006 - www.panaustralian.com

19 Oxiana webpage, description of Prominent Hill Project, 2007 – www.oxiana.com

20 Xstrata Zinc webpage, description of MRM Open Pit Project, 2007 – www.xstrata.com

Transforming Flowsheet Design with Inert Grinding - the IsaMill

J.D.Pease[1], D.C. Curry[1], K. E. Barns[1], M.F.Young[2], C.Rule[3]

[1]Xstrata Technology
Level 2, 87 Wickham Tce
Brisbane, QLD, 4000
Australia
Ph: +61 7 3833 8555

[2] Xstrata Zinc
Mount Isa, Qld,
Australia

[3] Anglo American Platinum Corporation
55 Marshall St, Johannesburg
South Africa

Key words: grinding, flotation, inert media

ABSTRACT

The IsaMill was developed for fine-grained ores that required at least double the grinding efficiency of ball or tower milling to be economic. This was achieved, but in practice, the benefits of using inert media (eg sand, slag, ore, ceramic) have proven to be at least as important as the higher grinding efficiency. Flotation selectivity and rate was improved for all particles, particularly fines. This allowed a dramatic simplification of the Mount Isa lead zinc flotation circuit – adding 6 MW of ultrafine grinding power *reduced* reagent addition and flotation volume and increased plant energy efficiency. This was quite unexpected.

This paper examines four orebodies that were enabled by inert grinding – McArthur River, George Fisher and Black Star Open Cut Mines (complex fine-grained lead zinc orebodies in Australia), and the Western Limb Tailings Retreatment plant (a PGM operation in South Africa). In a unique case history, the operating performance of the Mt Isa lead zinc concentrator is explained by size-by-size mineralogical data collected over 25 years, systematically explaining the impact of declining ore quality, and the effects of additional conventional grinding, inert grinding, and circuit and reagent redesign.

The mineralogical case histories are so compelling that it is argued that the advantages of inert grinding should not be confined to difficult, fine-grained orebodies. The availability of large-scale efficient inert grinding mills could have profound impact on circuit design for many orebodies.

INTRODUCTION

The IsaMill was developed by MIM (now Xstrata) to enable the development of the McArthur River Orebody in the Northern Territory of Australia. This ultra-fine grained ore needed a grind size of 80% passing 7 microns to produce a saleable concentrate. Such a fine grind was not economic with conventional ball or tower milling. Below about 25 microns the low energy efficiency and high media consumption of these mills is prohibitive. Critically, the high steel consumption and associated change in pulp and surface chemistry also seriously affects flotation metallurgy.

Figure 1 demonstrates the need for stirred milling to reduce power consumption for fine grinding of a pyrite concentrate. A grind P80 of 12 microns requires over 120 kwh/t in a ball mill using 9mm balls, but only 40 kwh/t in an IsaMill with 2mm media. The curve demonstrates the difference between “conventional” grinding and fine grinding. The Bond Work Index is developed for the flat “A” section of the curve, but does not apply to the exponential increase in grinding energy after the “knee”, section “B”. Here, very small changes in target P80 mean huge increases in power requirement, which cannot be achieved with conventional mills.

The key to efficient fine grinding is fine grinding media. A charge of 2mm media has 91 times more particles than the same volume of 12mm media; the same charge of 1 mm media has 730 times more particles. More particles means more breakage events, but to build a practical mill to use fine media requires three things:

- The small media particles need sufficient momentum to quickly break small ore particles. Slow stirring speeds may still grind efficiently, but the slow breakage rates will make the mill size prohibitive.
- An economic source of fine media.

- A practical means for the mill to retain media and pass target size product.

The IsaMill achieves these needs by combining high power intensity (280 kw/m3, compared with about 20kw/m3 for a ball mill), high stirring speed (22 m/s tip speed), and a centrifugal discharge to retain the media. The screenless centrifugal discharge allows low cost media to be used – eg granulated slag, silica sand, gravel fractions of the ore.

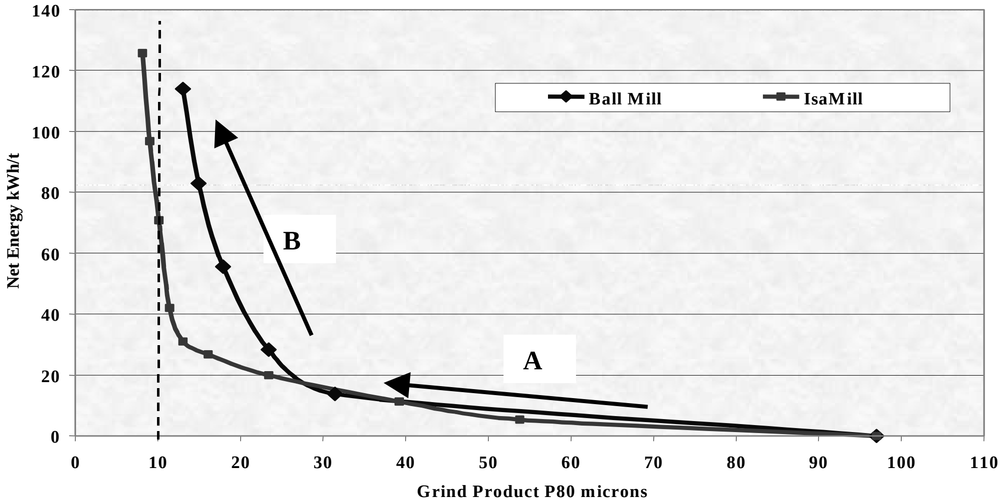


Figure 1: Grinding Energy vs Product size for a pyrite concentrate

The grinding mechanism and high power efficiency of the IsaMill is discussed elsewhere (Pease 2004). This paper will focus on another aspect of the technology that has been equally important in enabling orebodies – the impact of high intensity, inert grinding on flotation performance. Compared with steel grinding, inert grinding profoundly changes fines flotation, demolishing many common theories about fines behaviour. For example, at McArthur River, 96% of the individual particles recovered are less than 2.5 micron, and they are recovered at high grade and recovery using conventional cells. At Mt Isa, recovery from cleaner feed is above 95% for all size fractions from 1 micron to 37 micron, also using conventional (second hand) flotation cells (Pease 2004). Contrary to the common belief that “slimes don’t float”, the best performance, above 98% cleaner recovery, is in the 4 to 16 micron fraction.

A key question is – are the metallurgy gains of inert grinding confined to fines flotation? Until recently this was a hypothetical question – there was no practical technology for inert grinding at coarser sizes. This paper contends that recent developments have changed that – the availability of large scale (2.6MW) IsaMills and low cost ceramic grinding media have extended the gains in power efficiency and metallurgy to coarser grinds. In 1995 stirred milling “crossed over” from industrial manufacturing to the ultrafine mineral grinding. It is now ready for a more significant crossover to coarse mineral applications.

IMPACT OF STEEL GRINDING ON FLOTATION

Historically fine grinding applications use conventional ball or Tower Milling. As well as the high cost and low power efficiency, the chemical impacts of steel grinding on flotation offset the benefits of better liberation. Operations often respond to the poor flotation chemistry by using higher reagent additions or intensive conditioning to clean mineral surfaces, and pre-aeration steps to increase pulp potential and mineral hydrophobicity. These responses are expensive and only partly effective, and do not address the root cause.

In the early 1960s investigations by Ray and Formanek looked at beneficial effects on flotation of grinding lead-zinc ore in porcelain mills compared with grinding in iron mills or the addition of iron powder to porcelain mills (Kocabag, 1985). This work was confirmed by Fahlstrom (1960) and Thornton (1973) on chalcopyrite, galena and sphalerite ores. Later work by Greet (2004), Cullinan (1999 and 1999b), Pietrobon (2004), Grano (1994) and Johnson (2002) and Fleahy (1994) supported the benefits of inert grinding for a wide range of ores including nickel.

If a mineral or metal is immersed in water it assumes an electrical potential with respect to the water. The principal reactions are oxidation of the mineral to form metal ions and elemental sulphur, oxidation of xanthates to dixanthogens, and reduction of oxygen to hydroxyl ions (Figure 2).

Oxidative reaction:	$MS \rightarrow M^{2+} + S^0 + 2e^-$	oxidation of mineral sulphide
	$2X^- \rightarrow X_2 + 2e^-$	oxidation of xanthate
Reductive reaction:	$½O_2 + H_2O + 2e^- \rightarrow 2OH^-$	reduction of O_2

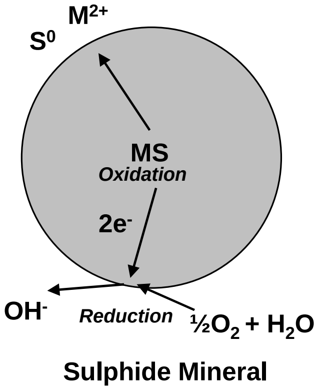


Figure 2: The anodic and cathodic domains in a mineral system

Sulphide minerals are semi-conductors. When they are brought into contact with steel media an electrochemical cell is formed (Figure 3). The steel media has the highest rest potential and is the anode; the sulphide mineral is the cathode.

Oxidative Reaction:	$Fe \rightarrow Fe^{2+} + 2e^-$	oxidation of steel media: anode
Reductive Reaction:	$2e^- + H_2O + ½ O_2 \rightarrow 2OH^-$	reduction of oxygen: cathode

Further reduction of Fe ions from solution results in iron hydroxide deposits on the mineral surface, as shown in Figure 3.

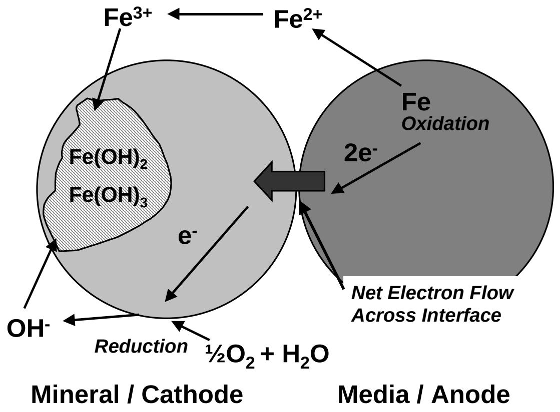


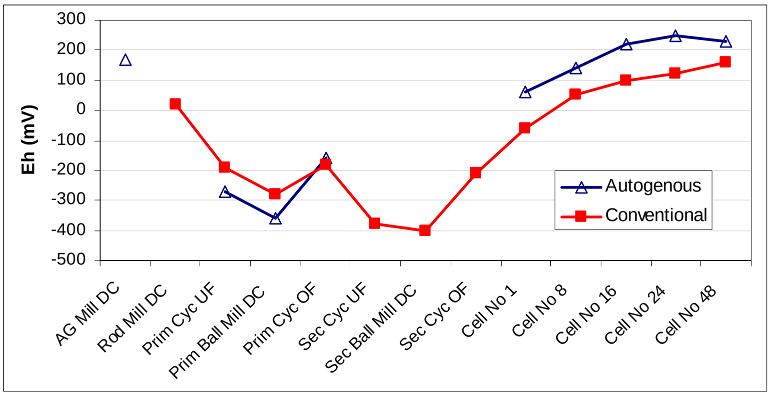
Figure 3: The galvanic cell in a mineral / media system

As a result, grinding in a steel media environment has several detrimental effects:

- ***The Eh effect***: the reducing environment lowers dissolved oxygen and Eh of slurry. Since collector adsorption is Eh dependent and may require oxidation of xanthates to dixanthogen, this will reduce floatability. Providing pre-aeration steps before flotation reduces this impact, but is unlikely to completely reverse it. This effect of grinding media on Eh and floatability has been well documented by Trahar (1984).

- Oxidation of steel grinding media causes ***iron hydroxide coatings on mineral surfaces.*** This reduces flotation selectivity in all sizes, but particularly for fine particles. This is even evident at the coarse sizes of autogenous grinding (figures 4-6). The impact is more significant in secondary and regrinds mills, as more fines are created and steel consumption is much higher

- Oxygen reduction on mineral surfaces promotes ***precipitation of hydrophyllic, insoluble metal hydroxides on the surface of sulphide minerals***. The impact on flotation is more pronounced for fine particles (Figure 6). Surface coatings can be offset by higher reagent addition, but at the cost of lower selectivity for clean minerals, as well as the higher reagent cost.



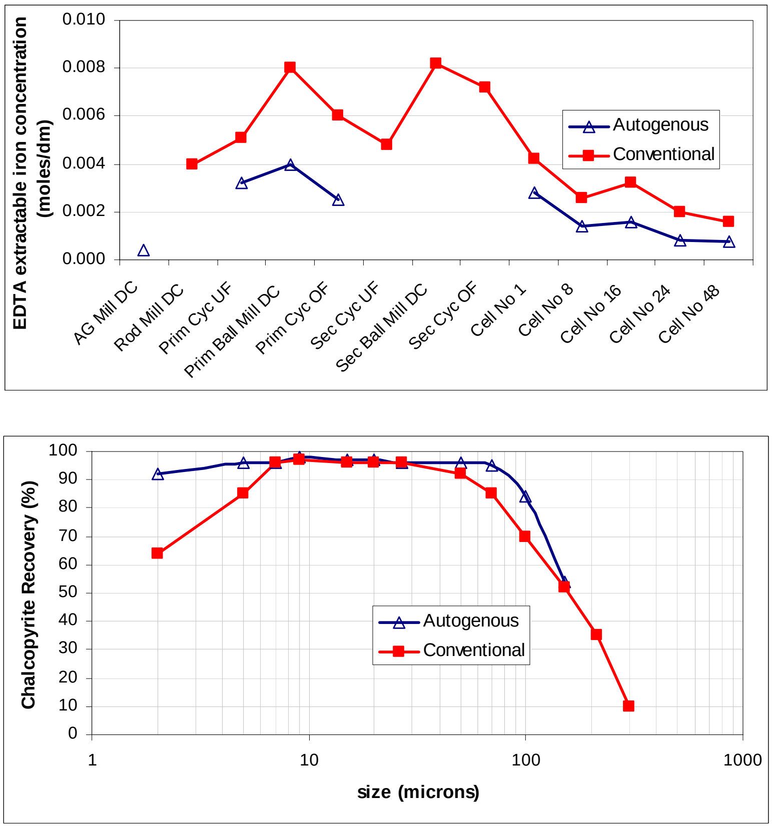


Figure 4-6: The effect of grinding media on flotation conditions and recovery (Grano et al 1994)

Some of the flotation impacts of steel media can be overcome by increasing pH and higher reagent addition. But flotation selectivity will still be low – some coarser minerals will have clean surfaces and will suffer from higher reagents, but the coated fines surfaces will demand them. Further, some flotation circuits require low pH (ie: pyrite/arsenopyrite/gold flotation circuits).

A much better solution is to address the root cause of the problem – keep all mineral surfaces clean by using inert media. While the benefits of high chrome media over forged steel media have been well documented, ceramic media has a much greater impact. Table 1 (Greet, 2004b) shows that changing from forged steel to high chrome media reduces the surface atomic composition of iron from 16.6% to 10.2%. Griding with ceramic media reduces this surface iron measure to below 0.1%. Considerable work has been done to demonstrate the flotation advantages of high chrome media. By contrast, little commercial based work has been done on fully inert media since it has not been a practical option until recently.

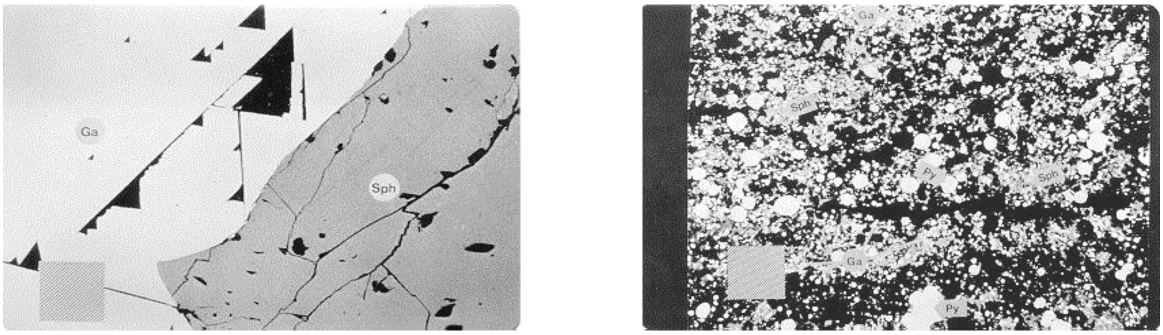
With steel media, the liberation benefits of grinding are offset by the chemistry impacts – a case of “two steps forward and one step back”. For grinding below 25 microns the chemical impact may be dominant – grinding is “one step forward, but two steps back”. The availability of large scale inert grinding allows operators to improve both liberation and flotation performance, getting full value from the installed grinding power.

Table 1: Composition determined via XPS, of the unetched surfaces of Rapid Bay Galena ground with different media (Greet 2004b, Cullinan 1999)

Media Type	**Surface Atomic Composition (%)**			
	O	**Pb**	**Fe**	**S**
Forged Steel	53.1	15.6	16.6	14.7
High Chrome	50.0	20.6	10.2	19.2
Ceramic	33.6	32.0	<0.1	34.4

CASE STUDY 1: MCARTHUR RIVER MINING (MRM)

The McArthur River orebody was discovered in 1955, with a resource of 227 Mt at 9.2% Zn and 4.1% Pb. However it remained undeveloped for forty years since existing technology could not economically treat the extremely fine grained minerals (Figure 7). The IsaMill was developed specifically to treat this orebody. It allowed economic regrinding to 80% passing 7 microns, fine enough to reduce silica in bulk concentrate to marketable levels.



Broken Hill Ore

McArthur River Ore

Figure 7: Different Grain Size of Broken Hill and McArthur River Ores (Grey Square is 40um)

The plant started mid 1995 with 4 IsaMills regrinding rougher concentrate. Media for the mills was screened ore gravel from the SAG mill discharge – a fully autogenous ultra-fine grind! , Two more mills were installed to increase production and recovery (in 1998 and in 2001). In 2004 the media was changed from ore gravel to screened sand – the higher efficiency of the sand increased mill capacity, and reduced wear on mill components at the higher throughput.

Table 2 shows production performance at MRM – very high concentrate grades are achieved at over 80% recovery for the fine grind. The concentrate sizing is P80 7 microns, and P50 2.5 microns. Looking at this from the perspective of a flotation bubble, 50% by weight means that 96% of the particles in MRM concentrate are finer than 2.5 microns. So 96% of the successful particle-bubble collisions at MRM happen for particles finer than 2.5 microns. This is achieved in conventional flotation cells – the selection criteria was to design for adequate lip length, then buy the cheapest cells available.

Fines float extremely well if they have fresh clean surfaces.

Table 2: Performance of McArthur River since commissioning

MINING			METALLURGY			
Year	**Tonnes**	**Head Grade**	**Tonnes**	**Zn Recovery**	**Con Grade**	
					%Zn	**%Pb**
1995/96	707,994	12.9%	759,519	66.4%	39.3%	11.2%
1996/97	1,035,222	14.4%	1,026,150	73.5%	43.5%	11.0%
1997/98	1,127,000	16.1%	1,139,000	74.3%	43.3%	11.9%
1998/99	1,222,238	16.4%	1,220,957	79.5%	45.0%	12.7%
1999/00	1,254,227	16.3%	1,262,639	80.9%	46.9%	12.0%
2000/01	1,226,499	15.4%	1,270,319	82.4%	46.8%	11.2%
2001/02	1,398,109	14.9%	1,404,539	82.7%	46.8%	11.1%
2002/03	1,505,306	12.7%	1,511,856	82,4%	46.6%	10.5%
2004	1,523,243	12.7%	1,579,762	80.0%	47.1%	10.3%

CASE STUDY 2: GEORGE FISHER OREBODY AT MOUNT ISA

The changes to the Mount Isa circuit as part of the “George Fisher Project” are detailed elsewhere (Young & Gao, 2000, Young, Pease & Fisher, 2000). In summary, the project involved adding a further 6 IsaMills, to regrind lead rougher concentrate to P_{80} of 12 µm, most zinc rougher concentrate to 12 µm, and a zinc regrind to P_{80} of 7 µm (Figure 8). Lead performance increased by 5% concentrate grade and 5% recovery (equivalent to 10% increase in lead recovery at the same grade). Zinc recovery increased by 10%, in two steps, and zinc concentrate grade by 2% (equivalent to 16% increase in zinc recovery at the same grade). The story of zinc metallurgy is told in Figures 8, 9 and 10.

The project predicted 5% higher zinc recovery (and no extra concentrate grade) due to extra liberation. Figure 9 shows this was achieved instantly. The surprise was the “second wave” of a further 5% zinc recovery increase and the 2% increase in zinc concentrate grade. This was because fines flotation improved after grinding finer.

It took about 6 months to discover how much better the fines could perform because we were so used to flotation after conventional grinding rather than after IsaMilling. Our three biggest mistakes were:

- Expecting to need a lot more reagents after IsaMilling due to the huge new surface area created. Some reagent additions were forecast to triple.
- Not taking the depressant (lime to pH 11) off the zinc cleaners.
- “Pulling” flotation harder because we thought flotation rates of the fines would be slower.

To our surprise, even though we introduced 6MW of extra grinding power, operating cost per tonne of feed did not increase. This was because of:

- Lower reagent additions
- Elimination of circulation loads between roughing and cleaning – a lot of power (and flotation capacity) is wasted in conventional circuits by pumping circulating loads of 100%-300%.
- Virtual elimination of spillage – due to new designs for pump boxes and pumps, the lower reagents, and especially the reduction in circulating loads.

The reduction in reagents was most unexpected, but perfectly logical in hindsight. During mill commissioning we increased reagents because we were increasing mineral surface area threefold. We were wrong. While more surface means more collector on the surface, this doesn't necessarily mean more collector in the pulp. If surface coatings mask mineral surfaces, then high collector in solution is needed to drive the diffusion through the coating to the surface. But if surfaces are truly clean – something that is never seen in a conventional steel grinding circuit but can be achieved by grinding in an inert environment – lower solution concentrations can achieve high surface coverage. Flotation after inert grinding is profoundly different – we had to forget everything we had learnt in our circuit and start with a "clean slate". Later we realised that the answers we needed were in the classic flotation texts. Bubble contact angles in clean mineral systems may not seem relevant to the plant operator who has badly altered mineral surfaces. But they are quite relevant in a system that quickly creates fresh clean surface.

The new circuit cut circulating loads from 200-300% to less than 50%. Liberated minerals with clean surfaces in a narrow size distributions respond predictably. They don't form large circulating loads of "undecided" particles. This creates a virtuous circle – lower circulating loads in cleaning means lower density, which gives better dilution cleaning. It also means less flotation cells are needed – eliminating a 100% circulating load doubles residence time. Mt Isa had to shut down some zinc cleaning capacity after installing the IsaMills.

The reduction in reagents and reduction in circulating loads almost eliminated spillage. This is another virtuous circle – returning spillage disrupts a circuit, creating new circulating loads and spillage.

Figure 10 shows the recovery by size in the zinc cleaning circuit at Mt Isa after IsaMilling (the recovery with respect to rougher concentrate, which is the feed to IsaMilling). Recovery is above 95% for all size fractions from 1micron to 37 micron. Recovery drops above 37 microns but there are very few particles here – these are composite that the circuit directs to regrinding to fine liberated, high recovery fractions. The highest recovery, over 98%, is in the 4-16 micron range. This size range is sometimes called "slimes", and it is often said that "slimes don't float". Indeed, after steel grinding they often don't.

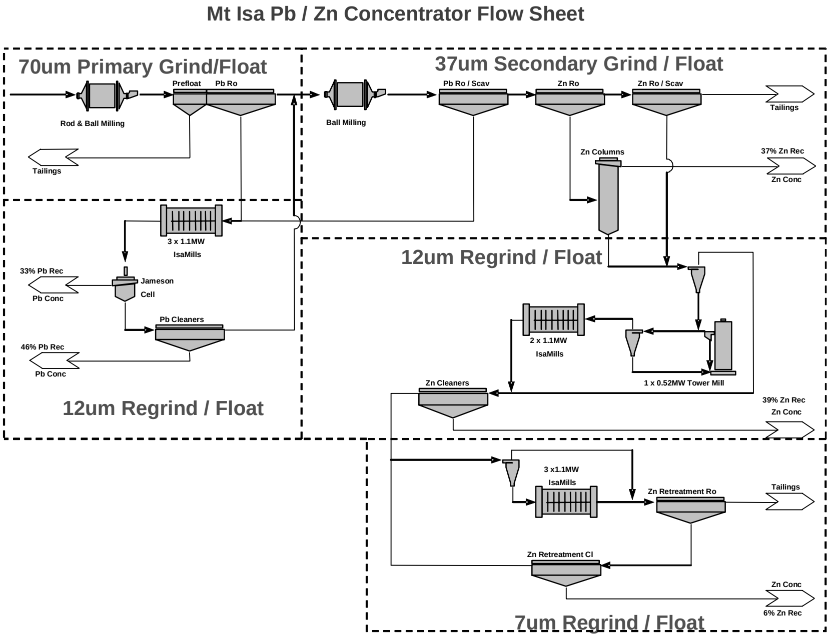


Figure 8: Mt Isa Pb/Zn Concentrator Flow Sheet

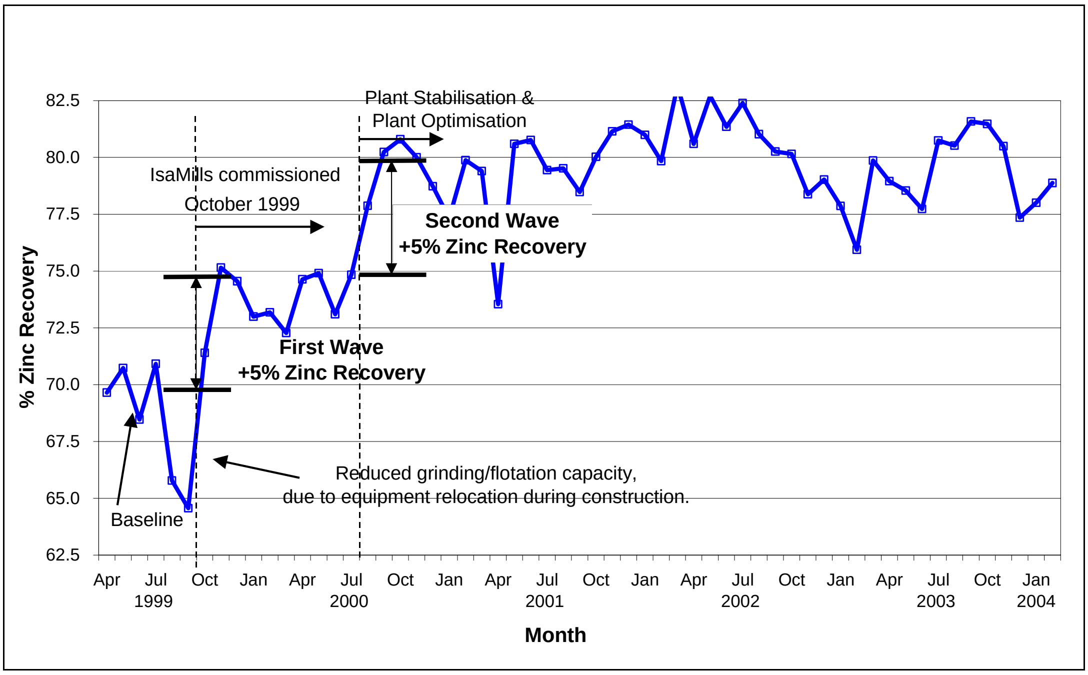


Figure 9: Zinc Recovery Increase from IsaMilling

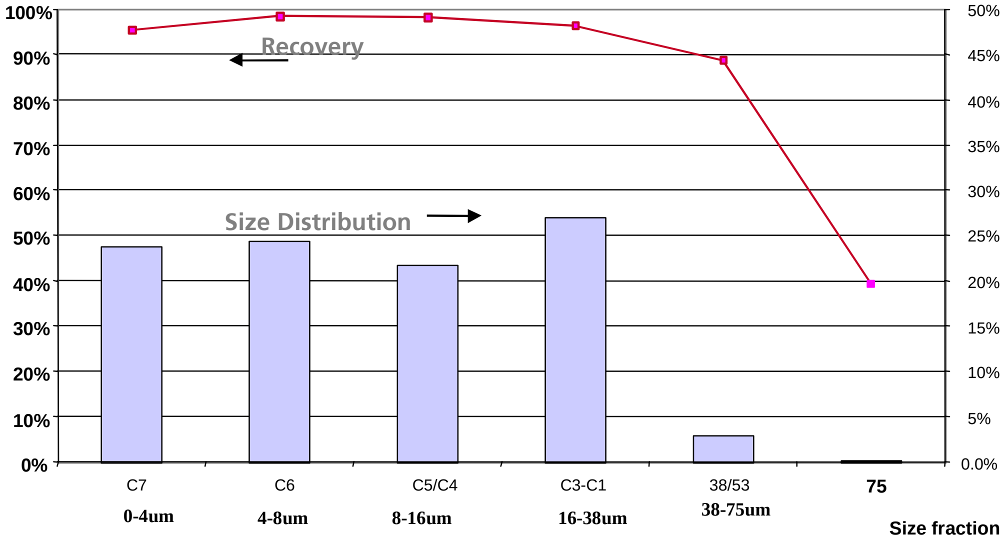


Figure 10: Mt Isa Zinc Recovery from Rougher Concentrate by Size

The Big Picture – Mineralogy, Liberation, Chemistry and Recovery Explained

Figure 11 is a graphic summary of the impacts of ore type, liberation, steel grinding, inert grinding and chemistry changes unfolding over twenty years at the Mt Isa operation. Each month Mt Isa collects inventory samples of all feed and exit streams. Each exit stream is sized and cyclosized, and size fractions are submitted for quantitative mineralogy. A huge amount of data is generated about different mineral classes in different size fractions. However one summary variable plotted in Figure 11 tells a compelling story. It plots sphalerite liberation in "recalculated feed" and sphalerite recovery. "Recalculated feed" is created by mathematically combining all the plant exit streams according to their relative tonnage. Therefore it captures the impact of all the grinding and regrinding stages of the circuit. The Mt Isa journey unfolds in four stages, as described below.

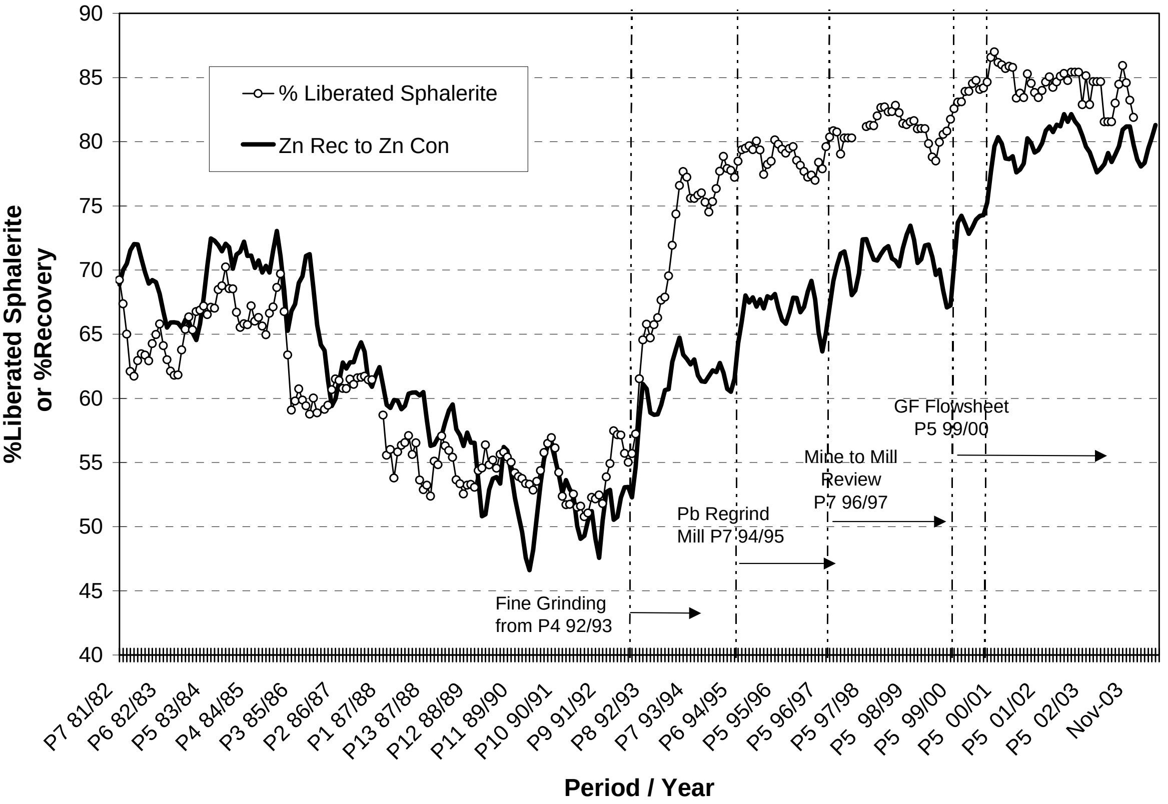
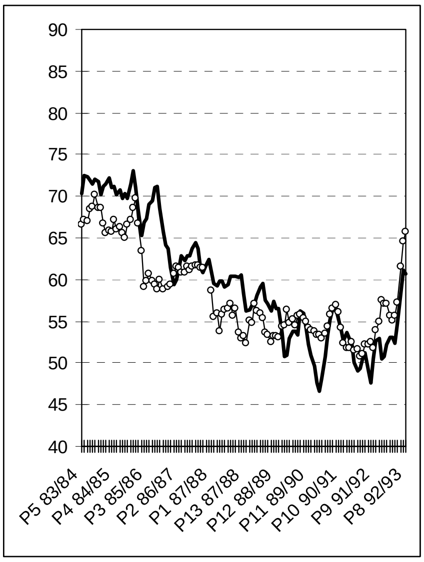


Figure 11: The Mt Isa Story: Zn Recovery and Sphalerite Liberation

Stage 1: The sickening decline

During the 1980's ore sources became increasingly fine grained. Liberation declined, and recovery inexorably declined with it. Like most operators under profit pressure we increased feed tonnage. This increased revenue but further reduced liberation. We tried dozens of reagents, dozens of circuit changes, and dozens of metallurgists. Finally we accepted we were just "shuffling the deck chairs on the Titanic" – reagents don't grind. A fundamental rule seemed to guide plant performance in spite of our metallurgical endeavours – recovery equals liberation plus 10%. A small amount of composite particles can be accepted into concentrate , until the quality constraint on impurities is reached.

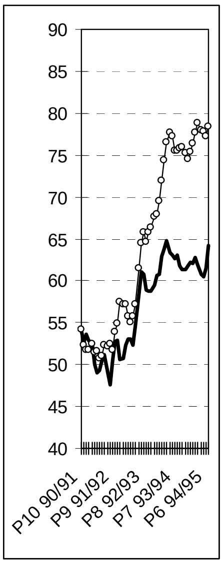


Liberation dropped by 15%

Recovery dropped by 20%

Stage 2: Use a bigger hammer

We had to increase liberation to increase recovery. The only available technology was conventional ball milling, and the then emerging Tower Milling. We installed 6MW of ball and Tower Mills, effectively doubling the grinding capacity. While we knew this would increase liberation, we also knew there was a downside – creating finer particles using so much steel media would create surface chemistry problems. As expected, this delivered two steps forward from liberation, and one step backward from surface chemistry. Liberation increased by 18%, while zinc recovery increased by 7%. The old rule “recovery is liberation plus 10%” had broken down, but this was the inevitable price of grinding fine with steel media.



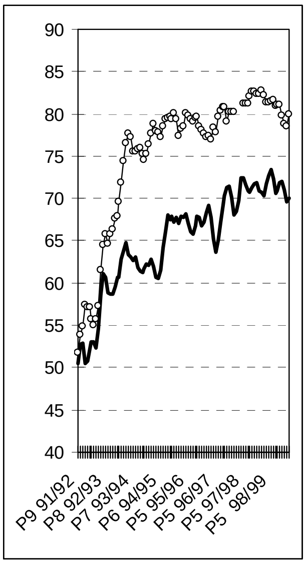
Liberation increased by 18%

Recovery increased by 7%

Concentrate grade unchanged
(not shown)

Stage 3: A pleasant surprise

The IsaMill technology for McArthur River was developed in the Mt Isa concentrator. The first installations were to regrind lead cleaner feed. The intent was to liberate some of the galena/sphalerite binaries to reject some zinc from lead concentrate. This was achieved, though the total effect on sphalerite liberation was small. However zinc recovery to zinc concentrate increased by 4%. This was because the clean surfaces after IsaMilling allowed us to improve flotation chemistry – we were able to reject fine liberated sphalerite particles that had previously misreported to lead concentrate. We had achieved what no amount of chemistry or circuit changes had been able to achieve in the past – increasing recovery increase without significantly changing liberation.

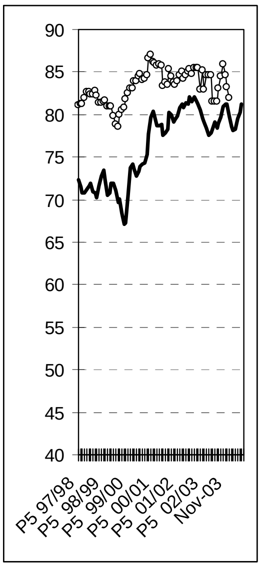


Liberation change negligible

Recovery increased by 4%

Stage 4: Cooking with Gas

Breaking the quandary between increasing liberation and harming surface chemistry was a revelation. With steel milling we had reached to stage of one step forward from liberation, but two steps back from chemistry. Suddenly a new option was available – improve both liberation and chemistry in the same step. We put 4 MW IsaMills into the zinc circuit. Liberation increased 4 %, but recovery increased by 10%. Concentrate grade also increased by 2% - we could have taken this as an extra 6-8% recovery, but the higher grade was more profitable. Making the higher grade had not been possible in the past.



Liberation increased by 4%

Recovery increased by 10%

Concentrate grade increased 2% (not shown)

Figure 10 shows that fine sphalerite recovery is over 95% after IsaMilling (ie: recovery of sphalerite from rougher concentrate). Figure 11 however shows total circuit sphalerite is still only 80% with respect to plant feed. Of the 18% sphalerite losses, 7% reports to lead concentrate, and 11% reports to tailing. In the tailings about half is in coarse composites and half in fine liberated particles. If a particle is too coarse, or too badly surface altered by steel grinding, then it doesn't float in the roughers, and doesn't get a chance to see the IsaMills.

Undoubtedly rougher performance would be better after inert grinding, but the technology was not available when we installed the extra secondary grinding. Mount Isa is now considering the feasibility of a large open cut operation which would necessitate building a new concentrator. Our design for that concentrator uses large IsaMills (3.3MW or bigger) in place of ball mills to grind rougher feed, and further mills to regrind rougher concentrate. This reduces footprint, reduces capital cost, increases energy efficiency, and will achieve higher recovery with lower reagent additions than the current circuit.

CASE STUDY 3: MT ISA BLACK STAR OPEN CUT

Surface resources at Mt Isa had long been a target for open cut mining. However the poor metallurgical response was always a barrier to production. Much of the ore is "transitional" between surface oxides and deeper primary sulphides. The transition ore is lower grade than primary ore, has fine grained mineralogy, and leaching has activated pyrite and sphalerite, leading to non-selective flotation. Constant attempts over the last 80 years failed to make the ore economic, with flotation unable to make smelter quality concentrates at any recovery.

The development of the IsaMills and the flowsheet to treat George Fisher ore changed this. The fine grinding achieves mineral liberation and cleans the mineral surfaces by attrition, and the combination of high intensity inert grinding and the correct water chemistry in flotation stops re-activation of unwanted minerals. The impact is shown by the grade recovery curve in Figure 12 - target concentrate grades can now be made at acceptable recoveries.

As a result, IsaMill inert grinding technology enabled production from the Black Star Open Cut resource. Production commenced in the first half of 2005, targeting 1.5M t/y to supplement underground production, produced from a mineral resource of 25Mt at 5.1%Zn and 2.7%Pb. This represents only a small portion of the potential open cut resources at Mt Isa. The success to date has led reassessment of the economics of the entire open cut resource and future production scale at Mt Isa.

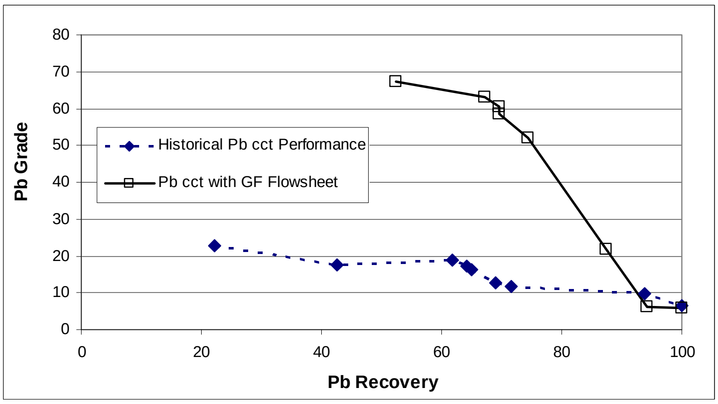


Figure 12: Pb Grade/Recovery Curve – ISA Lead-Zinc Transition Ore

CASE STUDY 4 : MERENSKY PLATINUM TAILINGS RETREATMENT PLANT

In 2001 Anglo Platinum assessed the retreatment of dormant tailings dams in the Rustenberg area in South Africa. These tailings contained economic amounts of Platinum Group Minerals (PGMs) if new technology could address two issues:

- The fine grained mineralisation of the PGMs in tailings (why it wasn't recovered first time)
- Surface oxidation and oxidation products which harmed flotation – some of the tailings were placed over 100 years ago.

Anglo Platinum and Xstrata Technology worked together to find an economic treatment route. To achieve economies of scale for the project the IsaMill was successfully scaled up from 1,000 kW to 2,600kW. This proved to be the enabling technology for this project due to:

- The ability to grind fine at low cost – the mill operates in open circuit, and uses cheap local sand as the grinding media.
- The clean mineral surfaces resulting from the inert grinding environment. This was crucial to achieve target grades and recoveries after regrinding.

Figure 13 shows the improvement made by an IsaMill regrinding rougher concentrate before cleaning (Buys et al, 2004). The mill increases flotation kinetics in cleaning, just as it does at Mt Isa and McArthur River. This is in contrast with the common observation that regrinding in a steel mill slows kinetics of all minerals.

Anglo Platinum commissioned the Western Limb Tailings Retreatment Plant in 2004. At the end of 2004 they concluded that (Buys, 2004):

- IsaMill technology was enabling for the WLTR project since it allowed acceptable concentrate grades to be made from oxidised slow floating tailings.
- Flotation kinetics improved after fine grinding due to both extra liberation and the removal of iron oxide surface coatings. Inert fine grinding of rougher concentrate was necessary.
- The scale up to the M10,000 IsaMill (from 1 MW to 2.6 MW) was successful.

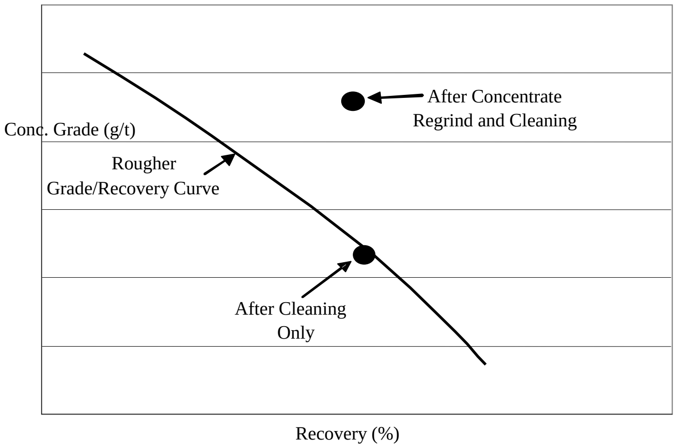


Figure 13: Improvement in Platinum Grade/Recovery After IsaMilling for Western Limb Tailings Retreatment

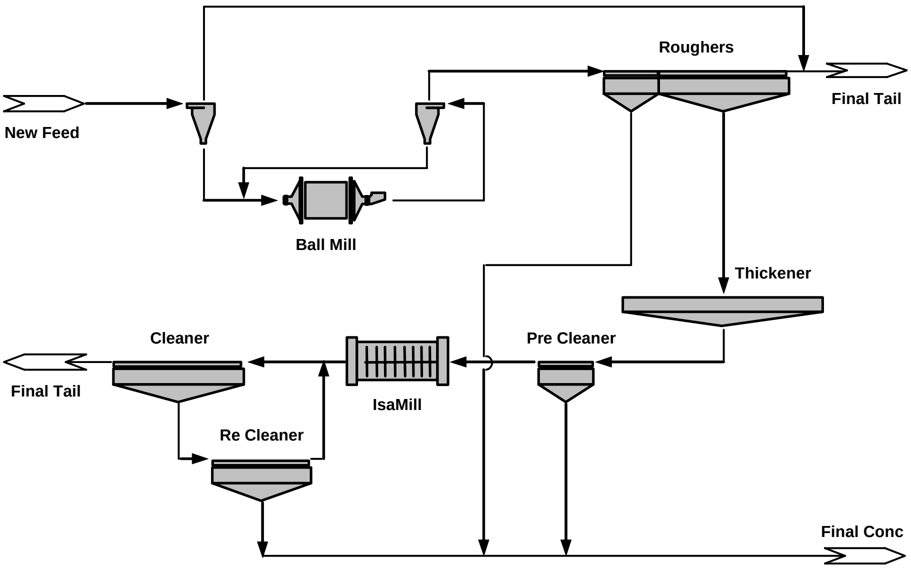


Figure 14: Western Limb Tailings Retreatment Flowsheet

INERT GRINDING BEFORE LEACHING

Just as the ability to grind to 10 microns in an inert grinding environment has been the enabling technology for the flotation of fine grained ore bodies so to this ability has been an enabling step for several hydrometallurgical technologies. The high surface area of fine particles means high leaching rates at relatively low temperature and pressure, reducing capital and operating costs. High intensity fine grinding also reduces the activation energy required to leach minerals by creating a highly stressed surface, reducing the crystalline nature to amorphous phases. This effect of mechanical (or mechanochemical) activation of minerals is well reported (Balaz, 2000; Juhasz and Opoczky,1990; Grelach et al 1989), and it means that minerals leach under much less aggressive conditions. The higher power intensity of grinding would enhance this effect. Several emerging leaching processes have been based on fine grinding of feed – the Activox process, the UBC/Anglo process, the Phelps Dodge Process, and Xstrata's Albion Process.

In flotation the chemical benefits of inert grinding can be dramatic. In leaching there is a lesser, but still important impact. Conventional grinding before leaching will directly input steel to the leach feed. This may require additional preoxidation of leach feed, and also result in higher reagent consumption. An IsaMill will soon be trialed to replace a ball mill regrinding concentrate before gold cyanidation. The ball mill is currently consuming 10t/day of steel media, which enters leach feed. This will provide a future case study of the impact of inert grinding on leach performance and reagent consumption.

THE CROSSOVER TO COARSER GRINDING

The principles learnt for fines flotation also apply for coarser particles (above 30 microns). The impact of steel on flotation is not as dramatic as it is for fines, but it is still there. This is why so much work has been done on the use of high chrome media. The principles of floating with clean surfaces, in narrow size distributions, with fast kinetics, low reagents and low circulating loads are vital to achieve good fines recovery, and will also improve flotation of coarser particles.

Until recently the prospect of inert grinding at coarse sizes was generally impractical. This has changed with IsaMills now operating at 2.6 MW, and the availability of low cost, high efficiency ceramic (MT1) developed specifically for stirred milling (Curry and Clermont, 2005). Pilot work on coarser feeds (eg 200 microns) consistently shows that IsaMills with ceramic media have significantly higher power efficiency and lower capital cost installations (eg open circuit) than conventional ball milling. The crossover to coarse grinding may be a “transformational” technology change, delivering lower capital, higher efficiency grinding, as well as better metallurgy.

The pilot predictions need to be demonstrated at full scale. The first industrial application of a 2.6 MW IsaMill with MT1 ceramic will be commissioned in late 2005, at Centerra Gold's Kumptor operation in the Kyrgyz Republic. This operation which produces 500,000 oz/y gold will use the IsaMill to grind rougher concentrate before leaching. This application will allow full scale comparison of grinding rates and size distributions achieved by SAG mills, ball mills and IsaMills.

Another full scale coarse grind application is currently under design, and is expected to be operating in late 2006. This will involve a 3.3 MW IsaMill using MT1 grinding rougher tailings in open circuit, from 150 micron to 55 micron. Based on current plant data, an 8 MW conventional ball mill in closed circuit would be required to match this grinding performance. Ball milling cannot be justified for this application due both to the high capital cost and the relatively poor metallurgy compared with inert grinding.

CONCLUSION

Necessity was the mother of invention for ultra-fine grained orebodies. They were simply intractable with conventional ball and Tower milling. High intensity stirred milling was the breakthrough that transformed the economics – low capital and installation cost, low media cost, high power efficiency, simple installations, sharp size distributions in open-circuit, and inert attrition of mineral surfaces.

It is most unlikely that only fine grained ores will benefit from these features. The rapid increase in scale to 3.3 MW enables large scale coarser grinding applications for IsaMills. Several projects are under design to transfer this technology to the mainstream.

The performance of these projects will answer the question: is this the next big thing in grinding?

REFERENCES

Balaz, P., 2000, Extractive Metallurgy of Activated Minerals, Elsevier.

Buys, S, Rule, C and Curry, D,. 2004, The Application of Large Scale Stirred Milling to the Retreatment of Merensky Platinum Tailings, CIM Montreal.

Cullinan, VJ., Grano, SR., Greet, CJ., Johnston, NW., Ralston J., 1999, Investigating fine galena recovery problems in the lead circuit of Mt Isa Mines lead/zinc concentrator Part 1: grinding media effects, Minerals Engineering vol12 No 2 (1999) 147-163.

Cullinan, V J, 1999b. Improving the Flotation Response of Fine Galena, PhD Thesis, University of South Australia.

Fahlstrom, PH., 1960, Discussions, Proceeding Int Min Process Congr, IMM, London, pp387

Fleahy J., Lane G., 1994, Grinding environment effects in the flotation of Kambalda nickel ores, 5th Mill Operators Conference, 43-50.

Grano, S.R., Ralston, J. and Smart, R.St.C., 1990. The influence of electrochemical environment on the flotation behaviour of Mt. Isa Mines copper and lead/zinc ores, Int. J. Miner. Process., 30: 69-97.

Grano S.R., Wong, P., Skinner, W., Johnson, N.W. and Ralston, J., 1994. The effect of autogenous and ball mill grinding on the chemical environment and flotation of the copper ore of Mount Isa Mines Limited, In: III Latin-American Congress on Froth Flotation (University of Concepcion Publ.), Concepcion, Chile, pp. 351-388.

Greet, CJ., Small, GL., Steiner, P., Grano, SR., 2004, The Maggotteaux Mill: investigating the effect of grinding media on pulp chemistry and flotation performance, Minerals Engineering 17 (2004) 891-896.

Greet, CJ., Steiner, P., 2004b, Grinding – The Primary Conditioner, Metallurgical Plant Design and Operating Strategies, AUSIMM, Perth Australia, pp 319-336.

Grelach, Gock, Gosh, 1989, Activation and Leaching of Chalcopyrite Concentrates with Dilute Sulphuric Acid, International Symposium of Hydrometallurgy.

Johnson, NW., 2002, Practical aspects of the effect of electrochemical conditions in grinding mills on the flotation process, Flotation and Flocculation – From Fundamentals to applications, Strategic Conference and Workshop, Hawaii, 28 July – 2 August 2002.

Juhasz , Opaczky, 1990, Mechanical Activation of Minerals by Grinding, Ellis Horwood Ltd.

Kocabag D., Smith M., 1985, The Effect of Grinding Media and Galvanic Interactions Upon the Flotation of Sulphide Minerals, in Proceedings Complex Sulphides – Processing of Ores, Concentrates and By-Products (eds: A D Zunkel, R S Boorman, A E Morris and R J Wesley), pp 55-81 (The Metallurgical Society: New York).

Pease, JD., Young, MF., Curry, D., Johnson, NW., Improving Fines Recovery by Grinding Finer, Centenary of Flotation Symposium, Brisbane, Australia, 2004.

Pietrobon, MC., Grano, SR., Greet, C., 2004, Matching laboratory and plant performance – a case study of Elura lead circuit, Pasminco Australia Ltd, Minerals Engineering 17 (2004) 811-824.

Thornton, E., 1973, The effect of grinding media on flotation selectivity, Proceedings 5th Annual Meeting of Canadian Mineral Processors, Dept of Energy, Ottawa, pp224-239

Trahar WJ, 1984 The Influence of Pulp Potential in Sulphide Flotation, Principles of Mineral Flotation, The Wark Symposium.

THE APPLICATION OF LARGE SCALE STIRRED MILLING TO THE RETREATMENT OF MERENSKY PLATINUM TAILINGS

Stephan Buys[1], Plant Superintendent
Chris Rule[2], General Manager - Concentrator Technology
Dan Curry[3], Technical Superintendent – Mineral Processing

[1]Anglo Platinum Management Services (Pty) LTD
Process Technology Division
Central Deep
Rustenburg
Republic of South Africa
+27 14 598 2312
E-mail: stephanb@angloplat.com

[2]Anglo Platinum Management Services (Pty) LTD
Process Technology Division
55 Marshal Street
Johannesburg
Gauteng
Republic of South Africa
+27 11 373 6998
E-mail: chrisrule@angloplat.com

[3]Xstrata Technology Pty Ltd
Level 2, 87 Wickham Terrace
Brisbane
Queensland 4000
Australia
Tel: +61 7 3833 8500
E-mail: dcurry@xstrata.com.au

Key Words:

Platinum, PGM’s, tailings, fine grinding, IsaMill, liberation, surface chemistry, fine flotation

ABSTRACT

The use of fine grinding (FG) in mineral processing is well established. The application of FG in the South African platinum industry is more recent and it's use in the recovery of PGM's from dormant tailings dams is unique. Anglo Platinum uses IsaMill FG technology at their Western Limb Tailings Retreatment project (WLTR) near Rustenburg in the North West Province of South Africa.

In December 2003, Anglo Platinum commissioned the first purpose built tailings re-treatment facility for PGM recovery in South Africa. The concentrator re-treats Merensky ore tailings from concentrators that operated early last century. The WLTR facility takes advantage of modern technology, such as FG, to economically recover PGM's from material historically considered as waste. The WLTR concentrator has a capacity of 4.8 Mtpa, and has been designed to easily expand to 10.8 Mtpa. The flow sheet includes recovery of tailings by high pressure water monitoring, ball milling, rougher flotation, rougher concentrate regrinding and cleaner /recleaner flotation.

The rougher concentrate is reground in an IsaMill, which is a stirred mill that operates with inert silica sand grinding media. For this project, Anglo Platinum required a stirred mill of a unit size not previously available. Xstrata Technology and Anglo Platinum collaboratively developed the largest wet, fine grinding mill available; the M10,000, 2.6 MW IsaMill.

The use of IsaMill technology was enabling for the WLTR project, as it allowed smeltable concentrate grades to be produced from the oxidised, slow floating tailings. The economics and practicality of fine grinding has fundamentally changed with the M10,000 IsaMill. Flotation is transformed by high intensity attritioning in the inert media environment, and the narrow size distribution of product in open circuit configuration. The use of conventional steel media grinding to improve fine particle flotation is limited due to the oxidation of liberated mineral surfaces inherent in these environments.

INTRODUCTION

During 2000 and 2001, Anglo Platinum conducted metallurgical and geological investigations into the retreatment of dormant tailings dams in the Rustenburg area. At the current market value of platinum and palladium, the concentration of PGM minerals in the dams represented a possible economic resource. Metallurgical test work identified a significant proportion of these minerals that could be recovered via fine grinding and flotation.

A pilot plant program was developed to confirm the effectiveness of fine grinding with respect to flotation recovery. The results from the program assisted with the proposed concentrator process design and samples were generated for vendor test work of critical equipment.

A collaborative design project between Anglo Platinum and Xstrata (then MIM – Mt Isa Mines) was initiated to run in parallel with the pilot plant tests and plant design. This project required a detailed design for a fine grinding IsaMill of 2,600 kW (3500 hp) capacity, to maximise economies of scale with this unit process.

METALLURGICAL TEST PROGRAM

Bulk Sample Collection

The Klipfontein Central tailings dam was selected to provide the bulk sample for pilot testing. A total of 56 holes were drilled, producing approximately 424 tonnes of sample. The resource for the Rustenburg section was calculated to be 186 Mt at 1.08 g/t PGM + Au. Over 4 Moz of platinum were contained within this resource. The Klipfontein Central bulk sample was representative of the total Rustenburg resource.

Test Program Detail

Using previous laboratory scale results in conjunction with certain process design criteria, the pilot scale program was developed to confirm the integrity of the concepts identified previously. Four campaigns were designed to provide the necessary information. The grinding circuits are summarised in Table 2.

Table 1 : Grinding Targets and Mill Type

Pilot Run #	Primary Grinding		Rougher Conc Regrind	
	Mill Type	**% -75 μm**	**Mill Type**	**% -25 μm**
1	1^0 Ball, 2^0 Ball	80	-	-
2	1^0 Ball, 2^0 Ball, 3^0 IsaMill	90	-	-
3	1^0 Ball, 2^0 Ball	80	IsaMill	90
4	1^0 Ball, 2^0 Ball, 3^0 IsaMill	90	IsaMill	90

A nominal grind of 80 % passing 75 μm was achieved by using a primary and a secondary ball mill (secondary mill in closed circuit with cyclone) for Runs 1 and 3. An IsaMill was used in open circuit to grind the cyclone product to 90 % passing 75 μm for Runs 2 and 4. All tests used rougher flotation with a residence time of 45 minutes. A cleaning /recleaning stage was used to increase the rougher concentrate grade to 50 g/t (final concentrate target). An IsaMill was used in open circuit to regrind the rougher concentrate to 90 % passing 25 μm prior to cleaning for Runs 3 and 4. The rougher and cleaner bank flotation tails reported to final tails. The recleaner tail reported to the cleaner feed. Reagent selection, addition rates and locations were based on results from laboratory scale test work.

Head Assay Analysis

The bulk sample was mixed prior to being fed to the pilot plant. The statistics in Table 2 indicate that this mixing process was effective, as very little variation in grade is evident. Samples were collected over three one-hour periods for each Run.

Table 2 : Head Assay Analysis

	3E (g/t)	Cu (%)	Ni (%)
Mean	1.17	0.027	0.076
Standard Deviation	0.04	0.003	0.010
% RSD	3.45	9.49	13.86

Pilot Plant Results

A summary of the flotation results, with respect to PGM, for each run is presented in figure 1. The overall recovery achieved for the four runs were similar with runs 1 and 3 achieving slightly better results. Final concentrate grade improves significantly by introducing a finer product to the cleaner flotation circuit.

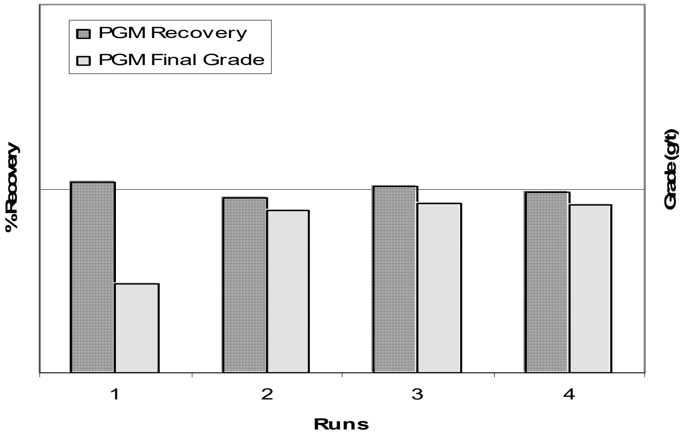


Figure 1 : Final PGM Recovery and Concentrate Grade

The final concentrate grade is dependent upon grind size. Table 3 demonstrates the particle sizes of the rougher and cleaner concentrate streams.

Table 3 : Rougher and Cleaner Stream Sizings

Run #	Rougher Feed	Cleaner Feed	Rougher Conc Regrind
	% Passing 75 µm	% Passing 25 µm	
1	77.0	50.3	No
2	93.4	85.4	No
3	76.6	83.4	Yes
4	94.6	88.1	Yes

The cleaner feed size distributions were similar for Runs 2, 3 and 4. Despite Run 2 not having a rougher concentrate regrind stage, the finer primary grind (rougher feed) resulted in a similar cleaner feed size as Runs 3 and 4 where regrinding was used.

Rougher Flotation

Figure 1 confirmed the laboratory scale observations that the recovery of PGM's from the tailings dams was highly dependent upon fineness of grind. At equivalent concentrate grade, PGM recovery was approximately 10 % higher with a primary grind of 90 % passing 75 µm, than with 80 % passing 75 µm.

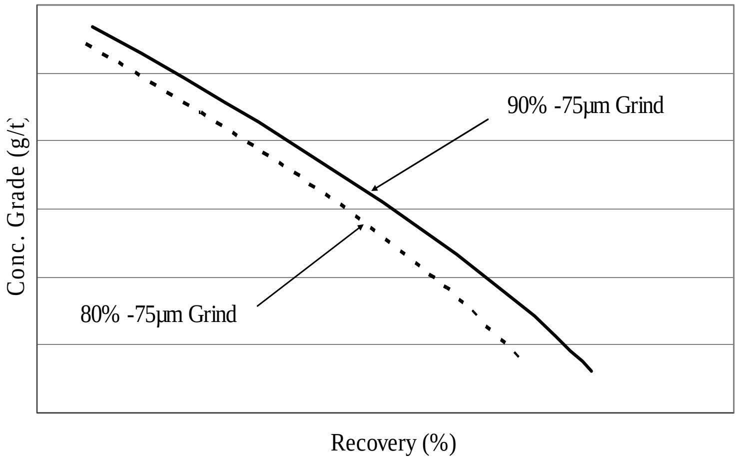


Figure 2: Rougher Flotation Grade /Recovery Curve

Whilst the flotation kinetics of PGM's are typically slow, it was significant that the rate of flotation increased with the finer grind. Mineralogical examination showed that the improved flotation performance at the finer grind was due to additional liberation of PGM's as well as the removal of iron oxide layers from base metal sulphide surfaces by grinding in the inert media environment of the IsaMill.

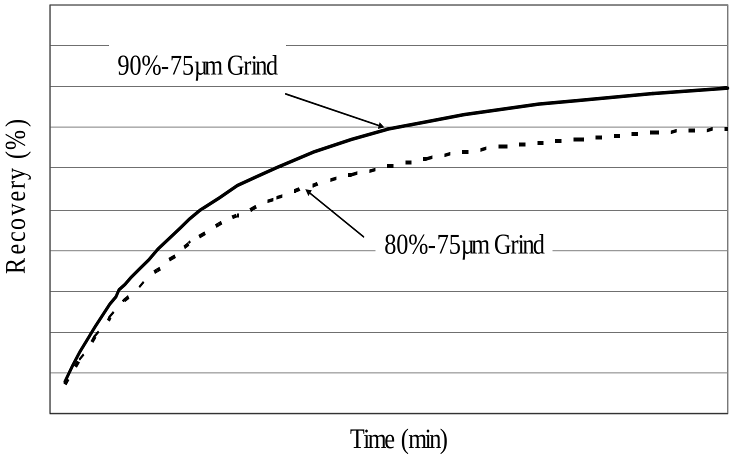


Figure 3 : Rougher Flotation Recovery /Time Curve

Cleaner Flotation

It was anticipated that the use of a conventional cleaner /recleaner circuit would enhance the grade of the rougher concentrate with minimal loss in recovery. However, it was not possible to shift the grade /recovery curve for Run 1 (80% passing 75 µm primary grind) as shown in Figure 4. As shown in Table 3, the cleaner feed size distribution was the coarsest of all Runs (50.3 % passing 25 µm). Limited success came from additional laboratory investigation to identify methods of upgrading the rougher concentrate using alternative reagents, pH modification and novel chemistry techniques. However, mineralogical examination showed that poor liberation and iron oxide coating of base metal sulphide surfaces were responsible for the poor upgrade potential of the rougher concentrate. Regrinding of the concentrate was required to improve PGM liberation and freshen-up the heavily oxidised mineral surfaces.

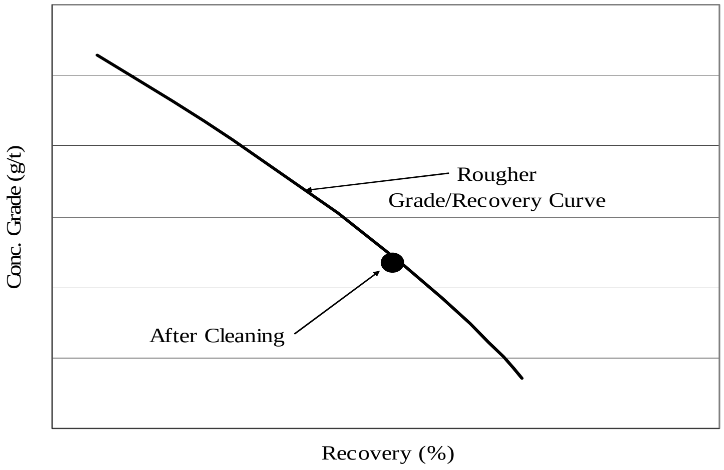


Figure 4 : Run 1 – Grade /Recovery for 80%-75µm Primary Grind

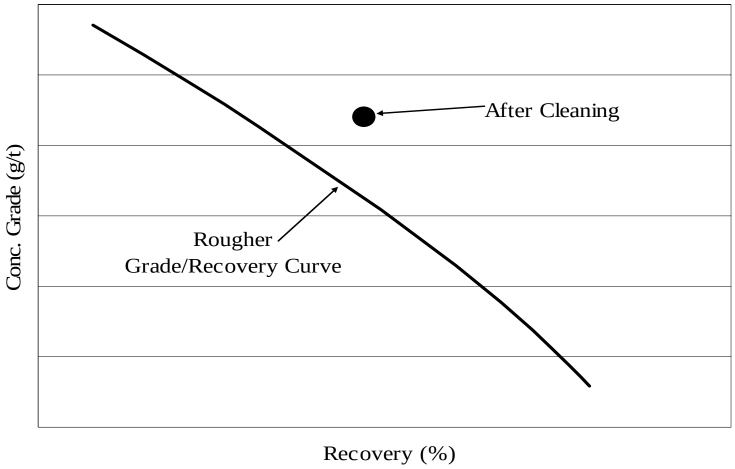


Figure 5 : Run 2 – Grade /Recovery for 90%-75µm Primary Grind

Run 2 adopted a finer primary grind of 90% passing 75 µm. The benefit of this to rougher recovery has been discussed, but the additional liberation and removal of base metal sulphide oxide surface coatings permitted significant upgrading in the cleaners. Figure 5 shows the shift of the grade /recovery curve.

Runs 3 and 4 used an IsaMill rougher concentrate regrind stage to test what improvements in cleaning could be gleaned from additional liberation and particle surface cleaning. With the coarse rougher concentrate size distribution produced in Run 1, it was expected that regrinding prior to cleaning would have the greatest benefit for this case. Figure 6 confirms this, with the grade /recovery curve shifted to the right of the rougher plot.

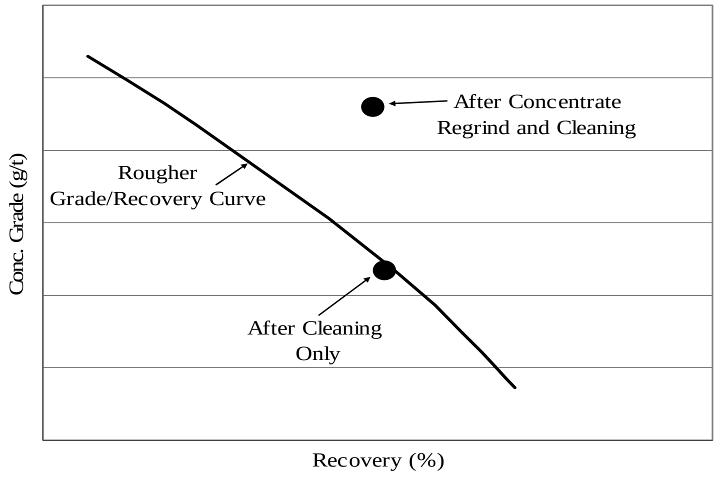


Figure 6 : Run 3 – Grade /Recovery After Rougher Concentrate Regrind at 80%-75 Primary Grind

The cleaner feed size after regrinding in Run 3 was 83.4 % passing 25 µm. This compares well to the cleaner feed size in Run 2 (finer primary grind) of 85.4 % passing 25 µm. As Run 2 and Run 3 both involved a stage of inert media grinding in an IsaMill (clean mineral surfaces), and possess similar size distributions it was expected that the overall flotation performance of these two cases would be similar. Figure 1 shows Run 3 exhibiting a marginally better final concentrate grade and recovery, despite having a lower feed grade and slightly coarser cleaner feed size distribution than Run 4. It is hypothesised that the better performance of Run 3 was due to IsaMilling applied immediately before cleaning, rather than before roughing (to produce a finer primary grind) in Run 2. It is likely that the clean, reactive mineral surfaces produced by the inert IsaMill environment oxidise during the 45 minute rougher flotation period. Application of IsaMilling after roughing as performed in Run 3, produces fresh mineral surfaces which are likely to be responsible for better cleaner flotation performance.

Run 4 used IsaMilling to produce a primary grind of 90 % passing 75 µm, and for regrinding the rougher concentrate. The cleaner feed size of Run 2 (finer primary grind only) was 85.4 % passing 25 µm, compared to 88.1 % passing 25 µm in Run 4. Only a small increase in fineness was produced by regrinding the rougher concentrate, but the flotation performance improved, showing that the cleaning potential is sensitive to rougher concentrate particle size distribution and mineral surface condition.

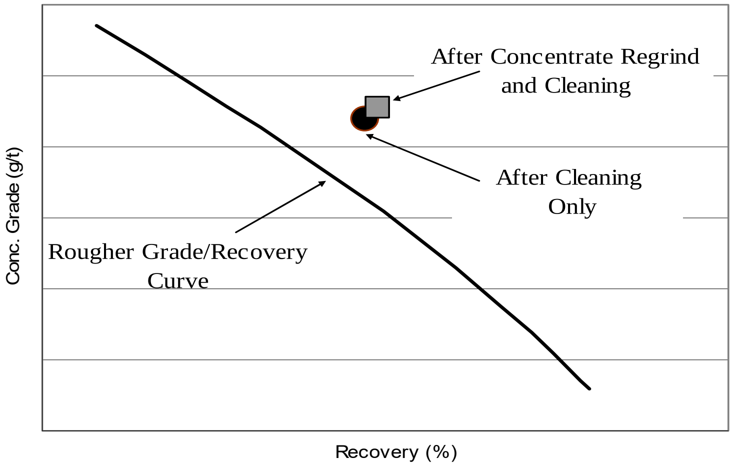


Figure 7 : Run 4 – Grade /Recovery After Rougher Conc Regrind at 90%-75µm Primary Grind

Pilot Plant Test Conclusions

The following conclusions were made from the pilot plant tests:

- The PGM grade and recovery targets can be met with the use of conventional flotation and IsaMill inert grinding.
- A primary grind of no less than 80 % passing 75 µm, and rougher concentrate regrind of no less than 85 % passing 25 µm is required to meet the grade /recovery targets.
- A finer rougher concentrate regrind is likely to improve cleaner flotation – flotation is sensitive to grind size.

Most importantly, the combination of improved liberation and inert media grinding is required to maximise flotation potential. The impact of changes in mineral surface chemistry (such as oxidation from iron in grinding media) to flotation recovery is generally much larger than the effect due to liberation alone (refer to AMIRA P336, Report P336/26). There are examples where the actual change in recovery due to increased liberation from grinding is much lower than predicted due to the negative impact of mineral surface oxidation. It is even possible that the overall flotation recovery decreases after regrinding due to surface chemistry changes, despite increasing mineral liberation (Frew, Davey and Glen, 1994).

PLANT DESIGN

The Western Limb Tailings Retreatment Project (WLTRP) simplified flow sheet is shown in Figure 8. A staged approach was taken with the design and installation of the circuit, to limit the capital exposure and project risk to the novel treatment of dormant PGM tailings. The project was divided into two Phases, with the detailed design and installation of Phase 2 dependent upon the operating knowledge of Phase 1.

The Phase 1 WLTR concentrator was commissioned in the fourth quarter of 2003. The design is based on 400,000 tonnes per month (4.8 Mtpa), in a single grinding /flotation line. Much of the Phase 2 civil work was put in place during construction of Phase 1, making the addition of Phase 2 a relatively simple task. Phase 2 would basically add a duplicate grinding /flotation line to take capacity to 900,000 tonnes per month, or 10.8 Mtpa.

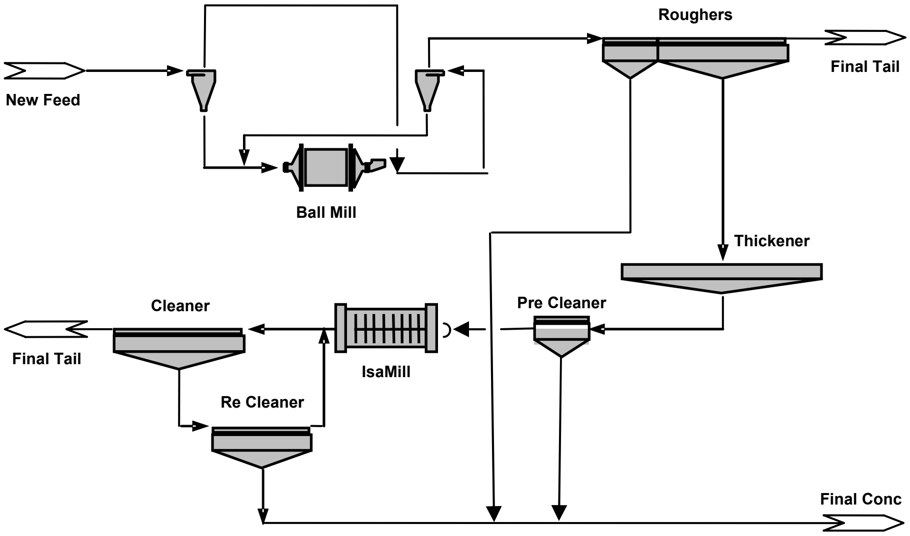


Figure 8 : Simplified WLTRP Flow Sheet

The deslime cyclone under flow reports to the single 7.3m x 10.06m, 10.5 MW (14,000 hp) ball mill in closed circuit with cyclones. The cyclone over flow feeds rougher flotation, which comprises 9 x 130 m^3 tank cells. Rougher tail reports to final tail. The first rougher concentrate is of smeltable grade, and is sent to final concentrate. The remaining rougher concentrate is thickened and then floated in a pre-cleaner prior to regrinding.

A single, open circuit, 2.6 MW (3,500 hp) M10,000 IsaMill grinds the rougher concentrate to a P_{90} of 25μm using local silica sand as grinding media. The IsaMill operates in open circuit and produce a narrow product particle size distribution, because of the internal classification system this technology uses. The ground product is then sent to a two stage cleaning circuit where final concentrate is produced, and cleaner tailings are discarded as final tailings. The cleaner circuit comprises of 2 x 10 m^3 pre-cleaners, 6 x 20 m^3 cleaners and 4 x 10 m^3 re-cleaners.

IsaMill Design and Scale-Up

The WLTRP design required grinding of a nominal 53 th^{-1} and maximum 65 th^{-1}. Test work demonstrated that a reduction of F_{80} = 75 μm to P_{90} = 25 μm required 35 $kWht^{-1}$ using -5 +3 mm silica sand grinding media from a local quarry. As grinding media, the local sand was extremely low cost and demonstrated a consumption rate of 50 g/kWh (1.75 kg/t). The purpose of the scale-up project was to use only one IsaMill. Considering the maximum duty, a 2.6 MW IsaMill with

10,000 litre grinding chamber was designed. This would be the largest fine grinding mill available to the minerals processing industry.

The design represented a significant challenge, as the scale-up of absorbed grinding power was more than 2.5 times the existing design. Historically, the scale-up process of the IsaMill to the 1.1 MW (M3000) model was based on conservation of power intensity. At the 2.6 MW scale, the media agitator (grinding disc) tip velocity would be excessive using this method. A 'constant tip velocity' method was developed which proved more complex than power intensity models, as power does not scale-up linearly with volume.

A variable frequency drive was designed for the first M10,000 installation to reduce process risk and permit testing with fine (regrind duty) and coarse (secondary grinding duty) media types, important to Phase 2 of the WLTR Project.

Component and materials selection was critical, and importantly the disc surface abrasion rate was kept within 3 % of the 1.1 MW mill's rate. As the M10,000 disc design was larger in diameter and thickness, the increased volume of rubber would mean a longer disc life than in the 1.1 MW mill, as abrasion rates are a function of area.

The larger diameter of the M10,000 required modifications to the design of the Product Separator; a centrifuging /pumping device that classifies the mill product (and retains grinding media). As centripetal acceleration decreases with increasing diameter (at constant tip speed), and pumping efficiency decreases with lower radial velocity, both the centrifuging and pumping actions of the Separator had to be re-designed. The new design focused on improving efficiency of the rotor suction region, rotor pump finger shape and product classification efficiency.

The disc tip velocity and power calculations have proven to be accurate. Figure 9 shows the correlation between calculated power draw, and observed power draw at different disc tip speeds during commissioning. Higher tip speed (and power draw) was not required, as the grinding efficiency was high and product size specification was met. Operation of the new Product Separator was successful above 76 % motor speed. Classification efficiency was poor at this speed, however the minimum operating design speed for the M10,000 was 20.5 ms^{-1} or 81 % output. Figure 11 shows how the product particle size distribution of the IsaMill is narrower than that of the feed. This is due to the classification effect of the Product Separator. The IsaMill was operating at 80 % output, or 20 ms^{-1} tip speed at the time of that survey.

Tip Speed m/s	Calculated Power kW	Observed Power kW	Motor Output %
19	1500	1400	76
20	1750	1700	80
21	2026	2000	84
22	2329	N/A	88

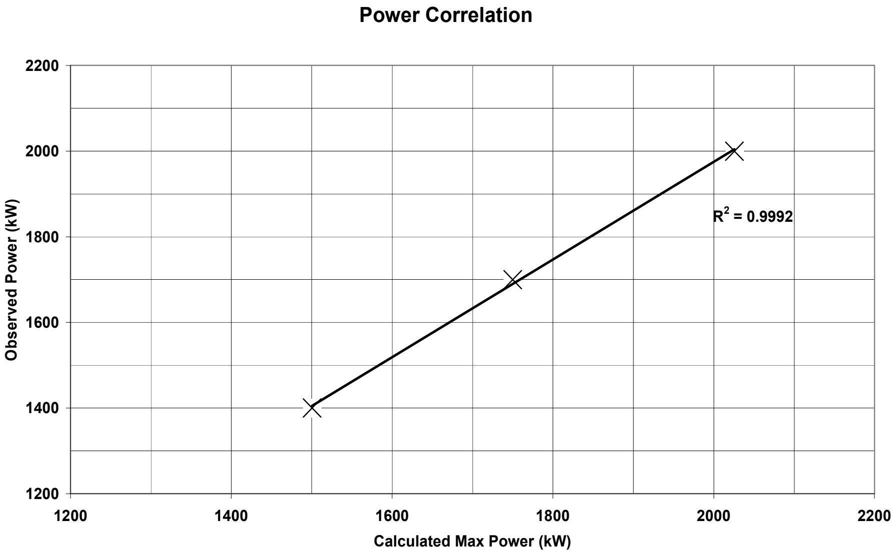


Figure 9 : Power Scale-Up For M10,000 IsaMill

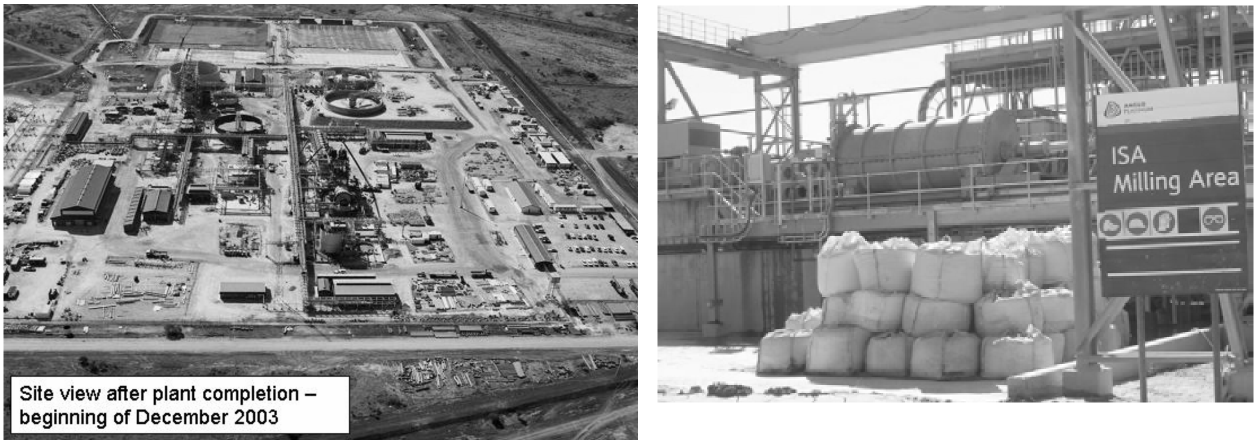


Figure 10 : WLTRP Site And IsaMill

FINE GRINDING AND FLOTATION CIRCUIT OPERATION

Commissioning

The IsaMill circuit was commissioned late December 2003, and began operation in early February 2004 once steady rougher concentrate feed was available. Figure 11 shows the first size survey data across the mill during commissioning. At this time, the feed sizing was finer than design (low primary mill feed tonnage) and of lower pulp density (1.13 kg/L, 17 % solids). As the IsaMill can operate successfully with low density feed, the product specification could still be met (P_{90} = 25 µm). Notably, the product particle size distribution is significantly narrower than the feed, which is a function of the internal classification of this type of mill.

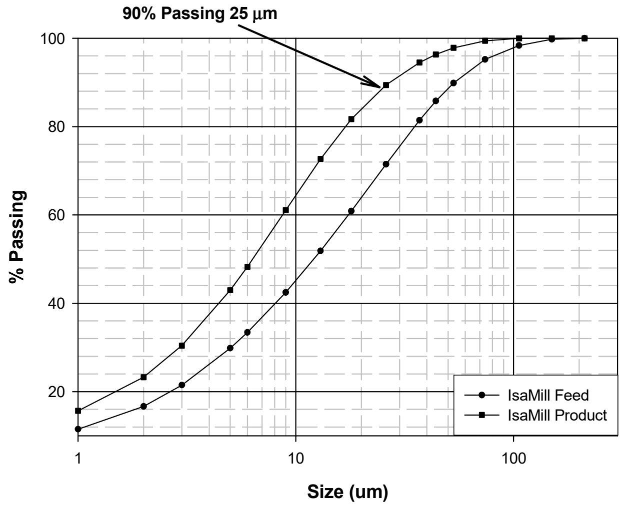


Figure 11 : IsaMill Size Distributions – Commissioning

The cleaner flotation banks had already been commissioned, so a qualitative comparison of froth characteristics before and after IsaMilling could be made. Figure 12 shows the observed difference between the recleaner froths. Without regrinding, the froth was pale, watery and barren; stability was poor and mass pull was very low. After regrinding, the froth was dark grey and heavy in mineralisation. Mass pull was higher and froth stability better.

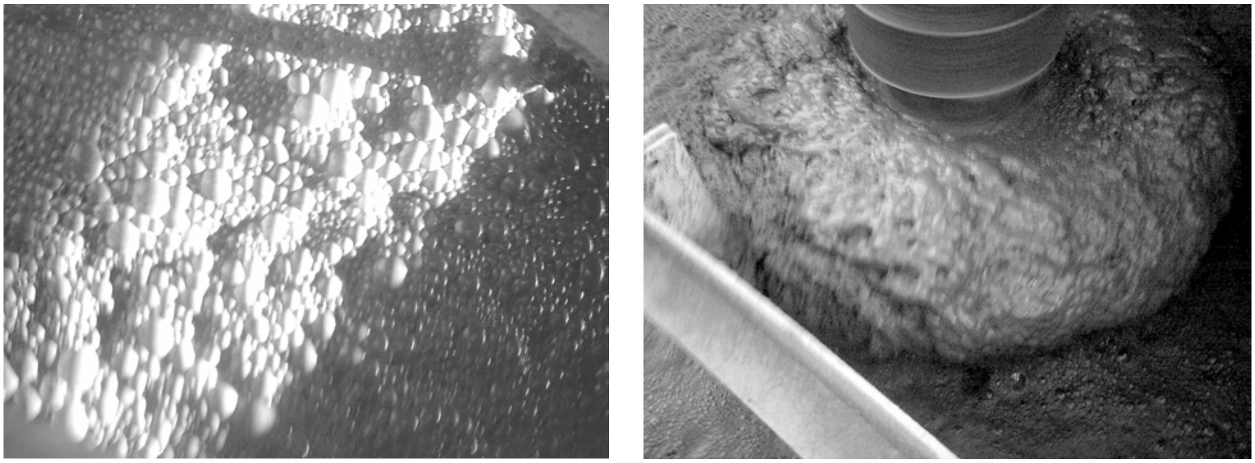


Figure 12 : Froth Appearance Without (Left) and With (Right) Regrind

Current Operation

The main focus after the commissioning of the newly built WLTRP was the optimisation of the main stream circuit as this is the key to ensure optimum PGM recovery from the recovered tailings material. Optimisation of the IsaMill and cleaner flotation circuit will be in full swing when the first phase is completed.

Operational data generated from the IsaMill and cleaner circuit has revealed that the unit is producing a P_{90} of 40µm compared to the design P_{90} of 25µm. The optimisation of this circuit is currently in progress. The first obvious change in the circuit was the increase in mass pull from the rougher circuit from a design of 8% to a current mass pull of between 10 to 12%. This is a 50% increase in feed tonnage to the unit. Consequently, the energy input is 14.5 kWh/t compared to design of 35 kWh/t which explains the coarser grind. Increasing the power draw and specific energy of grinding will form part of the cleaner circuit optimisation program.

Figure 1 indicates the average of the feed and product size distributions taken over a week period. The IsaMill produced a reduction ration of 1.6 from a F_{80} of 40µm to a P_{80} of 25µm. The current feed to the mill is much finer than originally anticipated, so the use of a finer grinding media (to increase efficiency) will also be investigated.

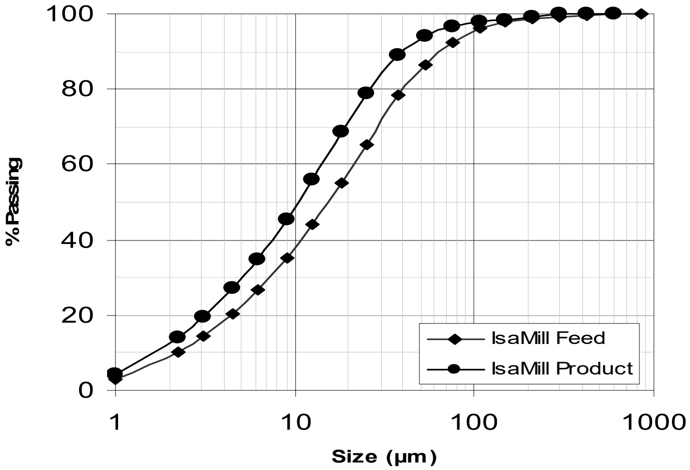


Figure 13: IsaMill Size Distribution

Fine grinding is a new and relatively unexplored field in the platinum industry and optimisation of PGM flotation after fine grinding will be the key to success. Anglo Research Laboratories has initiated several projects to investigate the optimisation of the flotation parameters to ensure maximum benefit from the IsaMill installation.

CONCLUSIONS

- The use of IsaMill technology was enabling for the WLTR project, as it allowed smeltable concentrate grades to be produced from the oxidised, slow floating tailings.
- The collaborative IsaMill design project between Anglo Platinum and Xstrata was successful and is a sound model for technology development.
- Flotation kinetics increased after IsaMilling. The improved flotation performance was due to additional liberation plus the removal of iron oxide surface coatings by grinding in an inert media environment.
- Rougher concentrate IsaMilling was required for the WLTRP, to liberate PGM's and clean-up heavily oxidised base metal sulphide surfaces.
- A staged approach was taken with the WLTRP design to limit capital exposure and minimise risk associated with the novel treatment of dormant PGM tailings.
- The design PGM grade and recovery targets were met using conventional flotation and IsaMill inert grinding technology.
- The IsaMill M10,000 scale-up (from 1.1MW to 2.6MW) was successful.

REFERENCES

Cope, A.J., et al, “Tailings Retreatement Project: Recovery Of 900 ktpm Tailings From Rustenburg Dormant Dams (Klipfontein & Watervaal Only)”, Process Technology Division, Anglo Platinum, 2002.

Durant, A.C., Buys, A.S. and Knopjes, L.M., “The Recovery Of Platinum Group Minerals From Dormant Tailings At RPM – Rustenburg Section. A Pilot Plant Study.”, Divisional Metallurgical Laboratory Report, Anglo Platinum, August 2002.

Frew, J.A., Davey, K.J. and Glen, R.M., “Effects Of Fine Grinding On Flotation Performance: Distinguishing Size From Other Effects”, AusIMM 5th Mill Operators’ Conference, Roxby Downs, October 1994.

Napier-Munn, T., et al, “The Methods And Benefits Of Fine Grinding Ores”, AMIRA Report P336/26, Chapter 7 and 8, March 1994.

Fine Grinding as Enabling Technology – The IsaMill

Pease, J.D. (1), Young, M.F.(2) , Curry, D.C. (1)
(1) Xstrata Technology
(2) Xstrata - Mount Isa Mines

Introduction

The new generation of stirred mills like the IsaMill has fundamentally changed the economics of fine grinding. This has made them enabling technology for several existing and planned operations, and has opened new fields of processing in hydrometallurgy. These opportunities are made possible by the unique combination of features of stirred mills :

- Very high intensity attrition grinding mechanism, suited to fines grinding
- Small media size, essential to increase grinding efficiency for fines
- The use of inert grinding media. This can deliver dramatic improvements to flotation kinetics and recovery, and improved leaching leaching rates and chemistry.

Stirred milling was developed for fine grained ores that required an economic grind to sub 10 micron sizes. The first examples were lead zinc deposits – McArthur River, George Fisher and Mt Isa Blackstar orebodies enabled by the IsaMill, and Century which uses the stirred mill detritor (SMD).

The original application for ultra-fine grained orebodies is a relatively small niche, but it is now clear that there will also be applications in coarser grinding applications, particularly when power efficiency, space, and flotation surface chemistry are important. Two features specific to the IsaMill that make it attractive for coarser grinds are :

- the internal product classifier, which allows low cost open-circuit installations with a sharp product size
- the large unit size (currently up to 2.6 MW) suitable for large scale applications.

Installations at Lonmin Platinum and Anglo Platinum are examples where the IsaMill was chosen for coarser grind applications because of the flotation benefits of inert grinding. The case study of the Anglo Platinum tailings retreatment plant shows that the 2.6MW mill was the enabling technology for the operation.

It is expected that the lower cost of fine grinding will also enable the economics of many leaching technologies operations, eg Activox, Albion Process. For example, in the Albion process atmospheric leaching of otherwise refractory minerals is feasible at fine sizes.

Stirred Milling Technology

Three features of stirred mills that transform the economics of fine grinding are :

- the high intensity attrition grinding environment
- the ability to use fine grained media (eg 1 mm) to suit to the fine grained feed
- the ability to use cheap natural products (local sand, slag, ore) as grinding media

These features distinguish stirred mills as fundamentally different from both ball mills and Tower Mills, as demonstrated by Tables 1 and 2.

	Mill Diameter (m)	Mill Length (m)	Installed Power (kW)	Mill Volume m^3	Power Intensity (kW/m^3)
Autogenous Mill	10	4.5	6400	353	18
Ball Mill	5	6.4	2600	126	21
Regrind Ball Mill	3.2	4.8	740	39	19
Tower Mill	2.5	2.5	520	12	42
IsaMill	1.3	3	1120	3	280

Table 1: Power Intensity of Different Grinding Devices

	Power Intensity (kW/m	Media Size (mm)	No. Balls / m^3	Surface Area (m
Ball Mill	20	20	95,500	120
Tower Mill	40	12	440,000	200
IsaMill	280	1	1,150,000,000	3600

Ball Mill is a 5.6m D x 6.4m L @ 2.6MW
Tower Mill is a 2.5m D x 2.5m L @ 520KW

Table 2: Mill Comparison of Media Size, Power Intensity, number of grinding media

The ability to use smaller media is probably the dominant impact on grinding efficiency. It dramatically increases the grinding surface area and the number of grinding “events”, essential to efficiently grind fine particles. Figure 1 shows the grinding power required to grind a sample of KCGM pyrite concentrate to different target P80 grind sizes, using a ball mill (with 9mm steel media) or an IsaMill with sand media. There is little difference at coarser sizes, but below 30 microns the advantage of stirred milling becomes dramatic. Ball milling simply cannot produce a 10 micron product at any practical power consumption. In this case the IsaMill has extended the economic range of grinding from about 20-30 microns to 10 microns – enabling technology if a 10 micron grind is needed, as it was for the KCGM cyanide leach.

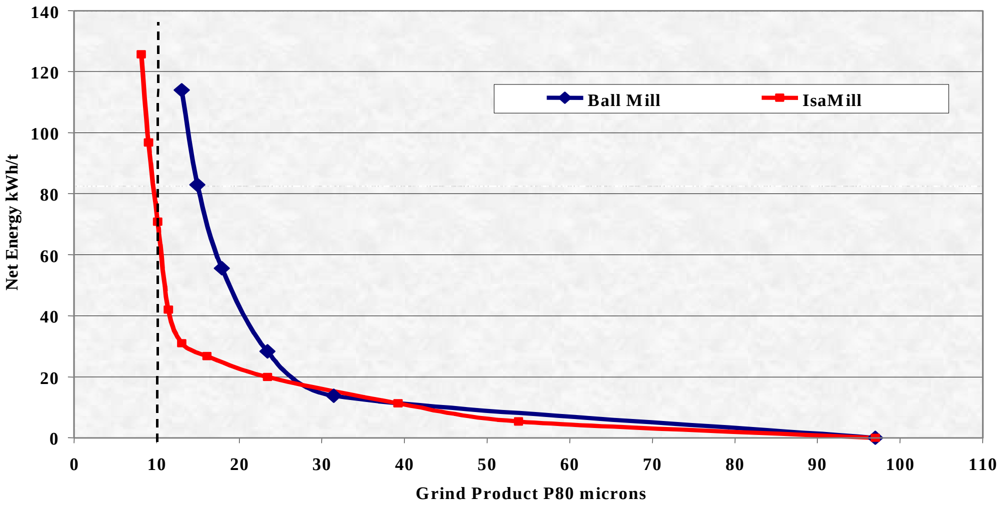


Figure 1 : **Grinding Power to produce various product sizes in a Ball Mill (9mm balls) and an Isa Mill (2 mm sand) (for KCGM pyrite concentrate)**

Chemistry Impacts

The use of inert grinding media gives a crucial advantage to stirred milling in fine flotation and leaching applications. Even if it were economic to grind to 10 microns in a steel mill with very small balls, the amount of iron in solution would almost certainly ruin downstream flotation or leaching processes. The chemical impacts of steel grinding have been well reported (Trahar, 1984; Frew et al 1994; Greet & Steinier 2004), and compete with the benefits obtained from better liberation. Many plant metallurgists still believe that “slimes don’t float”, in spite of the fact that between them, Mt Isa, McArthur River and Century produce over 1.5 Million tonnes a year of concentrate below 10 microns, at high recovery. At Mt Isa recovery in the zinc cleaners is above 95% in all size fractions from 1 micron to 38 micron. At McArthur River, 96% of individual particles recovered are under 2.5 microns (Pease et al 2004).

While high-Chrome media can reduce the chemistry impact, the cost is higher and the impact for fines is only marginal compared with inert media.

Fine Grinding Before Leaching

Unlike flotation, ***leaching*** applications do not suffer as much from the same surface chemistry impacts from steel media. The use of steel media, however, can still be detrimental to a leaching process. When fine grinding pyritic concentrates of precious metals, it is common to follow the fine grinding stage with a pre-aeration stage to remove active pyrite and pyrrhotite before cyanidation. Worn steel media in the ground pyrite can significantly increase the pre-aeration time needed. In a recent application of the Isamill, the existing regrind mill before gold leaching was consuming 10 t/day of steel balls. This reduced pulp Eh and extended residence time in subsequent pre-aeration and increased cyanide consumption in leaching.

Three mechanisms are important when fine grinding before leaching :

- ***the liberation impact*** – in simple cases the grinding is simply to expose fine grained minerals to leachant (eg exposing fine gold to cyanide). In this case dissolution of the host mineral is not needed.
- ***the sizing impact*** – “refractory” minerals often do react, but are passivated by reaction products forming a 2-3 micron “rim” on the particle. For a 30 micron particle this rim prevents the molecular transfer necessary to keep the reaction proceeding deeper into

the particle. But for a 9 micron particle this rim is sufficient for the mineral to disintegrate (Figure 2).

- ***The mechanical activation impact*** - the high energy intensity of fine creates a highly stressed surface, reducing the crystalline nature to amorphous phases (Figure 3). The surface defects act as electron transfer sites, accelerating the rate of surface oxidation reactions, and lowering the activation energy required to oxidise the mineral. This effect of mechanical (or mechanochemical) activation of minerals is well reported (Balaz, 2000; Juhasz and Opoczky,1990; Grelach et al 1989). It means that subsequent leaching of the minerals can take place under much less aggressive conditions, with a reduction in the capital cost of the leach plant.

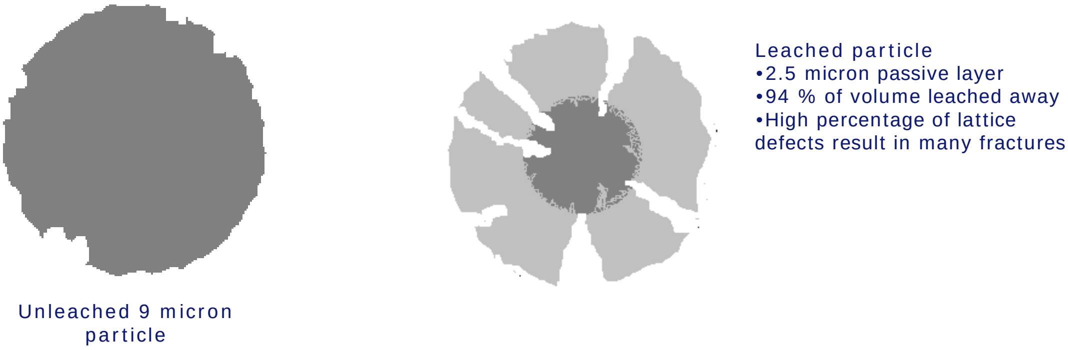
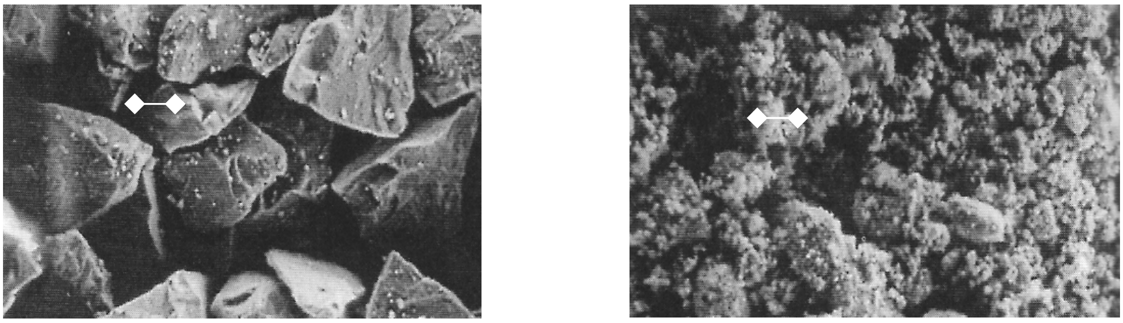


Figure 2: **The sizing impact of fine grinding. For a bigger particle, a 2.5 micron passive layer will prevent further leaching, but for a 9 micron particle it is sufficient for the mineral to be consumed.**



Chalcopyrite Concentrate prior to Grinding

Chalcopyrite Concentrate Following Ultrafine Grinding

Figure 3 : **Impact of intense grinding on surface appearance of Chalcopyrite, the stressed and fractured surfaces on the right leach faster and with lower activation energy, even at the same size particles (from Balaz, 2000).**

In practical situations all three effects of liberation, size reduction and surface activation occur together. Each increases leaching rate but it is difficult to distinguish the relative contributions. However the combined impact can be dramatic – eg in Xstrata's Albion Process, the Isamill grinds and activates minerals to a point where bacteria or pressure are no longer required, and leaching can be carried out in simple open tanks. The extremely high power intensity in the IsaMill compared with other grinding methods suggest it would enhance mechanical activation.

Peculiarities of Fine Grinding – tips for new players

Some aspects of fine grinding are not immediately intuitive to operators of conventional grinding. While there is nothing fundamentally different about small particles, some effects that are minor at coarser sizes become dominant at fine sizes. Some important tips for those designing fines circuits are :

- ***beware the "knee" of the signature plot*** : Figure 1 shows that stirred milling extends the practical range of grinding, but at some point the signature plot still goes "vertical" (the "knee" can be pushed finer by using smaller media). Sometimes clients tell us their target grind size is "about 8 or 10 microns" – but the power to get to 8 microns may be double the power to get to 10 microns.
- ***The importance of consistent sizing technique :*** this follows from the first point. A one micron difference between two sizing machines can change an estimated power draw by 50% ! To compare different grinding devices or media near the "knee" of the signature plot, it is essential that you use the same sizing machine (ideally operated by the same person), otherwise noise in the sizings will overwhelm the results.
- ***The importance of classification :*** every grinding operator knows that sharp classification is important for grinding efficiency. But this is difficult to achieve for ultrafine grinding. A sharp cut at 10 microns needs 2 inch cyclones – but no-one who has ever operated a cluster of 2 inch cyclones in a concentrator will want to do it again. As a result, operators usually choose bigger cyclones, but the operability comes at the expense of grinding efficiency. The solution offered by the IsaMill is to classify within the mill by the centrifugal product separator, which produces a sharper cut than fine cyclones. It also eliminates the extra capital and operating cost of closed circuit cycloning.
- ***Density and Viscosity impacts :*** stirred mills operate at lower pulp densities than conventional mills. The efficiency of the IsaMill is much less affected by density than conventional mills. While efficiency does generally still increase with feed density, the maximum density will be limited by viscosity, and viscosity effects are much more apparent for fine products. Though it is ore dependant, as a general guide sub 10 micron applications will be limited to about 45-50% feed solids.

Power and Energy Efficiency

Grinding energy is one of the major costs of mineral processing. Choosing the right grinding machine and the best media are certainly important. Some other important factors that are sometimes overlooked :

- Energy efficiency should be defined in terms of ***power per unit product*** recovered, not per tonne of ore. Well targeted grinding will improve recovery.
- The ***energy usage of all production steps*** should be considered. This includes energy in the blasting, mining, milling and smelting. Eg, higher concentrate grade will reduce smelting fuel and fluxing needs. It also includes ***the energy content of grinding media*** (eg steel balls versus local slag or sand), and flotation reagents (typically lower consumption after inert grinding).

- The circuit design should aim to apply ***the right grinding power in the right place***, on the smallest possible stream, and avoiding circulating loads (Pease et al 2004; Young et al 1997). The circuit needs to be designed to the size-by-size ore mineralogy.
- ***Efficient classification***. Grinding before flotation or leaching should narrow the size distribution by reducing the top size particles; it should not waste energy grinding already fine particles. A sharp grinding curve, characterised by a low ***ratio of P98 to P80*** is vital. This is demonstrated by Figure 4.

As an example, in the George Fisher circuit 6 MW of additional grinding power was installed, but total energy efficiency was increased. This was because the inert milling increased flotation rates, increased recovery, dropped circulating loads (less pump and flotation energy and less spillage rehandling), and increased concentrate grade (less fuel in smelters). The circuit design of applying efficient regrinding to small streams meant that up-front grind size targets could be relaxed. The energy content of IsaMill media is free (granulated smelter slag). As a result, even with the extra grinding power total unit cost per tonne of ore did not increase, yet grades and recoveries increased significantly (Case Study 2).

Figures 4 and 5 demonstrate why the energy efficiency of the IsaMill is so high. Unlike conventional grinding, the size distribution of the IsaMill product ***sharpens*** with additional grinding. This unique behaviour is because :

- There is no short circuiting – particles have to pass through 8 consecutive grinding chambers, then pass the centripetal field of the product separator before leaving the mill.
- The low volume/high intensity means a short average residence time in the mill (typically 90 seconds). So a particle can travel through the 8 grinding chambers and “see” the product separator within 90 seconds. Fines will exit the mill, coarse particles will be retained.

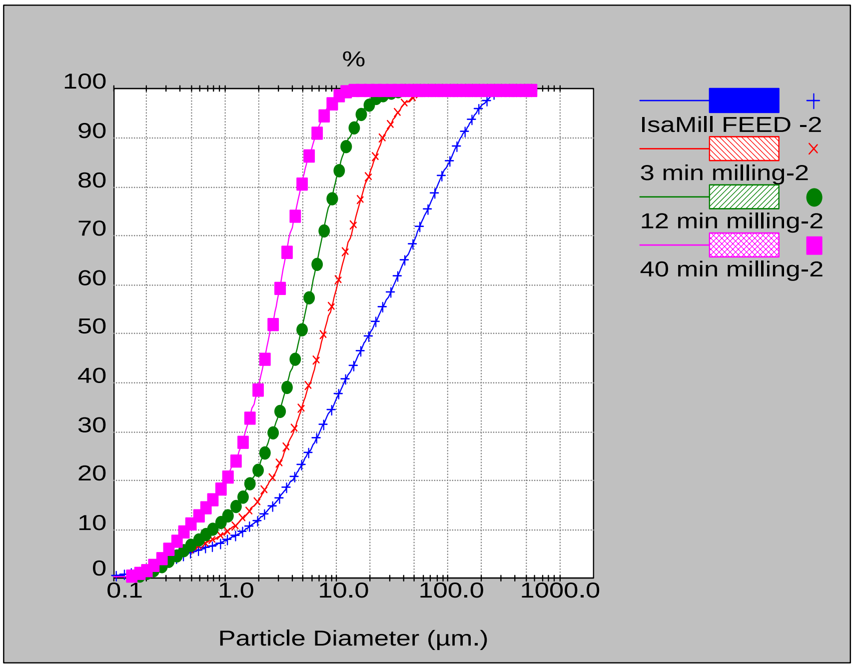


Figure 4: Increasing grinding in an open circuit IsaMill sharpens the product size

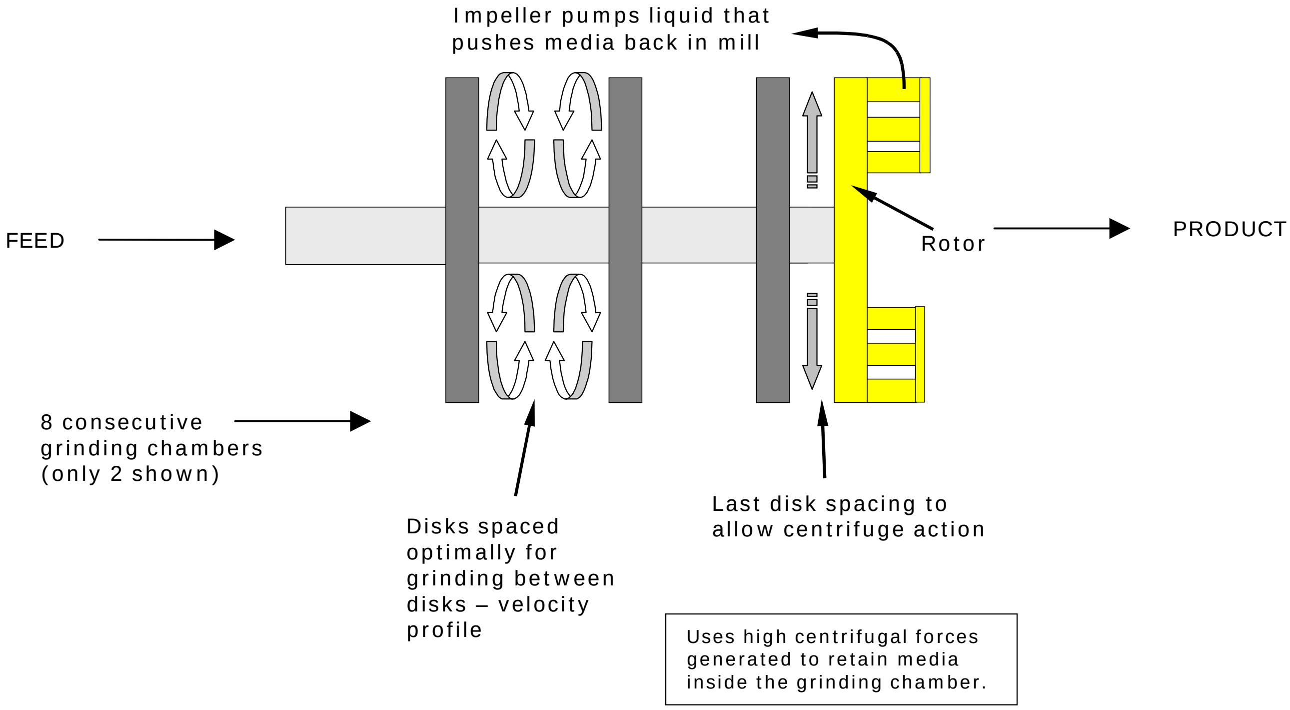


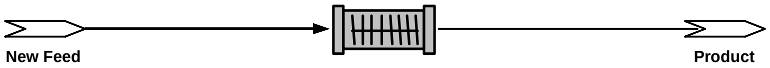
Figure 5: Grinding and Internal Classification Mechanisms in an IsaMill

Recent Developments in Stirred Milling

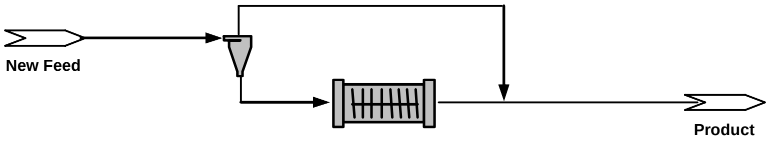
Stirred milling to enable fine grinding and flotation operations is well established - currently 21 MW of IsaMills are installed worldwide, and have produced over 10 Mt of concentrates at Mt Isa and McArthur River alone. Because these applications are in the "niche" of fine grained lead zinc ores, it is easy to overlook the potential for conventional grinding. Recent developments in IsaMills bring some crucial advantages to more conventional grind size:

- improvements in design and materials for ***wear parts*** – eg "slip-in" shell liners. IsaMills routinely achieve availabilities over 97% at McArthur River and Lonmin Platinum (lower at Mt Isa since mills are frequently taken off line to conserve ore).
- Taking advantage of the ***internal classifier in the circuit design***. In early installations (eg MRM) we took a "belts and braces" approach to classification, backing up the product separator with cyclones. In fact, cyclones reduce circuit performance, resulting in a flatter size distribution than produced by an open circuit IsaMill. The ideal IsaMill installation is shown in Figure 6, precyclone mill feed if necessary, but run the mill in open circuit.
- Developments in ***grinding media***. The product separator allows cheap local grinding media to be used (there is no screen to block if some media degrades). For example, Mt Isa operates on waste grannulated smelter slag, MRM ran for 6 years on autogenous ore chips, and Lonmin and Anglo use local sand at $US 0.08/tonne milled. Concurrently, there are new developments in manufactured media – higher cost, but low wear rates and much higher grinding efficiency in some applications. The availability of a standard product, with a choice of media size to suit the application, is important for the stirred mills to be accepted as a mainstream option.
- The successful scale up to the ***2.6MW mill***.

In combination, these developments mean that IsaMills may be a low cost, high efficiency alternative in some mainstream grinding applications. The low cost comes from simple installation – low footprint, low crane heights and loads, no need for closed circuit cyclone installations – an IsaMill installation is fundamentally different from a conventional grinding mill installation.



Recommended Configuration : Open Circuit



If Densification Of Feed Required : Pre-Cyclone Circuit

Figure 6: Recommended circuit configurations for IsaMilling, taking advantage of sharp internal classification.

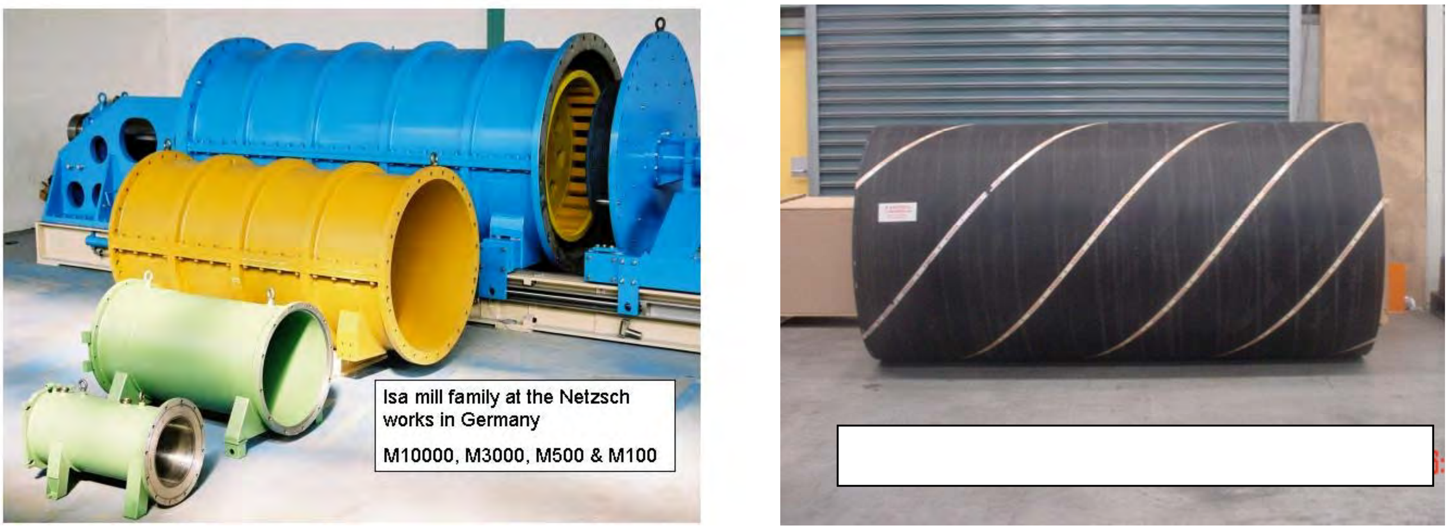
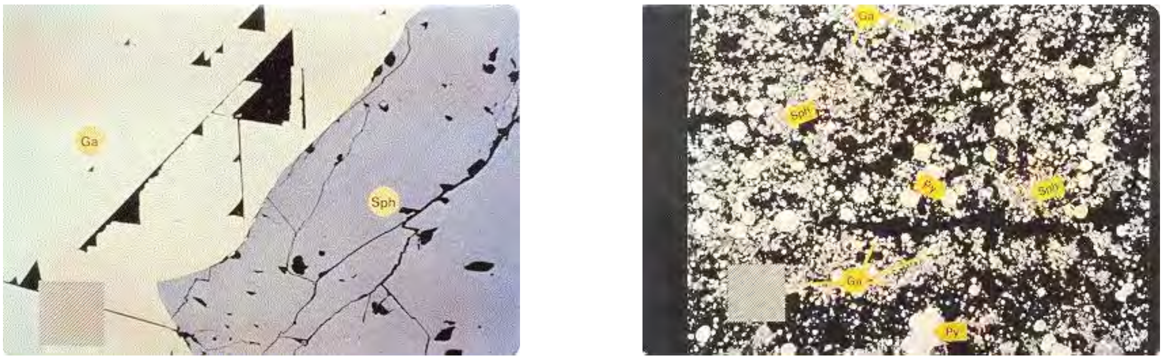


Figure 7: The IsaMill range (left); ‘slip in’ rubber shell liner

Case Study 1 : McArthur River Mining (MRM)

The McArthur River lead zinc deposit was the driving force behind the development of IsaMills. The orebody was discovered in 1955. It had a resource of 227 Mt at 9.2% Zn and 4.1% Pb, however no existing technology could economically treat the extremely fine grained minerals (Figure 8). The development of the IsaMill was truly enabling for this orebody. It allowed economic regrinding to 80% passing 7 microns, fine enough to reduce silica in bulk concentrate to marketable levels. Note that even at this size there is not adequate galena-sphalerite liberation to allow separate lead and zinc concentrates.



Broken Hill Ore McArthur River Ore

Figure 8: Different Grain Size of Broken Hill and McArthur River Ores (Grey Square is 40um)

The plant started mid 1995 with 4 IsaMills regrinding rougher concentrate. Media for the mills was provided by screening a fraction of ore gravel from the SAG mill discharge – a fully autogenous ultra-fine grind ! , Two more mills were installed to increase production and recovery (in 1998 and in 2001). In 2004 the media was changed from ore gravel to screened sand – the higher efficiency of the sand increased mill capacity, and reduced wear on mill components at the higher throughputs.

Table 3 shows production performance at MRM – very high concentrate grades and recoveries are achieved in spite of the ultrafine minerals. This disproves the view that “fines don’t float”. Consider the following perspective : a P80 of 7 micron means a P50 of 2.5 microns at MRM. While 50% of concentrate weight is finer than 2.5 microns, this means that 96% of individual particles recovered are less than 2.5 microns. Since flotation depends on individual particles attaching to bubbles, this means that 96% of the successful particle-bubble collisions at MRM happen for particles finer than 2.5 microns, into a high grade concentrate at high recovery. Fines float very well indeed after IsaMilling.

MINING			METALLURGY			
	Tonnes	**Head Grade**	**Tonnes**	**Zn Recovery**	**Con Grade**	
					%Zn	**%Pb**
1995/96	707,994	12.9%	759,519	66.4%	39.3%	11.2%
1996/97	1,035,222	14.4%	1,026,150	73.5%	43.5%	11.0%
1997/98	1,127,000	16.1%	1,139,000	74.3%	43.3%	11.9%
1998/99	1,222,238	16.4%	1,220,957	79.5%	45.0%	12.7%
1999/00	1,254,227	16.3%	1,262,639	80.9%	46.9%	12.0%
2000/01	1,226,499	15.4%	1,270,319	82.4%	46.8%	11.2%
2001/02	1,398,109	14.9%	1,404,539	82.7%	46.8%	11.1%
2002/03	1,505,306	12.7%	1,511,856	82,4%	46.6%	10.5%
2004	1,523,243	13.2%	1,579,762	80.1%	47.1%	10.4%

Table 3 : Performance of McArthur River since commissioning

Case Study 2 : George Fisher Orebody

The IsaMill technology for MRM was developed in the lead zinc concentrator at Mt Isa. It was clear that the technology would have benefits for the Mt Isa lead zinc orebodies, and it was to prove enabling for the George Fisher orebodies north of Mt Isa. While not as fine grained as MRM, components of George Fisher require a 7 micron grind to achieve acceptable concentrate grades and recoveries.

A circuit was designed to treat the mix of ores from George Fisher, Hilton, and Mt Isa lead zinc orebodies (Young et al, 2000). The circuit included eight 1 MW IsaMills, grinding lead rougher concentrate and intermediate zinc streams as shown in Figure 9. The principles of the circuit design were :

- ***Only grind the minerals you have to*** – this needs a thorough understanding of size by size mineralogical performance throughout the circuit. Recover what minerals you can at coarser sizes, then apply successively finer grinding and flotation stages to recover the finer grained minerals.
- ***Float in narrow size distributions and tailor the flotation conditions to suit*** – this is achieved in the staged grind and float circuit in Figure 9, with separate zinc recovery stages for 37 micron, 15 micron and 7 micron particles. A vital principle is to avoid circulating loads – if particles don't float in the 37 micron circuit, ***don't*** send them back to roughing, send them to regrinding and a custom designed circuit. The circuit may look more complex on paper, but in reality is much simper to operate.

In fact the benefits of the inert grinding and the staged flotation design were so profound that it took us 6 months to appreciate them. Figure 10 shows the immediate 5% zinc recovery gain after installing. This was the gain due to extra liberation of sphalerite, and was all we expected. The "second wave" of even higher benefits happened when we realised that the clean surfaces from inert milling, and the staged flotation circuit, fundamentally changed mineral behaviour. In spite of the finer grind we didn't need more reagent, we needed less. We didn't need more flotation capacity, we needed less – fine minerals floated quite fast in conventional cells when they had clean surfaces and the right reagent conditions. The net impact of the circuit changes was :

- Lead recovery increased by 5% and lead concentrate grade increased by 5%
- Zinc recovery increased by 10% and zinc concentrate grade increased by 2% (in economic terms equivalent to 18% recovery increase at the same grade).
- Unit cost per tonne of ore was unchanged in spite of 6MW of extra grinding power.

Figure 11 demonstrates the combined effect of the staged grinding and cleaning approach – high zinc recovery (+95%) in all size fractions from 1 micron to 38 micron.

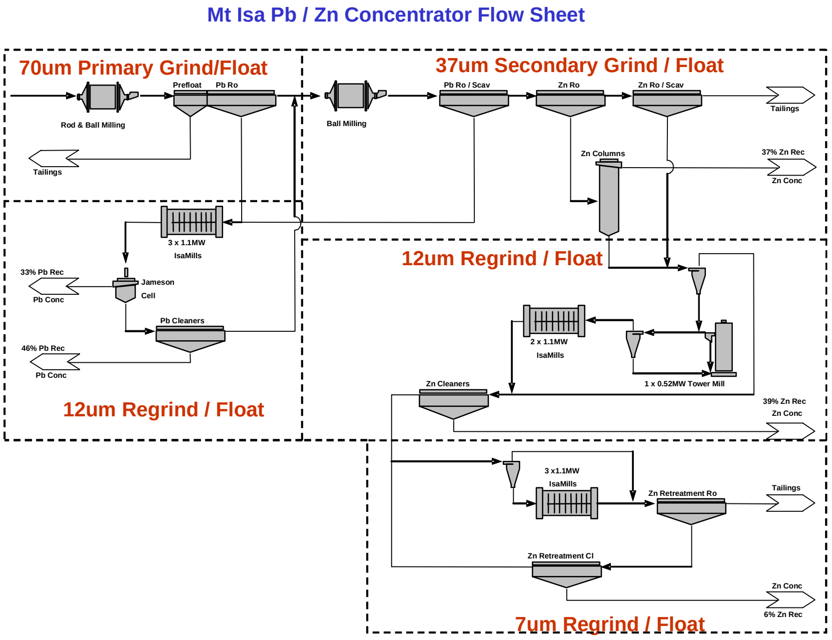


Figure 9: Mt Isa Pb/Zn Concentrator Flow Sheet

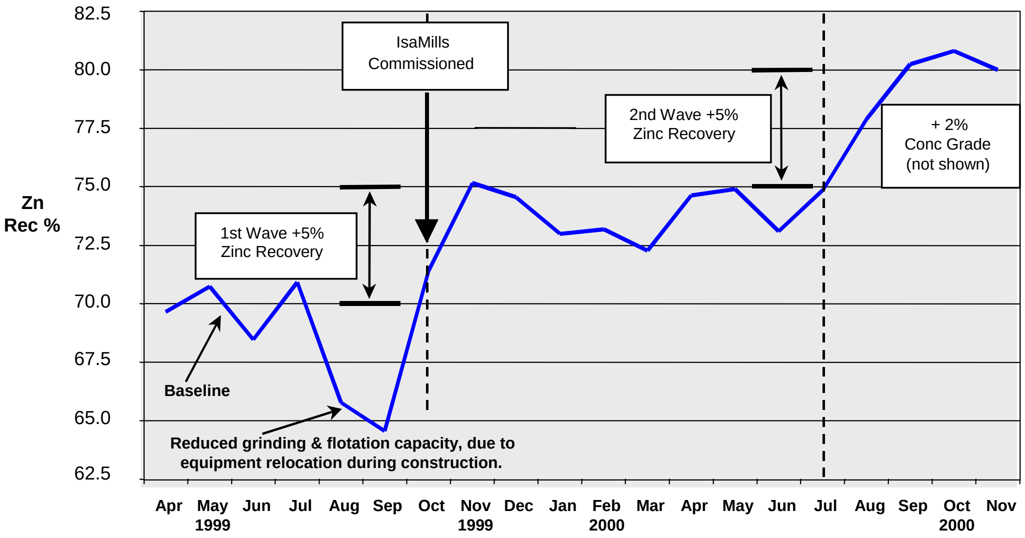


Figure 10: Zinc Recovery Increase from IsaMilling

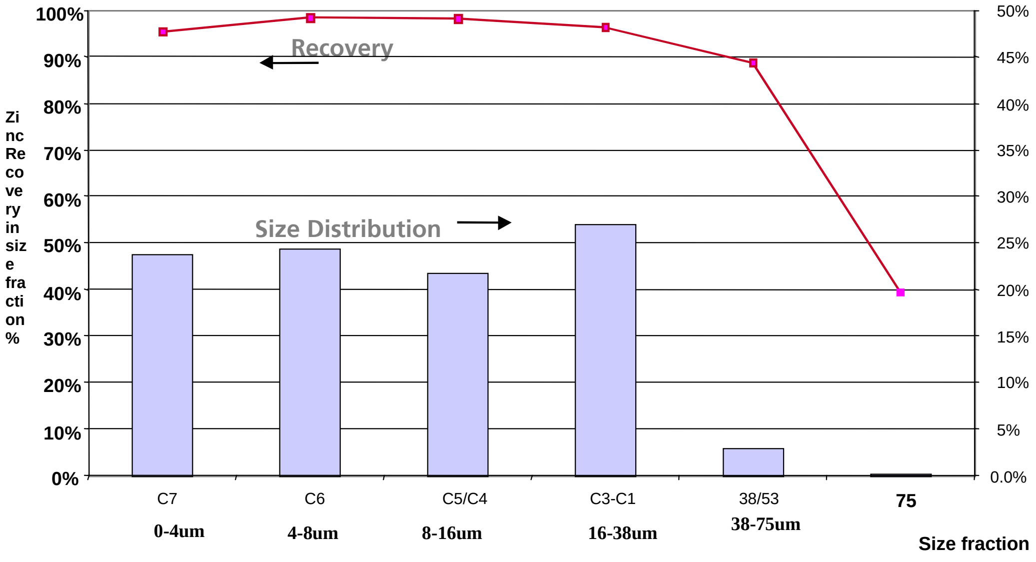


Figure 11: Mt Isa Zinc Recovery from Rougher Concentrate by Size

Case Study 3 : Mt Isa Black Star Open Cut

Surface resources at Mt Isa had long been a target for open cut mining. However the poor metallurgical response was always a barrier to production. Much of the ore is "transitional" between surface oxides and deeper primary sulphides. The transition ore is lower grade than primary ore, has fine grained mineralogy, and leaching has activated pyrite and sphalerite, leading to non-selective flotation. Constant attempts over the last 80 years failed to make the ore economic, with flotation unable to make smelter quality concentrates at any recovery.

The development of the IsaMills and the flowsheet to treat George Fisher ore changed this. The fine grinding achieves mineral liberation and cleans the mineral surfaces by attrition, and the combination of high intensity inert grinding and the correct water chemistry in flotation stops re-activation of unwanted minerals. The impact is shown by the grade recovery curve in Figure 12 - target concentrate grades can now be made at acceptable recoveries.

As a result, the IsaMills were the enabling technology that led to the approval of the Black Star Open Cut project at Mt Isa. Stripping commenced in 2004 and ore mining will commence in the first half of 2005, targetting 1.5M t/y to supplement underground production, produced from a mineral resource of 25Mt at 5.1%Zn and 2.7%Pb. This project represents only a small portion of the potential open cut resources at Mt Isa, the economics of which will also be reassessed once this project has been successful.

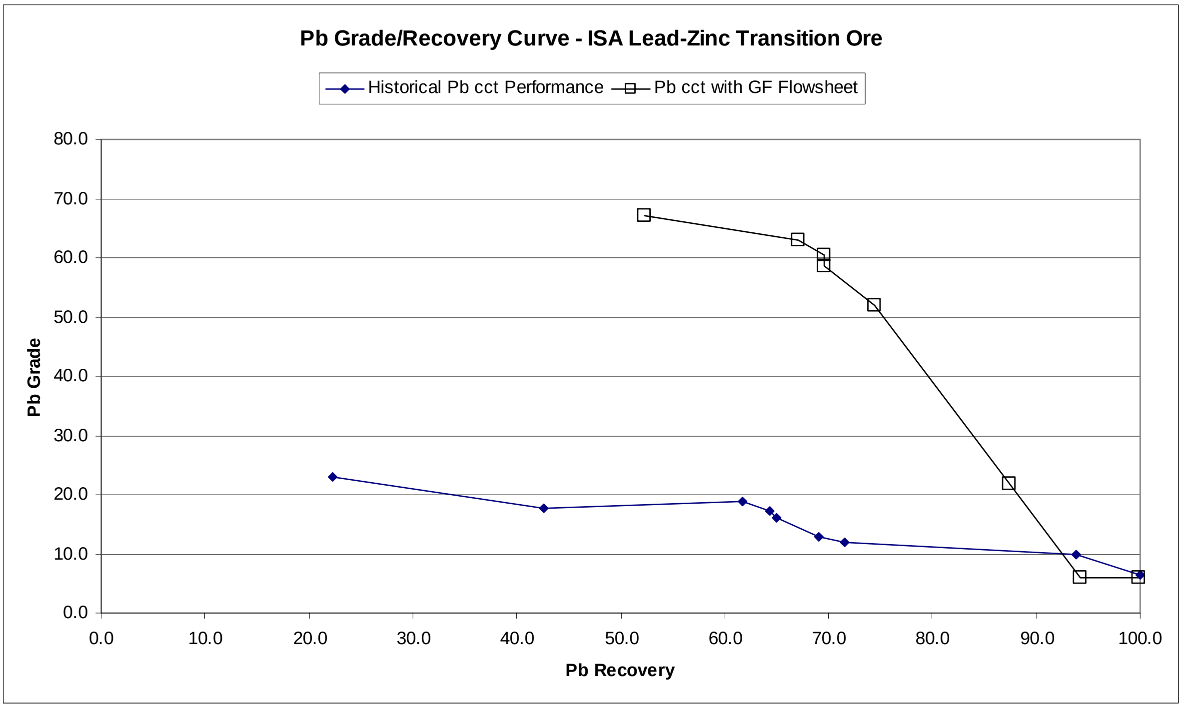


Figure 12: **Pb Grade/Recovery Curve – ISA Lead-Zinc Transition Ore**

Case Study 4 : Merensky Platinum Tailings Retreatment Plant

During 2001 and 2002, Anglo Platinum assessed the retreatment of dormant tailings dams in the Rustenberg area in South Africa. These tailings represented a possible economic resource with the new grinding technology. Two processing issues were:

- The fine grained mineralisation of platinum (why it wasn't recovered first time)
- Surface oxidation and oxidation products which harmed flotation – some of the tailings were placed over 100 years ago.

A collaborative project was undertaken by Anglo Platinum and Xstrata Technology to find an economic treatment. To achieve economies of scale for the project the IsaMill had to be scaled up from 1,000 kW to 2,600kW.

The program was successful, and the large IsaMill proved to be the enabling technology for this project due to :

- The ability to grind fine at low cost – the mill operates in open circuit, and uses cheap local sand as the grinding media.
- The clean mineral surfaces resulting from the inert grinding environment. This was crucial to achieve target grades and recoveries after regrinding.

Figure 13 shows the improvement that IsaMill regrinding makes on cleaning rougher concentrate (Buys et al, 2004). The mill grinds rougher concentrate, thereby reducing power input compared with targeting a fine primary grind. It was found that this also improved cleaner flotation compared with fine primary grinding before roughing. It is likely that the clean surfaces produced in the IsaMill would re-oxidise during the 45 minute roughing period. In contrast, fine grinding immediately before cleaning ***increased*** flotation kinetics in cleaning (in contrast to the common observation that regrinding in a steel mill slows kinetics of all minerals).

Anglo Platinum commissioned the Western Limb Tailings Retreatment Plant in 2004. At the end of 2004 they concluded that (Buys et al, 2004) :

- IsaMill technology was enabling for the WLTR project since it allowed acceptable concentrate grades to be made from oxidised slow floating tailings.
- Flotation kinietics improved after fine grinding due to both extra liberation and the removal of iron oxide surface coatings. Inert fine grinding of rougher concentrate was necessary.
- The scale up to the M10,000 IsaMill (from 1 MW to 2.6 MW) was successful.

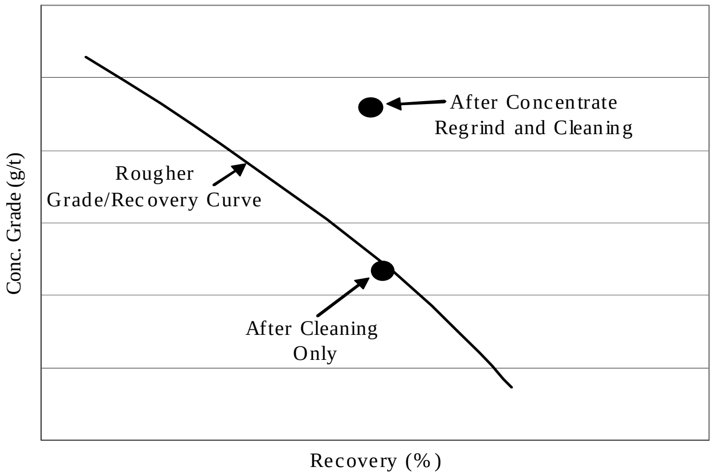


Figure 13: Improvement in Platinum Grade/Recovery After IsaMilling for Western Limb Tailings Retreatment

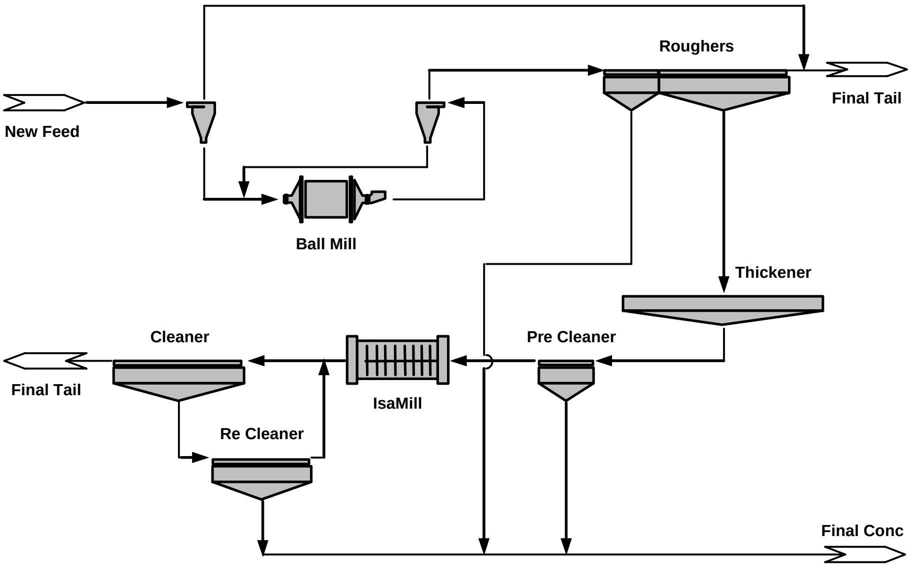


Figure 14: Western Limb Tailings Retreatment Flowsheet

Case Study 5 : Hydrometallurgical Processes and The Albion Process

The ability to efficiently grind minerals to 10 microns is an enabling step for several hydrometallurgical technologies. Fine grinding improves both kinetics and thermodynamics of leaching. The high surface area of fine particles gives high leaching rates at relatively low temperature and pressure, reducing capital and operating costs. Fine grinding also reduces the activation energy required to leach minerals. Several patented processes rely on fine leach feeds, eg the Activox process, the UBC/Anglo process (Driesinger and Marsh, 2002; Hourn and Halbe, 1999), the Phelps Dodge Process (Marsden and Brewer, 2003), and Xstrata's Albion Process (Hourn and Halbe, 1999).

In these processes, metals are leached from a sulphide concentrate by oxidation. Oxidation is typically achieved using ferric iron or oxygen. Fine grinding facilitates the action of both ferric iron and oxygen, making the mineral easier to leach. Fine grinding also ensures that the mineral disintegrates before the leaching surface is passivated by the deposition of leach products.

Fine grinding can also help leach precious metal from sulphide concentrates where oxidation is not required. Preferential breakage of minerals along grain boundary fractures, where occluded gold and silver often accumulate, can significantly improve precious metals recovery.

The Albion Process is a graphic example of the powerful combination of fine grinding plus leaching. Minerals traditionally regarded as "refractory" leach easily at atmospheric pressure and temperatures below 100 degrees when ground finely. Xstrata has demonstrated this process for leaching of sphalerite, chalcopyrite, pyrite, arsenopyrite, stibnite, pentlandite, cobaltite and enargite

Xstrata Technology's 1 t/d Albion Process pilot plant recently operated for 20 months treating McArthur River zinc concentrate, achieving 98% Zn recovery in leaching at 18 - 24 hrs residence time. Leaching was carried out at atmospheric pressure and 80 - 90 degrees. The pilot produced over 30 tonnes of full scale SHG zinc cathodes. Xstrata Zinc is currently undertaking a feasibility study for a full scale plant using this technology.

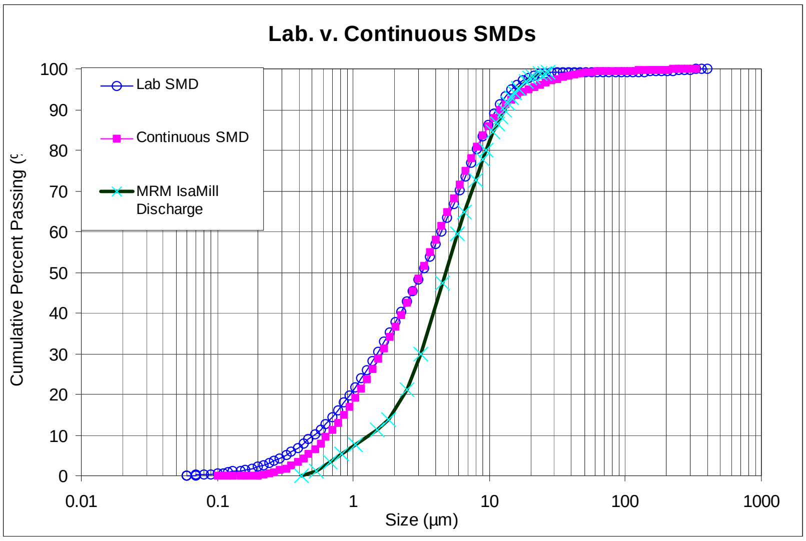
Advantages of the Albion Process are :

- ➢ The ability to treat a lower grade copper or zinc concentrate than conventional roasting or smelting. This may allow a lower ore cut-off grade. For MRM, the ability to leach a low-grade concentrate would increase economically recoverable zinc from 3.6 million tonnes of recoverable zinc to 11.5 million tonnes.
- Compared with conventional zinc refining, the Albion Process avoids either the high pressure autoclave leach, or the roasting step before leaching. The capital cost savings are estimated at $ 1000 per annual tonne of recovered zinc, with the roaster, acid plant, acid storage facilities and concentrate filtration and storage sheds eliminated.
- Either a Goethite or Jarosite stage can be used for iron control. Goethite is favoured since it precipitates as coarse particles, which are easy to filter and enviromentally stable. Any arsenic present in concentrate is fixed in residue as ferric arsenate, a stable phase for tailings impoundment. This could prove enabling for high Arsenic copper ores that cannot be treating by primary smelting methods.
- The simplicity of operating a grinding mill and rubber lined atmospheric tanks means that the Albion process can suit small operations, eg refractory gold operations in remote sites. Such operations would be unlikely to bear the cost and complexity of high pressure or bacterial leaching processes. The rapid leach kinetics typical of Albion Process

residues in cyanide leaching can mean that the Albion leach circuit can be retrofit using existing tanks. In such cases, the only substantial new equipment may be the Isamills.

The mechanisms of leaching fine particles (particle size and surface stress) highlight the importance of the grinding and classification circuit. A high intensity environment like the IsaMill would be expected to create highest surface activation. It is hard to quantify this effect separately, because the IsaMill also produces a much sharper size distribution than other mills, also contributing to higher leaching rate. This impact has not been quantified, because it is hard to distinguish it from the sizing effect. Figure *** above demonstrates the crucial importance of P98 in leaching. A 30 micron particle simply may not leach after the 3 micron passive layer has formed – so a P98 of 30 microns may give 2% lower recovery than a P98 of 19 microns, even for the same P80. Similarly, if the objective of grinding is simply to expose finely disseminated gold to a cyanide leach, then gold disseminated in 30 micron particles will be lost.

Figure 15 illustrates the different size distributions of two large scale industrial applications – IsaMills at McArthur River, and detritors at Century (Reemeyer, 2004). Both operations produce a similar P80, but the IsaMill produce a much finer P98. In flotation this difference is not dramatic, but it is for leaching. There is a vital conclusion for project economics – achieving a sharper size distribution (lower P98/P80 ratio) may mean that the same recovery can be achieved at a coarser grind size (measured by P80). In ultrafine grinding, this may mean a 30% lower power requirement (see Figure 1), or alternatively higher recovery at the same power consumption. This highlights the crucial importance of sharp classification in circuits before leaching. The sharp size distribution from open-circuit IsaMills is because each particle has to first pass through 8 consecutive grinding segments, then has to escape the centrifugal forces of the product separator before leaving the mill.



MRM plant IsaMills:

P80 9.5

P98 19.5

Century plant

P80 8

P98 38

Century Lab scale

P80 8

P98 19.5

Figure 15: Comparison of Sizings from MRM (IsaMill) and Century (SMD)

This effect has been demonstrated frequently in laboratory scale leaching. Table 4 and Figures 16 and 17 show the effect of different grinding technologies on sizing and copper recovery for chalcopyrite leached in the Albion Process. For the same P80, copper recovery varied by 3 % for different grinding machines, due to a combination of different P98 and different surface condition. A measure of the sharpness of the sizing curve is the ratio P98/P80. A value closer to unity means a sharper size distribution, and a higher leach recovery.

% Passing - microns	IsaMill	ECC Mill	GK Vibratory Mill	Metprotech Mill
98	23.1	34.4	42.8	51.90
95	17.44	26.33	30.61	33.40
90	12.31	18.6	19.44	23.41
80	9.11	9.56	10.2	8.95
50	4.85	4.71	5.12	4.10
40	3.76	3.55	4.03	3.31
30	2.66	2.21	2.41	2.31
20	1.94	1.86	2.02	1.96
10	1.42	1.32	1.51	1.61

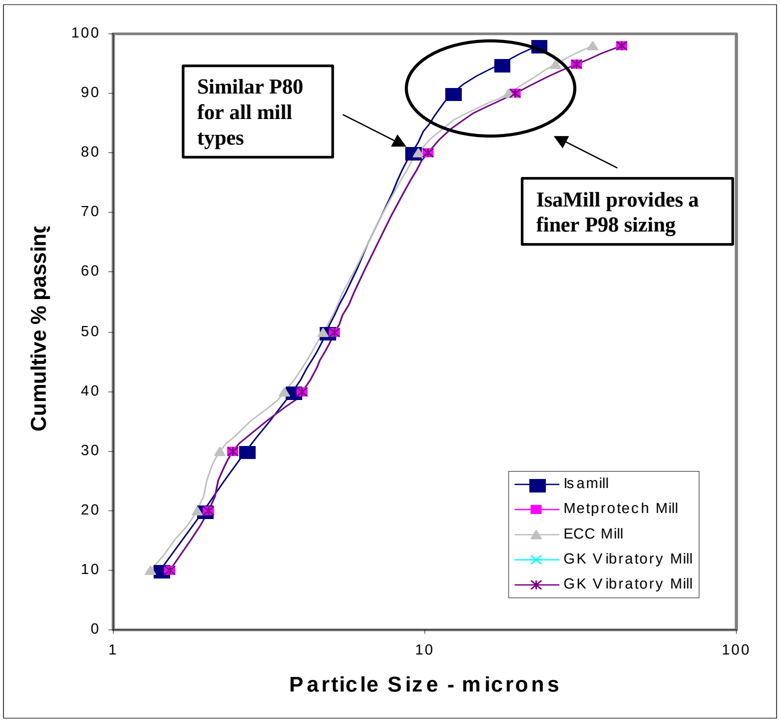


Table 4 and Figure 16:

Particle Size Distributions for Chalcopyrite Concentrate with Varying Grinding Mills laboratory grinding with different mills, with the same feed and same sizing device)

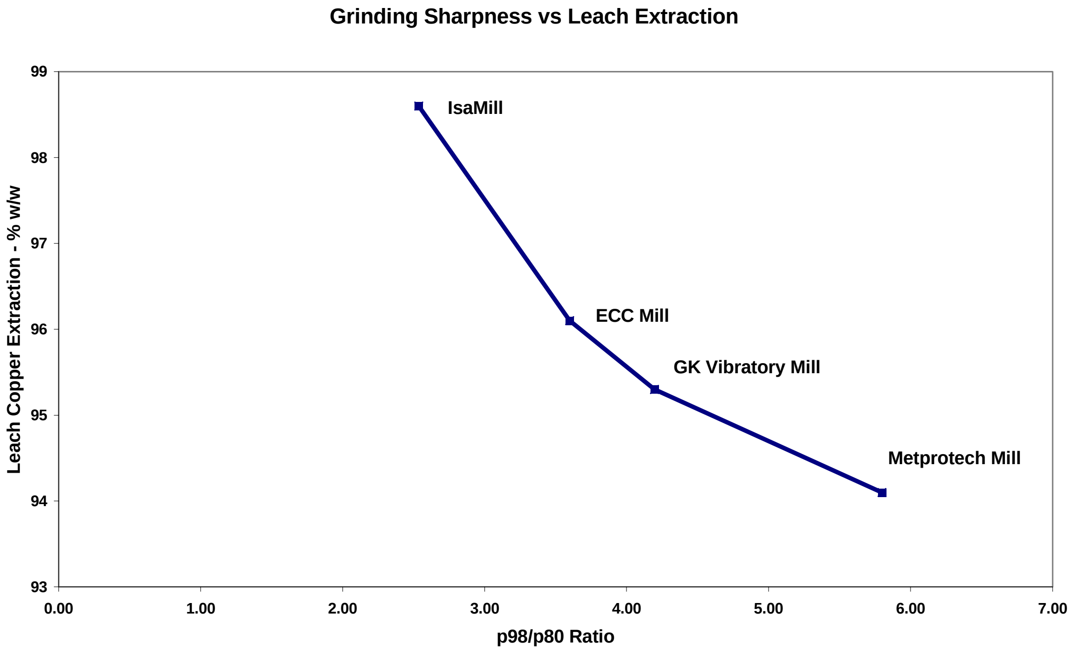


Figure 17: **P98/80 Ratio vs Leached Copper Extraction (Copper Bulk Concentrate)**

Of course, other grinding mechanisms can also produce a sharper cut by placing mills in series to minimise short circuiting, and/or by closed circuit cycloning of the mill discharge. But closed circuit cycloning is expensive and difficult, requiring a cluster of small cyclones to cut sharply at fine sizes. Further, a sharp cut in closed-circuit cyclones usually requires water dilution of cyclone feed – which may then necessitate a cost intensive dewatering stage before leaching. In this case the IsaMill can produce a sharp size distribution with two less unit processes, at lower capital cost, lower operating cost, and better energy efficiency.

REFERENCES

J D Pease, M F Young, D Curry and N W Johnson, 2004, *Improving Fines Recovery by Grinding Finer*, MetPlant 2004

M F Young, J D Pease, N W Johnson and P D Munro, 1997, *Developments in Milling Practice at the Lead/zinc Concentrator of Mount Isa Mines Limited from 1990*, AusIMM Sixth Mill Operators Conference

Trahar, *The Influence of Pulp Potential in Sulphide Flotation, Principles of Mineral Flotation*, The Wark Symposium, AusIMM 1984

M Hourn, D Halbe, 1999, *The NENATECH Process: Results on Frieda River Copper Gold Concentrate*

M Gao and K R Weller, K R, 1993. *Review of Alternative Technologies For Fine Grinding*, AMIRA Project

Juhasz and Opoczky, *Mechanical Activation of Minerals by Grinding*, Ellis Horwood Limited, 1990

Grelach, Gock and Ghosh, *Activation and Leaching of Chalcopyrite Concentrates with Dilute Sulphuric Acid*, Internaitonal Symposium on Hydrometallurgy, 1989.).

Balaz, P, *Extractive Metallurgy of Activated Minerals*, Elsevier, 2000

Driesinger and Marsh, *The Anglo American Corporation/University of British Columbia (AAC/UBC) Chalcopyrite Copper Hydrometallurgy Process*, ALTA Copper, 2002.

Marsden and Brewer, *Hydrometallurgical Priocessing of Copper Concentrates by Phelps Dodge at Bagdad Mine in Arizona*, ALTA Copper 2003

Reemeyer, L, 2004, *Personal Communication*

Frew, J A, Davey K J, and Glen R M, 1994, *Effects of Find Grinding on Flotation Performance: Distinguishing Size from Other Effects*, AusIMM Fifth Mill Operators Conference, Roxby Downs, Australia, 16-20 October 1994, pp 263-270

Young M F, Pease J D, and Fisher K S, 2000, *The Installation of the George Fisher Flowsheet in the Mount Isa Lead/Zinc Concentrator*, AusIMM Seventh Mill Operators Conference, Kalgoorlie, Australia, 12-14 October.

Greet C J and Steinier P, 2004, *Grinding – The Primary Conditioner,* Metallurgical Plant Design and Operating Strategies, Aus IMM, Perth 2004.

Buys S, Rule C, and Curry D, 2004, *The Application of Large Scale Stirred Milling to the Retreatment of Merensky Platinum Tailings,* CIM Montreal.

Improving The Efficiency Of Fine Grinding – Developments In Ceramic Media Technology

Curry D C[1]; Clermont B[2]
[1]*Technical Superintendent - Mineral Processing; Xstrata Technology*
[2]*R&D Project Manager – Comminution; Magotteaux International*

Abstract. The use of ceramic beads as grinding media within high intensity stirred mills (such as the IsaMill) is desirable to maximise the energy efficiency of these processing units. Using a ceramic media with properties tailored to high intensity stirred milling further increases energy efficiency and extends the practical operating range of these mills to coarser feed and product sizes. Historically, the economics of using ceramic media types in stirred mills in the minerals industry has not been attractive. Magotteaux International and Xstrata Technology cooperated in the testing and product development of a new ceramic media, known as Keramax MT1, designed specifically for the minerals industry. Comparative media consumption tests are described and show how the low wear rate and high efficiency of this new ceramic offers an economic alternative to existing media types. The paper proposes the sliding friction coefficient as a new tool in characterising stirred milling grinding media. The first application of this ceramic media will be regrinding a gold bearing, pyrite rich sulphide flotation concentrate in the industry's largest stirred mill; the M10,000 IsaMill.

1 INTRODUCTION

The high energy efficiency of stirred mills compared to ball mills is well understood within the industry. The use of tower mills as an energy efficient alternative to secondary and regrind ball milling became a standard flow sheet inclusion in the latter part of the previous century and is still common today. The modern high intensity stirred mills (such as the IsaMill) further extend the energy benefits of this technology by using higher agitation speeds and smaller media particles [1].

Media selection has a major influence on mill parameters such as energy efficiency, internal wear and operating costs. An inert grinding environment is beneficial to avoid mineral surface degradation and obtain downstream processing and cost advantages [2]. Ceramic media has a profound implication to these parameters and the availability of an economic ceramic media could give significant benefits to the users of IsaMills.

1.1 Grinding Media Types

To date, all IsaMill installations have taken advantage of the technology's ability to use a low cost, but relatively low quality grinding media such as silica sand, river pebble, smelter slag or fine primary mill scats (autogenous milling). Whilst the IsaMill produces high energy efficiencies compared to conventional milling when using these media types, these 'naturally occurring' materials handicap the technology in several ways. The energy efficiency is low compared to what is possible with higher quality media, such as ceramic based compounds. The angular shape and small grain size of the natural media types limit the size of media, and therefore size of feed that can be milled. For example, sand media typically breaks down to its natural grain size when exposed to the high intensity milling environment. Generally sands have grain sizes finer than 5 mm. This limits the feed size that the mill can treat. From a mill wear perspective, it is preferable to use a higher SG media to increase media forces rather than larger, low SG media. The ideal media type for high intensity stirred mills has consistent, reproducible characteristics as shown below, and is further detailed by Lichter and Davey [3]:

- Definite initial charge PSD and top up size
- Chemical composition
- Hardness (related to chemical composition and grain size)
- High sphericity
- High roundness
- Competency (mechanical integrity)
- SG (as designed for machine operation /ore breakage requirements)

A media type that can be manufactured to the ideal qualities is therefore desirable. This reduces power costs and extends the benefits of the technology to treatment of coarser feed sizes. Despite this, the use of manufactured media such as ceramic beads has generally been uneconomic, as the combination of low consumption rate, high energy efficiency and low unit cost have not converged. This has limited the application of such energy efficient grinding technology such as the IsaMill, and restricted its application to regrind and ultra fine milling only.

1.2 Keramax MT1 Development

Magotteaux International has developed a ceramic grinding media specifically applicable to high intensity stirred milling in the minerals industry. In cooperation with Xstrata Technology, the performance and cost effectiveness of the Keramax MT1 grinding media has been tested and verified using laboratory, pilot and full scale IsaMills. The first industrial application combining Keramax MT1 and IsaMill technology will be commissioned late 2005.

2 CERAMIC GRINDING MEDIA OVERVIEW

Two manufacturing processes may be distinguished in the production of ceramic media commonly used in fine grinding for non-contaminating applications:

- Sintered ceramic beads obtained by a cold forming of ceramic powder and by firing in high temperature kilns.
- “Fused” ceramic formed by electric fusion of oxides. The majority of these ceramic beads are named “zirconium silicate”.

Ceramic beads are usually classified according to their chemical composition and physical properties such as bulk density, hardness and fracture toughness. The bulk density has a large influence on the mill power draw. Hardness and fracture toughness could give an indication of the bead’s wear resistance.

The zirconium oxide beads are the highest quality grinding media with the highest initial cost. Keramax MT1 beads have been developed as an economic ceramic media for the mineral processing industry.

2.1 Keramax MT1 Properties

Keramax MT1 is a high density alumina grinding media with consistent microstructure to provide high resistance to wear and high energy efficiency. The media features a smooth bead surface which is ‘pearl’ like to touch. The surface properties suggest that the energy loss in grinding due to friction could be far lower than with other media and that the life time of the internal mill wear components will be increased.

Some low unit cost alumina media types have previously been tested in the IsaMill; both at laboratory and full scale. The consumption rate of these media types was very high (higher than silica sand) due to inconsistencies throughout the bead cross section. Some beads were very soft in the centre and others had air inclusions. The microstructure of MT1 is consistent throughout the cross section.

Table 1: Summary of ceramic bead compounds

Ceramic Beads	Chemical Composition
Alumina	$Al_2O_3 \geq 85\%$ - SiO_2
Yttrium stabilized zirconium oxide	ZrO_2 (95%) - Y_2O_3 (5%)
Cerium stabilized zirconium oxide	ZrO_2 (80%) - CeO_2 (20%)
Magnesium stabilized zirconium oxide	ZrO_2 (97%) - MgO (3%)
Zirconium silicate	ZrO_2 (69%) - SiO_2 (31%)
Aluminum silicate	Al_2O_3 (34%) - SiO_2 (62%)

Table 2: Summary of ceramic bead physical properties

Ceramic Beads	Bulk Density	Hardness (HV)	Fracture Toughness (K1C)
Alumina	2.0 – 2.1	1500-1700	3 - 5
Yttrium stabilized zirconium oxide	3.6 – 3.8	1300-1400	13
Cerium stabilized zirconium oxide	3.9 – 4.0	1100-1200	13
Magnesium stabilized zirconium oxide	3.2 – 3.4	900-1100	6
Zirconium silicate	2.2 – 2.4	600-800	3
Aluminum silicate	1.7 – 1.8	800-900	3

Table 3: Keramax MT1 compound

Ceramic Bead	Chemical Composition
Keramax MT1	Al_2O_3 (79%) – SiO_2 (6,5%) – ZrO_2 (14%)

Table 4: Keramax MT1 physical properties

Ceramic Bead	Bulk Density	Hardness (HV)	Fracture Toughness (K1C)
Keramax MT1	2.3 – 2.4	1300-1400	5 - 6

Table 5: Keramax MT1 cross sectional hardness

Location	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**Ave**	**SD**
HV	1308	1301	1308	1351	1322	1294	1366	1322	1396	1294	1326	34

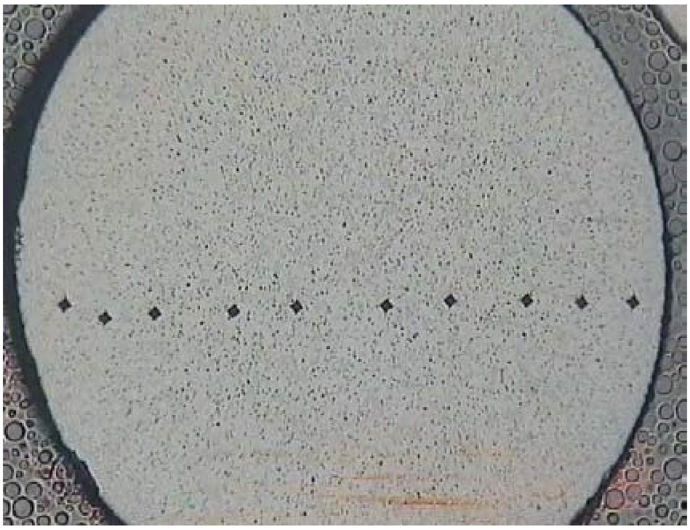


Figure 1: Keramax MT1 cross section – hardness measurement points 1-10, left to right

3 SURFACE PROPERTIES OF GRINDING MEDIA

Grinding media selection is usually focused around the parameters listed in 1.1. In testing of MT1 during product development, it was found that grinding efficiency was greater than other media types of similar size, shape factors and SG. In fact, grinding efficiency was better than could be predicted with stress intensity calculations. Of significance was the power trend that occurred during test work under the conditions of media-water and media-slurry. In all other media types ever tested by Xstrata Technology (including ceramics), an IsaMill will draw more power when operating with a media-water system than with a media-slurry system (ie motor power would always decrease when slurry was introduced to a mill operating with grinding media and water – in a start up situation for example). With MT1, the reverse occurs. It was hypothesized that slurry 'lubricates' a typical media charge as the frictional loss between media particles in motion is lower in the presence of fine feed slurry than media on media interaction in the presence of only water. Further to this, and considering the unique reverse trend with MT1, it was proposed that the media on media interaction of MT1 in a water environment resulted in minimal frictional loss, compared to what now would be considered an abrasive slurry environment with slurry being fed to the mill.

The appearance of the MT1 surface and observing how easily the media flows certainly supported the theory of lower frictional loss. Xstrata's Discrete Element Method (DEM) model was used to verify the sensitivity to energy distribution and power draw of surface friction coefficients.

3.1 Media Surface Sliding Friction Coefficient

Surface rolling friction (μ_r) does not have a measurable effect on energy distribution, shaft torque or media velocity and force, so was fixed at μ_r = 0.005 which is consistent with other DEM simulation work [4].

Two sliding friction coefficients (μ_s) were used in this simulation to illustrate the difference between media of mid and low μ_s. These were μ_s = 0.01 and 0.30 which compare well to measured μ_s of < 0.10 for MT1 and > 0.30 for silica sand and river pebble. All other media parameters were held constant.

Figure 2 shows the steady state flow patterns of media at a cross section near a disc surface, located at the shaft mid-point along the axial direction of the IsaMill. The colour represents the velocity of particles. Figure 3 shows the steady-state force distribution of the same media with colour representing total forces acting on media particles. The simulations use a mono-sized media charge of 4 mm diameter, in a 4 litre chamber and ignore the effect of slurry, so physical interactions are solely a function of media specification.

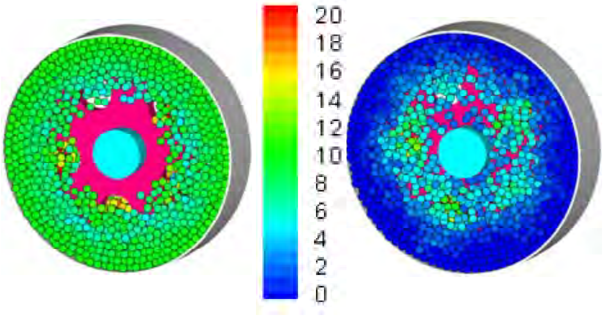


Figure 2: Snapshots of flow with left μ_s = 0.01, right 0.30 (units are 0.198 ms^{-1})

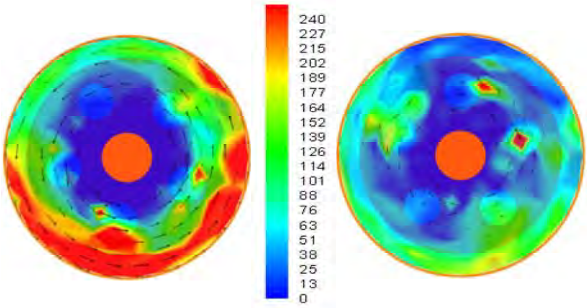


Figure 3: Snapshots of force distribution with left μ_s = 0.01, right 0.30 (units F = mg)

The velocity in both cases is highest near the disc hole meaning that the disc surface properties and holes are responsible for 'lift' of the media. Media with low sliding friction move to the periphery and have higher velocity. Media with high sliding friction stay closer to the shaft and possess lower velocity. The transfer of energy between the disc and media for particles with low sliding friction is more efficient.

Particles with low sliding friction have much larger forces than those with high sliding friction. Grinding should consume less energy when media types having low sliding friction coefficients are used.

Figure 4 shows the resultant torque acting on the IsaMill shaft for the two different media types. The IsaMill took longer to reach steady state with the low μ_s media but the stable torque was 14,000 torque units compared to 21,000 torque units with the higher μ_s media. In this case, the IsaMill operating with the higher μ_s would consume 50 % more power than the lower μ_s case. Again, this demonstrates that an IsaMill operating with a low μ_s media charge would have better energy efficiency.

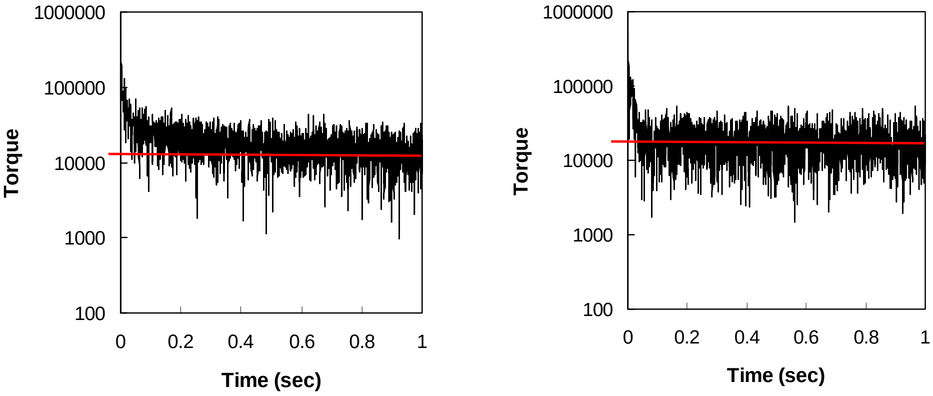


Figure 4: Total torque for left μ_s = 0.01, right 0.30 (units are 3 x 10^{-6} Nm)

4 TEST WORK

4.1 Laboratory

The aim of laboratory tests was to obtain comparative results of different grinding media in terms of relative consumption rate and grinding efficiency. Combining the energy and consumption data produces a kg/t consumption figure for each media type. This would present the first evaluation of the economic potential of MT1 in IsaMilling applications.

Tests were performed in Netzsch LME4 (IsaMill M4) machines at both Magotteaux's Belgian facility and Xstrata's Brisbane laboratory. This section describes test work performed on a gold bearing pyrite concentrate from the Eastern Gold Fields region of Western Australia.

The simplified test procedure is described below:

- Media is pre-conditioned in water before testing on slurry to simulate a conditioned charge. Media is dried, weighed and re-loaded in the mill. (Note the MT1 media charge did not lose any mass after 60 minutes of grinding in water).
- Pyrite concentrate slurry with a F_{80} = 170 µm is pumped through the LME4 in a single pass of ± 10 minutes. Energy, flow rate and pulp density are measured during this time.
- A sample is taken for particle size analysis using a Malvern Microsizer.
- This process is repeated a minimum of three times to produce size /energy data pairs.
- The test rig is then placed in closed circuit with the slurry stock, and operated for a further 60 minutes to maximise the accuracy of energy measurement and media consumption.
- The loss in mass is measured at the end of the test. Dividing this mass by total net energy, gives a g/kWh media consumption rate.
- A size value (eg P_{80}) is plotted against its respective energy value and a measurement of grinding efficiency is obtained (a Signature Plot).

The consumption rates of different media types on the pyritic concentrate, under identical grinding conditions are shown in Table 6. Keramax MT1 exhibited the lowest consumption rate, giving a relative wear ratio of 1. The relative wear ratios of the other media types are shown in the right hand column.

The size distribution of the media types (with exception of the nickel slag) were not representative of fully conditioned charges; most media types only had particles in the -4 +3 mm range as this was how they could be ordered from the suppliers. The absence of -3 mm media would lead to inefficient grinding to ultra fine sizes. Because of this, a coarser target product size was selected where all media types could produce a product of acceptable efficiency. A P_{80} = 47 µm was selected, as it met the above criteria for the narrowly sized media charges, and also presented as the only data point produced for Alumina 2 (due to a pump failure during testing, more data points were not produced for this media type). Therefore, the energy efficiency of all media types are compared at a P_{80} = 47 µm and shown in Table 7. All tests were performed in open circuit.

The next best media to MT1 required 50% additional energy to produce the same product size. Significantly, this media was an Australian silica sand which required less energy than the two alumina ceramic media types. The silica sand was however, coarser than either of the alumina media which would result in higher wear rates of internal mill components, so maintenance costs would need to be evaluated.

Using the available media consumption and energy data, the media consumption per tonne of concentrate treated can be calculated and is shown in Table 8.

It is clear that MT1 has economic potential in IsaMill applications, as the consumption per tonne milled in this test is at least an order of magnitude less than the next best media. The higher energy efficiency also has capital cost benefits, as a lower installed power can be selected to optimise project economics. For example, Table 9 shows the net power requirement for this concentrate type to treat an arbitrary 250 t/h. Once a specific project's power cost is factored in, the relative performance of MT1 further improves. Based on this outcome, further work at pilot scale was planned.

Table 6: Grinding media consumption rates - per unit energy

Media Type	Consumption Rate (g/kWh)	Relative Consumption
MT1 (-4 +3 mm)	15	1.0
Alumina 1 (-4 +3 mm)	128	8.5
Alumina 2 (-4 +3 mm)	295	19.7
Australian River Pebble (-4 +3 mm)	200	13.3
Australian Silica Sand (-6 +3 mm)	781	52.1
Ni Slag (-4 +1 mm)	1305	87.0

Table 7: Grinding media energy efficiency from F_{80} = 170 µm to P_{80} = 47 µm

Media Type	Specific Energy (kWh/t)	Relative Energy
MT1 (-4 +3 mm)	7.6	1.0
Alumina 1 (-4 +3 mm)	13.1	1.7
Alumina 2 (-4 +3 mm)	12.4	1.6
Australian River Pebble (-4 +3 mm)	27.9	3.7
Australian Silica Sand (-6 +3 mm)	11.2	1.5
Ni Slag (-4 +1 mm)	17.8	2.3

Table 8: Grinding media consumption rates - per tonne treated

Media Type	Consumption Rate (kg/t)	Relative Consumption
MT1 (-4 +3 mm)	0.11	1
Alumina 1 (-4 +3 mm)	1.68	15
Alumina 2 (-4 +3 mm)	3.66	32
Australian River Pebble (-4 +3 mm)	5.58	49
Australian Silica Sand (-6 +3 mm)	8.77	77
Ni Slag (-4 +1 mm)	23.23	203

Table 9: IsaMill net power requirement – illustration at 250 t/h treatment rate

Media Type	Net Power (MW)
MT1 (-4 +3 mm)	1.9
Alumina 1 (-4 +3 mm)	3.3
Alumina 2 (-4 +3 mm)	3.1
Australian River Pebble (-4 +3 mm)	7.0
Australian Silica Sand (-6 +3 mm)	2.8
Ni Slag (-4 +1 mm)	4.4

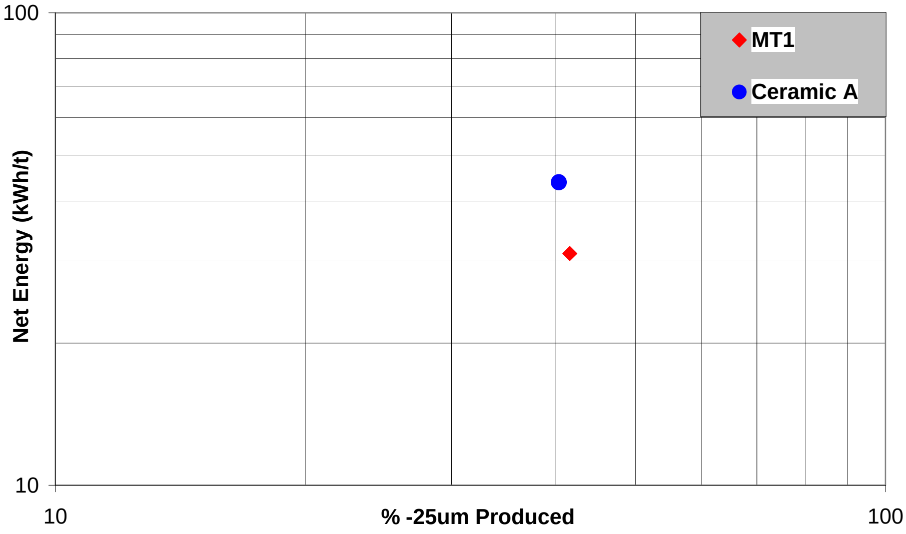


Figure 5: Specific energy vs size

4.2 Pilot

Pilot testing was conducted at a South African platinum concentrator treating MF2 tail (ie secondary ball mill /flotation circuit tailing) on UG2 ore in the Western Limb of the Bushveld Igneous Complex near Rustenburg. This site was selected because a full scale IsaMill is already operating in cleaner circuit regrind on silica sand which would permit full scale verification of the pilot results. Also, the M100 pilot IsaMill (55 kW) was located near this site which is an ideal mill for this work.
Four media types were selected, however only two were able to grind the extremely hard chromite in the plant tail. The two media types that failed to grind the feed were a silica sand and alumina ceramic (both having SG = 2.6 t/m^3). Neither of these media types are available in coarse enough sizes to break the chromite in the feed. 'Ceramic A' was of similar SG to MT1.

The simplified test procedure is described below:

- Equivalent volume of media is loaded into the IsaMill.
- The power draw is recorded, and maintained by pumping fresh media in with the feed slurry.
- Media consumption is calculated by recording the mass added during the test and by measuring the mass variation of the load before and after each test.
- Composite samples are taken every day for particle size analysis.
- All the operating conditions such as slurry flow rate, pulp density and mill speed are identical for each test.
- Each media type was tested for 5 days to generate sufficient data.
- The IsaMill was running in a continuous mode in a single pass operation.

Figure 5 plots the operating points of the IsaMill with the two media types during the test (specific energy against % passing 25 µm). Table 10 summarises the consumption and specific energy data per % -25 µm produced.

Ceramic A required 46 % more energy to produce the same product size, whilst consuming over 400 % more media in the process.

During the MT1 test, some media was crushed in the gland seal area of the mill which had worn during the Ceramic A test. The seal arrangement was subsequently modified to a similar design as the full scale mills. However, it was suspected that the observed 20 g/kWh consumption rate was high. This could be confirmed during full scale tests.

4.3 Full Scale

At the time of writing this paper, two full scale IsaMill tests with MT1 at different sites were underway. The sites agreed to perform full scale testing, based on the encouraging results from the laboratory and pilot test work. It is planned to publish the test results in the future, however the following conclusions can be made:

- The laboratory energy determination and media consumption rate tests scale up accurately.
- MT1 consumption in the UG2 ore test was inflated due to crushing of media. The actual consumption is < 10 g/kWh in the full scale test.

Table 10: Specific energy and media consumption rate

Media Type	Specific Energy (Per % -25 μm)	Consumption Rate (g/kWh)
MT1 (-3.5 +1.8 mm)	74	20
Ceramic A (-3.4 +1.7 mm)	108	81

Table 11: Kumtor – relative consumption and motor power draw @ P_{80} = 12.5 μm

Media Type	Relative Consumption	Relative Motor Power
MT1 (-2.2 +1.8 mm)	1.0	1.0
Ceramic B (-2.4 +1.4 mm)	4.6	1.3
Australian River Sand (-2.4 +1.2 mm)	9.2	2.2

5 FIRST INDUSTRIAL APPLICATION

The first application of MT1 and large scale IsaMilling is at Centerra Gold's Kumtor operation in the Kyrgyz Republic. Kumtor is the largest gold operation in Central Asia producing over 500,000 oz pa. MT1 media and the large M10,000 IsaMill enabled the highest energy efficiency possible on a small foot print. Due to the variation in (pyritic) rougher concentrate mass pull, a high intensity stirred mill with good turn down was required; the IsaMill is designed to operate between 1.3 and 2.6 MW power draw by varying the amount of media inside the mill.

The IsaMill will process an existing regrind mill product to ultra fine sizes of P_{80} < 12.5 μm.

Several media types were evaluated, including ceramics and silica sand. The lower energy efficiency of silica sand would have required the additional capital cost of a second mill. The media evaluation is summarised in Table 11.

6 CONCLUSION

Historically ceramic type grinding media has not been economically viable for mineral processing applications with high speed stirred mills such as the IsaMill, due to high operating costs. The availability of an economic ceramic media will provide considerable benefits to the users of IsaMills, and allow the energy benefits of the technology to be applied at coarser grind sizes.

Keramax MT1 ceramic grinding media was tested on various mineral concentrates and offers very high energy efficiency and extremely low consumption rates. The surface properties of MT1 contribute to the high energy efficiency of IsaMilling with this media type. DEM simulations explain the more efficient distribution of energy when using a grinding media with low sliding friction coefficient.

MT1 offers the design metallurgist a way of lowering capital and operating costs for large scale stirred milling projects. The foot print of IsaMill installations can be decreased because of the lower installed power requirement and smaller media handling system.

The Kumtor gold project will be the first to demonstrate the effectiveness of combining a high efficiency grinding media with large scale stirred milling.

REFERENCES

1. Gao, M., Young, M. & Allum, P. (2002) *IsaMill Fine Grinding Technology And Its Industrial Applications At Mount Isa Mines.* 34th Annual Meeting Of The Canadian Mineral Processors, Ottawa, Canada.

2. Pease, J. D., Curry, D. C. & Young, M. F. (2005) *Designing Flotation Circuits For High Fines Recovery.* AusIMM /SME Centenary Of Flotation Symposium, Brisbane, Australia.

3. Lichter, J. & Davey, G. (2002) *Selection and Sizing of Ultrafine and Stirred Grinding Mills*. SME Mineral Processing Plant Design Symposium, Vancouver, Canada.

4. Yang, R., Jayasundara, C., Yu, A. & Curry, D. (2004) *Numerical Simulation of the Flow of Grinding Medium in IsaMill*. 8th ICBMH Conference, Wollongong, Australia.

Collaborative Technology Development – Design And Operation Of The World's Largest Stirred Mill

Curry D C[1], Clark L W[2] and Rule C[3]

[1]*Technical Superintendent - Mineral Processing; Xstrata Technology*
[2]*Business Manager - Mineral Processing; Xstrata Technology*
[3]*Head of Concentrator Technology; Anglo Platinum Limited*

Abstract. Since late 2003, Anglo Platinum has been operating the world's largest stirred mill at their Western Limb Tailings Retreatment Project near Rustenburg in South Africa. The mill is a 2.6 MW, M10,000 IsaMill which was a development of the 1.1 MW M3,000 machine. Only 16 months was required from project kick-off to commissioning – a considerable undertaking in view of the large scale-up step. A novel collaborative development approach was taken between Anglo Platinum and Xstrata as a vehicle for robust technology advancement.

The IsaMill has operated successfully since commissioning in terms of product size, scale-up and mechanical availability. The M10,000 used new designs to reduce internal component wear rates compared to previous IsaMill designs and the relative operating costs of the large IsaMill models (M3,000 and M10,000) are discussed. Laboratory, pilot and full scale operating data is used to validate the success of this installation.

1 INTRODUCTION

Anglo Platinum commissioned the first large scale tailings retreatment facility in the South African platinum industry in December 2003. The facility (known as the Western Limb Tailings Retreatment Project – WLTRP) uses modern concentrator technology to recover PGM's from dormant tailings dams.

Pilot test work demonstrated the need for inert media regrinding of rougher concentrates before cleaning, to produce a smeltable concentrate grade. IsaMill technology was selected to regrind rougher concentrate to 90 per cent passing 25 µm. Due to the scale of the WLTRP, the regrind power requirement was significant at 2.2 MW absorbed. For this duty, Anglo Platinum and Xstrata developed a larger IsaMill; the M10,000.

2 COLLABORATIVE DEVELOPMENT OF TECHNOLOGY

2.1 Introduction

IsaMill Technology was first developed at Mt Isa in the mid 1990's when a major research program was undertaken by MIM as it was then known (now Xstrata) to enable fine grained ore bodies to be economically developed. The first installation of the technology was at the Hilton Concentrator in 1992. This machine was an M500 with a 205 kW motor. The machine that was ultimately developed as an economic full scale machine was an M3000 IsaMill, with a 1.12 MW motor.

Initially, two M3000 IsaMills were installed in the Mount Isa lead/zinc concentrator in 1994. This was then followed with four M3000 IsaMills in the McArthur River concentrator in 1995 [4]. These were required to achieve ultrafine grinding for liberation of fine grained minerals and then achieve good separation of the minerals by flotation, after ultrafine grinding.

The subsequent development of the 2.6 MW M10,000 in late 2002 meant that in just 10 years since the first unit was installed in an operating plant at Hilton Concentrator, the units have increased their capacity 13 fold from 205 kW to 2.6 MW and their volume has increased 20 times.

Further to this, and the focus on this section of the paper, the design, fabrication, installation and commissioning of the project for the capacity increase from 1.12 MW (M3000) to 2.6 MW (M10,000) took only 16 months; an incredibly short time frame.

The project was initiated during a range of meetings between Xstrata Technology (or MIM Process Technologies as it was then) and Anglo Platinum. The scale up was always contemplated by both Netzsch and Xstrata as a future plan, even during the early development of IsaMill at Mt Isa, however it was brought forward as a focused development by a need of Anglo Platinum. The project had an aim to develop a larger IsaMill that has a higher capacity than two M3000 Isamills for a similar capital and operating cost.

2.2 Driving Forces behind the Development

The reasons behind each company being involved in this collaborative development were all different. Anglo Platinum were trying to reduce capital cost for equipment and foresaw potential applications of large scale stirred milling requirements for whole of ore applications in their concentrators in the future. Anglo Platinum also had a current project with a specific application that deemed it necessary for an IsaMill with over 2 MW of grinding power.

Xstrata Technology had considered this development to be a natural expansion of the marketed product range. There were also a number of internal projects that were under consideration at the time which envisaged use of larger scale IsaMills. To have the technology developed in this way under a collaborative development with an external client was clearly seen as bringing benefit to Xstrata. Finally, for Xstrata the application of IsaMills was limited due to the size of the largest existing available machine (1.12 MW). For large concentrators with whole of ore, or high tonnage regrind applications, having a significant number of smaller M3000 machines was not considered desireable from a capital, operating and maintenance perspective.

For Netzsch, the long established IsaMill manufacturer, the appeal was more from a challenge to their technical expertise and improved demand for their products.

2.3 Collabarative Development Process

The Process used for the design of the M10000 development was very unique. It was a major departure from all previous technology developments which were performed in house within Xstrata, and given the developments underway within MIM at this time it was quite incredible that it actually happened.

The process used was to first crystallise the need of the customer. Considerable time and effort by both Anglo Platinum and Xstrata Technology was involved examining the application, the requirements for regrinding and the potential of IsaMill to handle the duty. The result from this was the “Defined Application”. The second step was for Xstrata Technology and Netzsch to examine the potential for the M3000 to be scaled up to meet the needs of the “Defined Application”. This was conducted by initially Xstrata and Netzsch to individually review the requirements and then come together for a meeting to discuss and agree on the scale up mechanism, and the recommended machine size.

The final part of the collaboration involved Anglo Platinum undertaking an Engineering Review of IsaMill Technology and this involved a visit and audit of Netzsch’s manufacturing and engineering facilities in Germany.

These three process steps resulted a signing of The Collaborative Development Agreement on October 5th 2002 between Anglo Platinum and Xstrata.

3 DESIGN

3.1 Project Description And Pilot Testing

Anglo Platinum investigated the retreatment of dormant tailings dams in the Rustenburg area in 2000. Given market value of platinum and palladium and US dollar exchange rates at the time, the concentration of PGM minerals in the dams represented a major economic resource. Metallurgical test work identified that a significant proportion of these minerals could be recovered via fine grinding and flotation.

The pilot scale program was developed to confirm concepts previously identified during laboratory scale testing [2]. The laboratory tests suggested that the recovery of PGM’s from the tailings dams was sensitive to the grind size presented to flotation. Pilot tests confirmed this, showing PGM recovery in rougher flotation was 10 % higher with a primary grind of 90 % passing 75 µm, than with 80 % passing 75 µm. Significantly, flotation kinetics increased with the finer grind and the additional liberation of PGM’s and associated base metal sulphides.

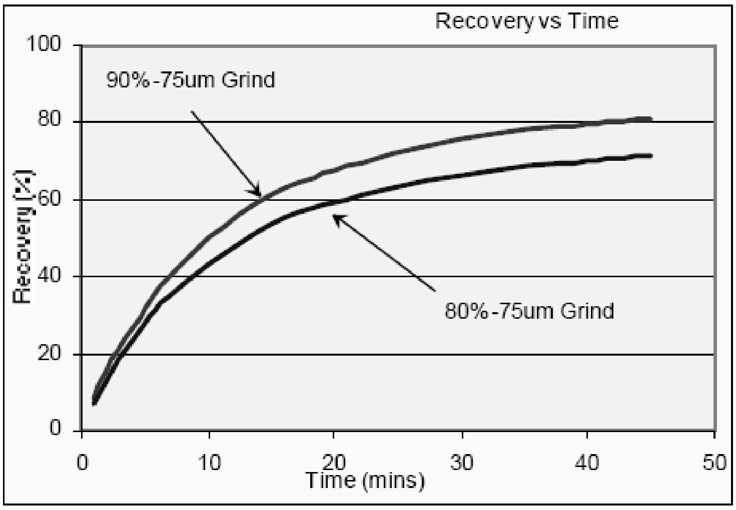


Figure 1: Flotation kinetics as a function of grind size

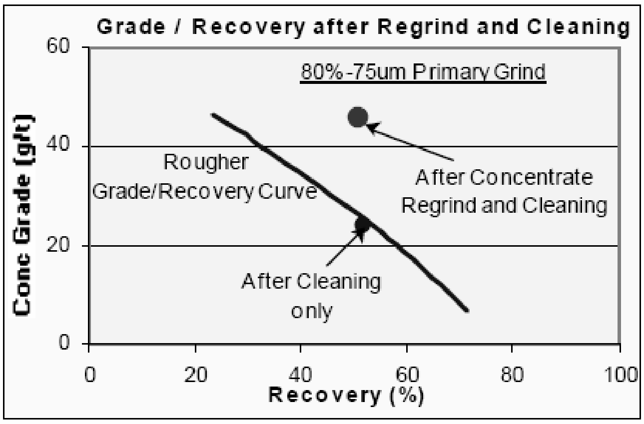


Figure 2: Conventional cleaning vs inert regrind and cleaning

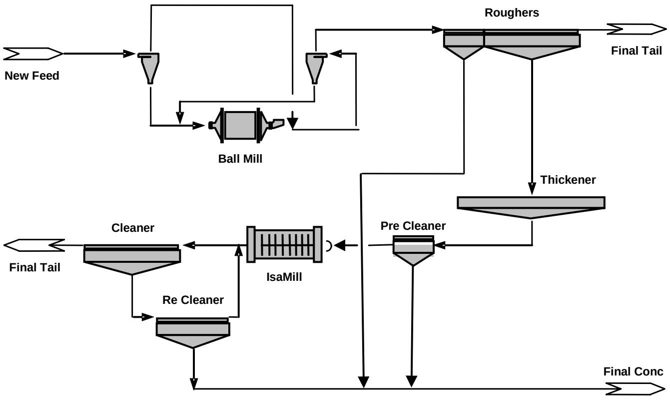


Figure 3: Simplified WLTRP flow sheet

Mineralogical examination showed poor upgrade potential of the rougher concentrate, due to insufficient liberation and oxide coatings on base metal sulphide surfaces. A conventional cleaner /recleaner circuit was not able to shift the grade /recovery curve, so regrinding of the rougher concentrate was employed to improve PGM liberation and freshen-up the heavily weathered mineral surfaces.

The main findings from the pilot plant results were [2]:

- The PGM grade and recovery targets could be met with the use of conventional flotation and IsaMill inert grinding.
- A primary grind of no less than 80 % passing 75 µm, and rougher concentrate regrind of no less than 85 % passing 25 µm is required to meet the grade /recovery targets.

IsaMill technology was enabling for the WLTR project, as smeltable concentrate grades could be produced from the oxidised, slow floating tailings.

3.2 Application

The WLTR concentrator has a capacity of 4.8 Mtpa, and has been designed to easily expand to 10.8 Mtpa. The flow sheet includes recovery of tailings by high pressure water monitoring, ball milling, rougher flotation, rougher concentrate regrinding and cleaner /recleaner flotation [1]. The process design called for an IsaMill nominal feed rate of 53 th^{-1} and maximum 65 th^{-1}. The production of a P_{90} = 25 µm required 35 $kWht^{-1}$ using -5 +3 mm local silica sand. The IsaMill product is fed to a two stage cleaning circuit to produce final concentrate [3].

3.3 IsaMill Design

Phase 2 of the WLTR Project (pending) requires the IsaMill to treat a far coarser feed size. Because of this, and to reduce process risk with the first installation, a variable frequency drive was installed.

The previous method of scale up (to the 1.1 MW M3,000 size) was based on conservation of power intensity. With the larger M10,000, the media agitator (grinding disc) tip velocity would be excessive using this method. A ‘constant tip velocity’ method was developed which proved more complex than power intensity models, as power does not scale-up linearly with volume.

The larger diameter of the M10,000 required modifications to the design of the Product Separator; a centrifuging /pumping device that classifies the mill product (and retains grinding media). As centripetal acceleration decreases with increasing diameter (at constant tip speed), and pumping efficiency decreases with lower radial velocity, both the centrifuging and pumping actions of the Separator had to be re-designed. The new design focused on improving efficiency of the rotor suction region, rotor pump finger shape and product classification efficiency [1].

Component and materials selection was vital to the operating cost competitiveness of the new mill. A key design aspect was to control disc wear, and the design proposed that the disc surface abrasion rate was within 3 % of the M3,000 mill's rate. As the M10,000 disc design was larger in diameter and thickness, the increased volume of rubber would mean a longer disc life than in the M3,000, as abrasion rates are a function of area and the power density per unit area would be lower.

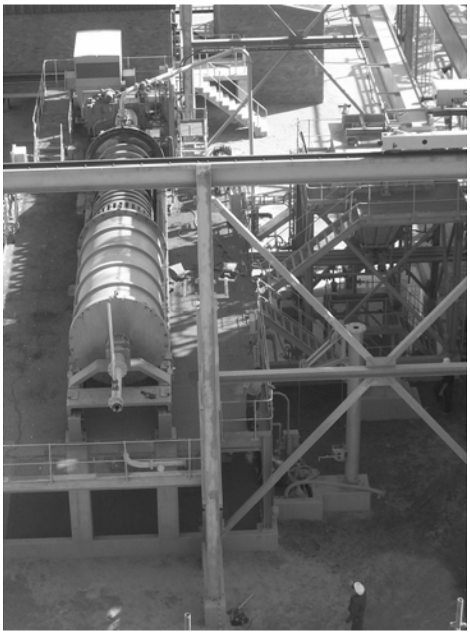


Figure 4: 22 x 11 m footprint of WLTR M10,000 IsaMill grinding circuit

4 OPERATION

4.1 Scale Up

The theoretical power draw can be calculated for a given volume, media filling and disc tip speed. With a fixed speed mill, the accuracy of the power calculation can only be verified at a single point. The power based scale up calculations could be verified over a wider range than usually possible, due to the use of a variable speed drive with the WLTRP M10,000. At the time of testing the power scale up, the IsaMill power draw had been reduced by operating at a lower media volumetric filling of 56 % (calculated). The IsaMill feed tonnage was well below design, so the mill was turned down to avoid over grinding. This explains why the motor power draw in Figure 5 is lower than normal. The actual motor power draw at different mill speeds is within a 2 % error band of the scale up calculation.

Test work on similar Anglo Platinum Merensky tailings material was conducted at laboratory scale in 2001. A range of tests were performed, at varying feed sizes and energy inputs. The grinding data for a laboratory test is compared to survey data of the M10,000 at similar energy input (Table 1). Allowing for minor differences in the ore samples, the milled products should exhibit similar particle size distributions as the mineralogy is broadly consistent and the same grinding media (-5 +3 mm local crushed silica sand) and specific energy were used.

The M10,000 exhibits marginally better top size control, and a narrower overall product PSD. The CSI (Coarse end Size Index - or P_{98} : P_{80} ratio) is 2.6 for the M10,000 whereas the CSI for the laboratory mill is 3.0.

The M10,000 scaled up accurately, both in terms of power and product PSD. Power efficiency appears equal to the laboratory mill, with control of the product PSD marginally better.

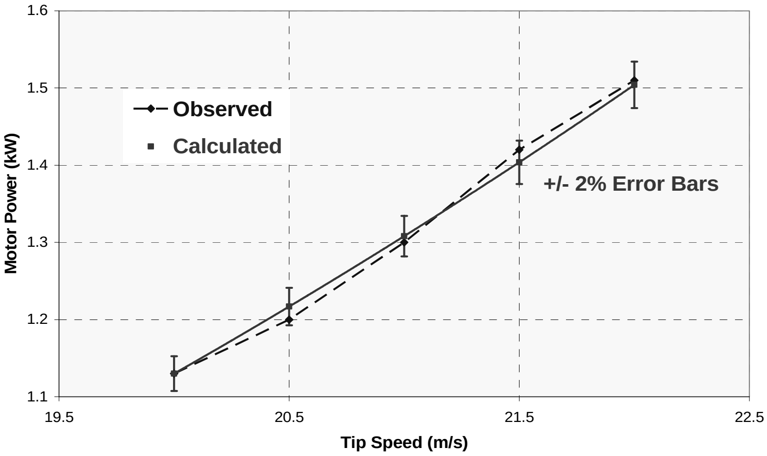


Figure 5: Theoretical vs actual motor power at varying disc tip speed

Table 1: Laboratory vs Full Scale Grinding Efficiency

IsaMill Model	Installed Power (kW)	Chamber Volume (L)	Specific Energy (kWh/t)	Pulp % Solids	P_{98} (μm)	P_{80} (μm)
M4	4	3.5	37	39	47.5	16.0
M10,000	2,600	10,000	37	42	42.5	16.5

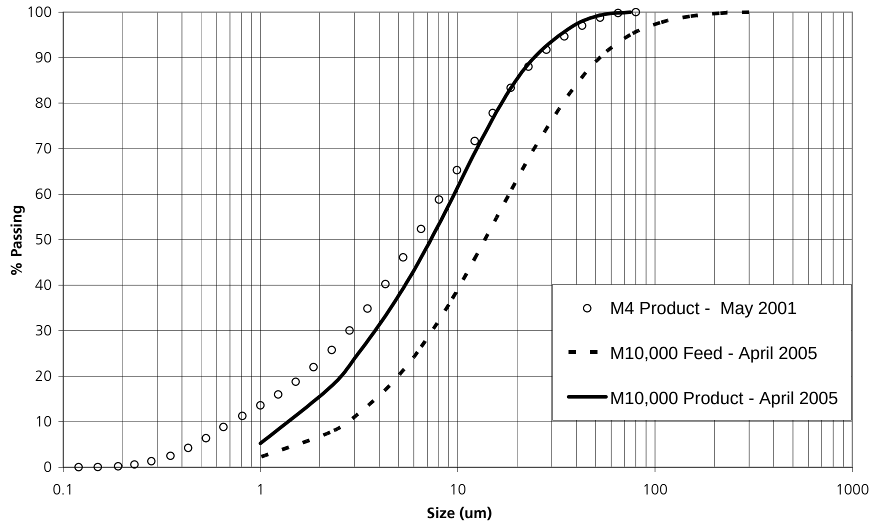


Figure 6: M4 vs M10,000 Product PSD

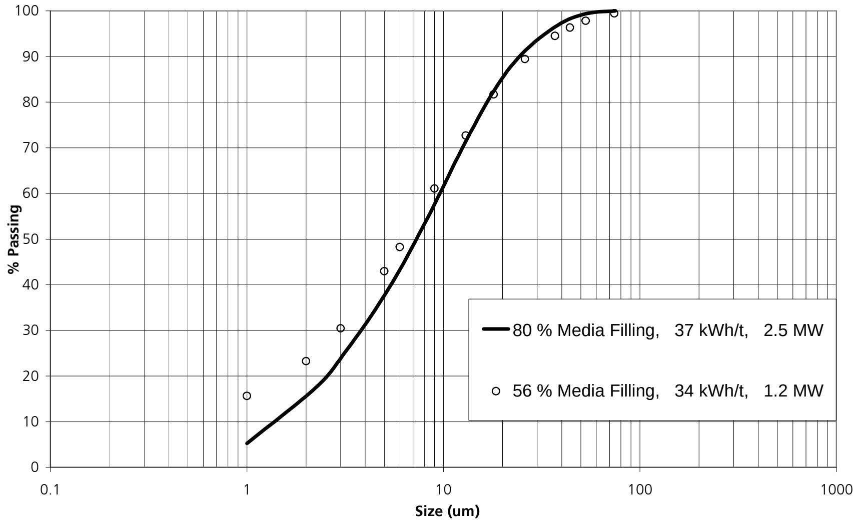


Figure 7: M10,000 turn down capability

4.2 Turn Down

Section 4.1 describes the M10,000 operating at low power (< 50% of installed motor capacity) by reducing the media load in the grinding chamber. This method of power control is common to all IsaMills. Decreasing power by reducing media load, or 'turn down' is necessary in single mill grinding lines to avoid over grinding (and over heating) during periods of low feed tonnage. Turn down would be important to the WLTRP, as the rougher flotation mass pull would fluctuate with head grade. Figure 7 shows the M10,000 operating at similar energy inputs, but greatly different power draws. Despite these extreme operating situations, the IsaMill produces the same P_{80} in both cases.

4.3 Flotation

A series of IsaMill on /off tests were performed between August and October 2004, to confirm the effect of fine grinding on flotation [5]. Figure 8 shows the impact of fine grinding on final concentrate grade, by plotting the IsaMill on /off data pairs over time. Only data pairs that represent a full 24 hour period (of

Table 2: Mean final concentrate and cleaner flotation tail grades

IsaMill Status	Final Conc (g/t)	% Change	Cleaner Tail (g/t)	% Change
On	57.3	+ 42 %	3.05	- 28 %
Off	40.4		4.26	

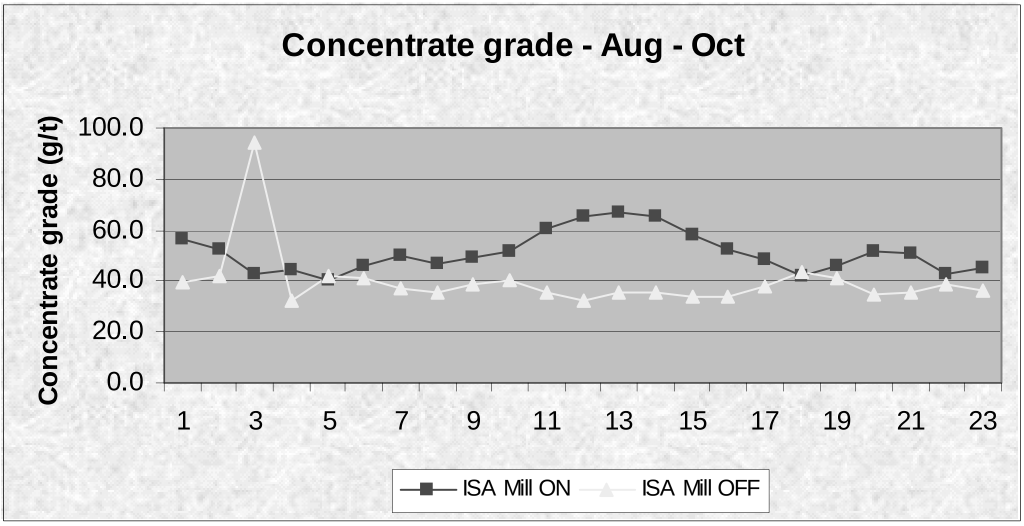


Figure 8: Impact of IsaMill on final concentrate grade

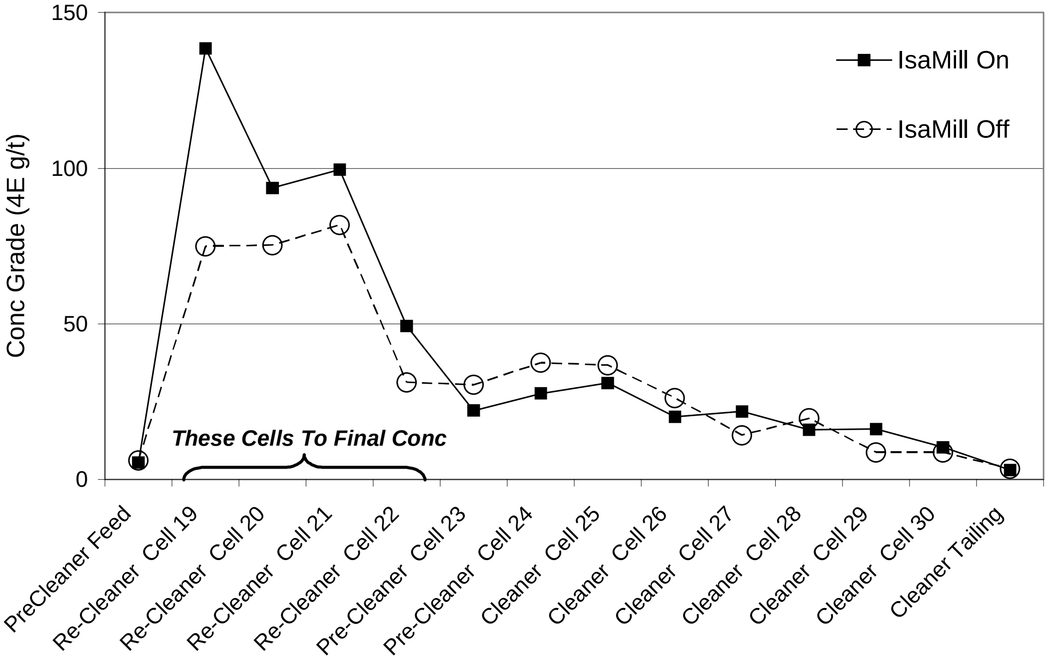


Figure 9: Down the bank survey – IsaMill On /Off

either on or off) are used. Table 2 lists the mean data from the test period, for final concentrate and cleaner tail grade. All PGM grades are built up 4E assays (Pt + Pd + Rd + Au). The effect of regrinding the rougher concentrate was in line with pilot plant findings as the IsaMill had to be operating for the plant to achieve target final concentrate grade (50 g/t 4E).

The higher flotation kinetics and improved PGM liberation that come with regrinding the rougher concentrate, resulted in higher concentrate grades in the recleaner circuit. As the recleaner concentrates report to final concentrate, the ability to maximise the grade in these cells is vital to meeting overall concentrate specification. The difference in recleaner concentrate grades with the IsaMill on or off is detailed in Figure 9 which shows a down-the-bank survey. Cleaner flotation grade and recovery is improved with IsaMilling.

4.4 Maintenance, Availability and Operating Cost

Being a completely new machine, the focus placed on commissioning activities was great. Representatives from Xstrata (Technology (Brisbane), Zinc (Mount Isa) and South Africa business units), Netzsch (Germany) and Siemens (Germany and Australia) were involved in the commissioning process. Despite having built contingencies upon contingencies in the planning stage, the end result was an uneventful and simple commissioning. Hot commissioning commenced on 15th December 2003, and was complete by the 18th December. The mill was then 'stood down' until the tailings dam reclaim mining and main plant ramp up were complete.

Following several brief shut downs in the first month of operation to inspect the mill internals, the M10,000 was put into a normal PM cycle of an 8 hour shut down every 8 weeks (or 1300 hours mill run time). These shut downs are used to rotate and /or shuffle the grinding discs to maximise their life. The disc life is 6000 hours, or 38 weeks (at current circuit availability). The first-fill shell liner is still in use, and was 30 % worn at the July 2005 shut down.

Down time for the IsaMill grinding circuit in the months June and July 2005 is shown below, and separated into stoppages related only to the IsaMill, and stoppages related to ancillary equipment in the circuit. Examples of the most common stoppages of ancillary equipment include pump trips and bypass events due to upstream disturbances that require the IsaMill feed thickener to be placed in recycle. The overall grinding circuit availability is greater than 95 %, with the IsaMill machine availability above 99 %. The month of July included a planned 8 hour IsaMill maintenance shut down.

As the M10,000 design was able to take advantage of hindsight and adjust several parameters from the M3,000 design, it was expected that the larger IsaMill would have a lower operating cost. Another platinum producer operates an M3,000 IsaMill only kilometres from the WLTR site, and uses the same grinding media, with similar feed and product sizes. This allows for the most accurate comparison in spares cost between the two machines. Table 4 shows that the smaller IsaMill spares cost is three times the cost per tonne milled and almost double the cost per kilowatt hour of the larger machine.

5 CONCLUSION

The use of a novel technology development method has been beneficial for Anglo Platinum and Xstrata. The design process was rapid, exhaustive in detail, transparent and above all successful. The M10,000 IsaMill has scaled up accurately and is operating below forecast cost and above budget availability. Most importantly, the down stream cleaner flotation results are matching those seen in pilot testing, whilst plant tests have confirmed the need for regrinding with the IsaMill prior to cleaning to achieve final concentrate grade.

Table 3: Down time analysis for June and July 2005

		June 2005	**July 2005**
A	Hours in month	720 h	744 h
B	Main plant running	711 h	744 h
C	IsaMill planned maintenance	0 h	8 h
D	IsaMill break down	0.5 h	3.1 h
E	Total IsaMill down time = (C + D)	0.5 h	11.1 h
F	Ancillary break down	28.2 h	27.9 h
G	IsaMill Availability = (B – E) ÷ B x 100	99.9 %	98.5 %
H	Overall Circuit Availability = (B - E - F) ÷ B x 100	96.0 %	94.8 %

Table 4: Normalised operating cost of M10,000 and M3,000 IsaMills

IsaMill Model	**Normalised Unit Cost**	
	/t milled	/kWh
M3,000 (with 1,100 kW motor)	1	1
M10,000 (with 2,600 kW motor)	0.33	0.55

REFERENCES

1. Buys, S., Rule, C. & Curry, D. (2005) *The Application Of Large Scale Stirred Milling To The Retreatment Of Merensky Platinum Tailings.* 37th Annual Meeting Of The Canadian Mineral Processors, Ottawa, Canada.

2. Durant, A.C., Buys, A.S. and Knopjes, L.M. (2002) *The Recovery Of Platinum Group Minerals From Dormant Tailings At RPM – Rustenburg Section. A Pilot Plant Study.* Internal Report - Divisional Metallurgical Laboratory Report, Anglo Platinum.

3. Cope, A.J., et al (2002) *Tailings Retreatement Project: Recovery Of 900 ktpm Tailings From Rustenburg Dormant Dams (Klipfontein & Watervaal Only).* Internal Report - Process Technology Division, Anglo Platinum.

4. Gao, M., Young, M. & Allum, P. (2002) *IsaMill Fine Grinding Technology And Its Industrial Applications At Mount Isa Mines.* 34th Annual Meeting Of The Canadian Mineral Processors, Ottawa, Canada.

5. Bedesi, N. (2004) *Comparison of Overall Plant Grades with ISA ON/OFF.* Internal Memorandum - WLTR, Rustenburg Platinum Mines, Anglo Platinum.

IMPROVING FINES RECOVERY BY GRINDING FINER

J.D.Pease – General Manager Xstrata Technology (previously Concentrators Manager, Mount Isa Mines)
M.F. Young - Senior Metallurgist, Mount Isa Mines
D. Curry – Technical Superintendent, Xstrata Technology
N.W.Johnson – Mineral Processing Chair, University Of Queensland

ABSTRACT

"Conventional wisdom" has deemed that fine particles have low flotation recoveries. Plant size -recovery graphs often have the classic "hill" shape - high recovery in the mid sizes and low recovery at the fine and coarse ends. Yet if mineral liberation is poor, low fines recovery may be because you don't grind fine enough!

This apparent paradox is explained by the old concept of "sand/slimes" circuits, which recognises the different flotation needs of fine and coarse particles. This concept is overlooked in the push for simpler circuits and larger equipment. Most plants now treat all particles together in a wide size-distribution. Reagent conditions are set for the dominant coarser particles, so fines are starved of collector. Worse still if there are significant mid-sized composites - often these have to be rejected in cleaning to achieve target concentrate grade. But the conditions which reject mid-size composites - collector starvation and high depressants – also reject fine liberated particles. In fact, fines flotation can be excellent when flotation chemistry is tailored to fines.

After first recovering fast floating liberated particles, correct grinding to liberate remaining composites is essential to increase fines recovery. Firstly, liberating composites allows lower depressant and higher collector additions, since composites will not dilute the concentrate. Secondly, finer grinding narrows the size distribution to flotation, allowing reagent conditions to be set to suit all particles. Additionally, fine grinding in an inert attritioning environment like an IsaMill removes surface deposits that may have made some fines slow-floating.

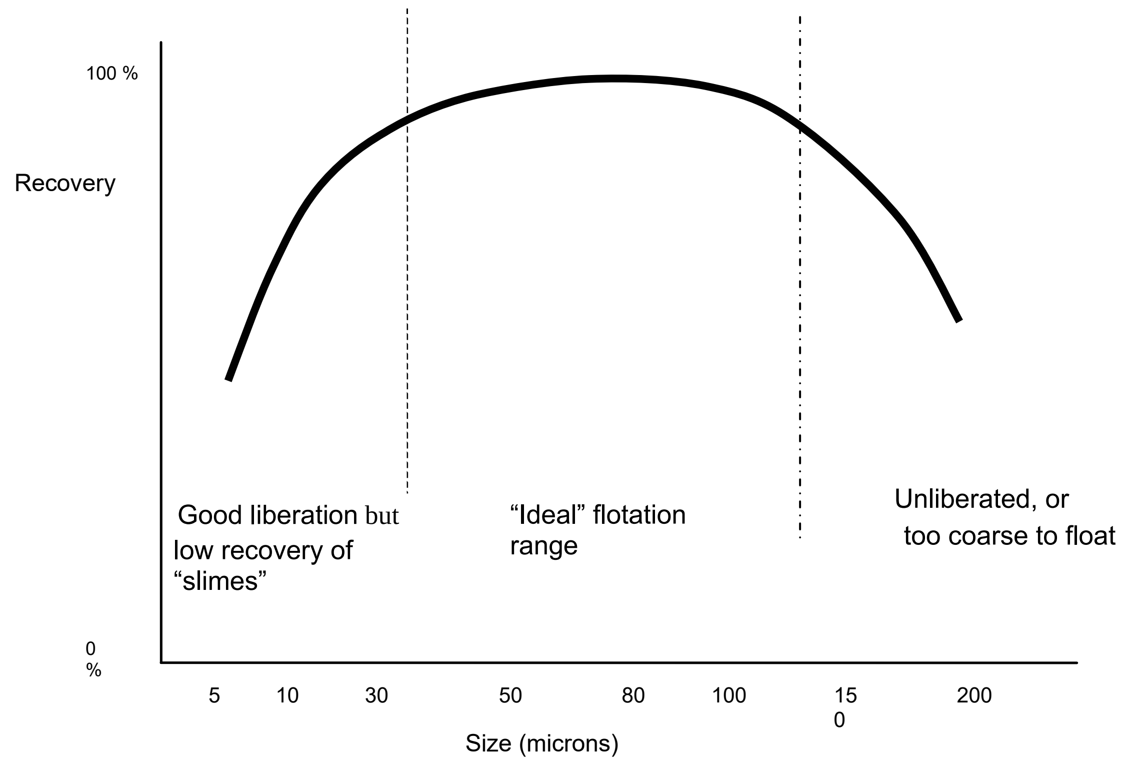
An excellent case study is the installation of IsaMills in the Mount Isa lead zinc concentrator to grind lead and zinc rougher concentrate to 12 microns and zinc cleaner tailings to 7 microns. Most plant losses had previously been in the sub 10-micron fraction, yet ultra-fine grinding increased plant recovery by 5% lead and 10% zinc. Circulating loads dropped, reagent additions dropped in spite of the much higher particle surface area, and the plant became much more operable and responsive.

INTRODUCTION : THE CONVENTIONAL VIEW

The conventional view of the flotation size-recovery curve is shown in Figure 1. There is a good reason for this view – if you sample almost any flotation plant you will produce a similar curve. The numbers speak for themselves – fine particles float poorly in most plants. Operators carefully avoid “overgrinding” and “sliming” of feed.

Yet this obvious conclusion is challenged by practice in some other plants. For example, Xstrata’s McArthur River Mine (MRM) produces 380,000 t/y of concentrate at a P80 size of 7 microns (µm), and a recovery of 82%. And Mount Isa Mines produces 260,000 t/y of lead concentrate and 350,000t/y zinc concentrates at P80 about 15 µm and over 80% recovery. The biggest problem for these plants is coarse particles – anything over 20 µm is considered as “gravel” which will reduce recovery.

Figure 1 : Typical Recovery vs Size Curve



RESOLVING THE PARADOX

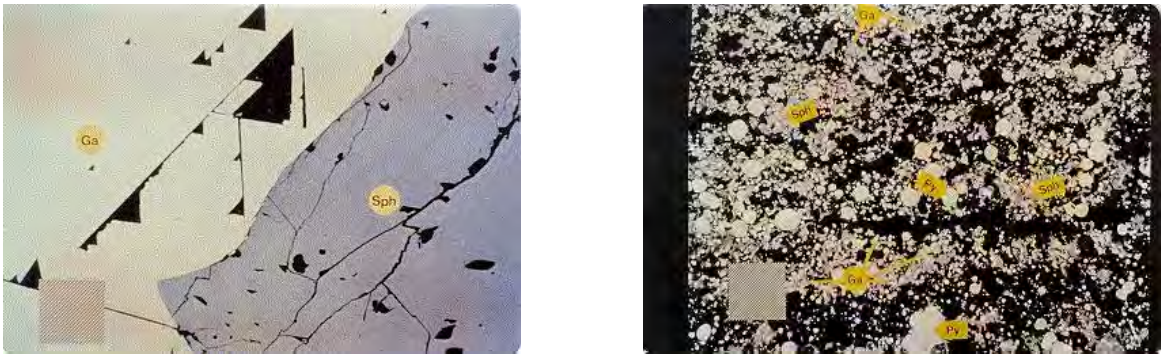
Why is it that most plants fear fines production, yet others rely on it? The answer lies in three areas:

- Mineralogy and liberation
- Operating constraints and strategy
- Flotation rates and surface area

Mineralogy: make no mistake, you don't want to grind fine if you don't have to. It is power intensive and expensive. But if you have fine-grained mineralogy, you have no choice. Figures 2 shows photomicrographs of Broken Hill ore and McArthur River (MRM) ore at the same magnification. Broken Hill will never have a liberation problem – its metallurgists can focus on simple grinding and flotation circuits and avoiding "overgrinding". But for MRM and Mount Isa, no amount of fiddling with circuit design or reagent testing will help (trust us, we wasted years of our lives trying!). For these ores, there is one question that must be answered before anything else: "Will this circuit change/new reagent increase mineral liberation? " If not, do yourself a favour and keep your money in your pocket.

Figure 2 : Different Grain Size of Broken Hill and McArthur River Ores

(both photos at the same magnification)



Broken Hill Ore

McArthur River Ore

SOME DEFINITIONS AND PERSPECTIVES

We often note that communication between different operating plants is confusing because we all use different definitions of "coarse", "intermediate" and "fine" particles. An operator of a "coarse" grained orebody may call minus 37 μm "slimes". To an operator at Mount Isa or McArthur River this is gravel. To them, anything above 20 μm is coarse, between 10 and 20 μm is intermediate, and less than 10 μm is fine or ultrafine.

We avoid using generic terms like "fines" in this paper. But if you hear someone saying that "fine" particles don't float well, ask them two questions:

- What do they mean by "fine"?
- Are they aware that in the last decade, Mount Isa Mines and MRM have produced over 10 million tonnes of concentrates at an average sizing less than 10μm, at over 80% recovery, in conventional flotation cells, with a simple xanthate reagent system?

Particle surface area per tonne increases rapidly as size gets finer. One tonne of 7 μm particles has 5 times the surface area of a tonne of 37 μm particles; for 2.5 μm particles the surface area triples again. This explains why grinding energy increases exponentially as grind size decreases (Figure 3), and why finer media with higher surface area is needed for grinding below about 30 μm. It also explains the higher collector need of fines.

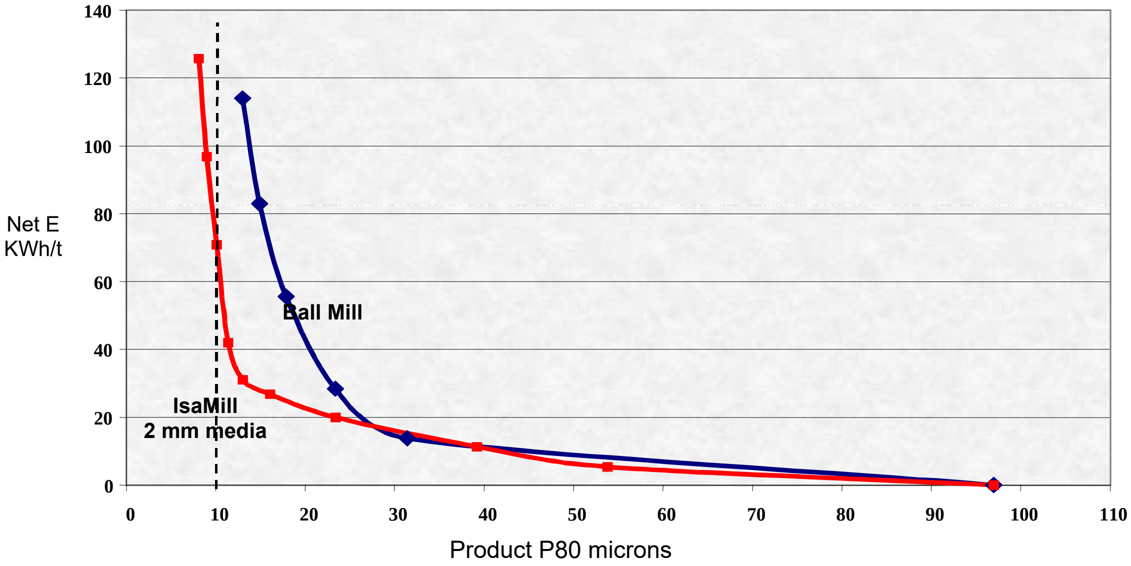
In this paper reference to a 7 μm particle means a single particle of that diameter. By convention a grind size of 7 μm refers to the 80% passing size (P80). For example, the regrind size at MRM is 7 μm, so only 20% of final product is 7 microns or coarser. Fifty percent weight of MRM concentrate is below 2.5 μm. Since flotation works on individual particles interacting with bubbles, consider this from a different perspective - 50% by weight less than 2.5 μm means that ***96 % of individual particles recovered at MRM are less than 2.5 μm !*** In spite of popular perception, fine particles float very well indeed.

ULTRAFINE GRINDING CIRCUIT DESIGN

If you have to grind below 25 µm, then you need to choose the right equipment. Three issues are particularly important:

- power efficiency
- classification within the grinding circuit
- the impact of grinding on flotation performance

Figure 3 : Grinding Energy versus Product Size for a Gold Ore



- **Power efficiency** is demonstrated by Figure 3, comparing the power required to grind a gold ore in a ball mill with 9 mm balls with an IsaMill with 2 mm media. The IsaMill is much more efficient below about 30 µm – to grind this ore to 15 µm would take 28 kWh/t in the IsaMill, but 90 kWh/t in a ball mill. Traditionally this has been attributed to the difference between attrition grinding and impact grinding. However by far the most important factor is media size, as shown by Figure 4, which shows the breakage rate in Tower Mills drops dramatically - the breakage rate for 20 µm particles is ten times lower than the rate for 40 µm particles. Even though the Tower Mill is full attrition grinding, practically it is constrained to using relatively coarse media, 9mm balls in this case. In contrast, the IsaMill (Netzsch mill in Figure 4) can operate with much finer media and much higher intensity of power input (Table 1), meaning the peak breakage rate occurs at 20 µm, and doesn't drop as quickly below that.

Figure 4 : Breakage Rates in Different Grinding Devices

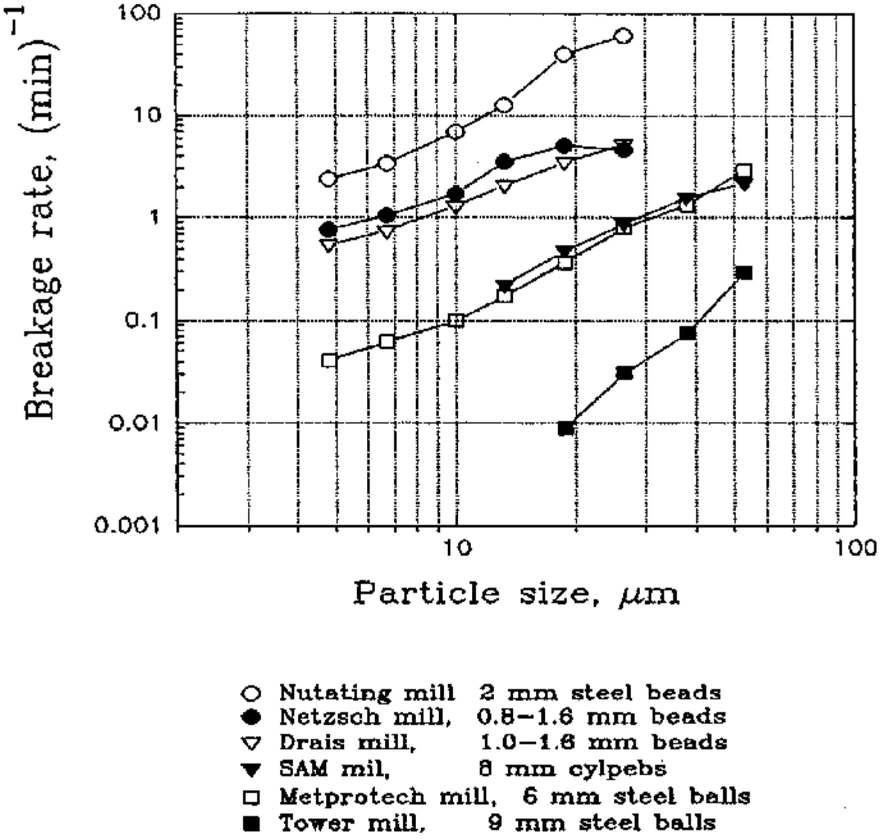


Table 1 : Comparison of Various Grinding Technologies Independent laboratory data

FEATURE	ISAMILL	TOWER MILLS	VERTICAL PIN MILLS
Grinding Intensity (kW/L)	0.54	0.005	0.15 - 0.18
Residence Time to 15 µm (min)	0.6	154	7 - 9
Power Usage to 15 µm (kWh/t)	17.4	59.6	37.5 - 39.0
Media Material	Various	Steel	Steel
Media Size (mm)	0.8 - 1.6	9 -12	6 - 8

Figure 4 and Table 1: Extracts from AMIRA P336, Gao M and Weller K, Review of Alternative Technologies for Fine Grinding, November 1993.

- **Good classification** is vital for power efficiency in ultrafine grinding, just as it is in conventional grinding. However it is not generally practical to use cyclones to close-circuit a grinding mill with a target below about 15 µm. To get good cyclone efficiency at these sizes requires small cyclones, eg two inch (50 mm) diameter or smaller. This is virtually inoperable on a large scale, so the circuit is either compromised (and less power efficient) by using bigger cyclones, or an alternative solution is needed. The IsaMill achieves this by the internal classifier mechanism, using the high centripetal forces generated inside the mill to classify the discharge, ensuring a very sharp product size without external cyclones. The very short residence time in the IsaMill also minimises "overgrinding", further contributing to the sharp product size distribution. As an added advantage this mechanism also retains fine media very effectively, meaning that low cost media can be used, eg local sand, or granulated smelter slag.

  A cautionary word to those designing circuits – the benefits of good classification on power efficiency and media retention does not show up in laboratory tests. These tests are done in batch mills, and many technologies will show the same power efficiency in a closed device. The crucial questions are, what is the power efficiency, media retention, and product size distribution in a full-scale continuous installation.

- Managing the **impact of grinding on flotation performance** is the third crucial factor in plant design. Even if you can accept the low power efficiency of a mill with steel balls, you may not be able to deal with its impact of surface chemistry. Consuming so much power in a steel environment means high retention time and lots of steel contamination. The resultant low pulp potential changes flotation behaviour, requiring additional reagents and reducing selectivity. One early response to this problem was to use High Intensity Conditioning (HIC), eg at Hellyer, to reverse the negative impact of Tower Milling on surface chemistry. Processes like IsaMilling are far more efficient by providing this high intensity as part of the grinding action, and grinding in an inert environment. Later we will show how IsaMills significantly improved the flotation behaviour of ultrafine particles at Mount Isa.

FLOTATION CIRCUIT DESIGN AND OPERATING STRATEGY

The view that “fine” particles don’t float is caused by circuit design and the constraints of operating strategy. Simply, flotation works best when applied to narrow size distributions. A 5 μm particle has 10 times the surface area of a 50 μm particle, and fundamentally different hydrodynamics. Yet often our circuit designs assume they will behave the same, and treat them together in flotation. Texts as old as Taggart described the benefits of “sand/slimes” splits into separate circuits. This simple concept has been largely ignored in the push for circuit simplification and larger flotation cells.

We are not advocating complicated flotation circuits. However if you have fine-grained minerals then you must design your circuit to suit the needs of fine particles, not coarse particles. The Mount Isa circuit developed into an excellent balance of the needs of different minerals, relying on several stages of grinding and flotation. The design principals are:

- **Don’t grind anything more than you need to**. Fine grinding is expensive – technically the best solution for Mount Isa would be to grind everything to 12 μm, but this would not be economic. Therefore stage grind and float to suit the mineral behaviour – at Mount Isa this means a 37 μm grind before roughing. Some mineral is liberated at this size and can go to concentrate. Other minerals in rougher concentrate need to be ground to 12 μm. Some of these are rejected in cleaning and need to be reground to 7 μm.

- **Float minerals in narrow size distributions** – this happens automatically with the staged grinding approach described above, and is assisted by the inherent sharp size distribution produced by the IsaMills.

- **Minimise circulating loads, and open-circuit as much as possible** – this is another automatic outcome of staged grinding. It is pointless to recirculate a composite particle unless you are going to grind it to liberation. If you do regrind it, you should now float it separately with similar sized particles.

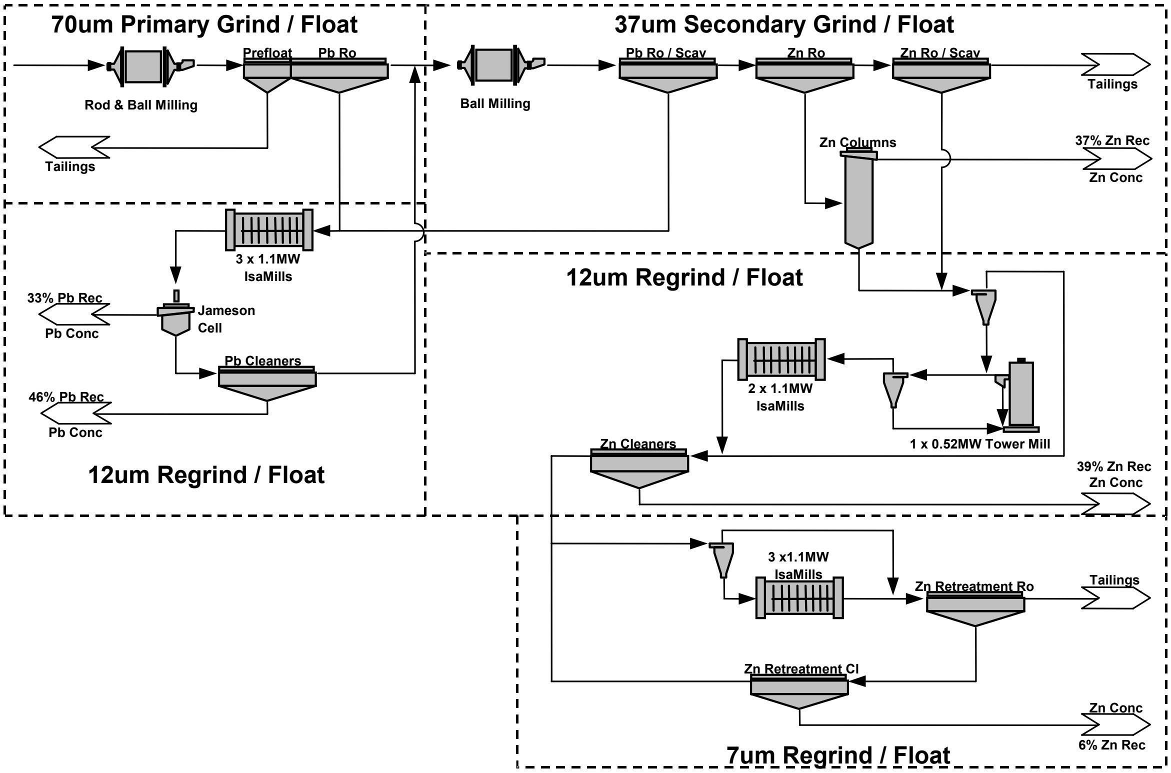
These principles can be seen in the simplified Mount Isa flotation circuit in Figure 5. Though the circuit may appear complicated, it is better than the alternatives of either:

- Grinding everything to 12 μm and floating together (too expensive)

- Recirculating regrind products and trying to float them with coarser minerals (causing poor performance of the reground minerals, high circulating loads and low recoveries).

Contrary to appearance, these developments at Mount Isa greatly simplified circuit operations. Lead recovery increased by 5% and lead concentrate grade by 5%, zinc recovery by 10% and concentrate grade by 2%. More surprisingly, reagent needs dropped, circulating loads dropped, and the circuit became far more stable. Flotation suddenly became as easy and predictable as the textbooks say it is!

Figure 5 : Mt Isa Pb / Zn Concentrator Flow Sheet



Overall Recovery: Pb = 79 %
Zn = 82 %

COMPETING FLOTATION RATES OF DIFFERENT PARTICLE SIZES

The profound impact of a narrow size distribution to flotation feed is explained by mineralogy and operating constraints. In a system with just pure liberated sphalerite and quartz, flotation could achieve good recovery in all size ranges, even though the “fines” have slower flotation rates. But in real circuits there are two crucial constraints:

- Other contaminant minerals such as pyrite and pyrrhotite also exhibit some floatability. If so, a “coarse” pyrite particle may have the same flotation rate as a “fine” sphalerite particle.

- Composite particles. To explain the problem with composites, imagine a simple 37 μm sphalerite-quartz binary. This particle has to be rejected since typically zinc concentrates must be less than 3% silica. The low collector and high depressant needed to reject the 37 μm composite will also depress the slower floating 10 μm liberated sphalerite particle.

Some further problems arise when floating coarse and fine particles together:

- It is no point depressing the 37 μm composite particle unless you can liberate it. While plants often “send it to regrind”, this is often a conventional ball mill or Tower Mill that has very low breakage rates on sub 30 μm particles. This causes high circulating loads of composite particles.

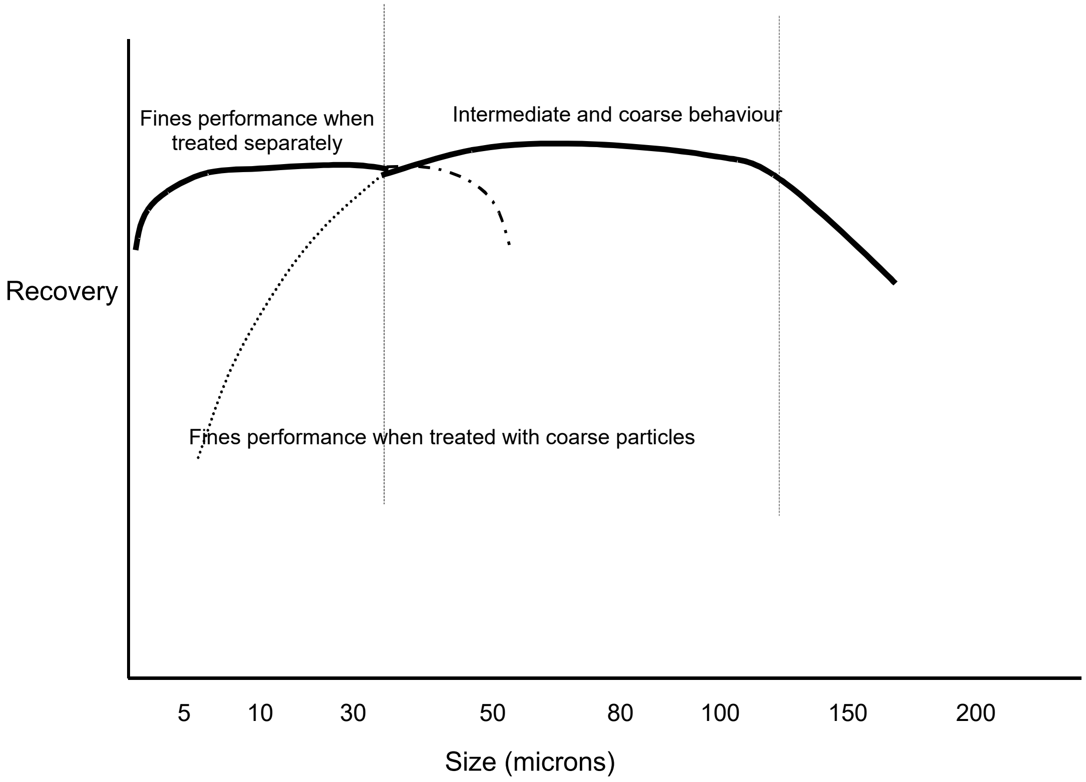
- The high circulating loads then take up volume and reduce residence time in roughing and cleaning. Since fine particles are slower floating, this drop in residence time further hurts their recovery.

- If high pH is used for depression, and lime is used to get high pH, then a surface chemistry problem is introduced. Circuit water can become super-saturated in Calcium ions. This leads to reaction with sulphate ions, which causes gypsum to precipitate on the nearest surface – usually a mineral particle. SEM work at Mount Isa before IsaMills showed that up to 80% of sphalerite surface was masked by gypsum. This has a more serious effect on sub 20 μm particles.

RECONCILING THE REALITY WITH THE PERCEPTION

The common perception is that “fines” don’t float, the reality is that they will in the right conditions. Figure 6 explains this conceptually – in most plants, sub 20 µm particles do perform poorly because they are mixed with coarser particles with much different needs. If these particles were floated in a narrow size distribution, flotation conditions can be tailored to them and they perform well. This explains the performance of the Mount Isa staged grind and float circuit.

Figure 6 : Conceptual description of staged grind and float circuit



CASE STUDY – MOUNT ISA LEAD ZINC CONCENTRATOR

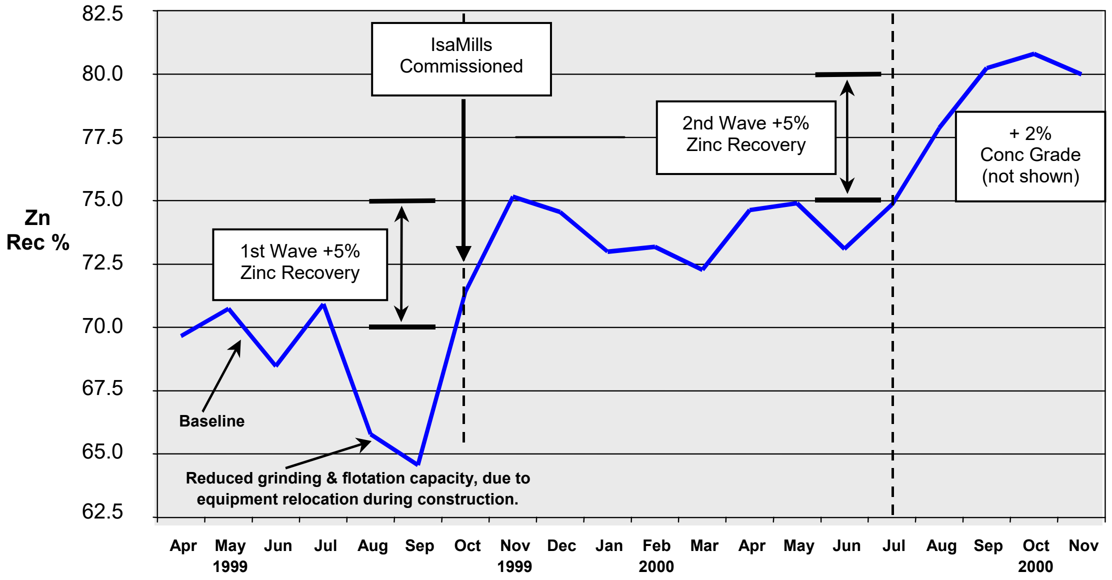
The changes to the Mount Isa circuit as part of the "George Fisher Project" are detailed elsewhere (Young & Gao, 2000, Young, Pease & Fisher, 2000). In summary, the project involved adding a further 6 IsaMills, to regrind lead rougher concentrate to P80 of 12 μm, most zinc rougher concentrate to 12 μm, and a zinc regrind to P80 of 7 μm (see Figure 5). Lead performance also increased by 5% concentrate grade and 5% recovery (equivalent to 10% increase in lead recovery at the same grade). Zinc recovery increased by 10%, in two steps, and zinc concentrate grade by 2% (equivalent to 16% increase in zinc recovery at the same grade). The story of zinc metallurgy can be told in Figures 7, 8 and 9.

The project predicted 5% higher zinc recovery (and no extra concentrate grade) due to extra liberation. Figure 7 shows this was achieved instantly. What surprised us was the "second wave" of a further 5% zinc recovery increase and the 2% increase in zinc concentrate grade. This was because fines performance improved after we ground finer.

It took us about 6 months to discover how much better the fines could perform because we were so used to flotation performance after conventional grinding rather than after IsaMilling. The three biggest mistakes we made were:

- We thought we would need a lot more reagents after IsaMilling. Since we we created a huge amount of new surface area we expected some reagent additions to triple.
- We didn't take the depressant (lime to pH 11) off the zinc cleaners.
- We thought flotation rates would be slower, so we "pulled" flotation banks harder to compensate.

Figure 7 : Zinc Recovery Increase from IsaMilling



This resulted in four fundamental changes:

- Circulating loads had dropped dramatically, because we had finally addressed the liberation problem. This created significantly more flotation capacity than we had expected, and reduced reagent need.

- We no longer needed depressant on the cleaners since there weren't many composites left. The liberated minerals "behaved" properly. This was a crucial step, since removing the lime dropped the calcium in circuit water, dropping gypsum formation and further increasing flotation rate of sub 15 µm particles.

- The narrow size distribution in each stage of flotation made flotation easy. Together with the low circulating loads and open-circuit design, the flotation circuit became steadier and simple to control.

- The inert grinding environment and attritioning in the IsaMill improved mineral behaviour and changed reagent needs. Our initial expectations were way off – we added far too much reagent, and in the wrong places.

Over the next 3 months we learnt to challenge every known "truth" about the behaviour of our circuit, and accept that grinding could improve mineral flotation behaviour rather than reduce it. Once we understood that we had to approach the circuit as a "clean sheet", we rapidly improved recovery by the 5% shown in the "second wave". In spite of adding 6 MW of grinding power to the circuit, we were amazed that unit costs ultimately did not increase:

- Overall reagent consumption remained the same in spite of the extra surface area – since we no longer had competing depressants and collectors battling with composites.

- Total power consumption did not increase as much as we thought, since circulating loads of over 100% simply disappeared. As a result we were able to shut down some flotation capacity.

- With lower reagents and no circulating loads, spillage was almost eliminated.

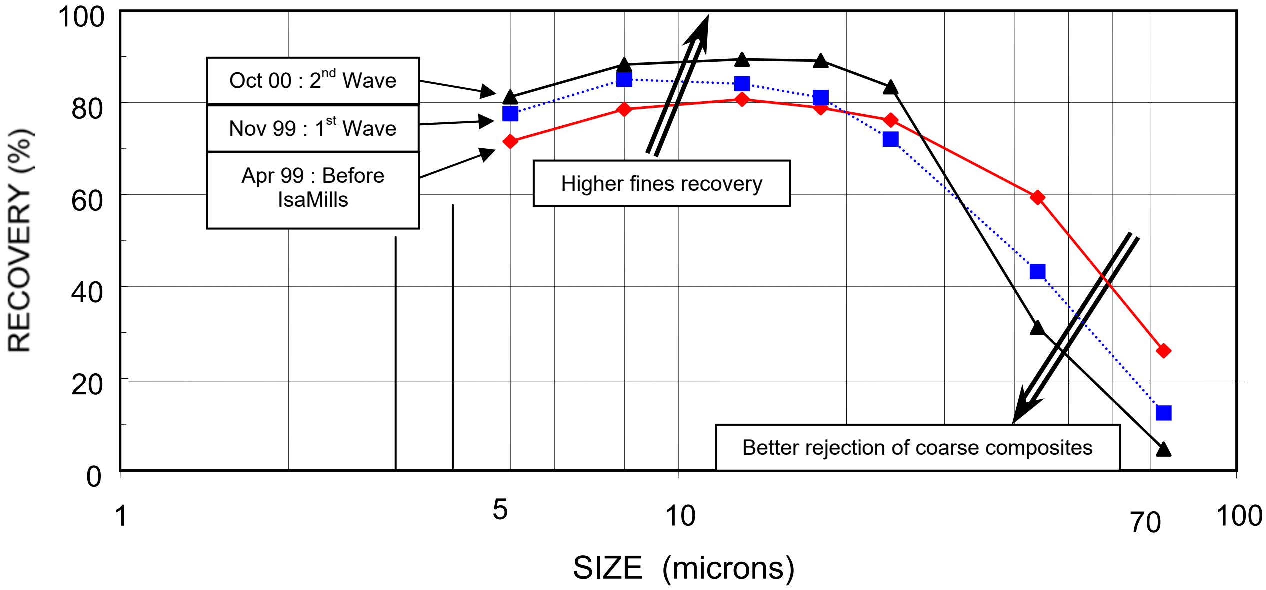
- The very steady nature of the circuit meant that simple stabilising control worked, keeping equipment at high efficiency.

- The IsaMill grinding media is free, granulated lead smelter slag that would otherwise be discarded.

The improvements in flotation are further described by Figure 8, which records size-by-size performance for sphalerite. Not only did the IsaMills put more material in the liberated sub 15 µm size range, they also improved recovery of all sub 25 µm particles. For example, recovery of 8 µm liberated particles increased from about 78% to about 88% after Isa Milling. Though recovery in the coarser fractions has dropped, this was what was needed to improve total recovery since:

- The coarser fractions had contained the composites, previously rejecting these composites also rejected fines.
- After IsaMilling, very fast floating coarser particles (ie liberated ones) were still accepted, but the slower floating ones were sent for regrinding and liberation.
- There was much less material in these coarse size ranges after IsaMilling – coarse composites were reground into finer size ranges with higher recovery (and higher grade since they were more liberated).

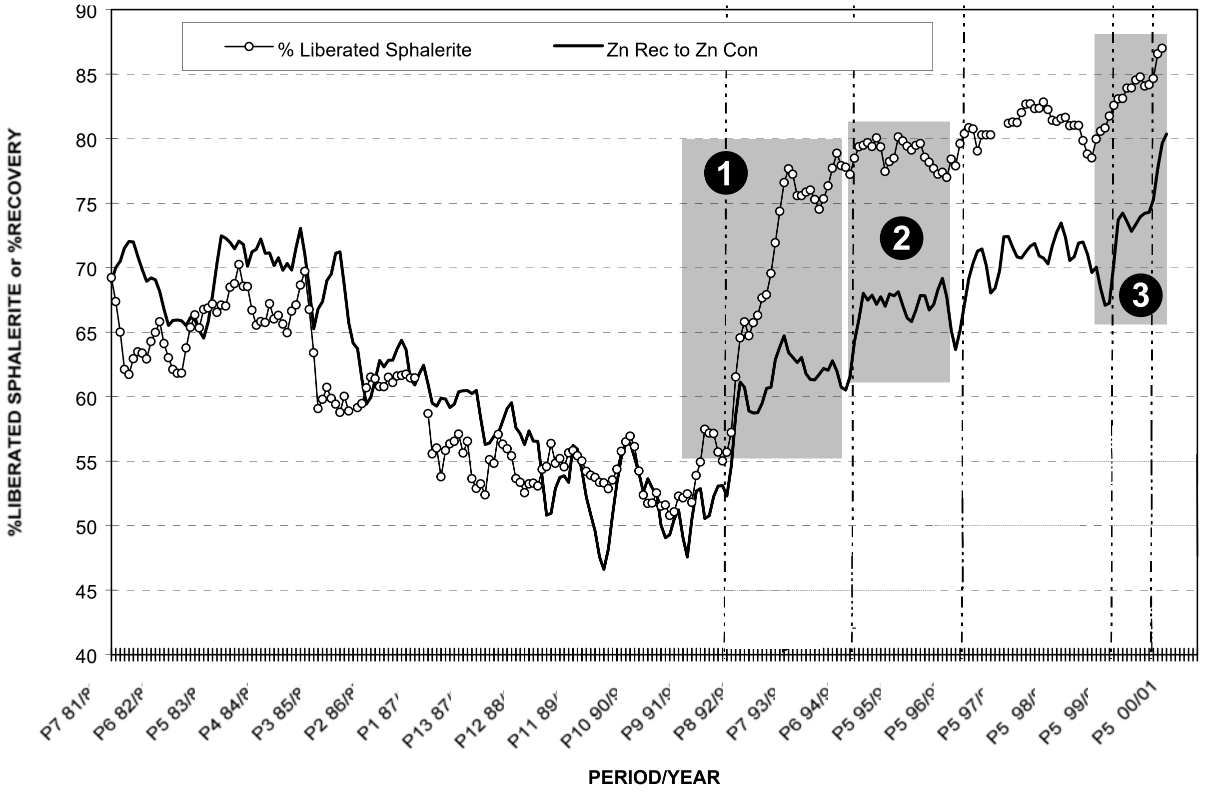
Figure 8 : Increased Fines Recovery After Fine Grinding in the IsaMills

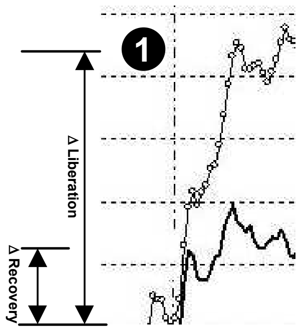


THE BIG PICTURE

The history of changes at Mount Isa is described by Figure 9, which considers sphalerite liberation and zinc recovery. As ore became finer grained in the 1980's, liberation dropped and recovery dropped accordingly. In 1992 we installed conventional ball and Tower Milling to improve liberation. Sphalerite liberation improved by 25% (from 55% to 80%), but recovery only improved by 10%, due to the extra difficulty of floating after grinding in ball and Tower Mills. In 1995 we installed the prototype IsaMills in the lead circuit. Not only did this improve lead performance, but zinc recovery also increased even though total sphalerite liberation was little changed. This was because the improvements from IsaMilling allowed sphalerite to be redirected from the lead concentrate to the zinc circuit. In 1999 the installation of additional IsaMills increased sphalerite liberation by a further 5%, but zinc recovery increased by 10% and zinc concentrate grade by 2% (net impact equivalent to 16% recovery increase at the same concentrate grade). This demonstrates the fundamental changes to fines flotation made possible by ultrafine grinding in IsaMills.

Figure 9 : Relationship Between Sphalerite Liberation And Recovery Over 20 Years



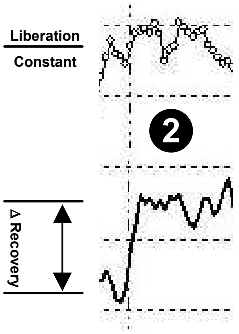


Ball & Tower Milling Added

Liberation increased by 20 %

Recovery increased by 5 %

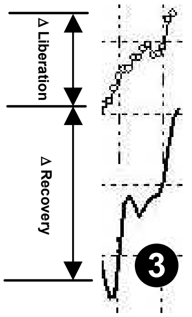
Recovery increased less than liberation due to negative impact of steel media on flotation of fines



First IsaMills Added in Lead Circuit

Sphalerite liberation constant

Zn recovery increased by 4 % due to better rejection from Pb conc after IsaMilling



IsaMills Installed In Zinc Circuit

Liberation increased by 7 %

Recovery increased by 10 %

Conc grade increased by 2 %

Recovery benefit higher than liberation increase because of improved flotation after IsaMilling

REFERENCES

Gao, M and Weller, K R, 1993.
Review of Alternative Technologies For Fine Grinding. *AMIRA Project P336, Report P336/20, November.*

Holder, R.K, 1994
Improvements in copper and silver flotation at Hellyer using high energy conditioning. *AusIMM Fifth Mill Operators' Conference, Roxby Downs, Australia, 16 – 20 October.*

Young, M F, and Gao, M, 2000.
Performance of the IsaMills in the George Fisher Flowsheet. *AusIMM Seventh Mill Operators Conference, Kalgoorlie, Australia, 12-14 October..*

Young, M F, Pease J D, and Fisher K S, 2000.
The installation of the George Fisher flowsheet in the Mount Isa Lead/Zinc Concentrator. *AusIMM Seventh Mill Operators Conference, Kalgoorlie, Australia, 12-14 October.*

Harbort, G., Murphy, A., Vargus, A., Young, M F, 1999.
The introduction of the IsaMill for ultrafine grinding in the Mount Isa Lead/Zinc Concentrator. *Extemin99, Arequipa, Peru, September.*

FINE GRINDING AND PROJECT ENHANCEMENT

L.W. CLARK - BUSINESS MANAGER – MINERAL PROCESSING XSTRATA TECHNOLOGY
B.D. BURFORD - SENIOR PROCESS ENGINEER XSTRATA TECHNOLOGY

ABSTRACT

Fine grinding has the potential to revolutionise the mineral processing industry due to the benefits attained through being able to economically grind finer to enhance particle liberation for improved flotation and to introduce a high degree of strain into the mineral lattice to improve leaching rates in hydrometallurgical processes.

The development of fine grinding technology has enabled mining operations to be developed at a number of mine sites previously considered uneconomic. It has also enhanced the performance of existing mining operations at a number of mine sites and provided the key such that leaching can be carried out under relatively mild conditions in low cost open tanks. This paper is a review of a number of case studies that demonstrate the benefit that fine grinding has provided to a wide range of applications.

The use of inert media has provided further improvements in circuit performance and recovery.

The Mt Isa operation, MRM operations and the Albion Process are reviewed in this paper.

INTRODUCTION

Fine grinding technology has come a long way in the last decade. The ability to grind ore down to sizes below 10µm has lead to the establishment of new mines such as McArthur River, Century and George Fisher mines, as well as improving recoveries of lead and zinc at the established Mount Isa deposits.

Fine grinding has also opened up the door to new leaching processes. Leaching has always been regarded as a simple process, but as orebodies have become more complex, and oxide deposits giving way to more refractory ore, more advanced leaching is required. One of the technologies that fits this duty is the ALBION Process, however unlike competing technologies that incorporate complex processing steps, the development of fine grinding has made this process a very simple and effective process.

Both processing routes was made possible by the innovation of the IsaMill and other fine grinding processes, and this paper investigates the impact of this technology on the processing of complex ores through mineral processing and leaching.

BACKGROUND – MOUNT ISA

Mount Isa is a long established mining town situated in North West Queensland. Lead carbonate ore was first found by John Campbell Miles in 1923, which later revealed one of the biggest lead/zinc sulphide deposits of its time. Mount Isa Mines Limited was formed soon after the initial discovery, and lead mining, concentrating and smelting operations began.

Over time, the easy to treat orebodies gave way to harder to treat orebodies, which had lower grades and more complex mineralogy. The two main lead/zinc orebodies found at Mount Isa were named Racecourse and Black Star, described as: (Pease, Young et al, 1997):

- "Black Star" orebodies are massive and are mined by open stoping at lower mining cost, and therefore have a lower cut-off grade. Generally this ore is finer grained and has considerably more fine-grained carbonaceous pyrite, whilst core replacement of pyrite by galena (atolling) is more common and at a more advanced stage.

- "Racecourse" orebodies are narrow and mined by bench stoping with a higher cut-off grade. Generally these orebodies are higher in grade for lead and silver, coarser grained and lower in pyrite. The pyrite is more the euhedral type than the fine-grained carbonaceous type.

The other major lead/zinc orebodies in the region, also owned and mined by Mount Isa Mines Limited, was the Hilton and Hilton North orebodies, 20km north of Mt Isa. The mine was later renamed as the George Fisher mine. The orebodies were similar to the orebodies at Mount Isa, with the upper part of the orebodies having characteristics of the "Black Star" ore, and the lower orebodies similar to "Racecourse" ore (Pease, Young 1997).

The presence of pyrite was described as two types. The euhedral pyrite was described as course grained, while the fine grained carbonaceous pyrite is present in the 5 to 30µm range containing natural carbon and is naturally floating, (Munro 1993).

CONCENTRATOR PRACTICE AT MOUNT ISA

Ore at Mount Isa was mined and treated through the Lead/Zinc Concentrator. There have been various concentrators built on the site. The current lead/zinc concentrator, No. 2, operating today, was installed in 1966, and produces lead and zinc concentrates through sequential lead, then zinc flotation. It also had a LGM circuit which was a zinc rich bulk concentrate where "low grade middlings" were processed into a saleable product.

As the operations at Mt Isa progressed, the Racecourse orebodies provided the bulk of the ore during the early days of mine development, being higher grade. However over time more of the Black Star material was processed. The introduction to mining the Hilton orebodies began in 1987, and a lot of this material was trucked into Mt Isa to be treated through the Lead/Zinc Concentrator, (a lead/zinc concentrator was built at Hilton, and was commissioned in late 1989 for processing some of the ore).

Along with the changing orebodies, and the resultant mineralogical changes, the mining rate was also increased. This is highlighted in figure 1 (Pease, Young et al 1997), where head grades of the concentrator feed dropped while the throughput rate increased.

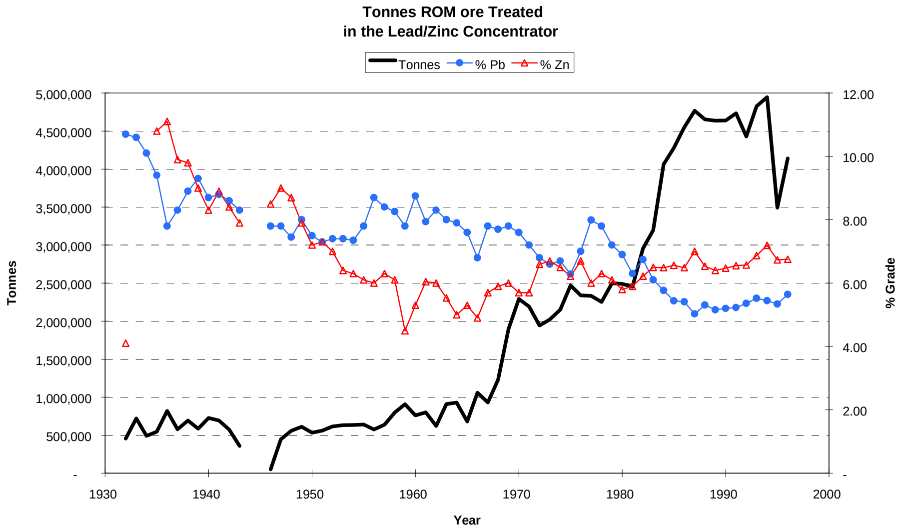
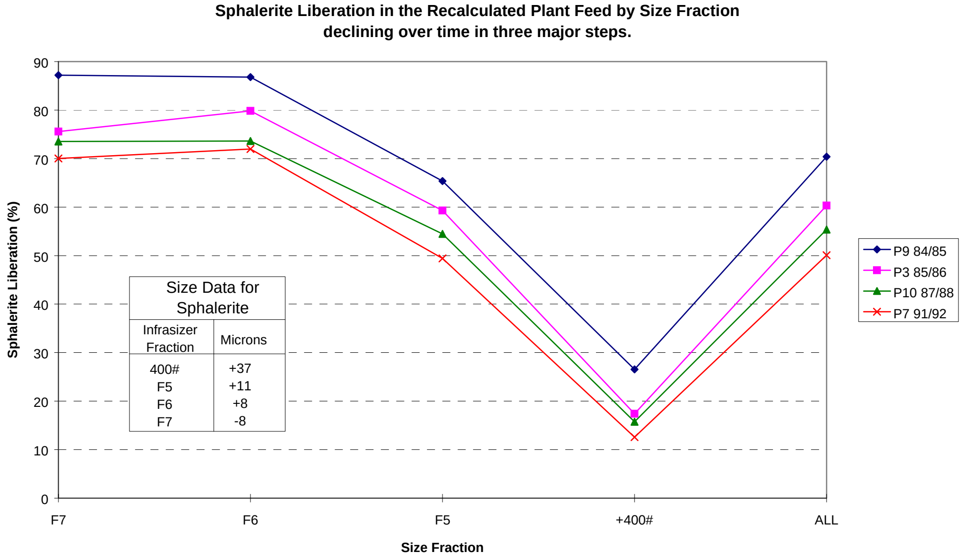


Figure 1 – Time vs Pb/Zn Head Grades

As more Black Star and eventually Hilton ore was being treated, the ore became finer grained and more complex. This was observed for all sizes of concentrator feed, having less liberated mineral species. Figure 2 shows the amount of sphalerite liberation per size class deteriorating over time, from 84/85 to 91/92.



Size Data for Sphalerite

Infrasizer Fraction	Microns
400#	+37
F5	+11
F6	+8
F7	-8

Figure 2 – Liberation Data

Also as more of the Black Star was mined, the proportion of carbonaceous pyrite compared to the euhedral pyrite increased. The carbonaceous pyrite is hydrophobic under most conditions and cannot be depressed easily, as well as consuming large amounts of reagents. The net effect of this change in terms of mineralogy is a greater concentration of iron sulphides in the concentrates (Pease, Young et al 1997). Any attempt made to depress the presence of iron sulphides in the concentrate such as the use of lime or dextrin, resulted in composites being depressed as well in the concentrate, which formed high recirculating loads in the circuit. This in turn limited cleaning and flotation capacity.

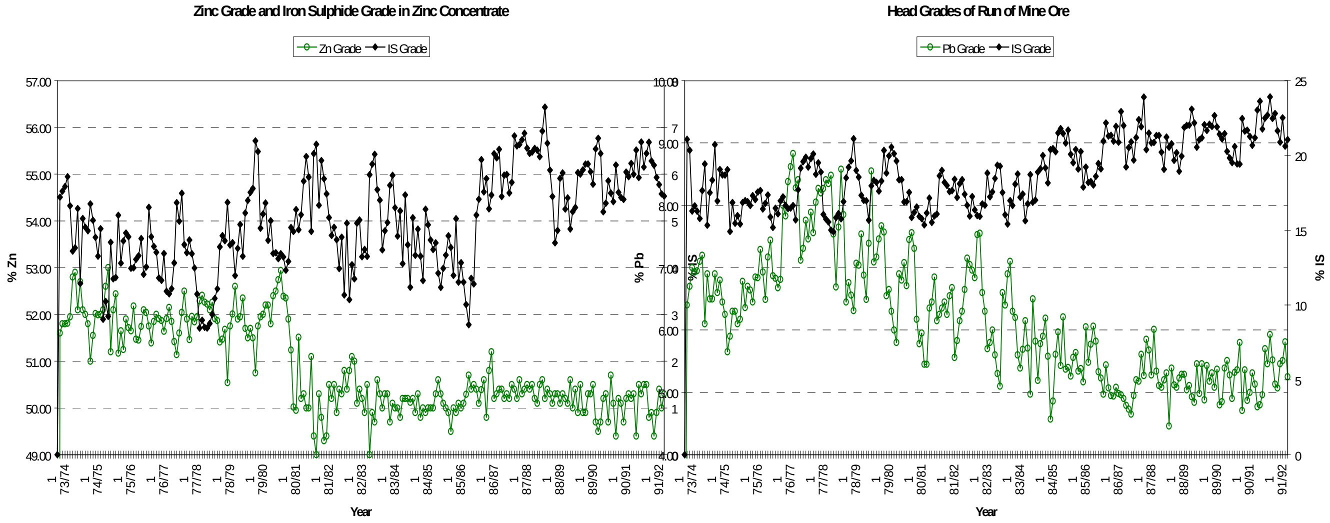


Figure 3 – Iron Sulphide Contamination of Final Concentrates

In summary the Lead/Zinc operation at Mount Isa was not recovering enough metal. The main concerns were:

- More fine grained ore
- Less liberation in the concentrator
- Greater presence of carbonaceous pyrite
- Increasing throughput

The net result was the detrimental impact of metal recovery. Figure 4 shows the gradual deterioration in zinc recovery following the decrease in zinc liberation.

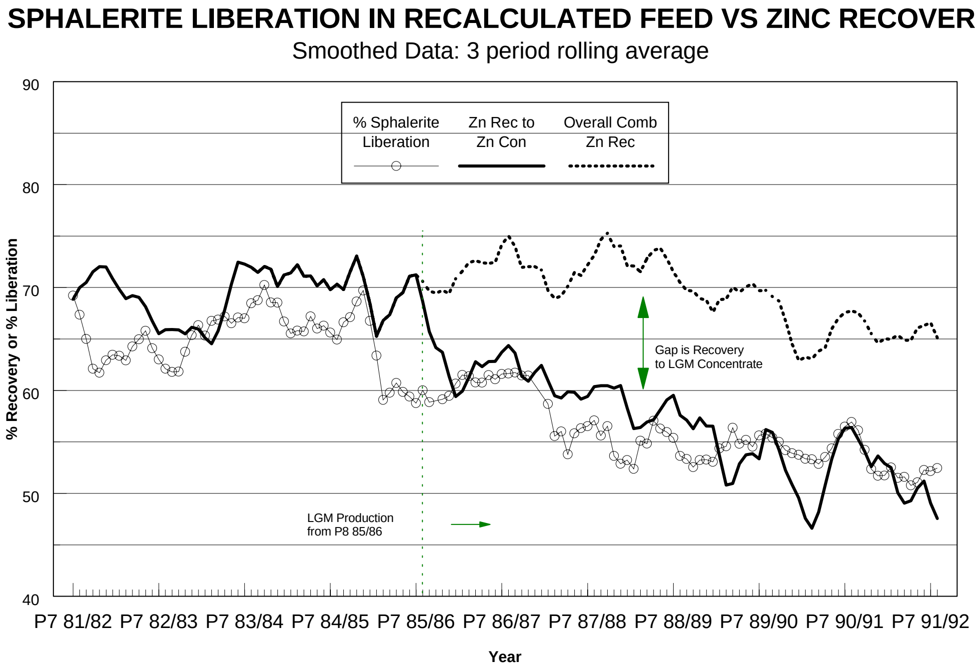


Figure 4: Zinc Liberation and Recovery

At the same time as this trend was occurring, MIM Holdings, the parent of Mount Isa Mines Limited, was embarking on processing the ore from the McArthur River deposit in the Northern Territory. This material was extremely fine grained and required grind sizes down to 7µm to liberate the particles. Obviously, new processing techniques were required to turn around the trend.

ADDRESSING THE LIBERATION AND SEPARATION PROBLEMS PRE 1994

At the start of the nineties there was very little equipment available for fine grinding in mineral processing, (fine being defined as below 25um at least). There was conventional ball milling, and tower milling, and the emergence of metprotech mills, but these were inefficient, as well as impacting on downstream flotation.

Never the less, Mount Isa Mines tried to address the impacts of deteriorating ore by implementing the following as described Pease, Young et al 1997:

- Tower mill regrinding installed in the LGM circuit (1991), dropping the p80 to 12um
- The “Fine Grinding Project,” which doubled grinding and flotation capacity and instituted a “cold” lead reverse cleaning circuit (1992). This project addressed both key issues, ie liberation in the zinc circuit and separation (of carbonaceous pyrite) in the lead circuit. The flotation feed dropped from a P80 of 80µm to a p80 of 37µm due to an increase in secondary grinding from 6.3 MW to 11.5 MW
- Improvement in liberation allowed circuit simplification, the increased use of conventional tools (eg high pH zinc circuit cleaning) and relocation of regrinding duties from the LGM circuit to the zinc retreat circuit.
- Generally, the application of process control became more effective with the improvements because of more achievable targets.

Details of these circuit changes were described in detail by Pease,Young et al 1997. A brief summary of the observations at this stage was that by grinding finer up front in the LGM circuit some of the liberation problems were being addressed. The liberated carbonaceous pyrite contamination of the concentrates were being improved by the reverse flotation in the lead circuit, the use of traditional depressants to knock out the pyrite, as well as the increased capacity to allow for the adequate separation separation.

Of the changes in the circuit it was reported that major gains had been achieved by the extra regrinding with some of the composite zinc that had reported to the LGM having been liberated enough to report to the zinc concentrate. This achieved an extra 5% of zinc recovery to the zinc concentrate. However, zinc liberation, had been increased to 20%, and higher recoveries were expected at this level of liberation than what had actually been achieved.

THE IsaMill DEVELOPMENT

In 1984 to 1986, fine grinding work had been conducted in the pilot plant at Mt Isa to investigate fine grinding circuits. Tower mill test work had also been investigated in 1991, with poor results due to the inability of the equipment to grind fine enough. The other drawbacks of these circuits was the high iron content in the mill charges and the resulting impact on the surface chemistry due its effect on the redox potential.

The introduction of a small stirred mill to the Mt Isa site could only be described as a major turning point in fine and ultrafine grinding in mineral processing. The mill, a ½ litre bench scale mill, resembled a milk shake maker, and initially used fine copper smelter slag to grind the ore and concentrate. It was provided by Netzsch, an experienced fine grinding equipment supplier, in the paint and food processing industry! Testwork on McArthur River Ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the pilot plant at Mt Isa.

In a relatively short time, the influence of the bench scale testwork lead to the development of the first pilot scale mill. Mill development is shown in the following table.

Time	Installed Mill
Jan 1992	LME100 model, powered by a 55KW motor used in pilot plant at Mt Isa
Nov 1992	LME500 model powered by a 205KW, then 250KW motor at the Hilton Concentrator, Mt Isa
Nov 1993	ISA1500 model powered by a 900KW motor installed in the Lead/Zinc Concentrator
Dec 1994	IsaMill M3000 (earlier known as 280/7)powered by 1.1MW motor installed in the Lead/Zinc concentrator
Dec 2003	IsaMill M10000 powered by a 2.6MW motor installed at Anglo Platinum Western Limb Tailings Retreatment Project

Table 1: IsaMill Development Timeline

The table highlights the major mill models that evolved, but does not attempt to show the detail of testwork that was involved with both the mill and flotation circuit downstream of the mill. To go into the development is a case study within itself. Needless to say there were hundreds of trials involved to develop the commercial mill, the M3000, including various component trials for disc materials, spacing of components, separator, liners, chamber lengths, plant and lab testwork, mineralogical investigations and examinations, and most importantly the development of the screenless discharge – ie. product separator.

It is important to note that in just 11 years since the first unit was installed in an operating plant at Hilton Concentrator, the units have increased their capacity 13 fold from 205KW to 2.6MW. Mill volume has increased 20 times in this time frame. This is a very rapid increase in capacity. In comparison, autogenous milling technology took 19 years to increase power draw just 6 times from 1940 to 1959, and this was for a technology that had been around since 1907 in the goldfields of South Africa, (Weiss 1985), (Metso 2004).

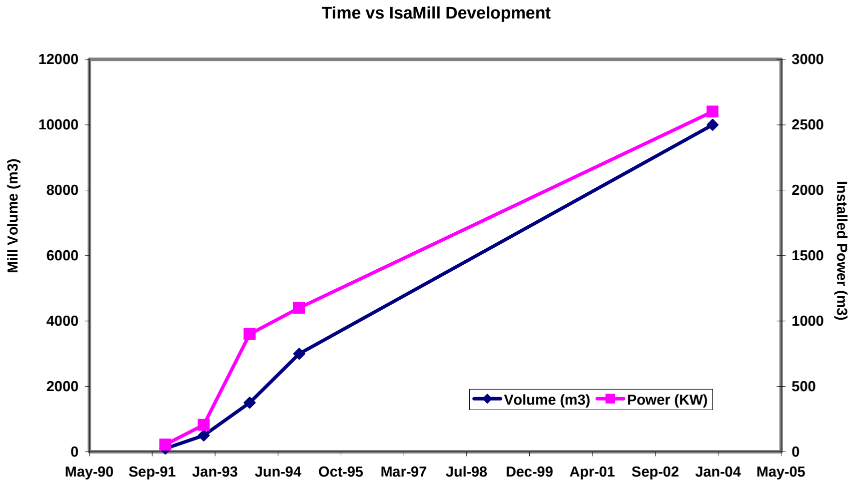


Figure 5: IsaMill Power and Volume Increase

The current model IsaMill, M10000, is powered by a 2.6MW motor, and has enabled large scale mineral processing applications to consider fine grinding as an economic option.

Once installed at Mt Isa and McArthur River, and the resultant success at the operations (as described later), MIM Holdings, allowed the IsaMill to be sold on a commercial basis. This still continues under Xstrata Technology, who actively promote the mill in mineral processing and leaching operations.

Today there are 27 IsaMills installed throughout the world. 14 are installed in Xstrata mines, McArthur River and Mt Isa, the rest are at non Xstrata mines. The mills have been installed by Xstrata Technology and Netzsch.

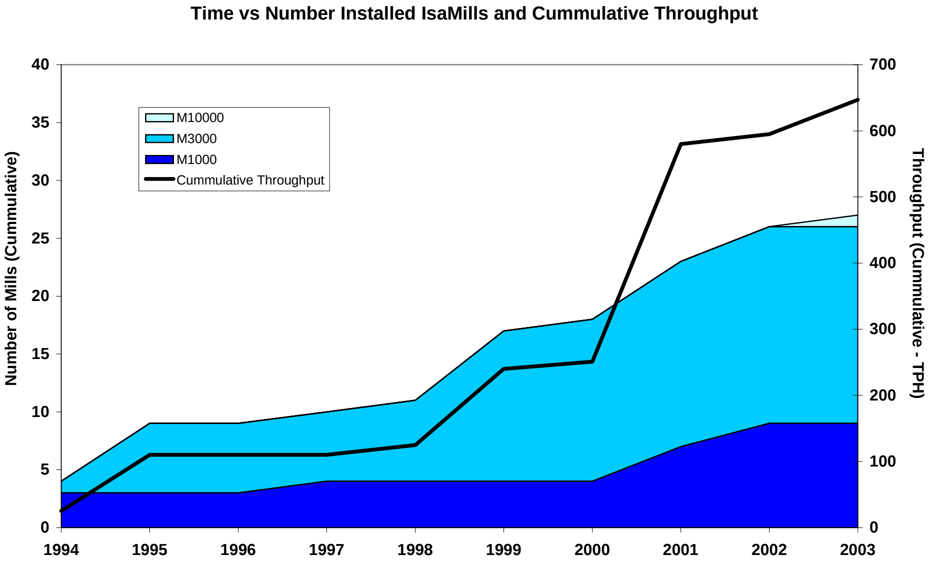


Figure 6: Number of Installed IsaMills (Not counting pilot and lab scale mills)

The technology, while developed to overcome serious liberation problems at the McArthur River and Mt Isa deposits, has become an industry standard in large scale fine grinding applications. It was probably due to its conception at a mine site, and the need to be reliable and robust at these operations that have made it accepted technology in fine grinding.

Of the 27 mills operating at mine sites, approximately 650T/Hr of material is now ground with the IsaMill, or **5.2M tonnes per annum**. The product sizes ranges from a P80 of 7µm to 25µm, for materials ranging from lead and zinc sulphides, platinum concentrates, industrial minerals, iron oxide and refractory gold concentrate. Such a new technology has been embraced by those operations that rely on the finer grinding to achieve metal recovery.

A summary of the Australian lead/zinc operations using IsaMills are as follows:

Operation	Material	Annual Treatment (T/Yr)	P80 (µm)	Recovery (% Metal)	Total Treatment Since IsaMill Installation (MT)
Mt Isa	Zn Cons	350,000	15	>80%	1.7 MT [A]
Mt Isa	Pb Cons	260,000	15	>80%	3.0 MT [B]
McArthur River	Zn/Pb Cons	380,000	7	82%	2.9 MT [C]

[A]: 1994 to June 04

[B]: 1999 to June 04

[C]: 1995 to June 04 – this figure equates to 4.9MT of feed to IsaMills

Table 2: IsaMill Pb/Zn Operations in Australia

In short almost **1MT of lead/zinc concentrate is produced by IsaMills every year in Australia alone**! The average of this material is well under 15µm, which before the development of the IsaMill would never have been economical to treat!

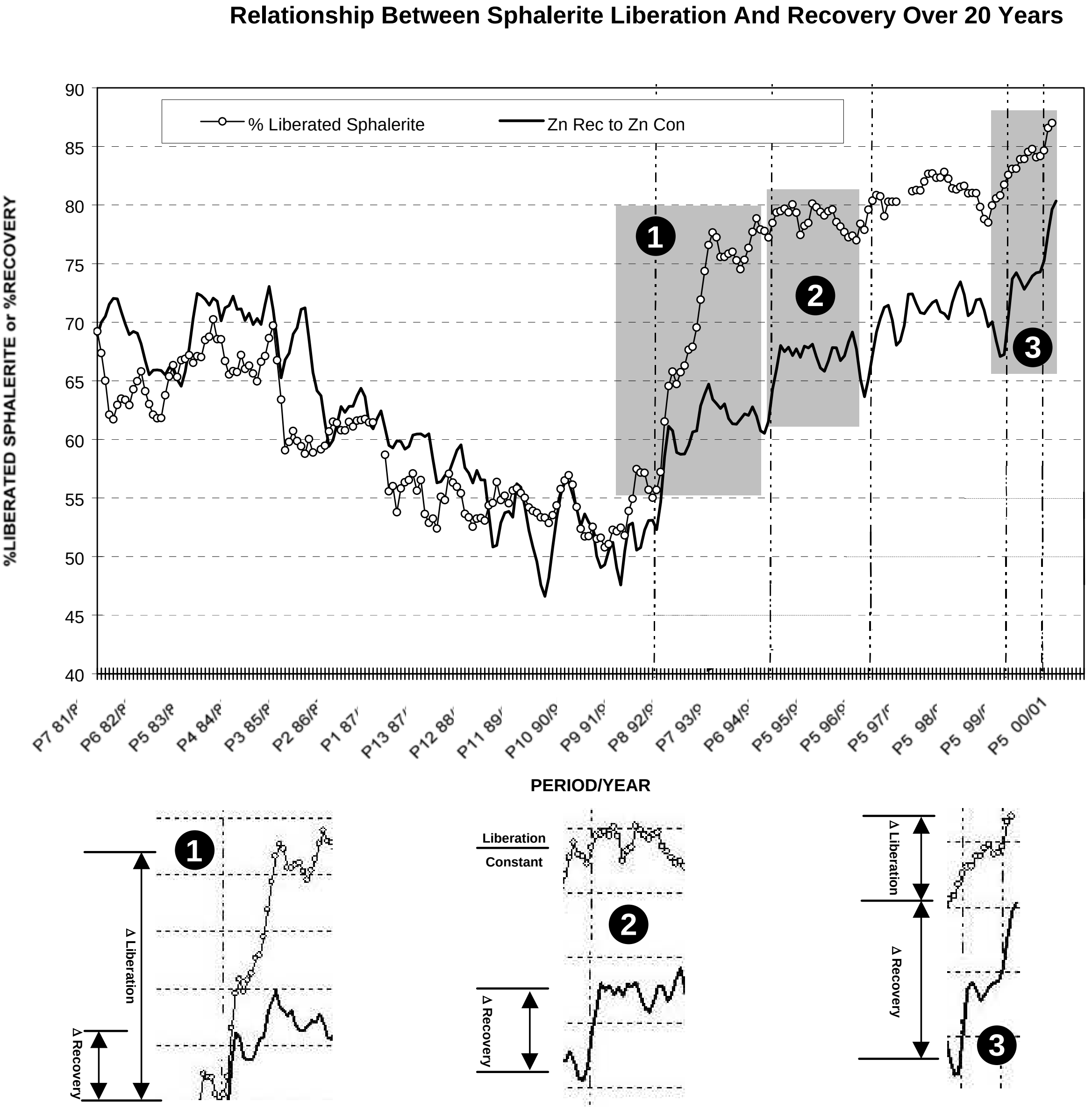
IMPACT OF IsaMills AT MOUNT ISA

As discussed earlier, the addition of more conventional secondary grinding and Tower Mill regrinding resulted in sphalerite liberation improving by 20% (from 55% to 75%), but recovery only improved by 5%. The lack of expected recovery was due to the extra difficulty of floating after grinding in ball and Tower Mills, ie the influence of iron media reducing the slurry.

In 1995 the first full scale prototype IsaMill (to become the M3000) was installed in the lead circuit. Not only did this improve lead performance, but zinc recovery in the zinc circuit also increased even though total sphalerite liberation had changed little. This was because the improvements due to grinding by iron free media from IsaMilling allowed sphalerite to be redirected from the lead concentrate to the zinc circuit. The lead circuit was able to operate more efficiently and separate the lead from the zinc, leaving the zinc to report to the following zinc circuit.

In 1999 the installation of additional IsaMills for the treatment of the George Fisher orebody, increased sphalerite liberation by a further 5%, but zinc recovery increased by 10% and zinc concentrate grade by 2%! This equated to a 16% recovery increase at the same concentrate grade. This demonstrates the fundamental changes to fines flotation made possible by ultrafine grinding in IsaMills with inert media compared to conventional means

This is well described in the following graph, figure 7, showing the dramatic turnaround in liberated sphalerite that was recovered to zinc concentrate. During this period ore quality still deteriorated, as more Black Star and George Fisher was being mined, containing the fine grained ore, as well as increasing levels of carbonaceous pyrite.



Ball & Tower Milling Added

Liberation increased by 20 %

Recovery increased by 5 %

Recovery increased less than liberation due to negative impact of steel media on flotation of fines

First IsaMills Added in Lead Circuit

Sphalerite liberation constant

Zn recovery increased by 4 % due to better rejection from Pb conc after IsaMilling

IsaMills Installed In Zinc Circuit

Liberation increased by 7 %

Recovery increased by 10 %

Conc grade increased by 2 %

Recovery benefit higher than liberation increase because of improved flotation after IsaMilling

Figure 7: Zinc Liberation and Recovery

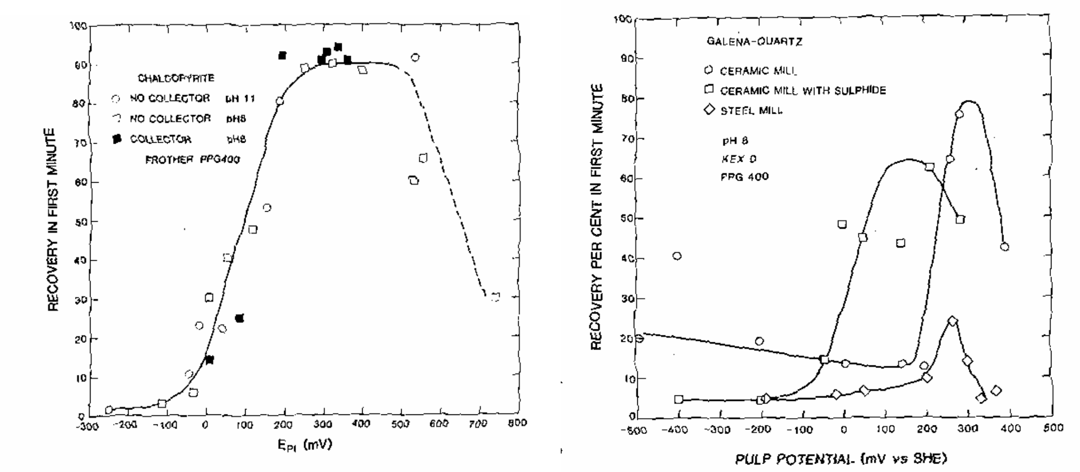
BENEFITS OF IsaMills

From the Mt Isa experience, size is important. It was important to get to 10µm to liberate the complex ore in the zinc circuit. The lead circuit was courser at 25µm. However even with the use of conventional mills and Tower Mills that could get to 25µm, the expected recovery was not achieved. It was only with IsaMills that the size and flotation performance improved. Why?

Three major benefits are gained from IsaMills over competing technologies reported by Pease et al 2004. They were:

- **Impact of grinding on flotation performance.**

  Milling using steel balls as media will effect the Eh as the balls reduce the pulp, creating negative redox potential. This reduces metal recovery, and will only improve as the flotation is oxidised. This is well documented as shown by work in figure 8 (Trahar 1984). The graph showing the impact of ceramic media is especially relevant to IsaMills as the sand media, like ceramics, are inert and has negligible effect on the pulp chemistry. Also conventional and Tower Mills generally have much higher residence times than IsaMills, resulting in lots of steel contamination. In these circuits additional reagents will be required, reducing selectivity or use innovative processes such as High Intensity Conditioning (HIC), eg at Hellyer, to reverse the negative impact of Tower Milling on surface chemistry. Processes like IsaMilling are far more efficient by providing this high intensity as part of the grinding action, and grinding in an inert environment.



Floatability-potential curve for chalcopyrite in presence and absence of collector.

Floatability-potential curves for galena at pH 8 after grinding in different environments.

Figure 8: Redox Potential vs Recovery

- **Power efficiency**

Even if size reduction could have been achieved by conventional milling, tower mills, or other stirred mill technology, only IsaMills provide efficient use of energy. Figure 9 compares the power required to grind a gold ore in a ball mill with 9 mm balls with an IsaMill with 2 mm media. The IsaMill is much more efficient below about 30 µm; to grind this ore to 15 µm would take 28 kWh/t in the IsaMill, but 90 kWh/t in a ball mill. Traditionally this has been attributed to the difference between attrition grinding and impact grinding. However by far the most important factor is media size, as shown by Figure 10, is the breakage rate. In Tower Mills this drops dramatically; the breakage rate for a 20 µm particles is ten times lower than the rate for 40 µm particles. Even though the Tower Mill is full attrition grinding, practically it is constrained to using relatively coarse media, 9mm balls in this case. In contrast, the IsaMill (Netzsch mill in Figure 11) can operate with much finer media and much higher intensity of power input (Table 3), meaning the peak breakage rate occurs at 20 µm, and doesn't drop as quickly below that.

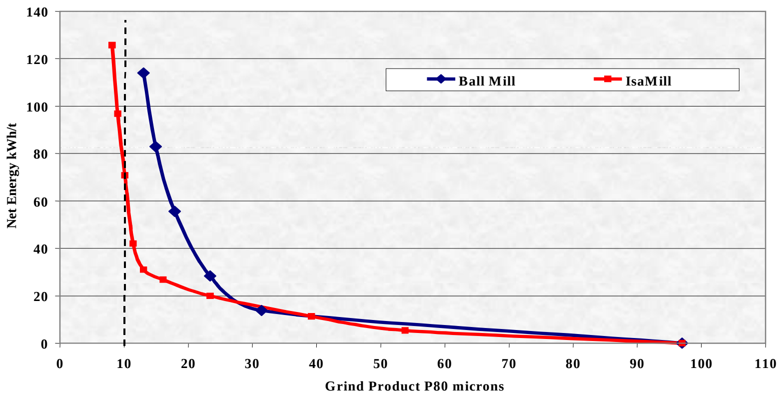


Figure 9: Efficiency of IsaMills Compared to Ball Mills

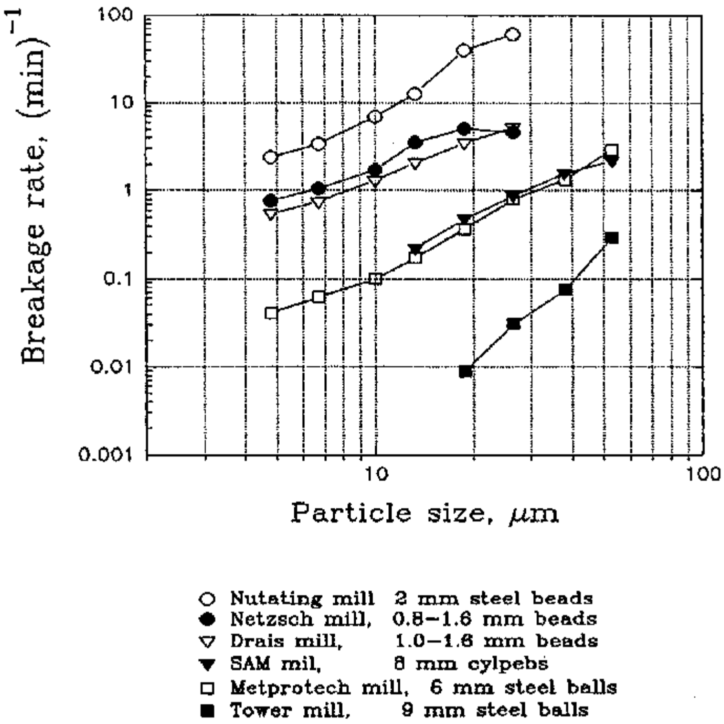


Figure 10: Particle Size vs Breakage Rate for Fine Grinding Mills

(Extracts from AMIRA P336, Gao M and Weller K, Review of Alternative Technologies for Fine Grinding, November 1993)

Comparison of Various Grinding Technologies
Independent laboratory data

FEATURE	ISAMILL	TOWER MILLS	VERTICAL PIN MILLS
Grinding Intensity (kW/L)	**0.54**	**0.005**	**0.15 - 0.18**
Residence Time to 15 µm (min)	**0.6**	**154**	**7 - 9**
Power Usage to 15 µm (kWh/t)	**17.4**	**59.6**	**37.5 - 39.0**
Media Material	**Various**	**Steel**	**Steel**
Media Size (mm)	**0.8 - 1.6**	**9 -12**	**6 - 8**

Table 3: IsaMill, Tower Mill and Vertical Pin Mill Comparison
(Extracts from AMIRA P336, Gao M and Weller K, Review of Alternative Technologies for Fine Grinding, November 1993)

The influence of the sand media greatly improves the chances of contact of the grinding medium and particle. It comes down to simple geometry; the smaller the size the bigger the surface area, and the more particles per volume. This creates a greater chance of contact, as highlighted in Table 4.

	Power Intensity (kW/m^3)	Media Size (m)	No. Balls / m^3	Surface Area (m^2/m^3)
Ball Mill	20	0.02	95541	120
Tower Mill	40	0.012	442321	200
IsaMill	280	0.001	1146496815	3600

Ball Mill is a 5.6m D x 6.4m L @ 2.6MW
Tower Mill is a 2.5m D x 2.5m L @ 520KW

Table 4: Mill Comparison of Media Size, Power Intensity

- **Good classification**

Like all grinding operations, good classification is vital for power efficiency in ultrafine grinding. But the problem posed by particles less that 15 µm is how is it possible to classify this low using conventional technology? In short it is difficult. It is not generally practical to use cyclones to close-circuit a grinding mill with a target below about 15 µm. To get good cyclone efficiency at these sizes requires small cyclones, eg two inch (50 mm) diameter or smaller. This is virtually inoperable on a large scale, so the circuit is either compromised (and less power efficient) by using bigger cyclones, or an alternative solution is needed. The IsaMill at the Mt Isa operations achieved the cut size of 10µm in the zinc circuit by using an internal classifier mechanism, figure 11. This device uses the high centripetal forces generated inside the mill to classify the discharge, ensuring a very sharp product size without external cyclones. At the same time it prevents the fine media from passing out of the mill meaning that low cost media can be used, eg local sand, or granulated smelter slag. Also the very short residence time in the IsaMill also minimises “overgrinding”, further contributing to the sharp product size distribution. This novel classification device replaced the conventional screen that was on the prototype mills in the early 90’s, as it was found impractical to screen out fine particles.

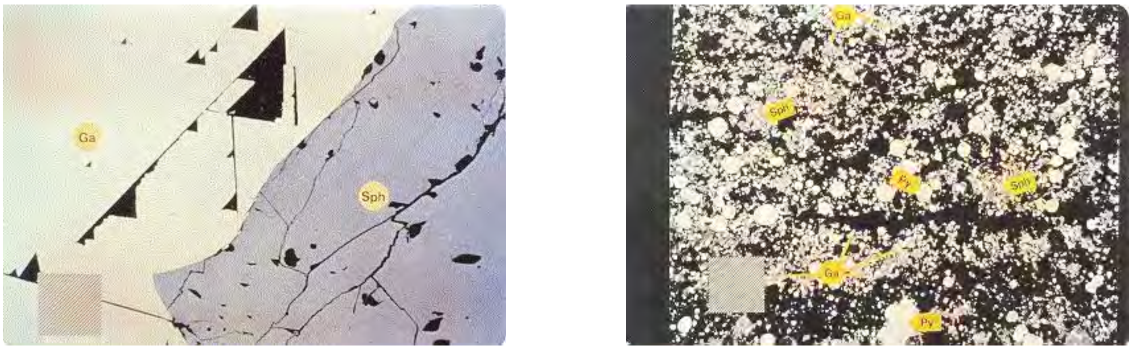
Figure 11 – Patented IsaMill Product Separator

IMPACT OF IsaMills at McARTHUR RIVER

As discussed earlier, the trialing of IsaMills at Mt Isa was to evaluate the use of this new technology for McArthur River. It was well documented the orebody was terribly fine, and was one of the reasons why the orebody had not been developed. It had been discovered in an exploration campaign in 1955 (Legge 1990), but the fine grain structure meant that conventional grinding couldn't liberate the grains fine enough. Figure 12 below, shows a sample of the ore compared to the courser grain structure of a Broken Hill sample (Pease, 2004)

Figure 12 : Different Grain Size of Broken Hill and McArthur River Ores

(Grey Square is 40um)



Broken Hill Ore

McArthur River Ore

However, the success of the IsaMill at the Mt Isa Lead/Zinc concentrator, meant that this deposit could be processed. The IsaMills were described as "enabling technology", and had the ability to grind down to 7µm to produce concentrate economically.

The plant started mid 1995 with 4 IsaMills, followed by another mill in 1998, and another in 2001. Circuit changes over the years have now resulted in a bulk concentrate being produced, where all rougher concentrate is presented to the IsaMills. A cycloning stage separates approximately $^2/_3$ of the rougher concentrate for regrinding by the IsaMills. It is interesting to note that while the P80 target is 7µm, approximately 50% of the final concentrate is less than 2.5µm. If this was viewed from a number of particles basis, it would mean **96% of the particles are less than 2.5µm report to the final concentrate**, (Pease 2004). This shows it is possible to grind fine and float fine particles.

INFLUENCE OF IsaMilling ON HYDRMETALLURGICAL APPLICATIONS

IsaMills are finding a potential use in fine grind leaching applications. It is currently a very important part in the Albion flowsheet, and has been looked at for other hydrometallurgical leaching processes for nickel and copper sulphides.

In the hydrometallurgical industry the Albion Process is one of the simpler processes. It is a process designed for the oxidative leaching of refractory base and precious metal bearing sulphide ores. The leaching occurs in conventional, non-pressurised reactors, which significantly reduces the capital cost compared to pressure and bacterial leaching processes.

What allows the leaching to be undertaken in relatively mild conditions is the addition of an IsaMill in the grinding circuit. The IsaMill produces mineral particles with a high degree of residual strain in the crystal lattice, and a very high surface area. This results in very high defect density within the individual mineral grains, resulting in the mineral being extremely active toward oxidation, (Hourn 1999). The conditions required to oxidise the mineral particle are less extreme than other leaching processes, with oxidation carried out at atmospheric pressure in agitated tank reactors. This is due to the mineral particle being highly fractured from the IsaMilling stage, which enables it to fall apart as it is leached. When the mineral is copper sulphide, the disintegration of the particles prevents the formation of the passive layer that prevents further leaching of the particles. The leach residence time is typically less than 24 hours, and the leach does not use any reagents other than acid, limestone and oxygen.

While other fine grinding technologies can be applied to the ALBION flowsheet, the IsaMill has the advantage of producing a very tight particle size distribution, as displayed in figure 13. The IsaMills are constructed with 8 grinding disc in a small chamber. Every disc acts as a separate grinding operation, therefore 8 disc implies 8 grinding operations. When this is coupled with the product separator, all chances of short circuiting are eliminated. However, other fine grinding practices have a tendency to produce a broad particle distribution as they short circuit some of the feed material.

The tight size distribution is an important factor in leaching operations. Where a flotation process can have slightly oversized particles, which can either report to concentrate or regrinding stages without any major effect on circuit recovery or grade, a leach circuit can be greatly impacted by the presence of large particles. Leaching processes need to have particles that are small enough to allow the leachant to fully leach the particle, otherwise the oversize will represent a recovery loss as the mineral hasn't had the opportunity to be leached. That is why P98 is important in leach circuits.

Work conducted by Hydrometallurgy Research Laboratory (now a part of Xstrata Technology) conducted several test using different fine grinding bench scale mills, leaching copper sulphide concentrate using the ALBION process. While all fine grinding methods produced a similar P80, the IsaMill produced the finest P98. When the products were then leached, the IsaMill produced an extra 3% of leachable copper compared to other processes. This is displayed in figures 14, where the ratio of P98/P80 is plotted for each grinding process. The ratio of P98/P80 is an indication of how tight the feed sizing is, ie the closer to 1 the sharper the cut.

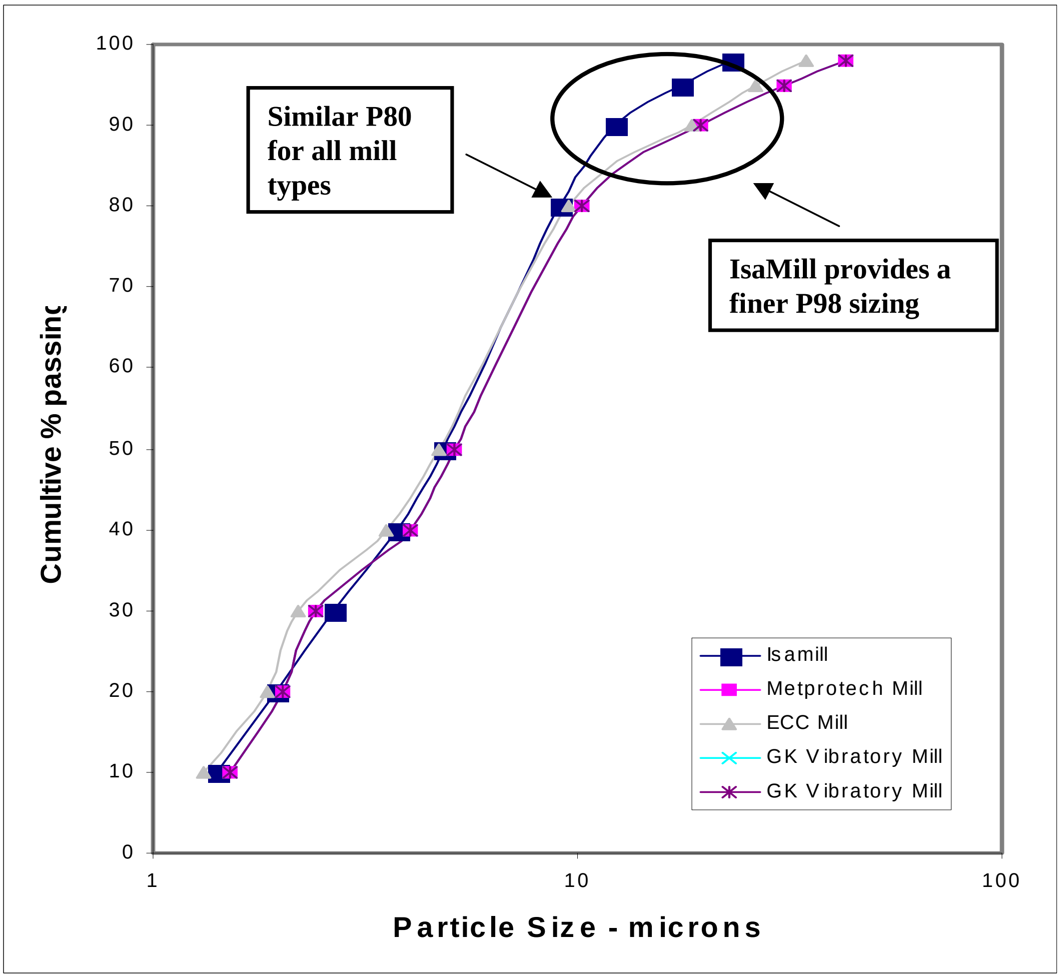


Figure 13 – Particle Size vs % Passing per Fine Grinding Method (Copper Bulk Concentrate)

% Passing - microns	IsaMill	ECC Mill	GK Vibratory Mill	Metprotech Mill
98	23.1	34.4	42.8	51.90
95	17.44	26.33	30.61	33.40
90	12.31	18.6	19.44	23.41
80	9.11	9.56	10.2	8.95
50	4.85	4.71	5.12	4.10
40	3.76	3.55	4.03	3.31
30	2.66	2.21	2.41	2.31
20	1.94	1.86	2.02	1.96
10	1.42	1.32	1.51	1.61

Table 5 – Fine Grinding Method vs Product Sizing (Copper Bulk Concentrate)

.

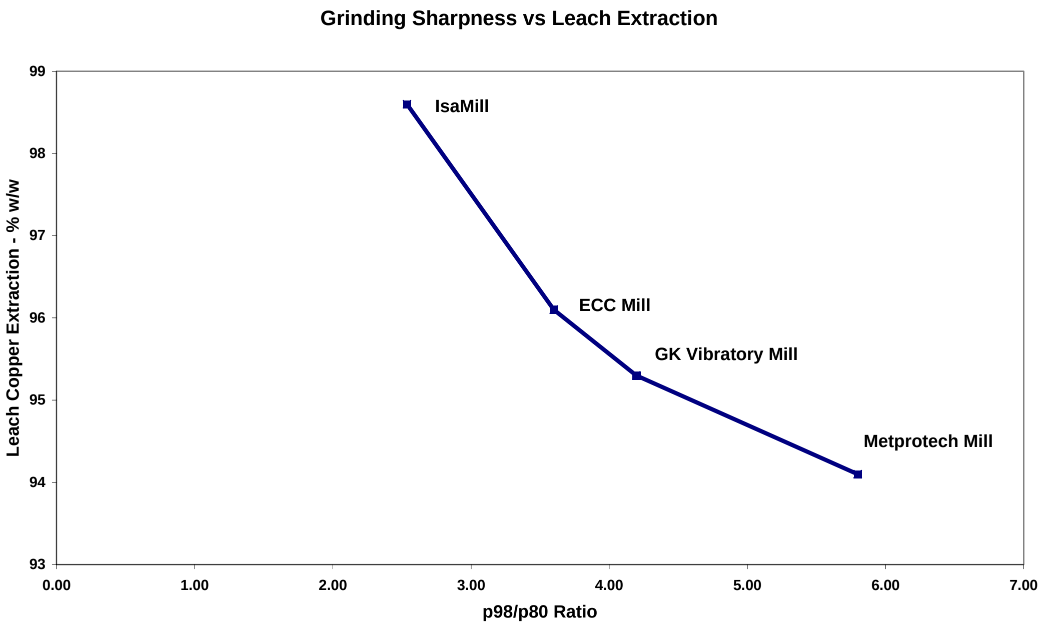


Figure 14: P98/80 Ratio vs Leached Copper Extraction (Copper Bulk Concentrate)

Finally, after the oxidative leach stage of the ALBION Process, the iron is precipitated out as goethite. This stage precipitates the iron and acid liberated during the oxidative leach, as well as neutralising any acid remaining in the slurry ahead of thickening and filtration. Goethite precipitates grow as coarse particles, considerably improving the settling properties of the finely ground leach residue.

The particle size distribution of the goethite residue is significantly coarser than the leach feed. A goethite precipitation stage also has the added benefit of precipitating any arsenic in the feed concentrate as a ferric arsenate, improving the stability of the arsenic phase in a tailings impoundment.

REFERENCES

J D Pease, M F Young, D Curry and N W Johnson, 2004, Improving Fines Recovery by Grinding Finer, MetPlant 2004

M F Young, J D Pease, N W Johnson and P D Munro, 1997, Developments in Milling Practice at the Lead/zinc Concentrator of Mount Isa Mines Limited from 1990, AusIMM Sixth Mill Operators Conference

Metso Minerals, Brochure 2001.1 06/02, Autogenous/Semi-Autogenous Mills

Trahar, The Influence of Pulp Potential in Sulphide Flotation, Principles of Mineral Flotation, The Wark Symposium, AusIMM 1984

P D Munro, 1993, Lead-Zinc-Silver Ore Concentration Practice at the Lead-Zinc Concentrator of Mount Isa Mines Limited, Mount Isa, Qld, AusIMM Monograph 19

Weiss, 1985, SME Mineral Processing Handbook – page 3C-57 to 34C-58

M Hourn, D Halbe, 1999, The NENATECH Process: Results on Frieda River Copper Gold Concentrate

P J Legge, C O Haslam, 1990, Periods of Lead-Zinc-Silver Exploration in Australia, AusIMM Monograph 17.

M Gao and K R Weller, K R, 1993. Review of Alternative Technologies For Fine Grinding, AMIRA Project P336, Report P336/20, November.

Ultra Fine Grinding - A Practical Alternative to Oxidative Treatment of Refractory Gold Ores

S. Ellis
Kalgoorlie Consolidated Gold Mines
Kalgoorlie, WA.

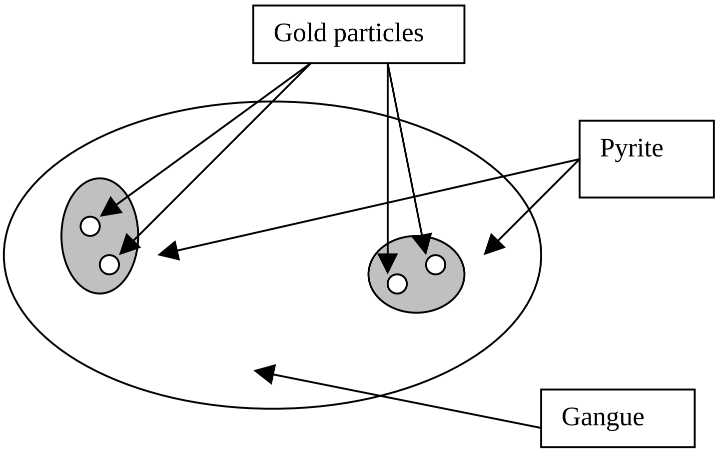
Abstract

Since early 2001, Kalgoorlie Consolidated Gold Mines (KCGM) has successfully operated an ultra fine grinding (UFG) circuit to supplement its roaster capacity for the treatment of the refractory gold flotation concentrate. A second UFG mill was installed in 2002 taking the total UFG capacity to over 20tph while grinding to 11-12 microns and achieving over 90% gold recovery. A development program in 2002/3 involving plant trials, pilot plants and laboratory testwork resulted in process improvements and a better understanding of the milling and gold leach processes that assisted in narrowing the gap between UFG and roasting . This paper details the operation of the ultra fine grinding process at KCGM as a non oxidative treatment for the extraction of gold from a refractory ores.

Introduction

The refractory nature of gold ores is often associated with gold finely disseminated in sulphide minerals, such as pyrite, at conventional grind sizes. Conventional milling can liberate the pyrite from the gangue allowing a low mass pyritic concentrate to be produced by a process such as flotation. However, direct leaching of the concentrate results in poor gold extractions as the cyanide lixivant is unable to contact the gold locked or included within the pyrite (Figure 1).

Figure 1 Pyrite Locked Gold Within an Ore



The traditional approach for such refractory material has been to liberate the locked gold by chemically destroying the pyrite through oxidation. Roasting, pressure oxidation, and bacterial oxidation all employ various degrees of temperature, pressure and catalysis to react the pyrite with oxygen to produce an iron oxide and sulphur by-products. This method efficiently liberates finely disseminated gold or gold in solid solution.

Whilst such oxidative reactions are metallurgically sound and are capable of achieving high metal recoveries, the environmental aspects of treating the reaction products can alter the economics of the process.

For example, capture and disposal of sulphur dioxide from the roasting of sulphides or the neutralisation of the acidic liquors from pressure oxidation, may add significant additional costs to the process. In certain cases, these additional costs may make an alternative process route more economically attractive.

Fine Grinding

An alternative, applicable to the liberation of disseminated gold from the host mineral, is to continue the grinding process to further reduce the particle size of the host mineral thereby exposing a part of the gold surface for contact with cyanide solution. A benefit of this technique is that the host mineral is not destroyed in an oxidative chemical reaction with the resultant problems of treatment of the reaction products. Such fine grinding, however, has proven to be increasingly energy intensive with each size reduction step. In pit blasting, primary crushing, secondary crushing, SAG and ball milling are all able to exploit natural fracture planes in the ore allowing breakage along these features. As progressive size reduction occurs, the naturally occurring minerals are liberated and a point reached where the crystal structure of the mineral has to be broken for further size reduction to be achieved. This may present a significant barrier to further breakage with higher power intensities required to achieve a breakage event.

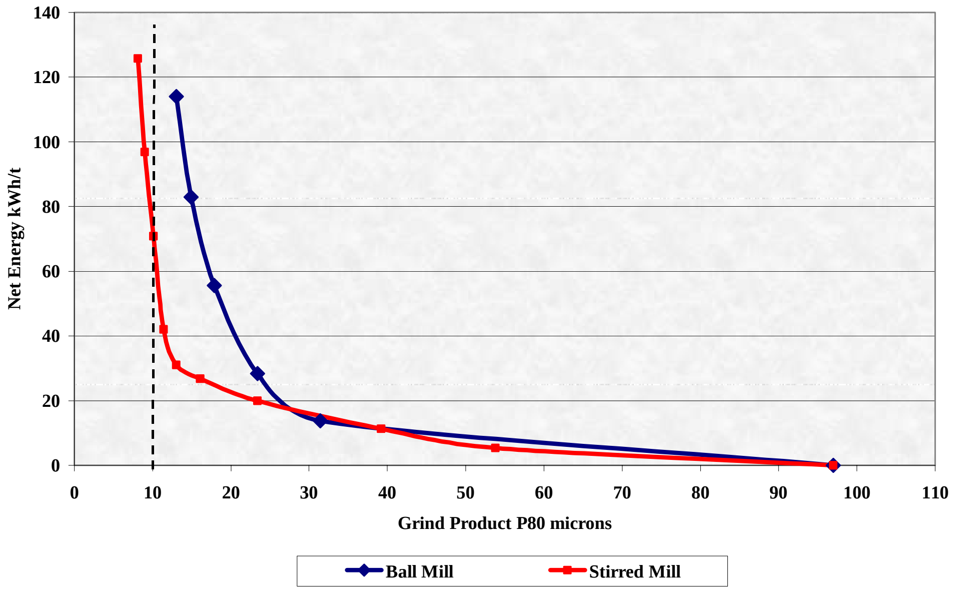
In the past, inefficiencies associated with conventional milling have made fine grinding unattractive to the mineral processing industry. The desired grinds for the harder minerals could only be achieved by prolonged milling with resultant low throughput and high power consumption. The use of smaller media in closed circuit can assist tumbling mills to achieve fine grinds but remain fundamentally limited in the manner in which they impart kinetic energy to the media as well as having large "dead zones" where little media movement occurs (Kalra,1999). As finer media is used, the kinetic energy imparted to the media lessens thus significantly reducing the available energy transfer in a media/particle contact event.

Ultra Fine Grinding

UFG mills overcome these limitations by the use of rotating stirrers inside a stationary mill shell. Ultra fine grinding mills have been in use for many years in a large number of every day applications such as pharmaceuticals, dyes, clays, paint and pigments before being used in the mineral processing industry. They usually fine grind in a range of l µm to 10µm, and impart a significantly increased surface area as well as other potentially desirable properties such as colour, ease of absorption into the blood stream and increased chemical reactivity.

Figure 2 compares the power consumption of a laboratory ball mill to a UFG mill in grinding KCGM concentrate.

Figure 2 Comparison of Grind Product Sizes



The use of UFG grinding in the minerals processing industry is a relatively new development being based on the smaller low mass, batch UFG mills being used by other industries for high value products.

The chief requirement of the minerals processing industry was a mill that could process quantities in the order of several tonnes per hour in continuous operation whilst maintaining cost efficiency in power and media usage.

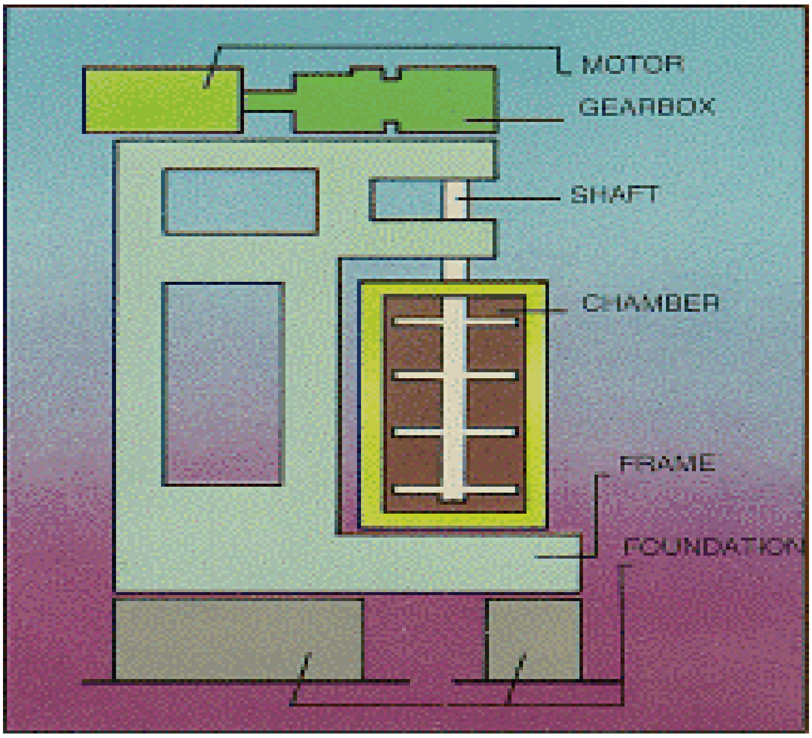
In achieving finer grinds, UFG mills use a finer media size (2-3 mm) than conventional milling (12-100mm) with a much higher installed power per mill unit volume (Table 1).

Table 1. Typical Mill Grinds and Power Intensities

Type of Mill	Typical Lower Grind Size P80 µm	Power Intensity kW/m^3
Ball Mill	**75**	**20**
Tower Mill	**20**	**40**
UFG Mill	**5**	**280**

Two basic types of UFG mills are available, the vertical stirred mill and the horizontal stirred mill. Both use rotating stirrers within a stationary mill shell to impart kinetic energy to a fine media charge (usually sand). The breakage mechanism is the same for the two mills, the differences being related to stirrer speed, method of media retention, and size of currently available mills.

Figure 3. Arrangement for a Vertical UFG Mill



Breakage and Particle Reactivity

As well as the resultant increase in the degree of liberation of the mill products, UFG also increases the surface area of the products enhancing the rate of downstream chemical reactions. The application of intensive non breakage stress events is believed to distort the mineral crystal lattice creating new defect sites which have high localised electron densities. These high electron densities facilitate the transfer of electrons to an oxidant thereby significantly increasing the rate of chemical reactions (Hourne and Halbe, 1999). This can result in a lowering of the activation energy for chemical reactions of the mill products and allow the reaction to proceed at lower temperatures and pressures than for the unmilled material.

This increase reactivity is demonstrated by the very high oxygen demand observed for pyrite after it is subjected to UFG .

The Activox and Albion processes make use of this phenomenon in their leach steps to achieve sulphide dissolution at reduced temperature and pressures to conventional pressure oxidation. However, for gold recovery, this increased rate of chemical reaction of the pyrite presents unwanted side reactions that can result in increased cyanide and lime consumptions.

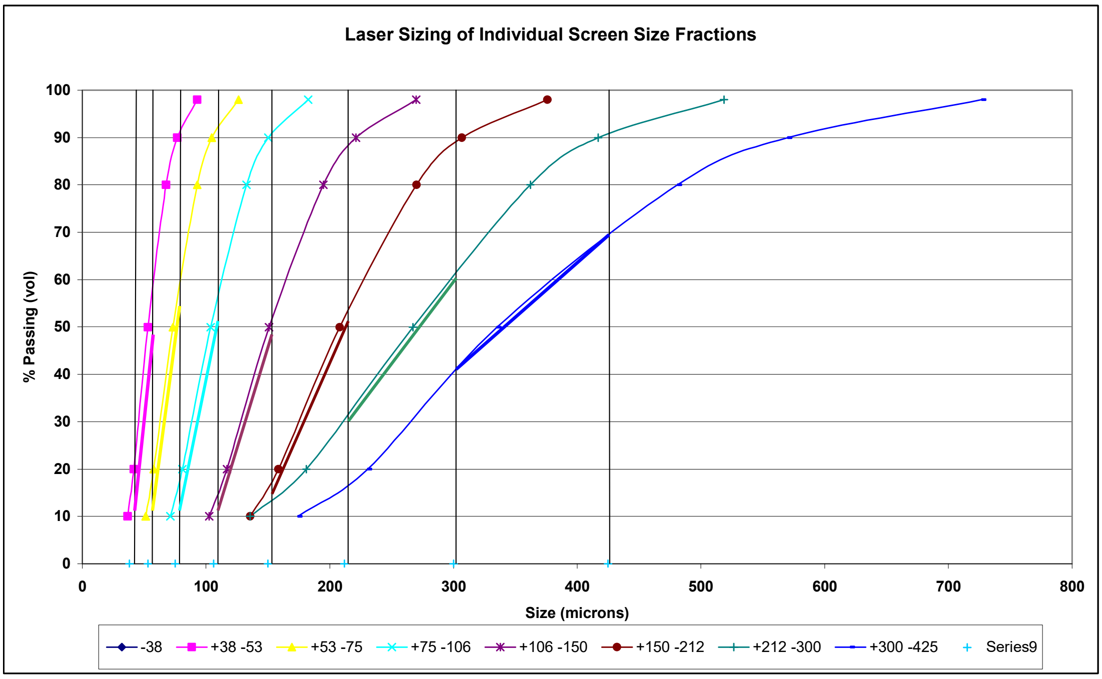
Sizing Measurements of the Product

Over time, laser sizing has gained popularity and has now become the de facto standard for fine particle size measurement. Limitations with the earlier laser machines have been overcome with the development of sufficient on board computing power to use the Mei Equation relating light scattering characteristics to particle size. The laser measures an average particle volume and converts this data to an average particle diameter. The laser method is fundamentally different from a screen sizing which measures the diameter of a particle that can pass through a screen aperature.

Testwork conducted at KCGM (Turton-White, 2003) showed that the source of the greatest error in carrying out a laser sizing measurement was the sub-sampling of the bulk sample to the very small sample presented to the machine. It was shown that samples of greater than 50µm required an additional sub-sampling step to ensure a reproducible measurement.

Caution should be exercised with the practice of screen sizing out the coarse particles and laser sizing the screen undersize and combining both sets of results. Figure 4 highlights the different results than can be obtained when size fractions from a dry screening are analysed by the laser.

Figure 4. KCGM Comparison of Laser and Screen Sizings



Method of Breakage

Stress Intensity has been examined as a critical determinant of the kinetic energy contained by the media in motion (Becker, 1997) and can be described in terms of the media diameter d, the stirrer tip speed v, and the media density ρ.

$$\text{Stress Intensity (Nm)} = d^3 \times v^2 \times \rho \qquad \textbf{Equation 1}$$

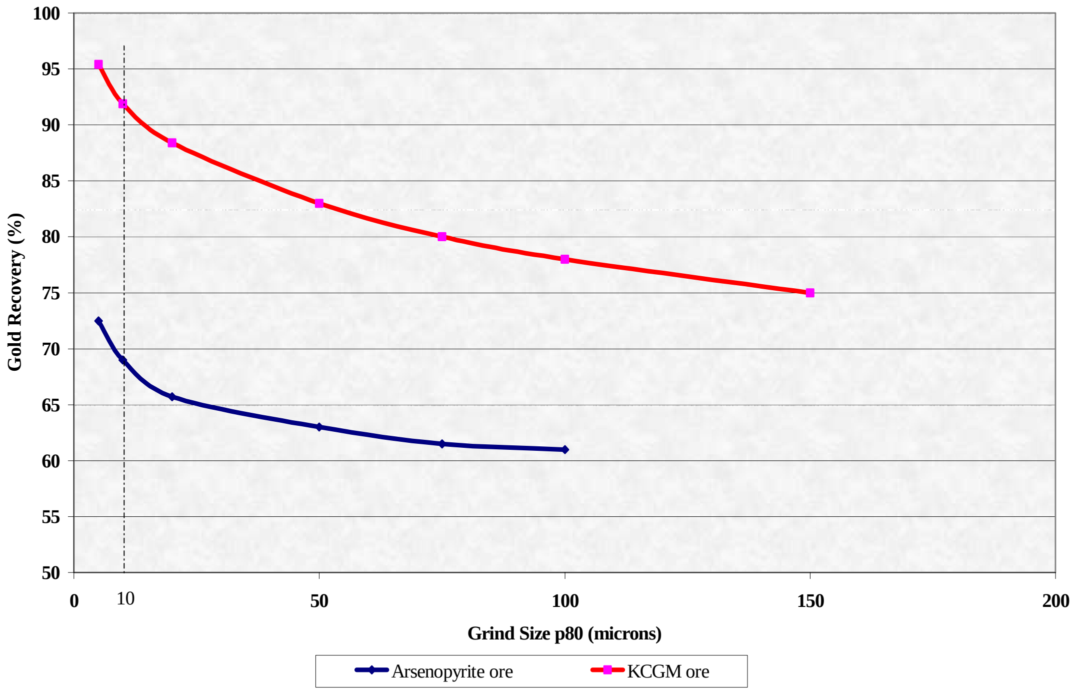
As shown in Equation 1, the Stress Intensity is related to the cube of the diameter of the media. For the same media type and mill speed, the Stress Intensity is increased by a factor of eight if the media size is increased from 3mm to 6mm. This highlights the importance of the inter-relationship between the top size of the mill feed, the selection of media size, and the wear on mill internals.

Kwade (1999) concluded that impact breakage rather than attrition was the main breakage mechanism in stirred mills for particles larger than 1µm. Recently Yue and Klein (2003) have examined breakage in successive grinding cycles and have found that a grinding limit exists where increasing grind time no longer results in particle breakage. They have postulated that this grinding limit is a function of the applied stress intensity and have noted that as this limit is reached, attrition rather than fracture becomes the chief breakage mechanism.

Sample Testwork

Not all refractory gold ores give a large recovery improvement after fine milling. Gold locked in arsenopyrite for example does not achieve the same gold recovery as gold disseminated in pyrite due to the smaller gold particle size of the locked gold (Figure 5).

Figure 5. **Gold Recovery of Arsenopyrite and Pyrite Ores**



Preliminary metallurgical testwork can be readily undertaken to determine the leach response to UFG. This involves the milling of a samples to a variety of nominated grind sizes and the leaching of the milled product. This not only determines the potential recovery but also gives a first pass economic assessment of the main economic drivers - power consumption, reagent costs, and gold recovery. Such preliminary data allows a comparison to be made with other processing routes.

Along with the determination of the grind/recovery curve, a mineralogical assessment of the gold deportment and particle size may provide additional information. Mineralogical scans such as QEMScan and MLA are able to give good data as to the mineral liberation and particle size of gold down to about 5µm. Below this size, an alternative method such as secondary ion mass spectrometry (SIMS) should be used.

Ultra Fine Grinding at KCGM

KCGM examined many concentrate treatment options that could provide an alternative to the roasting process in use. Chief among these were pressure oxidation, bacterial oxidation, and ultra fine grinding. An economic study carried out in 1997, determined that UFG had the superior Net Present Value return of the alternative options examined but was still a higher cost option than roasting as conducted at KCGM's Gidji site.

To progress UFG to the detailed engineering stage, further testwork was undertaken to define the flowsheet. A full pilot plant trial of the milling and leach process was carried out by Amdel (Adelaide) in 1999 which confirmed the recovery and power consumption indicated by laboratory scale testwork. The grinding of KCGM concentrate was amongst the highest energy consumers when compared to other UFG applications (Table 2); demonstrating that power consumption is very case specific.

Table 2. IsaMill Comparative Milling Data for Various Mineral Concentrates

Type of Ore	Lead Rougher Concentrate (PbS)	Zinc Rougher Concentrate (ZnS)	Lead/Zinc Rougher Concentrate	Copper Concentrate (CuFeS)	KCGM Gold Concentrate (FeS_2)
Feed F80 micron	25	30	65	120	120
Product P80 micron	10	10	10	10	10
Net Energy kWh/t	10	20	30	100	80
Throughput t/h	96	48	32	9.6	12.0

The work demonstrated that the recovery of gold from KCGM concentrate could be increased from 75% to 92% by grinding to 10µm.

Mill Selection

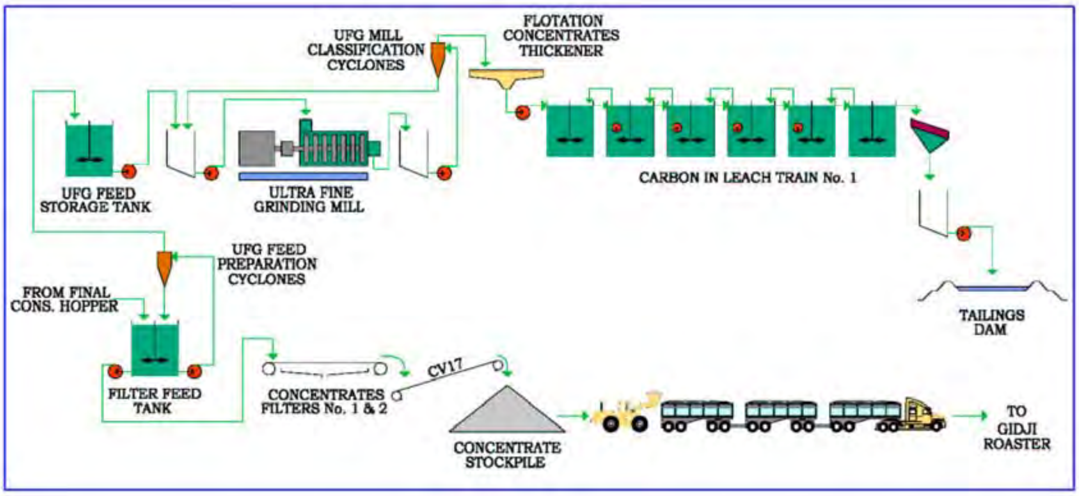
In the first instance equivalent samples of final concentrate were sent to all major mill suppliers for assessment. Results highlighted the lack of uniformity in the test methods and in the methods of sizing. No useful comparison could be made between results.

Site visits were useful in ascertaining the maintenance requirements of the mills, and general impressions of operators but with the variety ore types, hardness, and feed size to the mill any difference in metallurgical performance was difficult to gauge. From the preliminary assessment, the choice of mill was narrowed to the Detritor and the IsaMill. To discriminate between these mills, a pilot scale comparison was made by bring both the 18.5 kW pilot Detritor and the 55kW pilot IsaMill to site. The two mills performed similarly with no measurable difference in product quality or in power consumption under equivalent conditions. Given that no significant difference in metallurgical performance was apparent, KCGM opted for the IsaMill based on its mill size, low maintenance requirements, and its proven performance.

KCGM Flowsheet

For KCGM the main process risk lay with the unproven cyanide leaching of 10µm pyritic ore with acceptable reagent consumptions. In order to minimise this risk and to reduce capital costs associated with the process, KCGM determined to use single stage milling acknowledging that some inefficiency in power consumption and general mill and cyclone operation would result from the large size reduction step. To minimise the impact of the large feed size, a pre-treatment cyclone step was used to lower the feed size to the UFG mill by rejecting the coarser material to the cyclone underflow. The underflow was directed to the roasters for processing where the size was not of major significance. Whilst this approach assisted in reducing the feed size to the mill, it was subject to variations driven by the limited amount of suitably sized material in the concentrate. A further disadvantage lay in the pre-treatment cyclones directing an increased quantity of fine gangue material to the mill. The final flowsheet for the KCGM process is shown in Figure 6.

Figure 6. KCGM Flowsheet



The key design criteria for the process is shown in Table 3.

Table 3. UFG Key Design Criteria

Item	Design
Concentrate Processing Rate	10 t/h
Deslimed Concentrate Solids Size Distribution (F80)	120 microns
UFG Feed Prep Product Solids Size Distribution (P80)	50 microns
Initial Concentrate Gold Grade	40 to 50 g/t
UFG Feed Prep Product Solids Concentration	45 % to 55% w/w
UFG Product Solids Size Distribution (P80)	10 microns
Gold Recovery	92 %

Media Selection

The media size selected for the duty was based on the top size of the feed likely to be introduced to the mill. Pilot plant experience had shown that running too little media or too small a media size could result in a build up of unbroken feed top size in the mill. This build up in feed material within the mill led to a locking and centrifuging of the charge with a marked drop off in power draw. To ensure that the feed top size was adequately broken under all conditions, 6mm top size sand media was selected for the duty. Silica sand was selected due to its relatively low cost with initial supplies sourced from Northern New South Wales. Despite Australia wide searches for alternative supplies, only a few locations have been identified as having sand media of sufficient competency to achieve acceptable media consumption rates. Steel, smelter slag and ceramic medias were tested but rejected on the ground of cost, high consumption rates, and/or the introduction of deleterious materials (iron) to the leach process.

The use of a large media size (6mm) whilst ensuring top size particle breakage, resulted in severe wear on the mill internal components. Typically, wear associated with the leading mill disc, necessitated a mill stoppage every ten days compared to several months at other IsaMill installations running with 3mm media.

Media size is a critical determinant of the kinetic energy imparted to the media and is related to the cube of the diameter of the media. For the same media type and mill speed, the energy is increased by a factor of eight if the media size is increased from 3mm to 6mm. This highlights the importance of the inter-relationship between the top size of the mill feed, the selection of media size, and the wear on the mill internals.

Cyclones

Testwork had shown that closed circuit grinding with classification cyclones could significantly improve the mill throughput rate over open circuit. This was particularly apparent with the broad feed size distribution being presented to the mill as new feed. A key aspect of the KCGM flowsheet was the ability to provide a suitably sized product to the leach process for gold extraction whilst maintaining a suitable mill feed density (55%) with good fines rejection to overflow.

Closed circuit was also beneficial to the downstream gold leaching process in that the pyrite mineral carrying the gold have high SG's which preferentially reports to the cyclone underflow over an equivalent sized particle of a lower SG. This natural preference for high SG particles to report to the underflow, results in a further passage through the mill until such time as the reduced size counters the differential SG and allows passage out of the cyclone overflow and then to leach at a finer grind size.

KCGM selected 68mm cyclones for the classification duty on mill discharge. These cyclones have been essential in achieving a 10µm P80 product to leach but have presented a number of operational problems in spigots blockages and high wear rates.

With 10mm spigots, blockages were a constant problem with trash, scale, and sand media all contributing to blockages. Such blockages allowed a direct bypassing of mill discharge to leach with the resultant drop off in recovery. Significantly fewer spigot blockages have occurred since a vibrating screen was installed between the mill discharge and the cyclone feed hopper to remove this material. The main source of blockage nowadays is bridging of the spigot if the cyclone underflow becomes too high resulting in roping.

Thickening

The low pulp density of the cyclone overflow (8%) necessitated a thickening step to achieve the targeted 50% leach density. Testwork showed that the 10µm product would flocculate and settle rapidly at around 0.4 t/m^2/hr using the standard non-ionic polyacrylamide flocculant in use at a dosage rate of 140g/t. Plant experience confirmed these figures with thickener underflow densities of up to 55% being readily achieved.

Leaching

The leaching of sulphidic ores is often problematic with a number of side reactions likely to occur that can lead to high reagent consumption and poor gold recovery (Deschenes, 2001). The KCGM concentrate to be leached not only contained a high percentage of pyrite milled to a very fine particle size but also cyanide soluble copper (0.15%) as chalcopyrite. The presence of telluride gold (calaverite) also added to the leach complexity requiring specialised leach conditions.

In the laboratory testwork and early plant practice, a very high oxygen demand and slow leach kinetics were noted. Typical gold recovery of around 90% was achieved at a cyanide consumption rate of 15 kg/t. A leach development program was conducted in 2002 where intensive laboratory and pilot leach tests were carried out to improve gold recovery and lower reagent consumptions. The testwork resulted in a revised leaching regime where pre-oxidation was abandoned, lead nitrate was introduced at the start of the leach, and low dissolved oxygen levels (3ppm) maintained in the leach. It was also found that a high lime environment (30 kg/t) and use of carbon in leach (CIL) assisted the leaching of the gold tellurides.

By the use of these new leach conditions, the leach residence time was reduced from 72 hours to 24 hours, the cyanide consumption reduced from 15 kg/t to 8 kg/t, and average gold recovery increased by 2%.These improvements are shown in Figures 7, 8, and 9.

Figure 7. Improved Gold Recovery

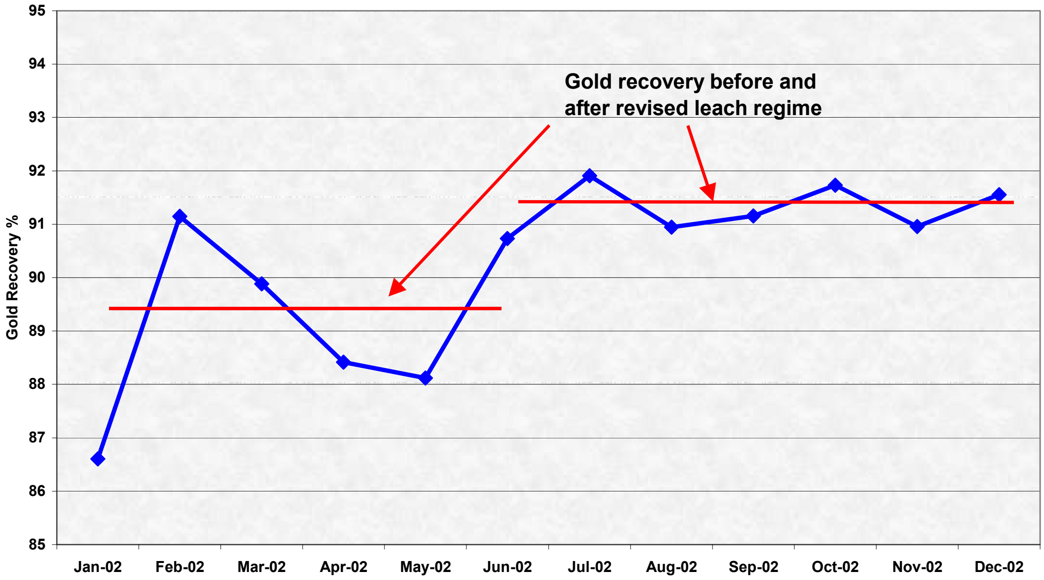


Figure 8. Reduced Cyanide Consumption

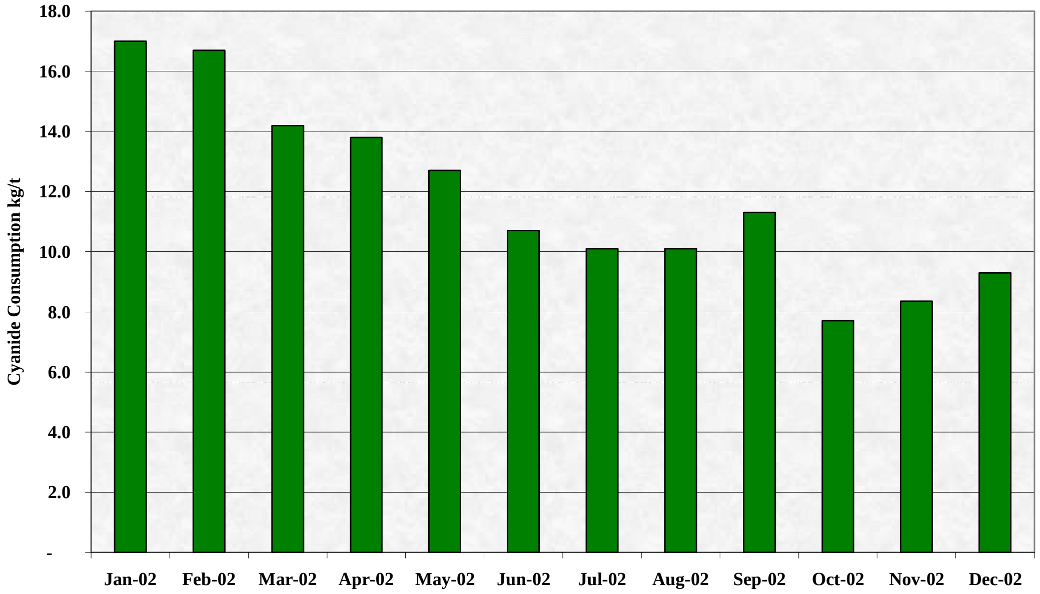
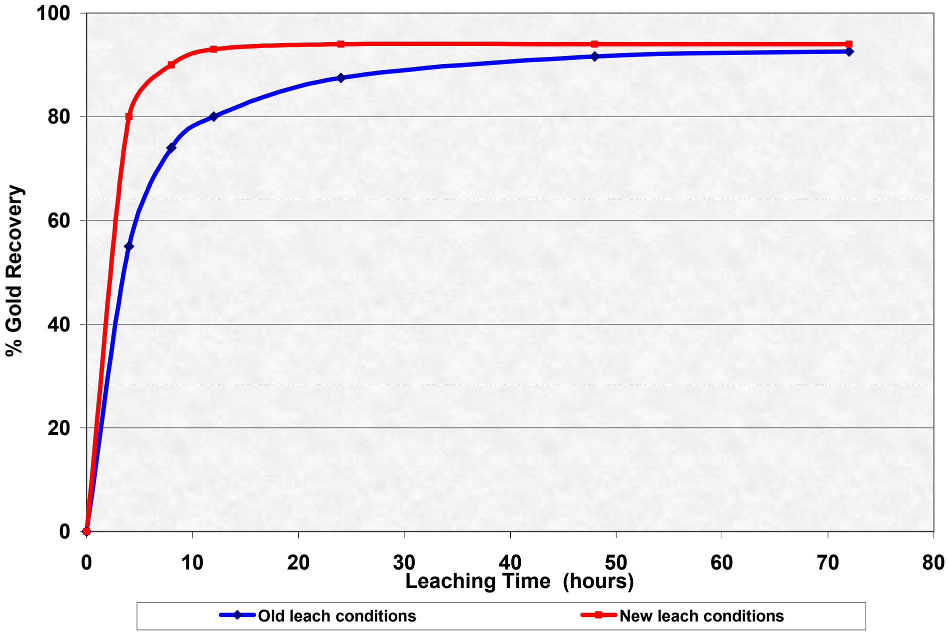


Figure 9. Improved Leach Kinetics



Despite the presence of an amount of sub 10µm material in the leach, no issue with the loss of carbon activity due to fine particles blocking pores in activated carbon has been apparent.

Costs

The capital cost of each of the KCGM UFG milling installation was $4.5 million. With ancillary equipment for the leach process the total project costs (including EPCM, owners management costs, and contingency) were around $6 million for each circuit.

Operational costs with a breakdown to the respective activity are shown in Figures 10 and 11.

Figure 10. Total UFG Processing Cost Breakdown ($62.02/t)

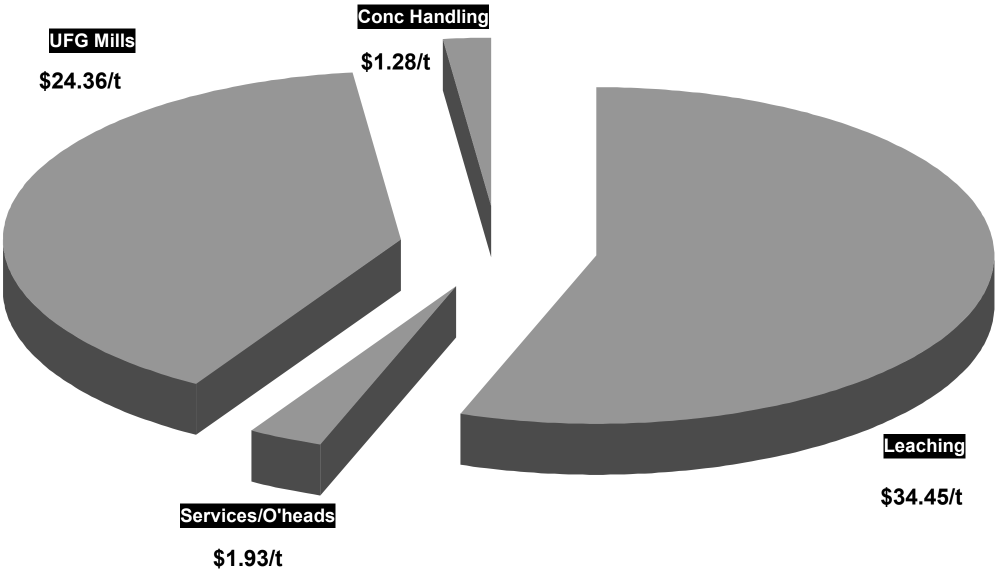
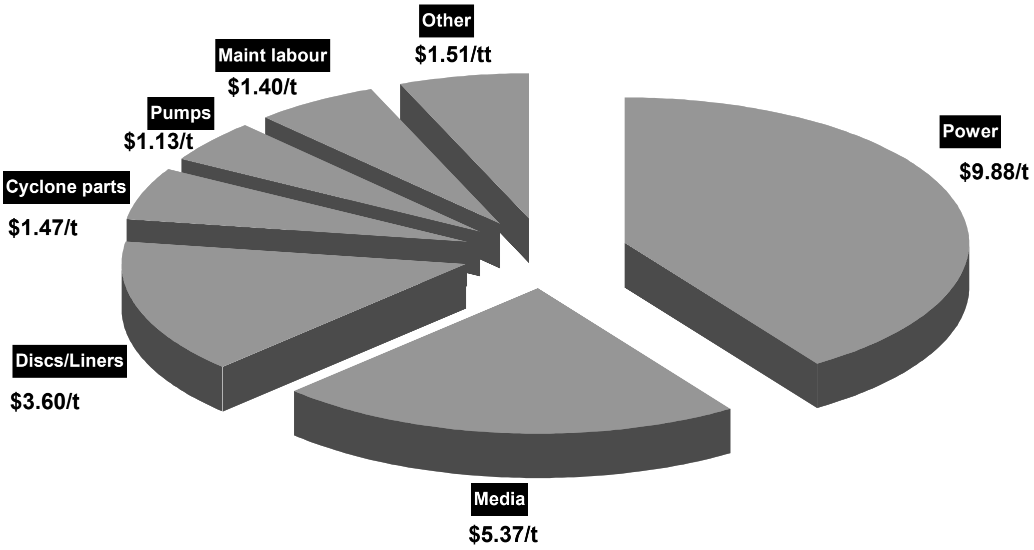


Figure 11. UFG Milling Cost Breakdown ($24.36/t)



The Future

With UFG being in operation for over two years at KCGM and having successfully proven the concept of economic gold extraction from refractory sulphide ore, albeit with lower gold recovery than roasting, the challenge is to further reduce process operating costs. The developmental work carried out has highlighted a number of areas where further improvements may be possible. Further work is in progress to quantify the benefits of introducing a primary milling stage. It is believed that a significant reduction in mill and cyclone wear components would result if a finer mill feed was produced as this would allow the use of a smaller media size thereby bringing wear rates in line with other similar mill users. Some improvement in power efficiency may also result from such an approach as well as some further improvement in gold recovery resulting from improved cyclone classification of a finer mill product.

The cost efficiency of ceramic media is likely to improve considerably in the future due to recent improvements in the quality and a lowering of the unit price with an increasing market size. As the number of UFG mills increases, the availability of sand of sufficient size and quality may be limited.

The application of ultra fine grinding to refractory gold ores has proven to offer a viable alternative to conventional oxidative processes in particular cases. Its application to a hydrometallurgical extraction route for other metals beyond gold is open and likely to lead to alternative process routes for these metals also.

REFERENCES

Alymore, M. G., Lincon, F.J., 1996, "Mechanical Milling of Auriferous Arsenopyrite and Pyrite," Journal of Alloys and Compounds, 242 pp. 22-34.

Becker, M., Kwade, A., and Schwedes, J., 1997, "Ultra fine Grinding of Raw Ceramic Materials and Wear of Grinding Beads in Stirred Media Mills", Aufbereitungs Technik 38 Nr8, pp 430-438.

Deschenes, G., Lastra, R. and Fulton, M., 2001, "Effect of the Mineralogy of Sulphide - bearing Gold Ores on the Performance of Cyanidation and its Control Variables", Proceedings of the Annual Meeting of the Canadian Mineral Processors. CIM, Ottowa, pp325-338,2002.

Ellis, S. and Gao, M., "The Development of Ultra Fine Grinding at KCGM", SME Annual Conference 2002, Preprint 02-072.

Enderle, U., Woodall, P., Duffy, M., Johnson, N. W., 1997, "Stirred Mill Technology for Regrinding McArthur River and Mount Isa zinc/lead ores," Proceedings of XX IMPC - Aachen, 21-26 September, pp. 71-78.

Hourne, M., Halbe, D. 1999, "The Nanatech Process: Results on Frieda River Copper Gold Concentrates," Proceedings of the Randol Gold Conference 1999.

Kalra, R., "Overview on Alternative Methods for fine and Ultra-Fine Grinding", IIR Conference, Crushing and Grinding, 1999.

Kwade, A.,"Wet Comminution in Stirred Media Mills - Research and its practical Application", Powder Technology 105, 1999, pp14-20.

Lichter, J. and Davey, G., 2002, "Selection and Sizing of Ultrafine and Stirred Grinding Mills", Mineral Processing Plant Design, Practice and Control.

Liddell, K. S., 1986, "Machines for Fine Milling to Improve the Recovery of Gold from Calcines and Pyrite," Proceedings of the International Conference on Gold. Vol. 2 pp. 405-417.

Turton- White, N.,2003, "Malvern Mastersizer Testwork: Investigating the Current Laser Sizing procedure for sizing Fine and Coarse Samples at KCGM, KCGM Report.

Warris, C. J., McCormick, P. G., 1997, "Mechanochemical Processing of Refractory Pyrite," Minerals Engineering, Vol 10, No 10, pp. 1119-1125.

Welham, N. J., 1997, "The Effect of Extended Milling on Minerals", CIM Bulletin, Vol. 90, No. 1007, pp. 64-68.

Weller, K. R. and Gao, M., "Ultra-fine grinding", AJM Crushing and Grinding Conference, Kalgoorlie, 27-28 th April 1999.

Yue, J. and Klein, B., "Particle Breakage in Horizontal Stirred Mills", CIM, 2003.

ISAMILL FINE GRINDING TECHNOLOGY AND ITS INDUSTRIAL APPLICATIONS AT MOUNT ISA MINES

Mingwei Gao[1], Senior Metallurgist
Michael Young[2], Metallurgical Superintendent
Peter Allum[3], Concentrators Manager

[1] MIM Process Technologies
Level 2, 87 Wickham Terrace,
Brisbane, Qld 4000,
Australia
PH: +61 7 3833 8500, Ext: 8505
E-mail: mingwei.gao@mim.com.au

[2] Mount Isa Mines Limited,
Mount Isa, Qld 4825,
Australia
PH: +61 7 4744 2011, Ext: 2926
E-mail: mfyoun@isa.mim.com.au

[3] Mount Isa Mines Limited,
Mount Isa, Qld 4825,
Australia
PH: +61 7 4744 2011, Ext: 2640
E-mail: pjallu@isa.mim.com.au

Key Words: IsaMill, ultra-fine and fine grinding, grinding media, product size distribution, regrind circuit, Jameson Cell, tower mill, maintenance, wear components, gland seal, power intensity

ABSTRACT

Tough economical conditions and high grade but fine-grained ore processing have forced the mining industry to look for more efficient processes. IsaMill Technology for ultra-fine and fine grinding is one such process employed at several mining operations in Australia including Mount Isa Mines Limited in Australia. Mount Isa Mines is a business unit of MIM Holdings Limited. MIM is an Australian-based international mining and mineral processing company with around 8000 employees worldwide. Major products include copper, gold, zinc-lead-silver, coal and technology sales.

The IsaMill is a horizontal stirred mill with sizes up to 3 m^3 chamber net volume and 1120 kW installed motor. It was jointly invented by Mount Isa Mines of Australia and Netzsch Feinmahltecknik GmbH of Germany for ultra-fine and fine grinding duties in the minerals industry. IsaMill operations started in 1994 at the Mount Isa Mines Lead/Zinc Concentrator and then at MIM's McArthur River Mine in 1995. These IsaMills have been used to liberate sphalerite and galena at Mount Isa and silica at McArthur River by grinding to less than 7 micron.

This paper discusses the IsaMill Technology and reports on the metallurgical performance of the IsaMills at Mount Isa Mines, including work done over the last 9 years to improve mill design, process efficiency and maintenance, which has resulted in the current state-of-the-art IsaMill Technology.

INTRODUCTION

The current lead/zinc deposits at Mount Isa Mines (George Fisher, Hilton and Isa) have an average grade of 10.5% of zinc, 7.2% of lead and 165g/t of silver. Fine grinding, in particular in the zinc retreatment circuit requiring a milled product P_{80} of 7 micron, is essential to treat the ores from the Mount Isa Mines deposits. Every one micron size reduction in the zinc retreatment circuit below a P_{80} of 10 micron improves the overall plant zinc recovery by 1%, resulting in an increase in revenue of about 1 million dollars per annum.

Eight IsaMills operate in the Lead/Zinc Concentrator at Mount Isa Mines, with two installed in 1994 and six commenced operation at the end of 1999. (Enderle et al, 1997; Johnson et al, 1998; Young et al, 2000). These are required for ultra-fine grinding to achieve liberation of fine-grained minerals, which improves separation of the minerals in the flotation circuits.

The process parameters and IsaMill designs have been continuously improved to increase the IsaMill grinding performance and the component life between maintenance periods. This paper reports the latest progresses in developing the IsaMill process and discusses the new features of the mills that have enhanced their performance.

THE ISAMILL TECHNOLOGY

IsaMill is a major development in grinding technology jointly invented by Mount Isa Mines of Australia and Netzsch Feinmahltecknik GmbH of Germany for ultra-fine and fine grinding duties in the minerals industry. The IsaMill was introduced by MIM over a period of 9 years, as there was no existing ultra-fine grinding technology that could be used or modified for use in the minerals industry. The IsaMill technology demonstrated its ability for very power efficient grinding to less than 10 micron. The major hurdles that required extensive investment in time and money were the high wear rates, characteristic of this type of technology, and the separation of media from the product. Other technologies use exotic materials in construction and screens for these purposes but they are not consistent with a low cost, large scale and reliable grinding technology. The result of MIM's development is a combination of smart component design and careful selection of rubber and polyurethane compounds to give long wear lives with relatively cheap materials.

IsaMills are currently operating in three mining operations in Australia. Namely, Mount Isa Mines, McArthur River Mine and Kalgoorlie Consolidated Gold Mines. A total of 15 IsaMills are in use at these operations. Figure 1 shows the eight IsaMills at the Mount Isa Mines Lead/Zinc Concentrator.

Figure 1: IsaMills in the Regrind Circuits at the Mount Isa Mines Lead/Zinc Concentrator

The IsaMill is a horizontal high speed stirred mill that operates with very high power intensities (up to 350 kW/m^3). In comparison, the power intensity of a ball mill is about 20 kW/m^3. The high power intensity enables the IsaMill to process fine particles at a high throughput that is essential for the economics of the minerals industry. The largest IsaMill currently available has a horizontally mounted grinding chamber shell of about 4 m^3 in total volume (or approximately 3 m^3 working volume). Inside the shell are rotating grinding discs mounted on a shaft which is coupled to a motor and gearbox. To allow quick and simple removal of the grinding chamber to expose the mill internals for maintenance purposes, the shaft is counter-levelled at the feed inlet end. The grinding discs agitate the media and ore particles in a slurry that is continuously fed into the feed port. A patented product separator keeps the media inside the mill allowing only the product to exit and simple control strategies based on power draw enable the IsaMill to produce a constant target product size. The invention of the product separator eliminates screens from ultra-fine grinding which delivers a process with the robustness required by the mining industry. The product separator uses the high centrifugal forces generated to retain the media inside the grinding chamber.

The IsaMill uses fine media, stirring it with very high disc rotation speed, which increases the probability and energy of media/particle collisions. The mill has been designed to break particles using the attrition/abrasion mode of breakage where particle surfaces are chipped repeatedly producing very fine sized progenies at relatively low power consumption.

The compact IsaMill operates under pressure with "stage-by-stage" grinding occurring between the high speed rotating discs, which ensures grinding events are evenly distributed throughout the grinding chamber. This homogeneous grinding mechanism results in significant improvements to the particle size distributions from the feed to the product. The IsaMill Technology grinds the particles requiring size reduction, without over-grinding material at or below the required P_{80}. The uniform grinding mechanism is also the reason why IsaMill scale-up is 100% direct from laboratory to pilot to full scale (Weller et al 1999).

Currently, all operating IsaMills are installed with 1120 kW (1500 hp) motors and a grinding chamber with a net volume of 3 m^3. A range of different size IsaMills below the 1120 kW motor model are also available. The schematic of the mill is shown in Figure 2.

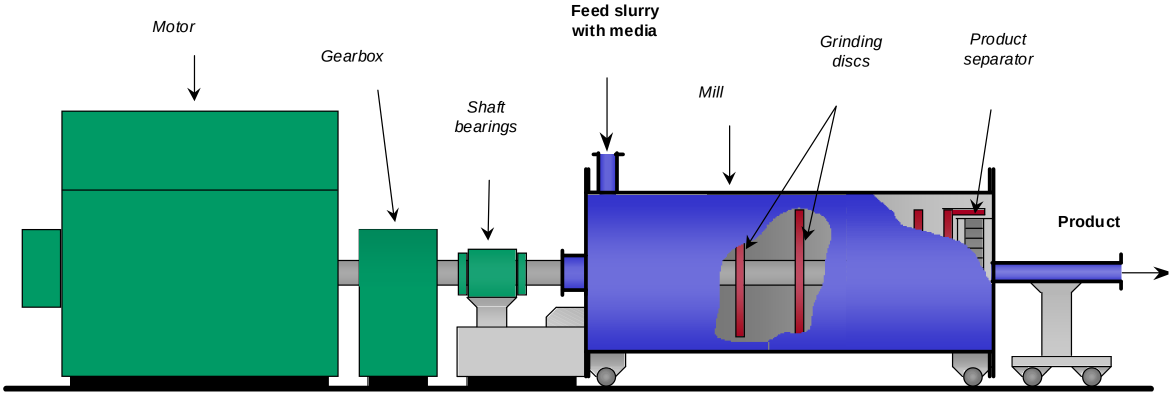


Figure 2: IsaMill Schematic indicating Major Components

MOUNT ISA LEAD/ZINC CONCENTRATOR FLOWSHEET

The process flowsheet of the Mount Isa Mines Lead/Zinc Concentrator is shown in Figure 3. The circuit has been through various stages of modifications to address the declining ore qualities in the last ten years (Young et al 1997). It contains three IsaMills in the lead circuit and five IsaMills in the zinc regrind and retreatment circuits.

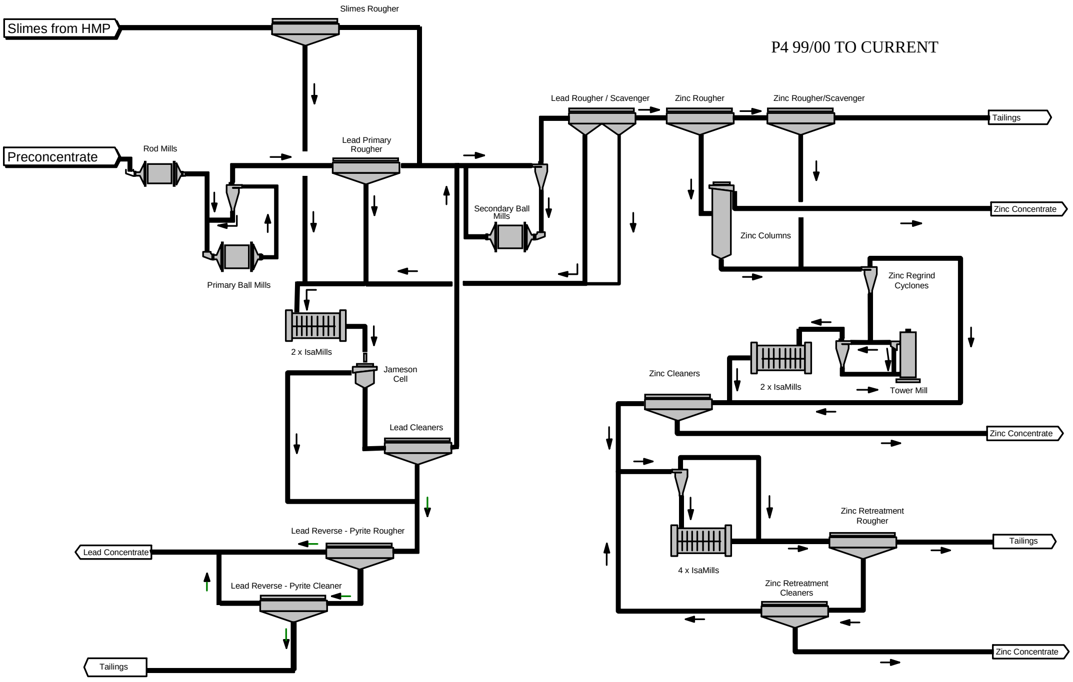


Figure 3: Mount Isa Mines Lead/Zinc Concentrator Flowsheet

The designed specifications of the IsaMills in the lead and zinc circuits are listed in Table 1. The product P_{80} sizing of 12 micron for the lead and zinc regrind IsaMills was required to liberate the galena and sphalerite from the gangue materials in the lead rougher concentrate, zinc column cleaner tailings and zinc rougher/scavenger concentrate. The zinc cleaner tailings were further reduced to a product P_{80} of 7.5 micron in the zinc retreatment circuit.

Table 1: Design Specifications of the IsaMills

	Lead Regrind	Zinc Regrind	Zinc Retreat
Feed F_{80}, micron	25	25	20
Product P_{80} , micron	12	12	7.5
Net Energy, kWh/t	10	25	45

The ores from lead/zinc deposits in Mount Isa Mines contain minerals with a range of textural associations (Bojcevski et al, 1998). These textural associations allow a portion of the minerals to be liberated at coarse grind sizings (P_{80} = 37 micron), a portion to be liberated at fine grind sizings (P_{80} = 12 micron), while the remainder require ultra-fine grind sizings (P_{80} = 7.5 micron) for liberation.

This range of textural associations allows the use of a flowsheet using staged grinding and staged flotation to remove liberated valuable minerals with a coarse grind, while composite particles are reground for further liberation. This reduces the overall grinding energy consumption of the flowsheet and decreases the flotation capacity required, as coarse particles have faster flotation kinetics than fine particles.

THE ISAMILL PERFORMANCE

Lead Regrind Circuit

The original lead regrind circuit was part of the bulk lead/zinc retreatment circuit that consisted of a tower mill operating in closed circuit configuration with 6 inch Kreb cyclones as shown in Figure 4. The cyclone overflow was processed by Agitair flotation cells. This circuit had problems of over-grinding the galena-bearing minerals as the cyclones returned the smaller but heavier galena particles and coarser but lighter gangue particles to the underflow for further grinding. The effect of this phenomenon is illustrated in Figure 5.

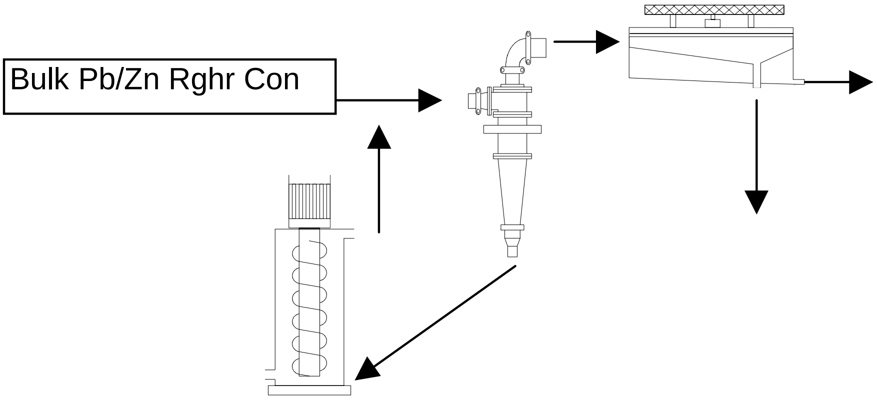


Figure 4: Closed-Circuit Tower Mill for Treating Bulk Lead/Zinc Concentrate

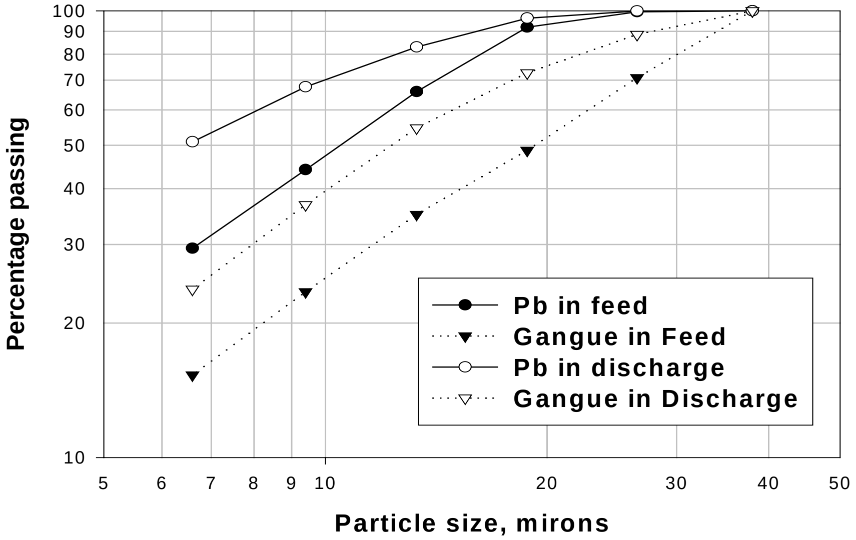


Figure 5: Component Size Distributions of Mount Isa Mines' Tower Mill Feed and Discharge

With the introduction of the IsaMills in 1994, better liberation was achieved. This eliminated the bulk lead/zinc concentrate regrind circuit (Young et al, 1997). In 1999, three IsaMills configured in open circuit were installed in the lead regrind circuit, followed by a Jameson Flotation Cell. This new circuit is shown in Figure 6.

As part of the new circuit the recycle stream in the milling stage was eliminated to ease the problem of over-grinding the galena minerals. The new circuit also incorporated a Jameson Flotation Cell that generates finer bubbles to process the IsaMill product that has a P80 sizing finer than 12 micron. The immediate benefit from this simple circuit modification was to increase the overall lead recovery by 5%.

Another feature of the new circuit is the use of inert lead granulation slag as grinding media in the IsaMill. This improves the pulp chemistry and hence flotation selectivity, compared with using 12 mm steel balls in the tower mill.

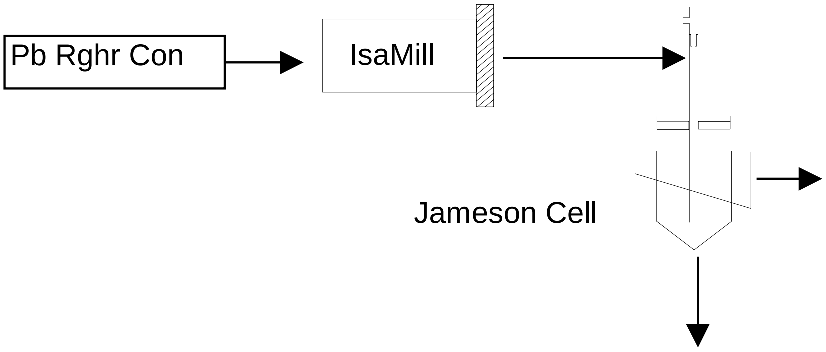


Figure 6: Lead Regrind Circuit

Zinc Regrind Circuit

Figure 7 shows the flowsheet of the zinc regrind circuit. It includes a pre-cycloning stage, a closed-circuit tower mill and two open-circuit IsaMills that operate in parallel. The feed to the pre-cycloning stage includes the zinc column cleaner tailings and the zinc rougher and scavenger concentrate. As well as removing particles already at or below the required P_{80}, the pre-cycloning stage also increases the feed density to the regrind circuit.

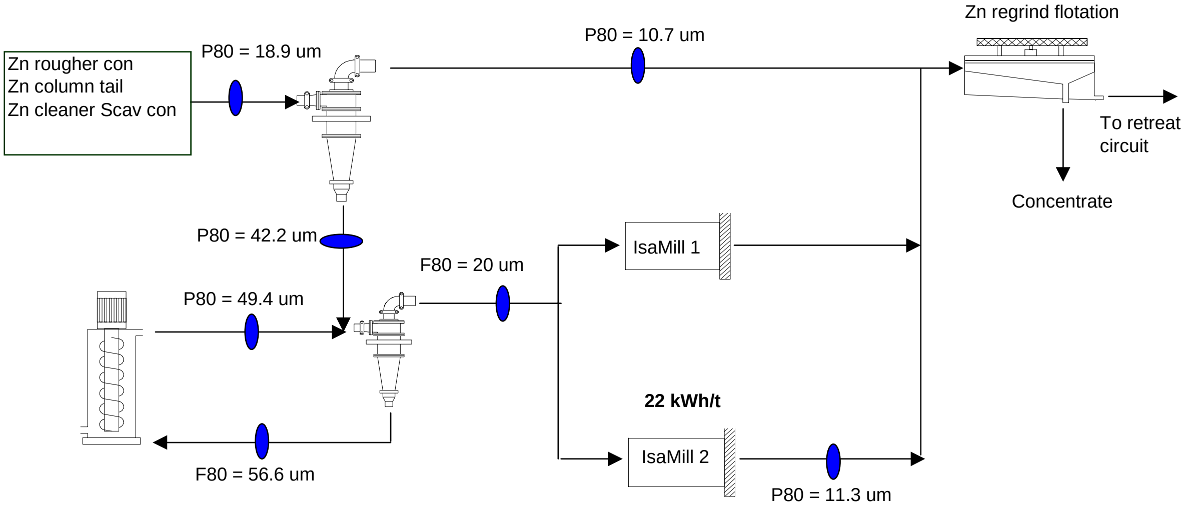


Figure 7: Zinc Regrind Circuit

Figure 7 shows that the IsaMills in the zinc regrind circuit reduced the feed F_{80} of 20 micron to a product P_{80} of 11.3 micron with a net energy consumption of 22 kWh/t. In comparison, the size reduction across the tower mill was only from a feed F_{80} of 56.7 micron to a product P_{80} of 49.4 micron.

Zinc Retreatment Circuit

Figure 8 shows the zinc retreatment circuit. There are three IsaMills in the circuit. The feed to the pre-cyclones in the retreatment circuit includes the zinc cleaner tailings and the zinc retreatment cleaner tailings. These streams are pre-cycloned to remove material already at or below the required P_{80} and to increase the feed density to the IsaMills. The IsaMill product combines with the cyclone overflow and feeds to the zinc retreatment rougher flotation cells.

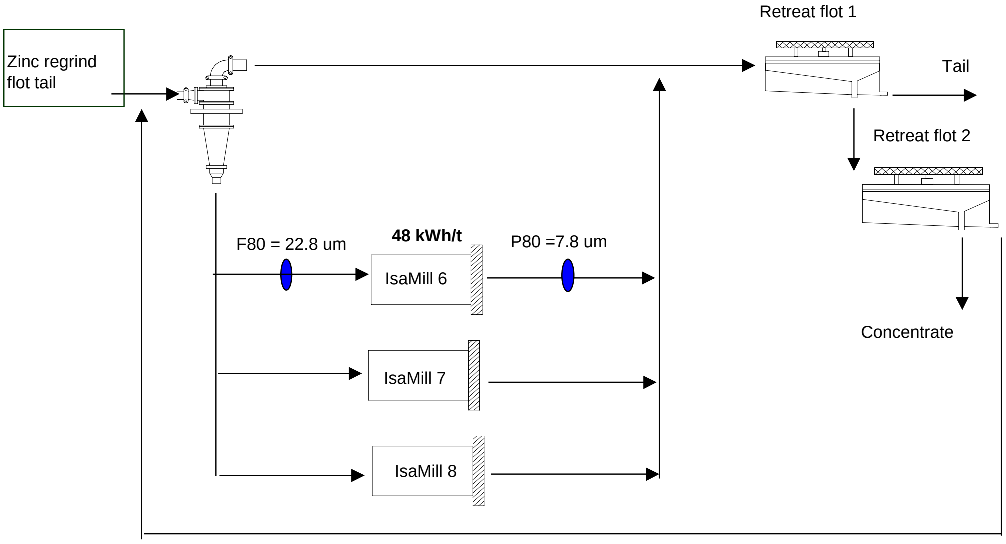


Figure 8: Zinc Retreatment Circuit

The feed to the zinc retreatment IsaMills is extremely hard to grind as it contains a large amount of composite particles with silica content as high as 30%. This is because the liberated valuable minerals have been removed in previous stages of flotation. Figure 8 shows the performance of one of the zinc retreatment IsaMills. It reduced a feed F_{80} of 22.8 micron to a product P_{80} of 7.8 micron with a net energy consumption of 48 kWh/t.

ISAMILL VERSUS TOWER MILL

The existing tower mill was in operation since 1991, firstly in the lead/zinc bulk concentrate regrind circuit, then in the original zinc regrind circuit. It was re-located to the current zinc regrind circuit in 1999 as the first stage of zinc regrinding. The purpose the tower mill was to produce a finer IsaMill feed, so the available IsaMills could be operated in open circuit with a fine lead slag media of 1 to 2 mm. This fine media sizing achieves the best power efficiency for products below 10 micron.

Using the survey data collected from the zinc regrind circuit, a comparison of the IsaMill versus the tower mill was possible. Figure 9 shows the size distributions of the feed and products for the tower mill. The difference in the 80% passing sizings of the feed (F_{80}) and the products (P_{80}) was in the order of 5 micron. Many surveys of the same tower mill in previous applications have also indicated that when being fed with a feed F_{80} finer than 50 micron, the tower mill was capable of a size reduction of only 5 to 10 micron..

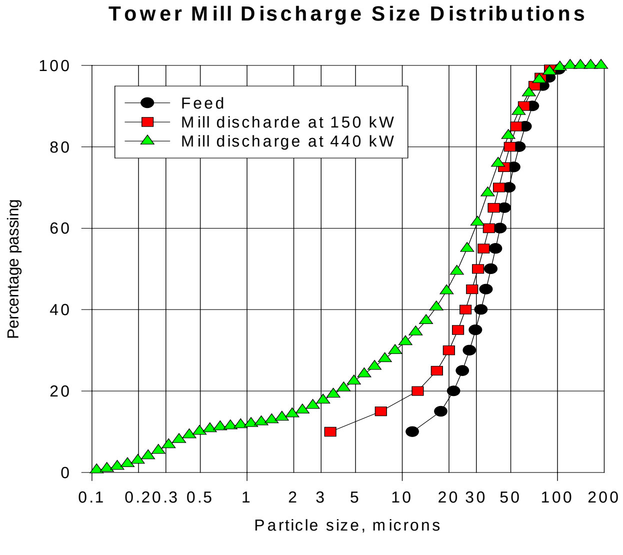


Figure 9: Feed and Product Size Distributions for Zinc Regrind Tower Mill

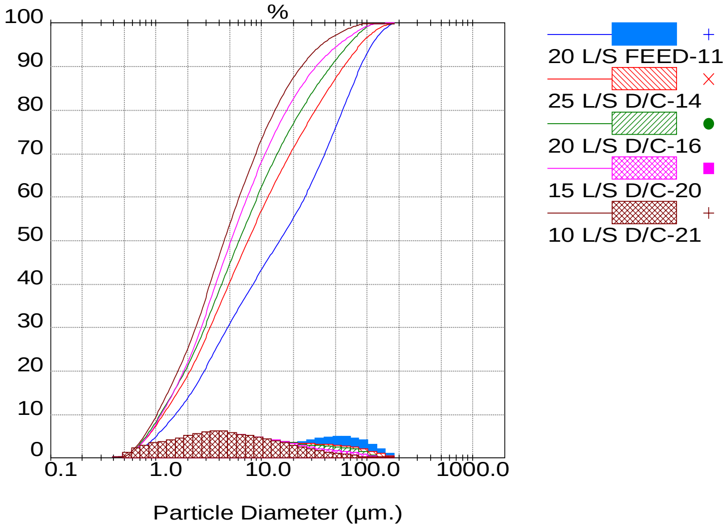


Figure 10: Feed and Product Size Distributions for Zinc Regrind IsaMill

In comparison, as shown in Figure 10, the size reduction across the zinc regrind IsaMill was from a feed F_{80} of 55 micron to a product P_{80} as fine as 12 micron, depending on the feed flow rate.

The excellent size reduction performance with the IsaMills is a result of using small media in the order of 1 to 2 mm. The advantage of operating with smaller media size is that it provides an increased number of media particles in the mill for grinding. For example, 1172 spheres of 1 mm media occupy the volume of only one 12.5 mm ball. Therefore, an IsaMill using 1 mm media will have 1172 times more media particles per mill volume available for grinding compared to a tower mill using 12.5 mm balls.

The IsaMill is also more efficient in using small media by stirring them with very high disc circumferential speed (21m/s), increasing the probability and energy of media-particle collisions. The IsaMill consists of a horizontal stationery mill chamber operating under a pressure of 100 to 200 kPa. This grinding mechanism ensures the grinding events are evenly distributed throughout the mill chamber. In comparison the tower mill is pressurised only by gravity, which results in the grinding events occurring mainly near the base of the tower mill's vertical grinding chamber.

Another major difference in the performance of the IsaMill and the tower mill is the profile of the product size distributions as indicated in Figures 9 and 10. The tower mill product has a long tail at the finer end of the particle size distribution due to over-grinding of the fines. The IsaMill on the other hand grinds mainly the coarser part of the feed size distribution and produces a minimum amount of new fines. This ability of selectively grinding coarser particles in the feed is ideal to liberate the valuable minerals from the inter-locked particles and maximise their recovery in the down stream beneficiation processes. The tighter size distributions produced by the IsaMill allow the flotation circuit to operate more efficiently, since the flotation feed is more uniform. Optimising the flotation circuit is also made easier with a feed having a consistent particle size distribution. The IsaMill Technology efficiently grinds the particles requiring size reduction, without over-grinding material at or below the required P80.

EFFECT OF SLURRY FEED SIZING

Over the years of operating the IsaMills at Mount Isa Mines, it has been found that the slurry feed size distribution is important for the IsaMill performance when using local smelter slag as the grinding media. A finer feed size would normally help to reduce the energy consumption for a finer product when operated in open circuit. When the feed size becomes coarser the media sizing should be increased to provide the momentum and media-particle size ratio needed for breaking larger particles.

To demonstrate the effect of feed size, the IsaMill performance with feed F_{80}s of 35 and 20 micron are compared in Figure 11. Reducing the IsaMill feed F_{80} from 35 to 20 micron enabled the IsaMill to produce a finer product P_{80} of 11.3 micron compared to 14.4 micron for the coarser feed. Also the net energy consumption was reduced by about 30% from 33 to 22 kWh/t.

The less efficient grinding when processing the coarser feed was due to the size of the fine slag media. Figure 12 shows the size distribution of the slag media collected during the survey. It contained about 50% fines below 0.5 mm, which is at the lower limit of media size for effective grinding in the IsaMill. A finer media normally helps to produce a finer product but has limited ability to cope with a coarser feed.

For an IsaMill operating in open circuit configuration with a fine grinding media, the feed size needs to be controlled operationally to achieve target grind sizes at maximum power efficiency.

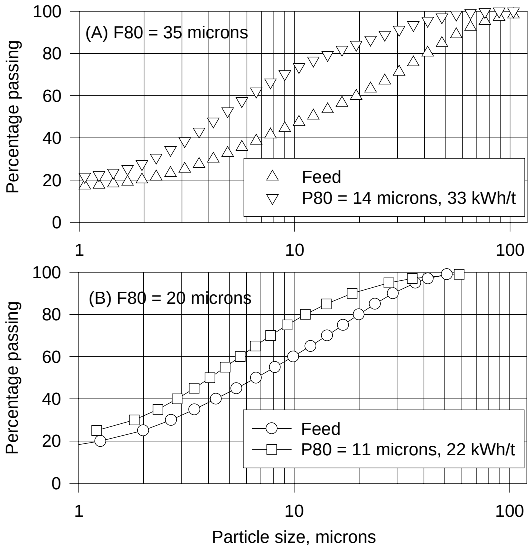


Figure 11: Effect of Feed Size on Zinc Regrind IsaMill

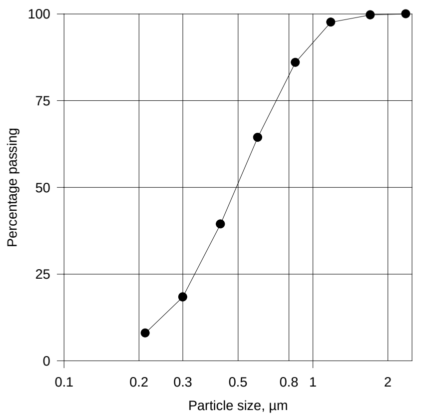


Figure 12: Size Distribution of Fine Lead Slag Media

EFFECT OF MEDIA SIZING

The tower mill circuit prior to the zinc regrind IsaMills is required to maintain the IsaMill performance when the feed size becomes coarser. Another option is to coarsen the slag media size used in the IsaMill to process the coarser feed. If an ultra-fine product is required with both coarse feed and coarse grinding media then the IsaMill should be operated in closed circuit to maintain power efficiency.

Trials with different lead slag media sizes were conducted in one of the zinc regrind IsaMills, with unscreened slag containing about 50% fines below 0.5 mm and screened slag containing about 8% fines below 0.5 mm. Figure 13 shows the size distributions of the screened and unscreened slag.

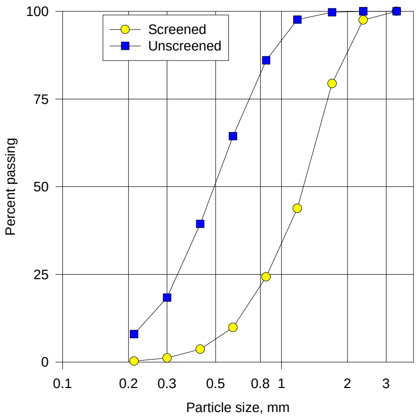


Figure 13: Comparison of Screened and Unscreened Media Size Distributions

Figure 14 shows the plant survey data using the same feed material, but with screened and unscreened slag as the media. The feed F_{80} was 30 micron in both cases. The figure shows that screened slag media achieved a P_{80} of 15 micron using less than 30 kWh/t, compared to unscreened slag which required more than 40 kWh/t for the same product size.

Selection of the grinding media size is undoubtedly one of the key criteria for achieving the maximum IsaMill power efficiency.

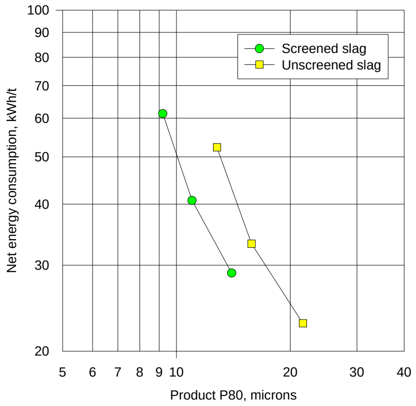


Figure 14: IsaMill Performance with Screened and Unscreened Lead Slag Media

NEW DESIGN FEATURES OF THE ISAMILL

The initial prototype IsaMill installed at Mount Isa Mines in 1993 had a grinding chamber volume of 1500 litres and operated with 5 grinding discs. The purpose of this mill installation was to aid in the design of the full-scale production IsaMills. The first production mills installed at MIM's Mount Isa and McArthur River mines operated initially with 6 discs. Subsequently the number of grinding discs was expanded to seven and then to the current standard of eight discs. The net mill volume of the production mills with eight discs was 3000 litres.

Continuous development and improvement of the IsaMills from 1994 to the present has lead to a design providing superior grinding performance, increased component life, and decreased downtime during routine maintenance periods.

With the developmental focus of the IsaMill being minimum operating costs and maximum availability, the equipment has been designed for fast and straightforward maintenance. Locating all major components at ground level with the shell components on rails results in very easy access to the wear items. The major mechanical items such as motor and bearings are standard components with typical maintenance programs and life expectancy.

Internal Wear Components

The materials of construction and design, and method of fabrication of all IsaMill internal wear components have improved since the first installation. Over 20 different types of wear materials have been tested. These tests determined the characteristics that components need for long wear life.

The internal steel surfaces on the original production IsaMill grinding chambers and end flanges were cold rubber lined. These wear components were required to be sent to the nearest rubber lining workshop after a certain amount of operating time for relining. This meant that a spare set of these steel components had to be held in stock.

To avoid this requirement components of the feed and discharge end flanges were redesigned, so that only a replaceable rubber wear component had to be held in stock. These rubber components simply slide on to the steel surfaces and then are bolted in place. This allowed improvements in the fabrication technique of these rubber components, which lead to an improvement in wear life.

The final challenge in terms of wear components was the IsaMill grinding chamber. The latest IsaMills were designed with a split shell (Figure 15), in which a replaceable slip-in liner could be installed. The IsaMill shell can be unbolted, opened into two halves and the worn shell slip-in liner removed and a new one installed. This removes the need to send the shell away to be cold rubber lined and the need to have spare shells to install when one is worn. All wear components are fabricated off-site then stored on-site until required. All components are steel backed with rubber or similar coating. These are relatively low cost, light and easy to handle and store, but still provide long wear lives.

Figure 15: IsaMill with a Split Shell Design

Shaft Sealing System

In addition to the internal wear components of the IsaMill, the shaft sealing system also requires routine maintenance. This system prevents slurry and grinding media discharging from the point where the rotating shaft enters the grinding chamber. The gland sealing system can be adjusted externally while operating, similar to a slurry pump, and can be maintained when the discs are removed from the shaft during scheduled maintenance.

The commercial full-scale IsaMills use gland packing and gland water sealing for sealing the shaft. The design of this gland sealing system was a great improvement over several of versions trialed during the scale-up phase of the IsaMill development program. The gland life of these trial systems was shorter than desired, especially since the life of the internal wear components (grinding discs and shell liners) had been extended.

Improvements were made to install an independent gland water supply system that increased the gland life. The new design for the gland system also includes an increased bearing spacing to decrease the deflection of the shaft at the gland, and introduces water flushing on the slurry side of the gland to reduce the amount of slurry and media in this area. The new gland system has been working well since installed about 2 years ago.

Mill Access

Figure 16 shows the support system for the IsaMill. The mill grinding chamber shell and discharge end flange are both fixed to trolleys which are mounted on four wheels that sit on rails. The IsaMill has hydraulic rams on the rails under the wheels. These rams are used to move the grinding chamber and discharge end flange along the support rails so that the internal components of the mill are exposed for maintenance. The original installations had fixed hydraulic power packs on each IsaMill, which tended to get covered in dust and other debris, thus requiring more than expected maintenance. The new IsaMill design has a portable hydraulic power pack that plugs into the rams on the mill when maintenance is required and is then stored in a clean area after maintenance has been completed.

This quick and simple system to access the mills is achievable since the internal shaft is counter-levelled at the feed inlet end. In other words there are only bearings at the feed inlet end of the mill and at the discharge end the shaft is free and does not protrude through the end flange.

Figure 16: IsaMill Support System

MAINTENANCE PROGRAM

Maintenance of the IsaMill is as simple as routine maintenance for a slurry pump. As previously mentioned the internal rotating shaft is counter-levered at the feed inlet end. This means that the discharge end flange and grinding chamber can be removed to expose the mill internals without having to remove bearings, shaft etc. The discharge end flange is firstly unbolted from the grinding chamber and the trolley fixed to the flange is then connected to the hydraulic ram and driven out along the rails. Next the chamber shell is unbolted from the feed inlet flange and is also moved along the rails using the hydraulic ram to fully expose the mill internals, including shaft, grinding discs and product separator.

To remove the grinding discs the locking nut at the end of the shaft is undone so that discs can be slid off the shaft. New discs are then slid on the shaft. The bolt-on wear components on the feed and discharge end flanges and the shell liner can then be replaced, if required. After replacing any worn components the chamber shell is moved back into position using the hydraulic ram and rebolted to the feed inlet flange. The discharge end flange is then wheeled up and bolted against the chamber shell. The IsaMill is now ready for to be put back into service.

The maintenance schedule is built around the grinding discs. The average life of a set of discs at Mount Isa is about 4000 hours. The other components last several cycles of disc replacement. The lives of all the components are continually being improved to decrease downtime and maintenance costs.

The maintenance program at Mount Isa Mines is currently based upon a 1,500 hour operating cycle. At the end of each cycle the IsaMill is shutdown and opened for inspection and replacement of wear parts if required. The shutdown has a total duration of 8 hours.

The inspection has the following procedure:

- Shutdown of mill
- Flushing of mill followed by dumping of mill contents
- Removal of end flange bolts
- Removal of Shell
- Inspection of discs, shell liner, product separator and gland seal for wear
- Replacement of any worn discs
- Replacement of gland seal if required
- Re-assembly of shaft, discs and spacers
- Replacement of shell and end-flange
- Recommencement of operation

On each third operating cycle a 16-hour shutdown is undertaken for detailed inspection of bearings and tolerances etc.

At Mount Isa Mines the shell liner, product separator and end-plate linings have typical wear lives of 18 months. The shaft sleeves (disc spacers) are low wear items and their life significantly exceeds that of the other mill internals.

The gland seal packing is comparable to that used in slurry pumps and can be expected to have a similar life. Clean and consistent water supply will ensure maximum life. Recent design changes as mentioned above has increased gland life by at least a factor of two. Gland packing life is estimated to be at least 6 months.

At Mount Isa Mines the overall IsaMill availability on an annual basis is 99%.

BENEFITS TO PLANT METALLURGY

The installation of eight IsaMills since 1994 in the Mount Isa Mines Lead/Zinc Concentrator, along with other circuit modifications, has increased the liberation of galena and sphalerite and increased the metal recoveries at the target grades.

Figure 17 shows the concentrator zinc recovery performance for the last twenty years with recent operation reaching the highest zinc recoveries. Significant liberation of valuable minerals as a result of using the IsaMill technology is one of the most important technical breakthroughs that has contributed to the improved plant metal recoveries.

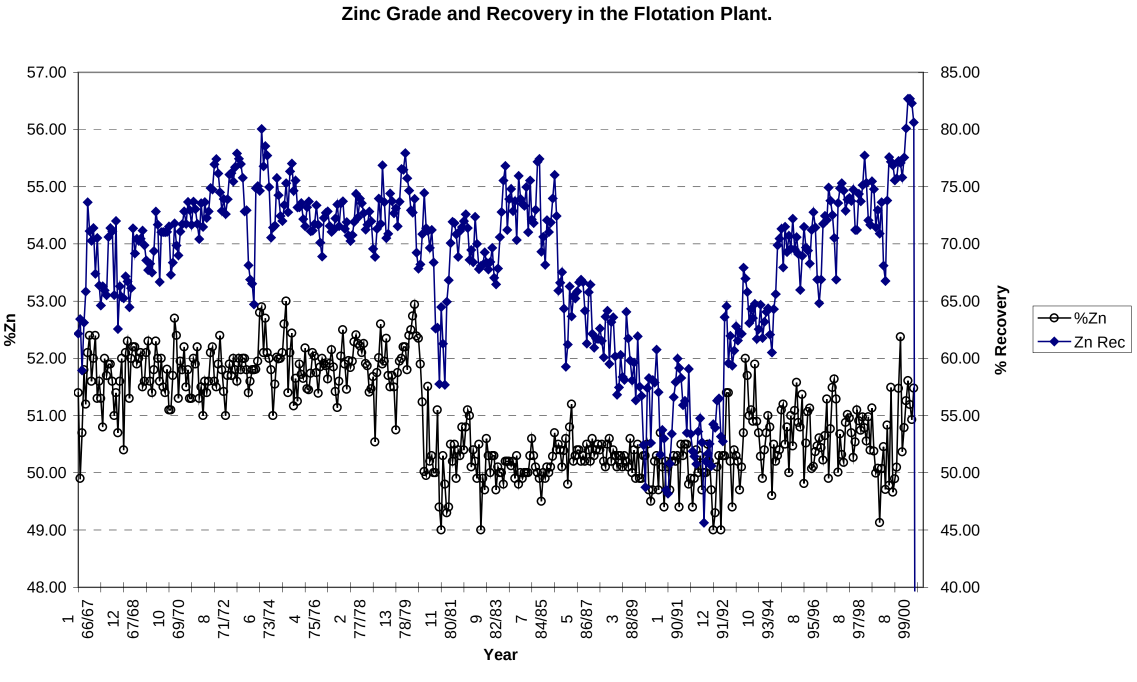


Figure 17: Lead Recovery at Mount Isa Mines' Lead/Zinc Concentrator since 1980

SUMMARY

IsaMill technology was successfully developed and operated at Mount Isa Mines over the last nine years. Currently 15 full-scale IsaMills operate in three mining locations in Australia, economically producing products as fine as 7 micron.

Compared to other fine grinding technologies, the IsaMill has high power input per mill volume (up to 350 kW/m^3) and operates under pressure (100 to 200 kPa). These two features enable the small grinding media of a few millimetres to work effectively to process a wide range of feed sizes and to produce the required fine products.

Media selection is important for the IsaMill operation. Mill scale-up data can be obtained from laboratory data but design of IsaMill circuits needs to be conducted carefully. Feed size and media size effects as well as variations in the circuit need to be appreciated.

The new IsaMill design has increased the grinding performance and improved the maintenance cycle giving greater component life and ease of maintenance.

ACKNOWLEDGMENT

The authors wish to acknowledge the management of Mount Isa Mines Limited and MIM Process Technologies for permission to publish this paper. For this work, the important input of many employees at Mount Isa Mines Limited and MIM Process Technologies is acknowledged, as well as the contribution of their colleagues at Netzsch Feinmahltecknik GmbH.

REFERENCES

Bojcevski, D., Vink, L., Johnson, N.W., Landmark, V., Mackenzie, J., and Young, M.F., “Metallurgical Characterisation of George Fisher Ore Textures and Implications for Ore Processing”; AusIMM Mine to Mill Conference, Brisbane, Australia, pp.29-41, 11-14 October 1998.

Enderle, U., Woodall, P., Duffy, M., and Johnson, N.W., “Stirred Mill Technology for Regrinding McArthur River and Mount Isa Zinc/Lead Ores”; XX International Mineral Processing Congress, Aachen, Germany, Vol. 2, pp. 71-78, September 21-26 1997.

Johnson, N.W., Gao, M., Young, M.F., and Cronin, B., “Application of the IsaMill (A Horizontal Stirred Mill) to the Lead/Zinc Concentrator (Mount Isa Mines Limited) and the Mining Cycle”; AusIMM Annual Conference, Mount Isa, Australia, pp. 291- 297, 19-23 April 1998.

Young, M.F., Pease, J.D., Johnson, N.W., and Munro, P.D., “Developments in Milling Practice at the Lead/Zinc Concentrator of Mount Isa Mines Limited from 1990”; AusIMM Sixth Mill Operators Conference, Madang, Papua New Guinea, pp 3-12, 6-8 October 1997.

Young, M.F., Pease, J.D., and Fisher, K.S., “The George Fisher Project to Increase Recovery in the Mount Isa Lead/Zinc Concentrator”; AusIMM Seventh Mill Operators Conference, Kalgoorlie, Australia, 12-14 October 2000.

Weller, K.R., Gao, M., Bowen, P., “Scaling-Up Horizontal Stirred Mills From a 4-Litre Test Mill to a 4000-Litre IsaMill”; Powder Technology Symposium, Pennsylvania State University, PA, USA, September 1999.

IsaMill Ultrafine Grinding for a Sulphide Leach Process

by
Harbort, G(1)., Hourn, M(2)., Murphy, A(1)

(1) MIM Process Technologies
(2) Hydrometallurgical Research Laboratories

ABSTRACT

Fine grinding mills have improved in design and efficiency in recent years, allowing major opportunities for treatment of materials where liberation to grind sizes below fifteen microns are required. The successful development of the IsaMill, a horizontal stirred mill, has produced equipment capable of grinding the larger tonnages which exist in mineral processing operations, to product sizes below ten microns. Initially developed for use with base metal flotation circuits, significant test work conducted in 1998 shows that major economic gains can be achieved by producing finely ground material for leaching.

IsaMill development and operation is reviewed. Results from test work to produce a feed stock for both straight cyanidation and sulphide leaching are discussed, with emphasis on energy consumption, product size and leachability. A number of options for ultrafine grinding and leaching are also discussed.

INTRODUCTION

In operations where flotation products are produced as a saleable concentrate, decreasing liberation size may result in decreased metallurgical performance due to gangue impurities and lower recoveries. Ultra fine grinding is gaining significant acceptance as a cost effective means to provide an optimum grade/recovery response in flotation.

In some gold leaching operations, ores contain gold in close association with sulphide minerals such as pyrite and arsenopyrite. Ultrafine grinding provides an effective method of liberating physically locked gold, or for producing a feedstock that is amenable to oxidative processes.

ISAMILL DEVELOPMENT

Decreasing liberation size and increased amounts of refractory pyrite in Mt Isa's lead/zinc ore resulted in a gradual decrease in concentrator metallurgical performance over time. Work had been conducted at Mt Isa between 1975 and 1985, involving regrinding to ultra fine sizes to increase liberation, using conventional grinding technologies. It was found that the conventional technologies had a very high power consumption to achieve the required sizing and the flotation performance was worse than expected due to contamination from the iron media used.

In 1990, there was no accepted technology for regrinding economically to ultra fine sizes in metalliferous operations. Test work in 1990 and 1991 indicated that high speed horizontal mills could efficiently grind to a product of 80%

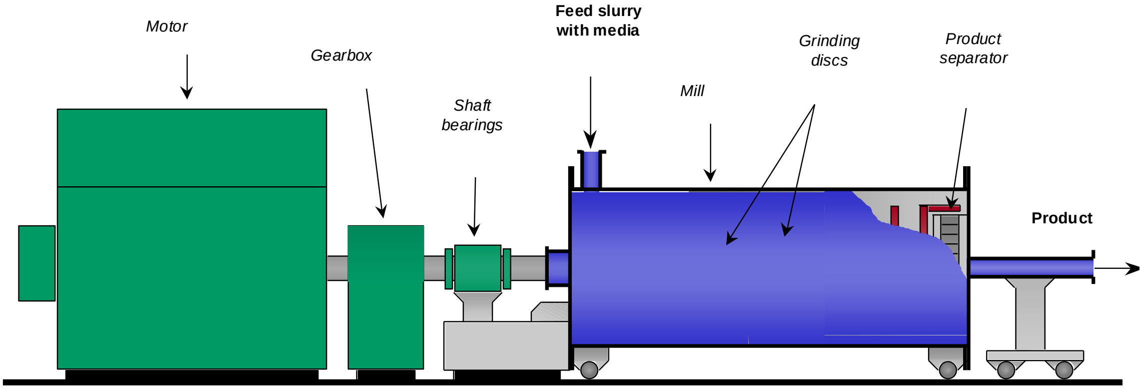


Figure1. The IsaMill.

passing 7 microns at laboratory scale and subsequently provide a major increase in metallurgical performance. To make an ultrafine grinding mill capable of treating the tonnages required at Mt Isa, a program of development was undertaken between Mount Isa Mines Limited in association with NETZSCH-Feinmahltechnik GmbH (NFT).

Scale-up was tested using trial installations at the Hilton and Mt Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill (1.1MW) was installed in the lead/zinc concentrator. This has allowed a suitable, low cost grinding media to be proven in operation and provided a system for separating grinding media from product, without the disadvantages of screens or sedimentation zones. Furthermore, it has allowed development of cost effective wear materials. A diagram of the 1.1MW IsaMill is shown in Figure 1.

In 1998 the rights for commercialisation of the IsaMill where transferred from Mount Isa Mines Limited to MIM Process Technologies and under an exclusive agreement with NFT, on the 14th of December, 1998 the IsaMill technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. There are currently two 1.1MW IsaMills operating in Mount Isa, treating lead concentrate and five 1.1MW IsaMills operating at the McArthur River site, treating bulk lead/zinc concentrate. A further six 1.1MW IsaMills will be installed in Mt Isa to treat zinc concentrates from the George Fischer Mine, commencing June, 1999.

APPLICATIONS

Ultrafine Grinding for Flotation

Several 1.1MW IsaMills have been installed in both the lead and zinc cleaner flotation circuits at Mount Isa.

Operations on lead rougher concentrate have shown that a throughput of 73tph per mill could be achieved at 45% solids. The net energy consumption for reducing a feed of 20 microns to a product 80% passing twelve microns was 6kWh/t. Figure 2 shows the improvement in silica rejection with and without the IsaMills in operation.

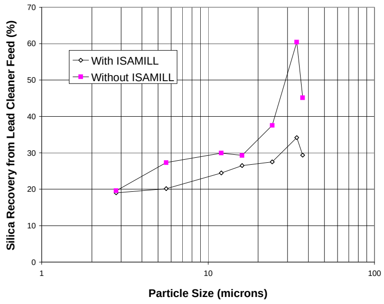


Figure 2. Improved silica rejection from concentrates.

IsaMills were also installed to treat zinc retreatment circuit cyclone underflow. From a feed of 80% passing 46 microns a product of ten microns could be produced with a specific energy of 65kWh/t. Figure 3 shows the improvement in the zinc grade/recovery curve after ultrafine grinding with the IsaMill.

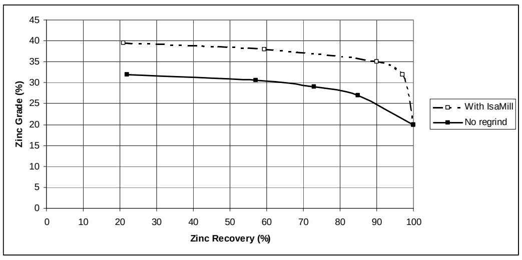


Figure 3. Improvement in the zinc grade/recovery curve after ultrafine grinding.

Of specific interest in the Mount Isa operations is the wide choice of grinding media available for use in the IsaMills. Mt Isa has variously used heavy media plant reject and lead or copper smelter slag for grinding media. These materials had previously been considered waste and provided an extremely inexpensive grinding media. In addition they are inert, having no effect on flotation chemistry, unlike steel media.

Ultrafine Grinding for Direct Cyanidation

Many refractory gold bearing ores contain gold in close association with sulphide minerals, such as pyrite and arsenopyrite. Refractory gold may be present in several forms, ranging from fine free gold housed on boundaries between mineral grains, to gold that is in solid solution with the sulphide matrix. Techniques used to recover gold from refractory ores range from ultrafine grinding for improved liberation of physically locked gold, to oxidative processes where the sulphide matrix is wholly or partially destroyed through chemical oxidation.

The effectiveness of ultrafine grinding as a method of liberating physically locked gold is illustrated in Table 1. The data in Table 1 is also displayed graphically in Figure 4.

A sample of pyrite ore, grading 4.59 g/t gold and 4.85 % sulphur was finely ground in a laboratory scale IsaMill to 80 % passing 20 and 10 microns respectively. The finely ground product was then leached in a conventional agitated cyanide leach test for a period of 48 hours at pH 10, and a free cyanide level of 500 ppm.

80 % passing size - microns	***% gold recovery***	***NaCN cons kg/t***	***Specific Energy Requirement – kWh/t***
74	23.4		
20	39.5	1.44	17
10	70.2	1.73	42

Table 1 .The Effect of Ultrafine Grinding on Gold Recovery from Pyritic Ore.

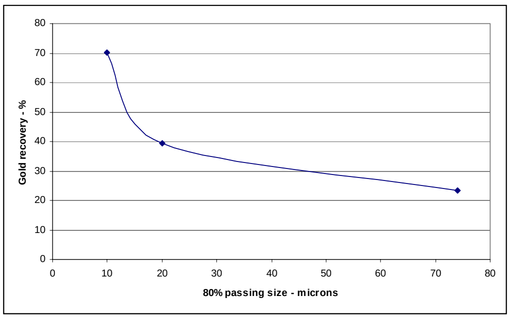


Figure 4. Gold Recovery versus Particle Size for a Refractory Gold Ore

Ultrafine grinding resulted in a significant improvement in the gold recovery through cyanidation. Typical gold recoveries for the ore sample, when ground to 80 % passing 75 microns in a conventional ball mill, were in the range 20 – 25 % w/w. The gold recovery from the ore after grinding to 80 % passing 20 and 10 microns improved to 39.5 and 70.2 %, respectively.

<u>Ultrafine Grinding for Chemical Oxidation.</u>

Not all refractory gold ores, however, will show an improvement in gold extraction through fine grinding alone. Many ores require oxidative breakdown of the mineral sulphides to achieve sufficient liberation of gold for recovery through cyanidation. The most common oxidative processes used to recover gold from refractory sulphides are pressure leaching, bacterial leaching and roasting.

Pressure leaching involves the oxidation of sulphide minerals in acidic solutions at elevated temperatures and pressures, where the aggressive leaching conditions are used to improve the kinetics of the oxidation process.

In bacterial leaching, microbes that occur naturally in acidic mine waters are utilised as a catalyst to increase the rate of mineral oxidation. The catalytic effect of the bacteria allows bacterial leach plants to run at considerably lower temperature and pressure than pressure leaching plants.

Roasting involves the reaction of sulphide minerals with hot gasses containing oxygen, at temperatures in the range 500 - 800 ^{o}C. Sulphur dioxide gas, a byproduct of the roasting process, is usually captured for production of sulphuric acid.

In recent years, there has been a considerable amount of development carried out on processes which couple ultrafine grinding and oxidative leaching in an effort to reduce the capital costs associated with Pressure Leaching, Bacterial Leaching and Roasting. One such processes is MIM Holdings' ALBION Process.

CASE STUDY

Comparative fine grinding and oxidative leaching testwork was carried out on a sample of pyrite concentrate using MIM Holding's proprietary ALBION process and two common laboratory scale ultrafine grinding mills. The two mills tested were the IsaMill and a Vertically Stirred Mill (VSM). A head analysis of the pyrite concentrate is listed in Table 2.

Element	***Assay***
Fe	28.5
S	23.6
Au	50
Ag	18

Table 2. Head Analysis of Pyrite Concentrate Sample used in the Fine Grind/Oxidative Leach Testwork.

Samples of ground concentrate were produced from both mills at a range of particle size distributions. The ground concentrate samples were then leached under the conditions typically specified for the ALBION process, which involves oxidation of sulphide minerals at atmospheric pressure in conventional agitated tanks. Cyanide leaching of an oxidised concentrate produced from each type of mill was also carried out.

Experimental Setup

The laboratory scale IsaMill consisted of a horizontal milling chamber, drive motor and 6 disc radial impeller. The mill was fitted with a 4.0 kW variable speed drive and Yokogawa power meter to determine the specific energy requirements of each grind. A diagram of the laboratory ISAMILL set up is shown in Figure 5.

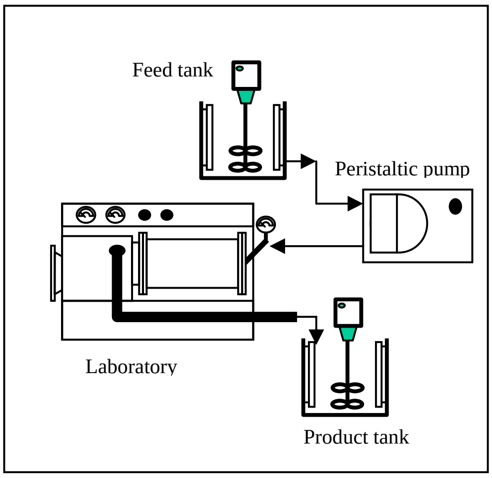


Figure 5. Diagram of the IsaMill Experimental Setup

The pyrite concentrate sample was pumped by variable speed peristaltic pump from a feed tank to the IsaMill feed port. Slurry exited the mill through a separator located before the mill mechanical seal, which allowed slurry to pass but retained media within the mill.

Ground slurry was discharged from the mill through a slurry discharge port and was collected in a sealed product tank. Samples of ground slurry were collected for the leaching tests, and each of the slurry samples was filtered and the filter cake stored frozen prior to the leach tests. The mill media was 0.8 – 1.2mm steel shot.

The Vertically Stirred Mill consisted of a 10 litre chamber and pin style impeller. The mill shaft held seven radial arms, each 160 mm long and 10 mm in diameter. The mill motor was a 2.7 kW variable-speed drive, fitted with a 3 phase AC inverter for power consumption measurements. The mill media was 2.3 – 6.6 mm steel shot. A diagram of the laboratory Vertical Mill set up is shown in Figure 6.

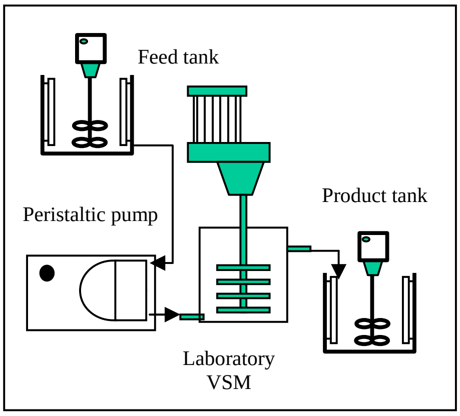


Figure 6. Diagram of the Vertically Stirred Mill Experimental Setup

Feed slurry was pumped through the mill feed port located at the base of the mill, and was discharged from the mill overflow port. Ground slurry was collected in a sealed product tank. Each of the slurry samples was filtered and the filter cake stored frozen prior to leaching.

Results.

The results of the ultrafine grinding and oxidative leaching tests are listed in Table 3. Figure 4 shows the effect of particle size on the extent of sulphur oxidation achieved for the two different types of fine grinding mill.

The IsaMill was the most efficient of the two mills tested, grinding to 80 % passing 8.61 microns at a specific energy input of 64 kWh/t. In comparison, the Vertically Stirred Mill required 105 kWh/t to achieve a product at 80 % passing 8.2 microns. The relationship between specific energy and particle size for both of the mills is shown in Figure 7.

The specific energy curves for the two mills were similar for product sizes down to 80 % passing 25 microns, at which point the slope of the curve for the Vertically Stirred Mill increased significantly relative to the IsaMill. The curves in Figure 4 show that the IsaMill was much more efficient at grinding in the particle range below 12 microns.

Mill Type	Specific energy Input – kWh/t	80 % passing size	% sulphur oxid.
Head	0	108.39	24
IsaMill	11	51.9	61
IsaMill	19	36.4	76
IsaMill	31	17.94	91
IsaMill	41	13.64	96
IsaMill	52	11.71	96
IsaMill	64	8.61	96
VSM	12	41.5	74
VSM	24	26.7	88
VSM	75	10.68	96
VSM	105	8.2	96
VSM	148	7.54	94

Table 3. Results of ultrafine grinding and oxidative leaching of a pyrite concentrate sample.

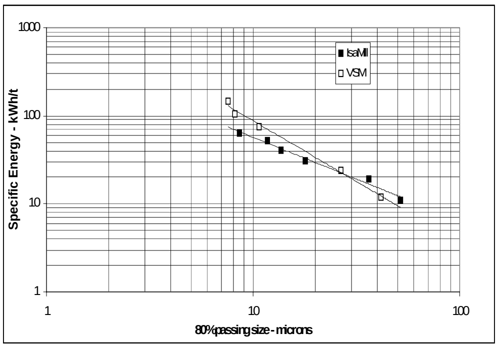


Figure 7. Comparative Milling Curves for the IsaMill and Vertically Stirred Mill.

In terms of breakage rate, the mill residence time to achieve a specified grind showed a similar pattern to that of specific energy. The relationship between mill residence time and particle size for both of the mills is shown in Figure 8. The residence time for the two mills were similar for product sizes down to 80% passing 25 microns and below this point the IsaMill required significantly less residence time to achieve a specified grind. At the target grind of 80% passing ten micons the IsaMill required 12.49 minutes grinding time, compared to 18.46 minutes residence time for the Vertically Stirred Mill.

The superior efficiency of the IsaMill in grinding to below 12 microns was believed to be due to the smaller media used in the mill. It is well known that the efficiency of any fine grinding process improves with finer media, and the mechanics of the IsaMill allow for agitation of finer media at higher tip speeds than is possible for a vertically stirred mill.

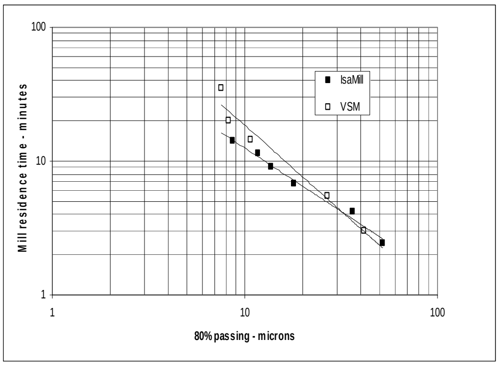


Figure 8. Comparison of residence time ***for the IsaMill and Vertically Stirred Mill.***

The degree of oxidation achieved from leaching the concentrate samples did not show a large variation with type of mill, as indicated by the closeness of the oxidation versus particle size curves in Figure 5, below. Both mills achieved a maximum of 96% oxidation at approximately 80% passing 15 microns.

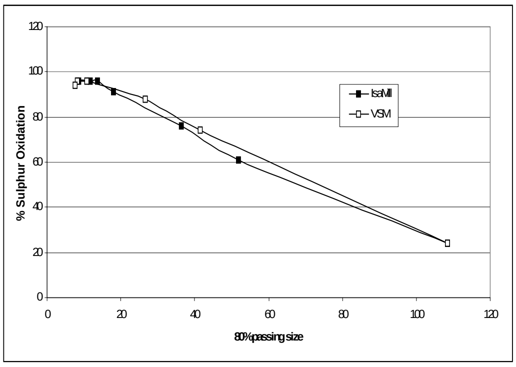


Figure 9. The Effect of Particle Size on Sulphide Oxidation for the two Mill Types

CONCLUSIONS

The use of ultra fine grinding in combination with cyanidation or chemical leach processes is a viable alternative to traditional expensive treatment of refractory gold concentrates.

A laboratory scale case study comparing an IsaMill (a horizontal stirred mill) and a vertical stirred mill has shown that a more energy efficient outcome is likely with the former.

In addition, a superior residence time performance of the IsaMill suggests a more compact, lower cost mill, when installed.

Developments in Milling Practice at the Lead/Zinc Concentrator of Mount Isa Mines Limited from 1990.

M.F. Young[1], J.D. Pease[2], N.W. Johnson[3] and P.D. Munro[4].

Submitted for:

AusIMM Sixth Mill Operators Conference, 6-8 October 1997

Madang, Papua New Guinea.

ABSTRACT

The Lead/Zinc Concentrator of Mount Isa Mines Limited processes complex fine grained ore from the Isa and Hilton silver-lead-zinc orebodies, producing lead concentrate, zinc concentrate and (until 1996) a low grade middlings (LGM or bulk) concentrate.

Metallurgical performance declined dramatically during the 1980's because of declining ore quality, as the ore became both finer grained, and contained increasing amounts of refractory pyrite. Developments in milling practice to restore performance focused on two areas: liberation and separation. Increased mineral liberation was achieved by more than doubling grinding and regrinding capacity to increase sphalerite liberation. This successfully recovered an extra 20 per cent zinc metal to zinc concentrate, which previously reported to final tailing, lead concentrate or LGM concentrate. There was also a small increase in galena liberation, increasing lead recovery to lead concentrate and reducing contamination of the zinc concentrate by lead. The increased sphalerite and galena liberation also significantly reduced the production of LGM concentrate.

The second area of development was to improve the separation of galena and sphalerite from gangue minerals. This was achieved by circuit rationalisation, better understanding of water chemistry leading to an improved reagent scheme, and improved process control. These changes both improved performance and simplified the circuit, giving better and steadier concentrate grades and recoveries.

The combination of increased liberation, improved separation, and circuit simplification has dramatically increased the metallurgical performance of the Lead/Zinc Concentrator when treating very complex fine grained ore.

1. Metallurgical Superintendent, Lead/Zinc Concentrator, Mount Isa Mines Ltd, Mount Isa Qld 4825.
2. Lead/Zinc Concentrator Manager, Mount isa Mines Ltd, Mount Isa Qld 4825
3. Minerals Processing Research Manager, Mount Isa Mines Ltd, Mount Isa Qld 4825
4. Principal Engineer, Metallurgy, MIM Holdings, 410 Ann Street, Brisbane Qld 4000

INTRODUCTION

Processing of Mt Isa silver-lead-zinc ore commenced in 1931 at the No.1 Concentrator, initially treating a mixture of oxidised and sulphide ore, but by 1935 treating only sulphide ore (Kruttschnitt et al, 1939). Over the years the flowsheet was developed to improve metallurgical performance of the fine grained and difficult to treat Mount Isa ore (Kruttschnitt et al, 1939, Cunningham, 1953, Challen et al, 1968, Davey and Slaughter, 1970). Over the period 1952 to 1960, ore reserves were increased substantially, and the decision was made to increase the treatment rate, firstly by modernising and expanding the existing No.1 Concentrator, and secondly by constructing a new (No. 2) Lead/Zinc Concentrator (Challen et al, 1968).

The No 2 Concentrator was commissioned in June 1966 and total silver-lead-zinc ore treatment was transferred from the No. 1 Concentrator in May 1967. Various improvements in the 1970's (Bartrum et al, 1977) were followed by the installation of larger flotation cells in a single circuit conversion (Johnson et al, 1982). A major increase in throughput occurred in 1982 with the commissioning of a Heavy Medium Preconcentration plant to reject 30 per cent of waste ore before flotation (Fiedler et al, 1984). From 1987, ore from the nearby Hilton mine was introduced to supplement the Mt Isa ore. By 1992, treatment rate had reached 5 Mt/y, of which 30 per cent was from the Hilton Mine and 70 per cent from Isa Mine.

During the 1980's, the increase in throughput and decline in head grade were exacerbated by a significant increase in ore complexity. This resulted in a severe liberation problem, with a finer mineral liberation grain-size than the plant grinding capacity could achieve, and a worsening separation problem caused by increasing amounts of naturally floating carbonaceous pyrite. Metallurgical performance declined dramatically as the plant did not have technologies to deal with either problem. This paper chronicles the change in ore characteristics, along with the highly successful technological changes to return good metallurgical performance with the more complex ore.

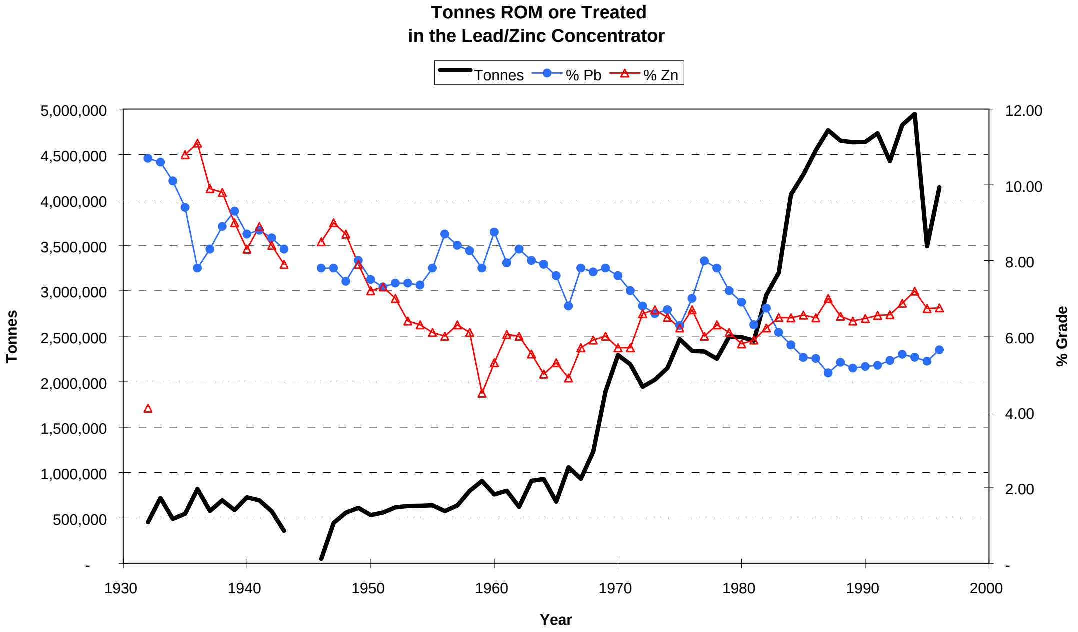


Figure 1.

The history of processing silver-lead-zinc ore at Mount Isa Mines has been one of steadily increasing tonnage and declining head grades, as shown in Figure 1. This was exacerbated by increasing complexity of ore.

MINERALOGY

The mineralogy of the silver-lead-zinc orebodies can be characterised as fine intergrowths of galena and sphalerite with both sulphide and non-sulphide gangue. The main silver material is freibergite which is intimately associated with the galena. The non-sulphide gangue is quartz, dolomite and carbonaceous shale. Minor amounts of chalcopyrite are present. The sulphide gangue is both pyrite and pyrrhotite, with pyrite predominant. Pyrite is present as two distinct types; firstly, as normal relatively coarse grained euhedral pyrite, and secondly, as fine grained (5 to 30um) spheroidal pyrite. This second type is refractory and contains elemental carbon, sometimes forming atoll rims on galena. The carbonaceous pyrite is hydrophobic under almost any conditions and is the dominant sulphide diluent in both lead and zinc concentrates (Davey et al, 1970, Munro, 1993).

Feed to the Lead/Zinc Concentrator consists of both Isa and Hilton ores. Isa ores can be classified into two broad categories :

- The upper Isa orebodies, No's 1,2 and 5, referred to as "Black Star" orebodies. They are more massive and are mined by open stoping at lower mining cost, and therefore have a lower cut-off grade. Generally this ore is finer grained and has considerably more fine-grained carbonaceous pyrite, whilst core replacement of pyrite by galena (atolling) is more common and at a more advanced stage (Davey et al, 1970).

- The lower Isa orebodies, No's 7 to 14 and Rio Grande are referred to as "Racecourse" orebodies. These are more narrow and mined by bench stoping with a higher cut-off grade. Generally these orebodies are higher in grade for lead and silver, coarser grained and lower in pyrite. The pyrite is more the euhedral type than the fine-grained carbonaceous type . Though this ore has a higher mining cost per tonne, it is the more profitable ore since it has the highest grade and the best metallurgy. The gradual displacement of this ore by Black Star and Hilton ores is the reason for the continual decline in ore quality experienced in the Concentrator.

The "Racecourse" and "Black Star" categories are used to describe the mineral types in the ore and its metallurgical performance, as well as its geological location.

Hilton ore has been treated through the Isa Lead/Zinc Concentrator since 1987 and is divided into two similar categories. The upper orebodies, No's 1,2 and 3 ("Black Star" type ore), are more massive in size and thus allow for open stoping, contain more fine grained, naturally floating carbonaceous pyrite than the lower orebodies and contain more pyrrhotite than Isa orebodies. The lower orebodies, No's 4 to 7 ("Racecourse" type ore), are narrower and deeper orebodies, mined by bench stoping, with more euhedral pyrite and non-sulphide gangue dilution than the upper orebodies. The silver minerals in the Hilton orebodies are less associated with galena than in the Isa orebodies, hence silver recovery to lead concentrate is lower than from the Isa orebodies. Hilton orebodies also contain a wider range of silver minerals.

The two microphotographs show the difference in complexity between coarse grained, high grade ore (Figure 2) and the fine grained ore with refractory fine pyrite dilution (Figure 3) (Riley and McKay, 1976). Both of these samples were taken from the No 5 orebody at Mt Isa. Over the years, as the tonnage has increased and the head grade declined, more of the ore feeding the Concentrator has been of the Figure 3 type and less has been of Figure 2 type.

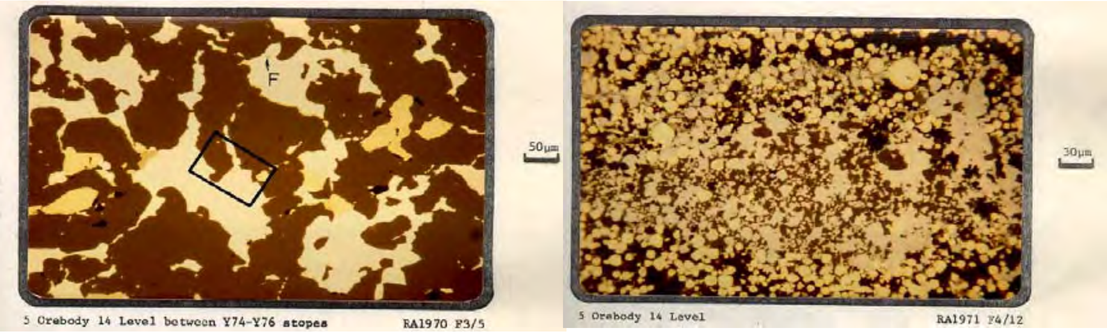


Figure 2. **Figure 3.**

Microphotographs of Mount Isa No 5 orebody showing the different grain sizes and complexities that occur in the orebodies at Mount Isa (Riley and McKay, 1976).

The mineralogy at Mount Isa presents two distinct problems that affect metallurgical performance – achieving adequate mineral liberation during grinding**,** followed by separation in flotation. Clearly the ore in Figure 3 needs much finer grinding to achieve equivalent mineral liberation. While the main separation problem is the increase in refractory pyrite, finer grinding to solve the liberation problem increases the difficulty of separation.

Ore Type Performance

The impact of the more difficult separation because of ore type on flotation performance is demonstrated by Figures 4 and 5 : at the same grind size and ideal laboratory conditions, lead performance can vary from 60 per cent Pb concentrate grade at 90 per cent recovery (characteristic of the best “Racecourse” ore) to 15 per cent Pb concentrate grade at 50 per cent recovery (characteristic of the worst “Black Star” ore) (Figure 4). Similarly, at the same (fine) grind size, zinc recovery at target concentrate grade can vary by 20 per cent (Figure 5).

The laboratory flotation tests shown in Figures 4 and 5 were conducted to evaluate ores using a laboratory flowsheet similar to current plant operation. Crushed ore was initially ground to P80 = 37 um, followed by a lead rougher, rougher concentrate regrinding to P80 = 15 um and final lead concentrate production by three stages of dilution cleaning. Lead rougher tailing and first cleaner tailing were combined to feed the zinc rougher, with zinc rougher concentrate regrinding to P80 = 15 um and final zinc concentrate production by three stages of dilution cleaning. The flotation tests were conducted with a very fine regrind size (15 um) to maximise liberation, since less regrinding gives less liberation, and hence both lower zinc recoveries and more zinc contamination of the lead concentrate.

Lead circuit laboratory flotation performance (Figure 4) varies widely for different ore types depending on the refractory nature of the ore and the type and content of the iron sulphides. Different ore types yield significantly different performance. Since different ore types are being mined at any one time from many sources, so the mixture of ore types feeding the concentrator is continuously changing. This causes the performance of the plant to be continuously changing in the absence of intervention by the control room operator.

This leads to two effects in the short term operation of the concentrator:

- there is a change in performance and the operator is unable to tell if it is an ore change or another input change (eg: reagents, mechanical failure, uncontrolled input); and
- the operator makes a controlled change and the performance changes in an expected or unexpected manner. Was the effect from the operator's change or an ore change ?

These issues also need to be addressed when trying to improve plant performance.

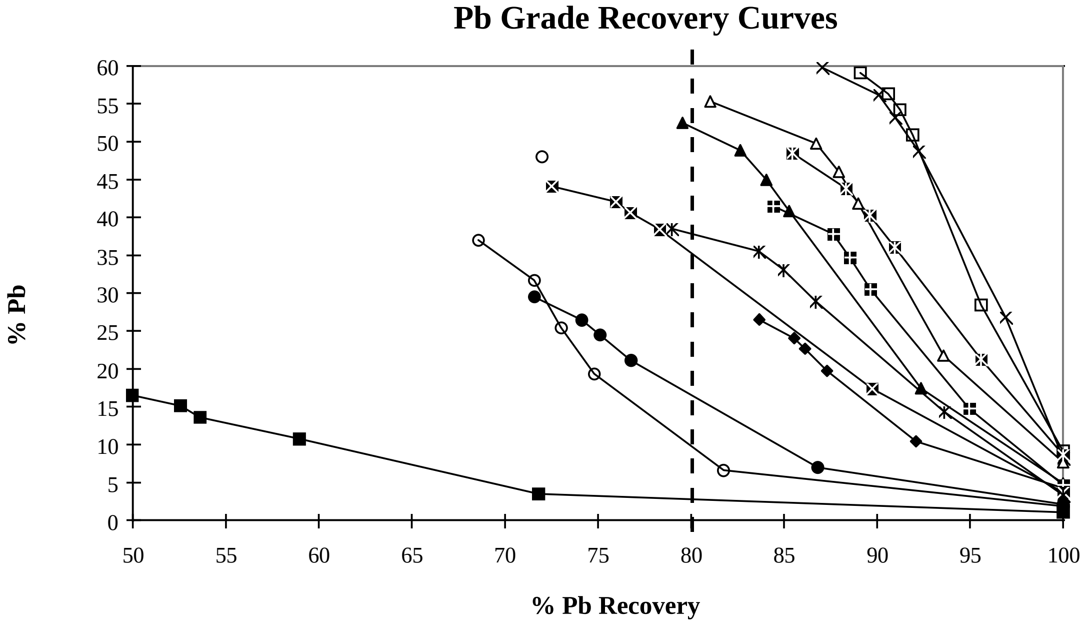


Figure 4 - Lead circuit laboratory flotation performance of different ore types.

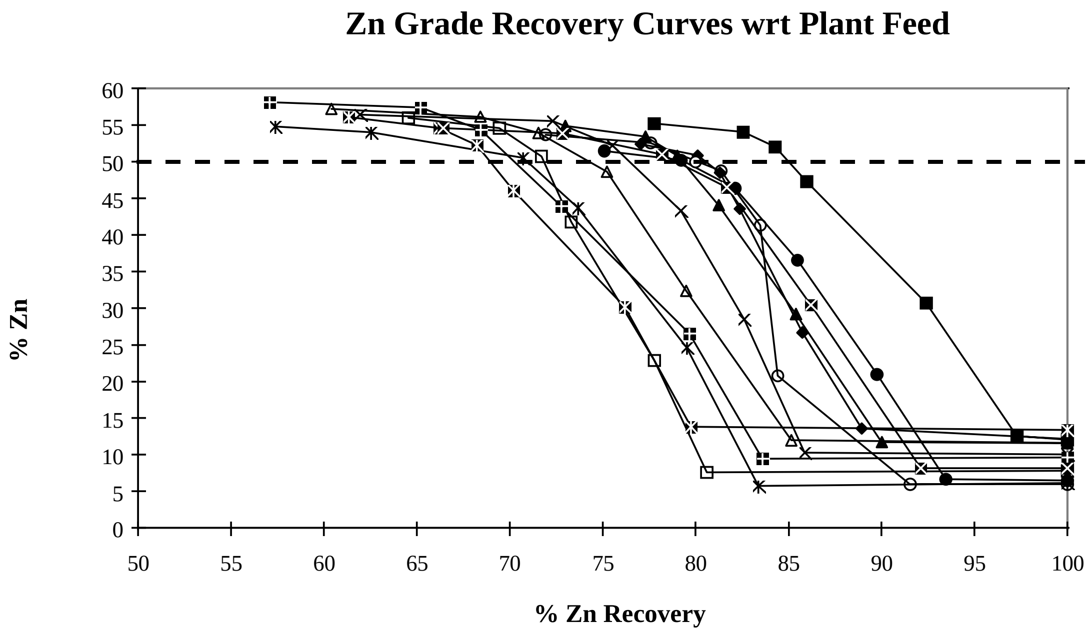


Figure 5 - Zinc circuit laboratory flotation performance of different ore types.

Zinc metallurgical performance (Figure 5) varies over a smaller band, due to low zinc losses in the lead circuit and the ability to be more selective against pyrite.

The methods used in the laboratory have shown good agreement with plant performance and provide confidence in using laboratory testwork to predict plant performance.

A SCIENTIFIC APPROACH TO THE PROBLEM

The rapidly deteriorating metallurgical performance in the 1980's was attributed to continual changes in ore mineralogy. Until the nature of these changes were fully understood, the response consisted of an endless circle of circuit changes, reagent changes, operator changes, metallurgist changes and so on. Fortunately, this was a brief, (though tense) period. It was clear that the solutions could only come from a rigorous scientific approach based on the mineralogy.

Fortunately, good scientific tools were in place to understand the nature of the changes. The two fundamental tools were size-by-size mineralogical analysis and liberation analysis. These two tools were applied to routine plant inventory samples, detailed plant surveys and laboratory and pilot plant testwork. The data combined to provide a unique mineralogical profile of plant performance, which captured both the decade-long decline in performance, and the results of the step changes in improvement.

Routine Analysis of Plant Inventory Samples

Inventory samples of plant products are taken and assayed every shift for metallurgical accounting purposes. Great care has been taken with the design and operation of inventory samplers to ensure there is no size or assay basis. Shift samples are combined into weighted daily composites, which are further combined into weighted monthly composites. In addition to chemical assays, the monthly composite samples are subjected to the following analyses:

- Screen sizing to 37 um, followed by fine sizing (infrasizing until 1992/93, then cyclosizing after 1992/93. The cyclonizer part of the sizing is extended to C7 by a precyclone, followed by collection of the normal C1 to C5 cyclosizer fractions and then to C6 by a centrifuging of the minus C5 fraction. This procedure allows extension of the size analysis to finer sizes, as well as collection of the finest sample. The C6 fraction is typically 4 to 7 um for sphalerite (Johnson, 1992).
- Chemical assay of all size fractions, providing a fully sized mass balance for the plant each month.

- Liberation analysis of size fractions. Until 1992/93 this was done by manual point counting and afterwards by QEM*SEM (Quantitative Evaluation of Materials by Scanning Electron Microscope). This provided a full size-by-size monthly mineralogical mass balance of plant operations.

Plant Surveys

In addition to monthly balances, more detailed information was obtained from occasional full or part plant surveys. The surveys are carefully designed to provide a complete snapshot of the operation, with a full mass balance including cyclone splits and down-the-bank flotation performance. All samples are assayed, and selected samples sized and analysed mineralogically.

Laboratory and pilot plant testwork

Laboratory and pilot plant testwork was used to test and identify solutions to problems, the size and quantity of potential performance improvements and the flowsheets required to achieve performance gains. As before, all samples were assayed, and selected samples sized and analysed mineralogically.

THE LIBERATION PROBLEM

The first step change in sphalerite liberation occurred in July 1980 when target zinc concentrate grade was dropped from 52 per cent Zn to 50.5 per cent Zn to maintain recovery above 70 per cent (Johnson et al, 1982). This change caused the adoption of the rigorous mineralogical approach to quantify future ore changes. Figure 6 shows a graphic summary of the changes to sphalerite liberation after 1980.

In Figure 6, ‘sphalerite liberation’ represents the percentage of sphalerite in plant feed which has been liberated before exiting the plant in either concentrate or tailings. This is achieved by the tonnage-weighed mathematical combination of all plant products to form a recalculated plant feed, which represents the total effect of all grinding and regrinding in the plant. A sphalerite grain is considered liberated if it is more than 90 per cent sphalerite in two-dimensional analysis (Gottlieb et al, 1994).

From 1984 to 1991, sphalerite liberation dropped from almost 70 per cent to just over 50 per cent. This was attributed to finer grained ore, although the recalculated feed sizing coarsened from P80 = 50 um to P80 = 80 um because of increases in throughput with no extra grinding power. A decrease in sphalerite liberation causes a drop in zinc recovery, since the maintenance of zinc concentrate grade at 50.5 per cent Zn allows no additional lower grade composites in the concentrate.

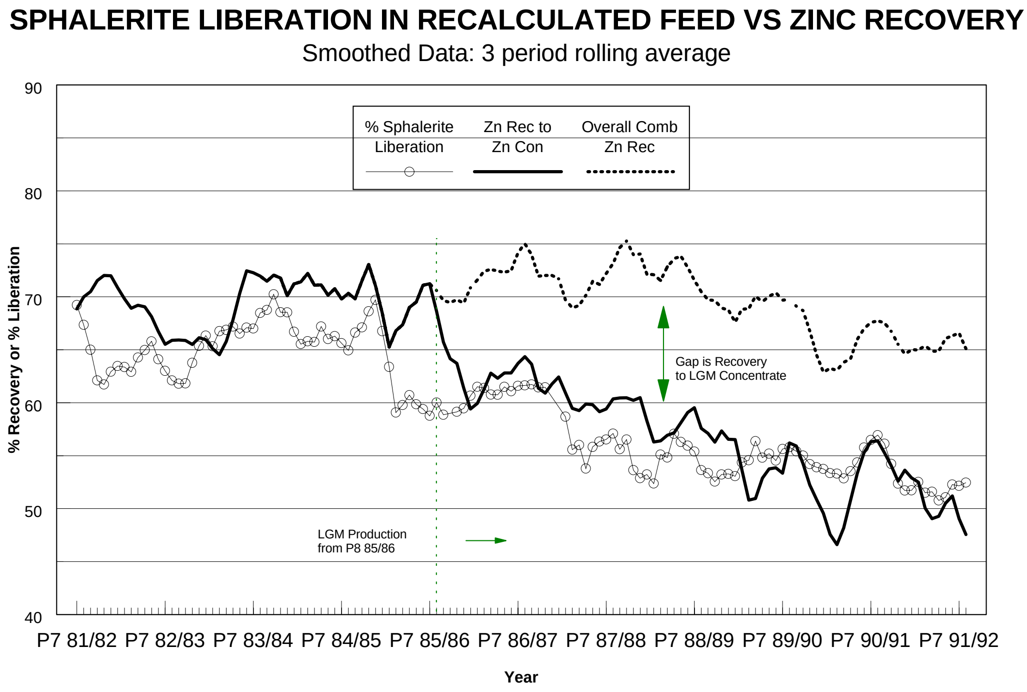


Figure 6.

There were two possible responses to the liberation problem; either grind finer to increase liberation, or place low-grade middlings into a new, lower grade concentrate. Because of the high capital cost of additional grinding, production of a Low Grade Middlings (LGM), or bulk concentrate became necessary in 1985/86, to maintain total zinc recovery. This concentrate typically assayed 13 per cent Pb and 34 per cent Zn, compared with zinc concentrate which had to contain 50 per cent Zn and less than 3 per cent Pb. Figure 6 shows the effect of the LGM Concentrate on total zinc recovery. The difference between the zinc recovery to zinc concentrate and overall combined zinc recovery is the zinc recovery to the LGM concentrate.

As liberation continued to drop, recovery to zinc concentrate fell with concomitant increases in LGM concentrate production. By 1988, total zinc recovery had to decline further as the target 34 per cent zinc in LGM concentrate was unattainable and the LGM concentrate market had become saturated.

In hindsight, production of LGM concentrate distracted management from the true severity of the problem, since zinc recovery was still quoted as over 70 per cent until 1989. This was really a misrepresentation, since only 55 per cent was recovered to zinc concentrate, with 15 per cent to LGM concentrate. It should also be noted that at the beginning of LGM production, revenue was high because of good contract terms for LGM concentrate. As production rose, contract terms declined until zinc in LGM concentrate was worth less than half that of zinc in zinc concentrate.

Size by size analysis

During the 1980’s the sphalerite liberation declined in three major steps (as shown in Figure 6):

- Mid 1985 from 70 per cent to 60 per cent,
- Mid 1987 from 60 per cent to 55 per cent, and
- Early 1991 from 55 per cent to 50 per cent.

Figures 7 and 8 show the recalculated plant feed sphalerite liberation (by size fraction) and the zinc recovery to zinc concentrate (by size fraction) for selected months over the period 1984/1985 to 1991/1992. Size fractions displayed in Figures 7 and 8 are infrasizer fractions (F7, F6 and F5) and sieve size fraction +400# (+37um). The unsized sample is shown as ALL.

It can be seen from the sphalerite liberation by size (Figure 7) that sphalerite liberation in all size fractions decreased almost uniformly. Consequently, zinc recovery by size also decreased uniformly

(Figure 8). Figures 7 and 8 combine to show that all size fractions were becoming more complex and difficult to liberate, not just the coarse size fractions.

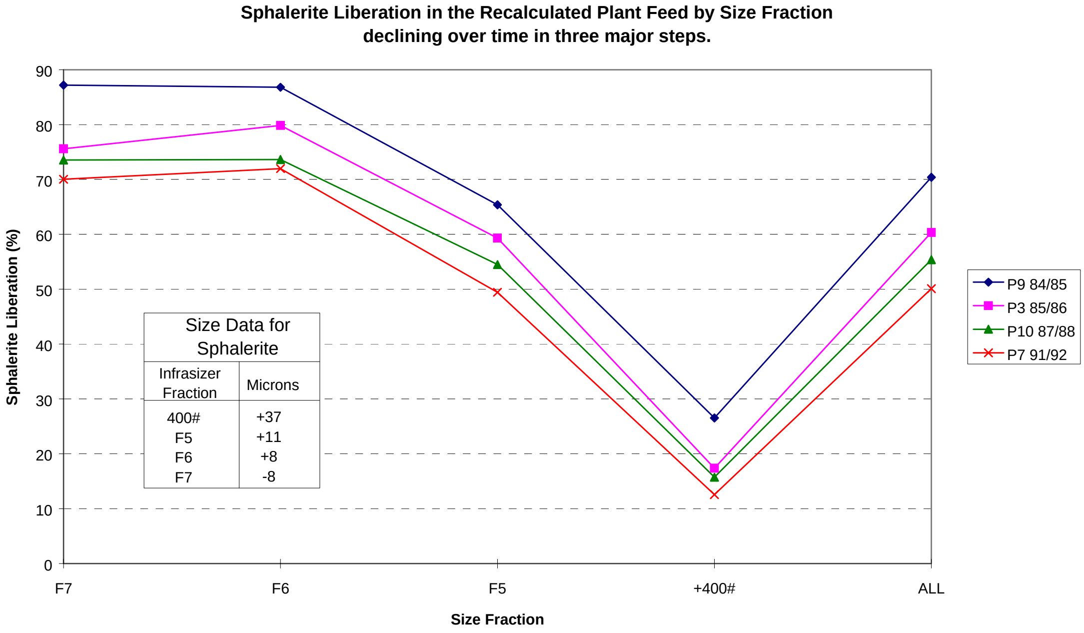


Figure 7.

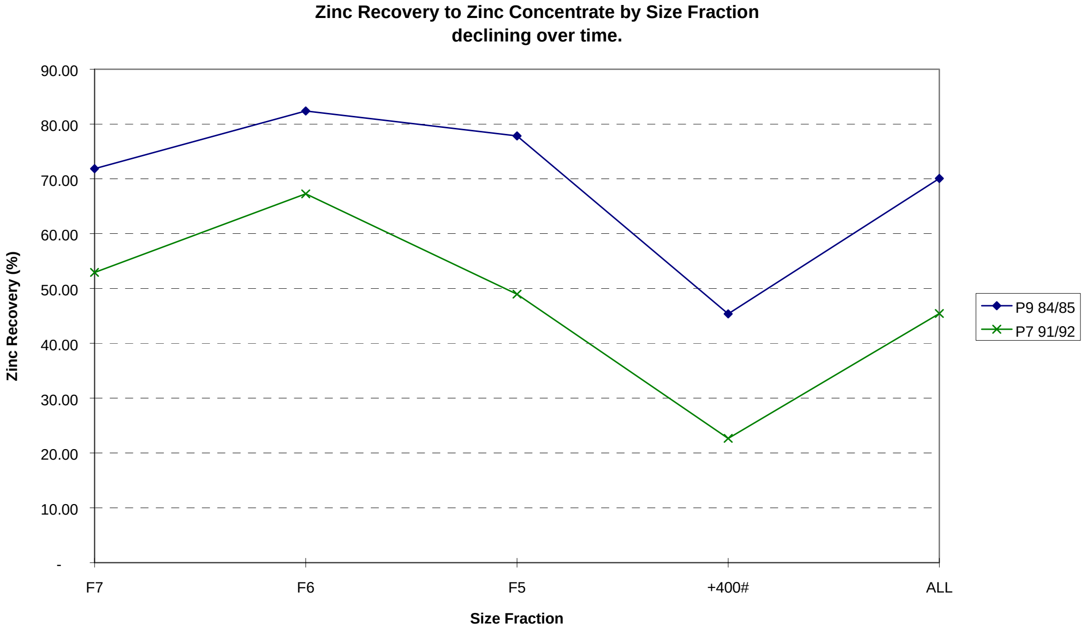


Figure 8.

THE SEPARATION PROBLEM.

While declining liberation posed the most serious problem for zinc metallurgy, decreasing separation efficiency posed the greatest problem for lead metallurgy, and was a secondary issue for zinc metallurgy. The separation problem was caused by increasing amounts of fine grained, carbonaceous pyrite. As shown by Figure 9, lead head grade declined during the 1980's with a concomitant increase in iron sulphides. This was exacerbated by an increasing proportion of the pyrite occurring as the naturally floating carbonaceous type, rather than the "well behaved" euhedral pyrite.

Carbonaceous pyrite is hydrophobic under almost any flotation conditions and consumes large quantities of reagent, making flotation difficult to control. Figure 10 shows the changes to lead metallurgy from 1973 to 1990, with lead concentrate grade and recovery both decreasing and the high viability of lead concentrate grade. Although the natural floatablility of the carbonaceous pyrite results in a greater impact in the lead circuit, Figure 11 shows that there was also some impact on the zinc circuit. For the same concentrate grade 1980-1992, iron sulphide content rose from 4 per cent to 6.5 per cent. This also contributed to falling zinc recovery, since the higher iron sulphides content in zinc concentrate left less room for composite particles containing sphalerite.

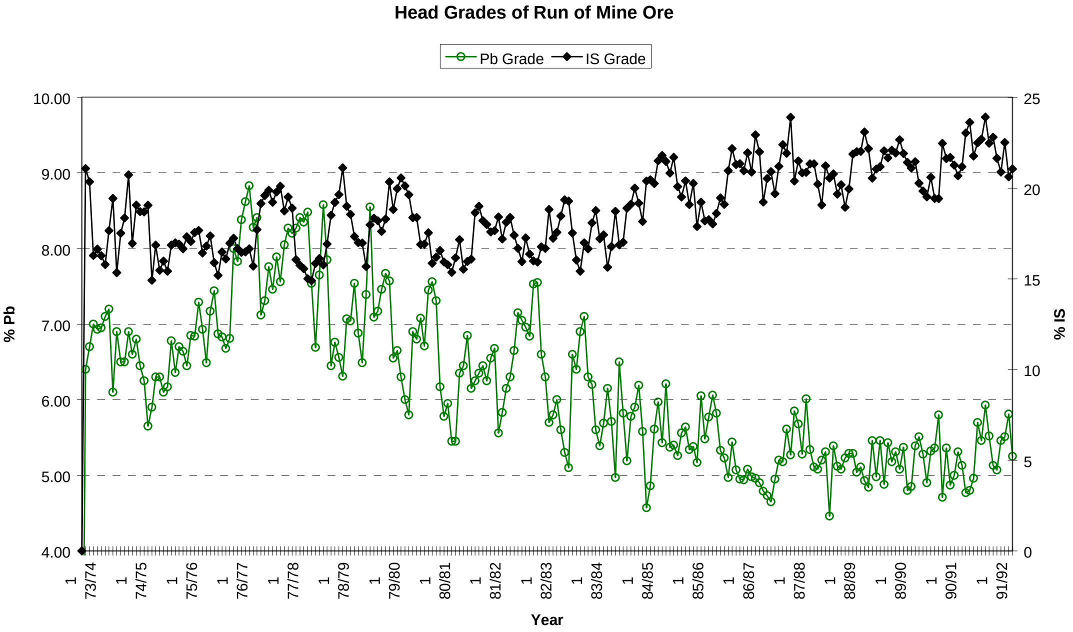


Figure 9.

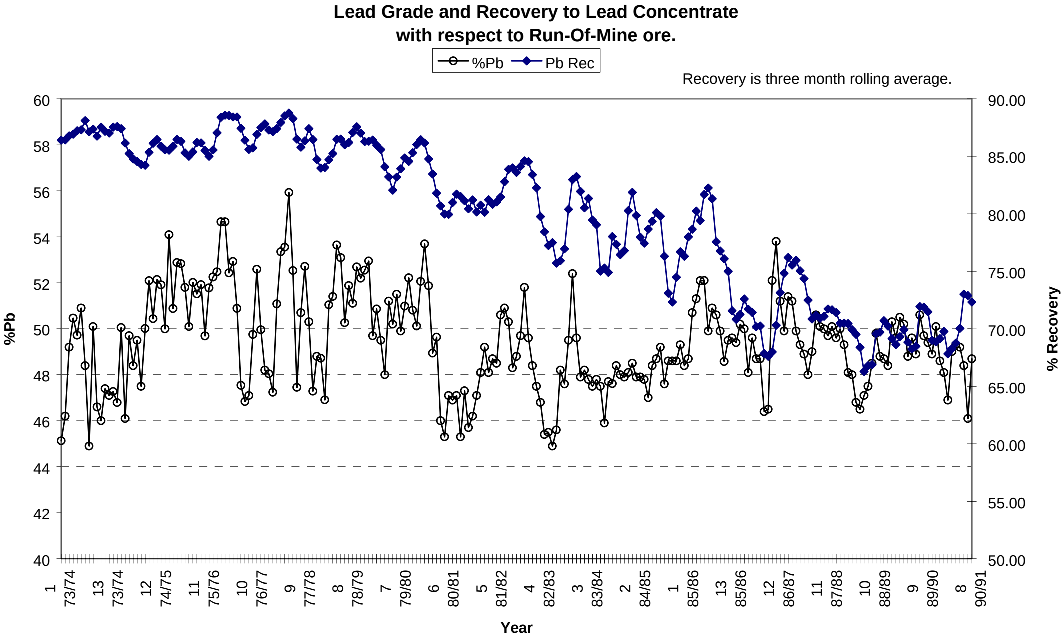


Figure 10.

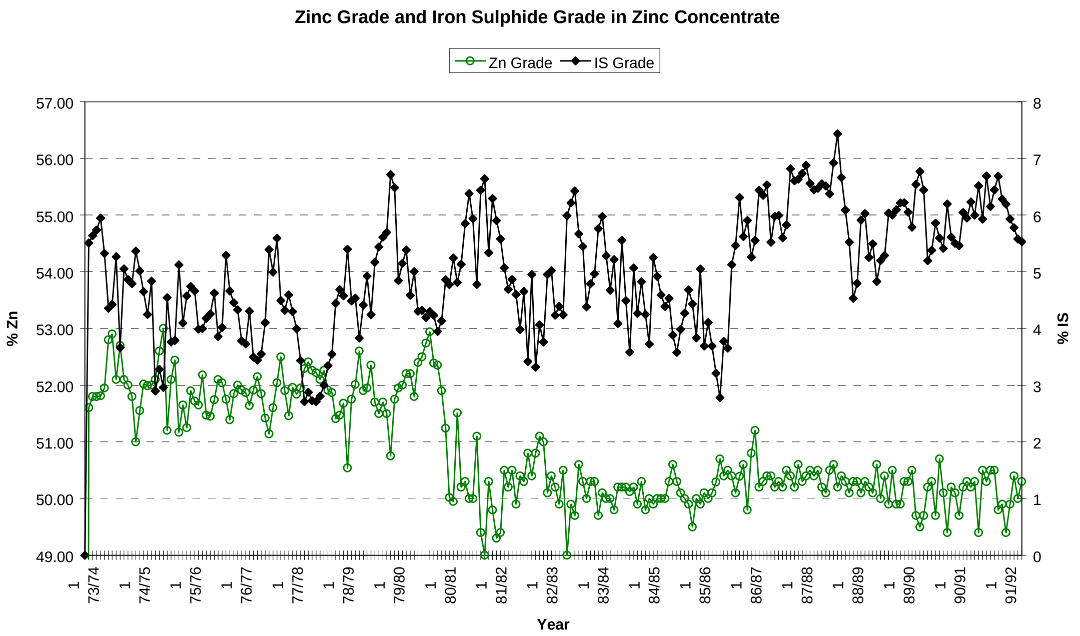


Figure 11.

The falling lead concentrate grade also posed a serious problem for the lead smelter. Smelter throughput was limited by sinter plant sulphur removal capacity. Increased pyrite in lead concentrate caused lead grade to decrease and sulphur to increase (Figure 12), reducing smelter throughput.

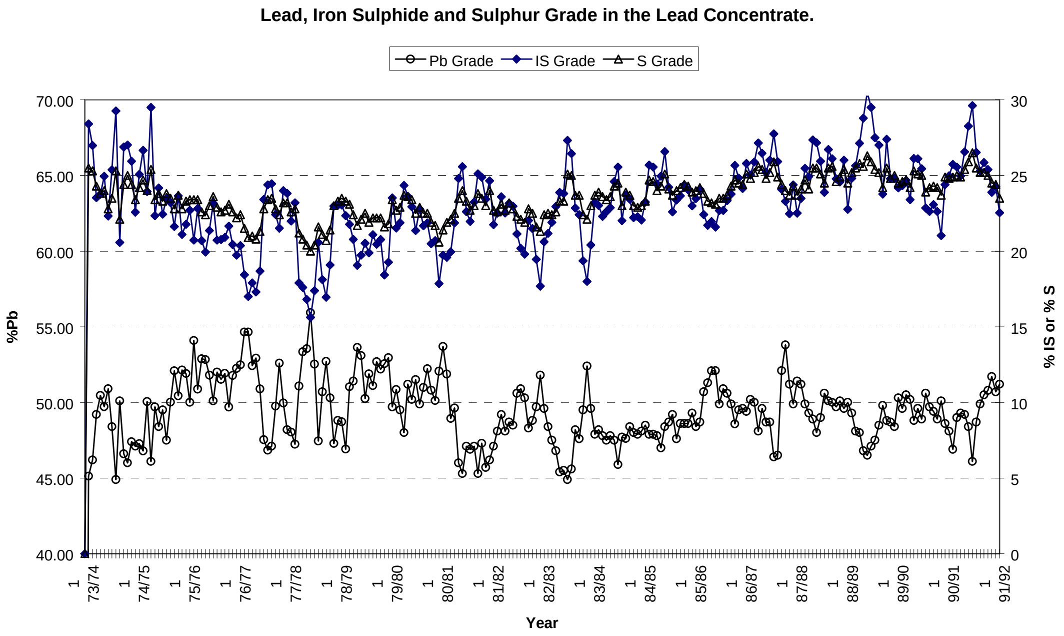


Figure 12.

Numerous circuit and reagent alternatives were trialed to minimise the impact of the carbonaceous pyrite. After several decades of work, the most effective response remains a dextrin depressant at natural pH (7.5 to 8). Other changes instituted in the 1980's to improve lead circuit separation included:

- Installation of preflotation before lead roughing which operated between December 1986 and February 1987. Preflotation concentrate was cleaned before discarding at an assay slightly higher than head grade for lead, zinc and silver. Preflotation operation was stopped partially because of recovery loss and partially due to the resultant unstable operation of subsequent lead roughing and cleaning. Without preflotation, the majority of lead circuit xanthate addition is added to mill feed, but with preflotation the addition is made to rougher feed. Addition of the xanthate to rougher feed appeared to cause instability in the lead circuit, particularly in summer with high pulp temperatures (+45 deg C).
- Use of a “high-low split” lead cleaning circuit during 1991 and 1992 (Figure 13). This circuit collected a high grade concentrate from the first part of the lead cleaners, whilst a low-grade concentrate from the second part of the cleaners was sent to further cleaning in a Jameson cell, the latter producing a high-grade concentrate and a low-grade tail. High-grade concentrates were sent to the lead smelter whilst the low-grade Jameson cell tailing was stored until smelter capacity was available.
- Installation of a Heavy Medium Plant (HMP) slimes roughing and cleaning circuit in 1988. These “slimes” are generated in the mining process, represent about 15 per cent of the lead in the flotation feed, and have different reagent requirements to normal feed (Grano et al, 1988). A flowsheet including roughing at pH 9 (with lime), zinc sulphate, and cleaning in a Jameson cell was developed, producing a lead concentrate from HMP slimes that averaged over 60 per cent Pb, compared with an estimated 45 per cent Pb when included in conventional feed. The separate HMP slimes roughing and cleaning contributed a two per cent increase in overall lead concentrate grade.
- The LGM concentrate circuit from 1986 assisted by directing difficult lead middling streams into a low grade concentrate. This raised lead concentrate grade, and helped match concentrator and smelter capacity by directing some metal away from the lead concentrate.

In the case of zinc separation, little improvement was possible because of the poor liberation. Use of traditional iron sulphide depressants (eg. lime and dextrin) became severely restricted because of uncontrollable circulating loads of composites and fine free sphalerite (Johnson et al, 1982). In turn, the circulating loads consumed limited cleaning and retreatment flotation capacity. Attempts to improve zinc circuit selectivity included :

- Operating the LGM circuit to provide an exit for the most difficult composite streams, eg zinc scavenger concentrate and zinc cleaner tailing;
- installation of new column cleaning capacity in the zinc retreatment and LGM circuits (Espinosa-Gomez et al, 1989 and Espinosa-Gomez and Johnson, 1991);
- substitution of lime with dextrin in zinc cleaning, with later restriction of dextrin additions to minimise circulating loads;
- trials of a number of supplementary collectors which promised, but did not provide, increased selectivity; and
- use of a hot reverse cleaning circuit developed in the pilot plant. However, by the time this approach was developed it was clear that circuit simplification was the priority rather than the addition of new equipment.

The changes made during the 1980’s were individually effective in achieving performance improvements, however the rate of improvement did not match the ore type rate of decline. Further, the succession of small changes had dramatically increased the complexity of the combined lead, zinc and LGM circuits as the changes had treated symptoms rather than the underlying mineralogical cause. By 1992, the Concentrator had 14 exit streams: 8 concentrates and 6 final tailings (Figure 13) and suffered operational difficulties with respect to stable operation.

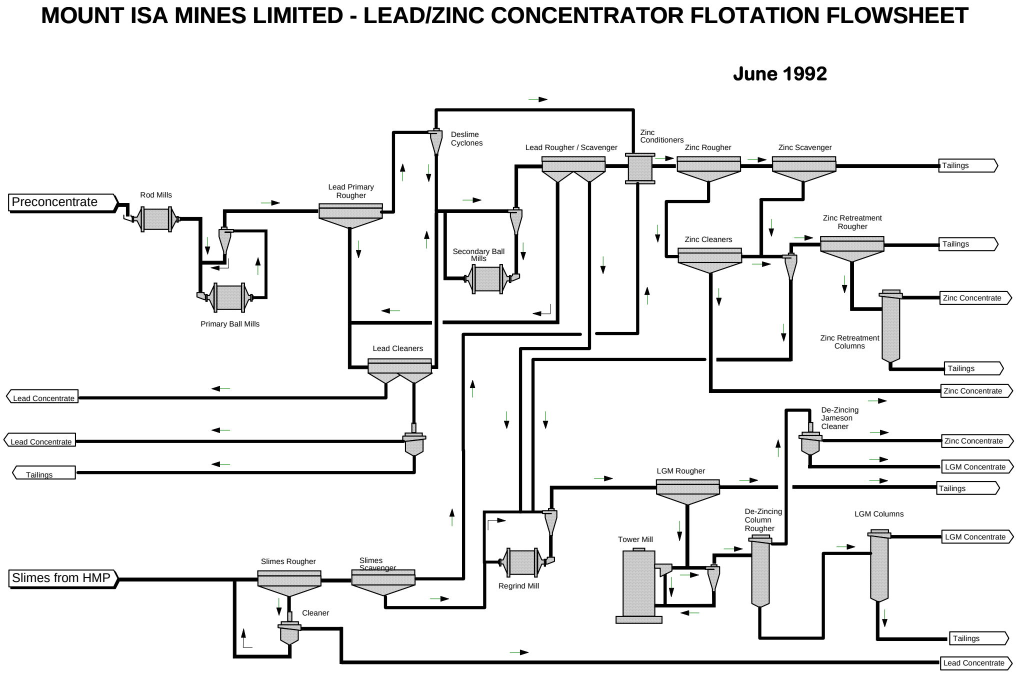


Figure 13. - Flowsheet after the tower mill had been commissioned but before the installation of increased primary grinding and flotation capacity (14 Products).

SOLVING THE UNDERLYING PROBLEMS - THE 1990'S

1990's metallurgical performance improved dramatically as several projects addressed the underlying mineralogical issues :

The projects were;

- Tower mill regrinding installed in the LGM circuit (1991).
- The "Fine Grinding Project," which doubled grinding and flotation capacity and instituted a "cold" lead reverse cleaning circuit (1992). This project addressed both key issues, ie liberation in the zinc circuit and separation (of carbonaceous pyrite) in the lead circuit.
- Improvement in liberation allowed circuit simplification, the increased use of conventional tools (eg high pH zinc circuit cleaning) and relocation of regrinding duties from the LGM circuit to the zinc retreat circuit.
- New ultrafine regrinding technology at Mt Isa was introduced for both zinc and lead regrinding (1994).
- Generally, the application of process control became more effective with the improvements because of more achievable targets.

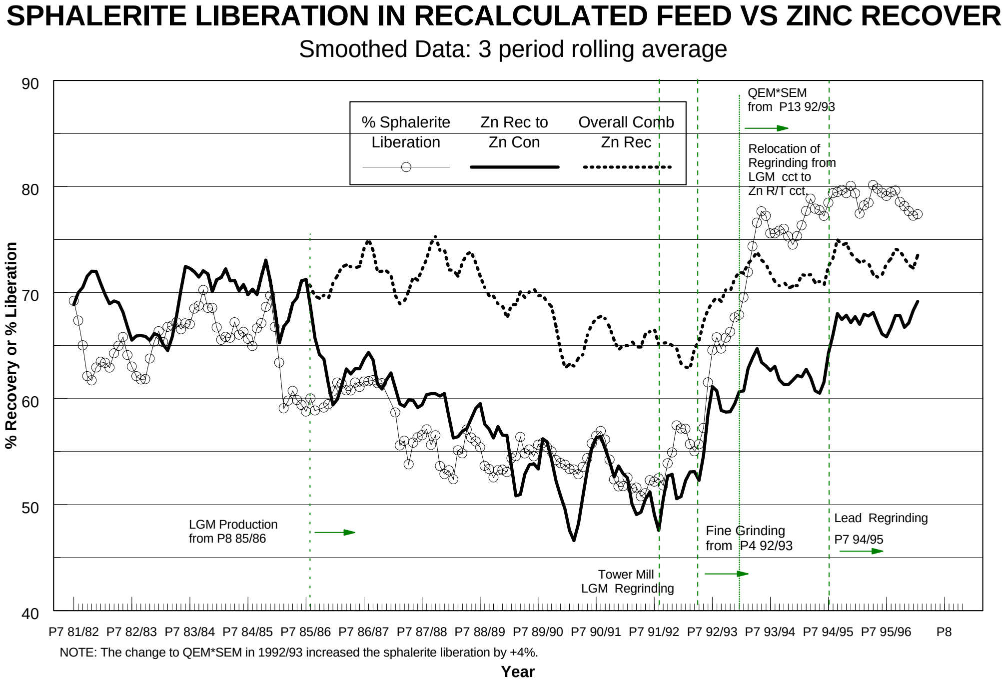
The fine grinding project

While it was well-known prior to 1992 that finer grinding was required, the high capital cost prevented the acquisition of the requisite additional grinding equipment. Conversion of the Copper Concentrator to SAG milling in 1991 solved this problem by releasing two 5m by 6.1m, 2.6 MW ball mills. A project was approved to install these mills for secondary grinding in the Lead/Zinc Concentrator, along with a 520 kW Tower Mill for additional regrinding in the LGM circuit. The finer feed sizing and more dilute pulps necessitated extra flotation capacity, which resulted in the installation of two banks of nine Dorr Oliver DO600 cells for lead secondary roughing and scavenging, and three banks of 12 Dorr Oliver DO600 cells for zinc roughing and scavenging. Existing flotation cells were used to provide additional roughing, retreatment, or cleaning capacity. In summary, the new equipment provided the following changes :

- An increase in primary and secondary grinding power from 6.3 MW to 11.5 MW. Recalculated feed sizing dropped from P80 = 80um P80 = 37um microns;
- increase in regrinding power from 0.75 MW to 1.27 MW. Regrinding size in the LGM circuit dropped to P80 = 12um; and
- a doubling of flotation capacity.

The Tower mill was commissioned in December 1991, and the two secondary mills in October and November 1992. Figure 14 shows the effect on sphalerite liberation and zinc recovery. The key features are:

- Sphalerite liberation increased by 25 per cent by 1993;
- an increase in total zinc recovery of eight per cent. More importantly, the amount of zinc reporting to the low value LGM concentrate was reduced, yielding an increase in zinc recovery to zinc concentrate of over 15 per cent. These recovery gains not only achieved feasibility estimates almost immediately after commissioning, but also quickly exceeded the same estimates.


SPHALERITE LIBERATION IN RECALCULATED FEED VS ZINC RECOVERY

Smoothed Data: 3 period rolling average

Figure 14.

An interesting feature of Figure 14 is that sphalerite liberation exceeded recovery quite significantly. Traditionally, ‘Johnson’s rule of thumb’ stated that combined zinc recovery equalled sphalerite liberation plus 10 per cent, reflecting the amount of diluents that could be tolerated in a zinc concentrate of 50.5 per cent Zn and an LGM concentrate of 34 per cent Zn. All the liberation gains of the Fine Grinding Project were not converted to recovery, since more minerals were now in the difficult to separate size ranges (eg 20 per cent of sphalerite is now less than 4 um). This does not imply that liberation is no longer an issue. Indeed, the pursuit of increasing levels of liberation since the Fine Grinding Project was installed has been a major theme of development. However, it creates an environment where pulp chemistry and flotation separation are now more productive areas of research.

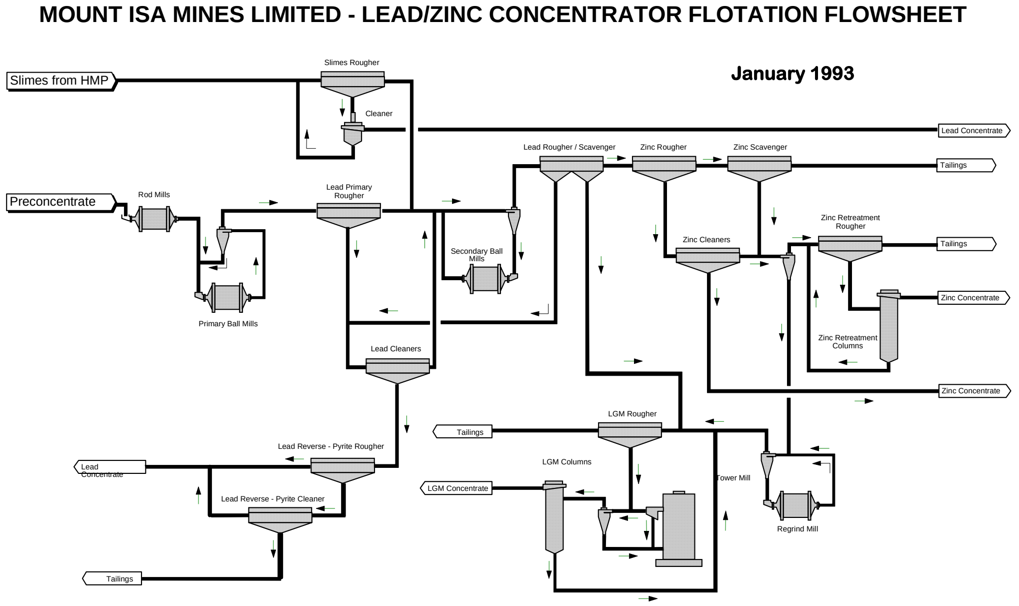


Figure 15. - Flowsheet after the installation of increased grinding and flotation capacity (nine products).

The Fine Grinding Project and Separation Improvements

The ‘cold’ lead reverse cleaning circuit was installed at the same time as the Fine Grinding Project to remove carbonaceous pyrite from lead concentrate. Conventional lead cleaner concentrate is raised to pH 12 with lime to depress galena but not carbonaceous pyrite. A pyrite concentrate is floated, cleaned, and discarded (Figure 15). The pyrite concentrate assays around 30 per cent Fe, 32 per cent S, and 19 per cent Pb. Typically, the reverse cleaning trades off one per cent Pb and 1.2 per cent Ag recovery for each one per cent increase in lead concentrate grade (and accompanying 0.5 per cent lower Fe and 0.3 per cent lower S). The maximum upgrading capacity of the circuit (because of physical constraints) is 4 per cent Pb. Operation of this circuit is intermittent, depending on current ore type, metal prices, and smelter performance. The circuit’s major advantage is the provision of independent control of lead grade/recovery decisions. In the conventional cleaners there is very little ability to trade lead recovery for grade, since the lead cleaner tailing assay has to be kept low to keep galena out of the zinc circuit. A process control system varies the reverse cleaning circuit air addition to control a setpoint lead concentrate grade. This gives the lead smelter a much steadier grade concentrate, while minimising the recovery loss of galena. The circuit is shut down when better ores are encountered. The improved lead concentrate quality resulting from this circuit (by decreasing the iron sulphides) is shown in Figure 16.

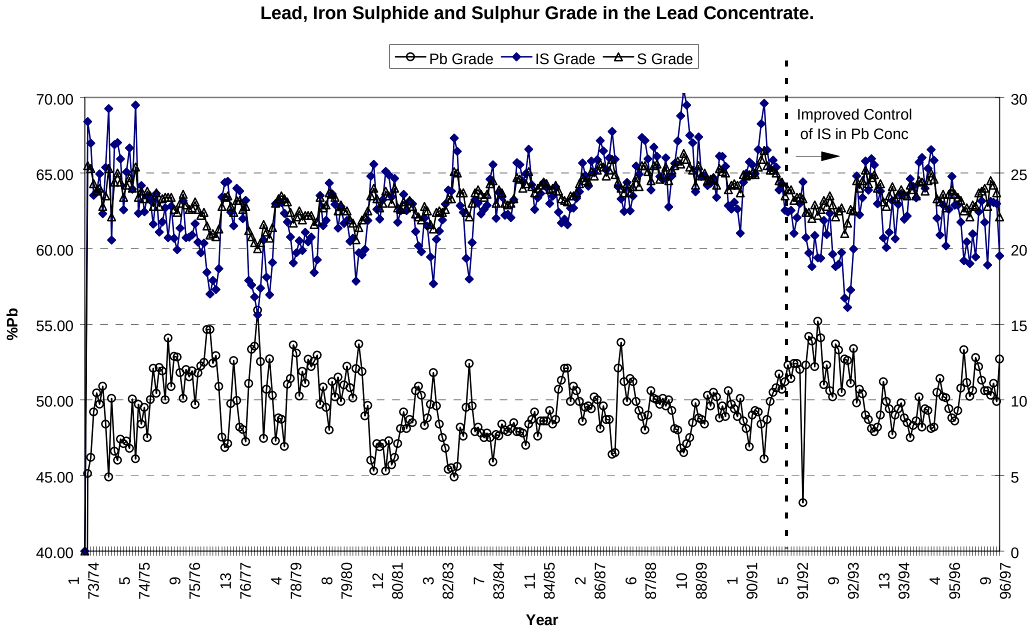


Figure 16.

FURTHER IMPROVEMENTS

In the two years after the implementation of the Fine Grinding Project, further performance gains were made as the circuit was adjusted and simplified. Effectively, operating personnel had to "unlearn" many of the circuit rules essential when poor liberation and insufficient flotation capacity were the root of many problems. The most significant of these were:

- Reintroduction of high pH zinc cleaning using lime. This had been abandoned prior to the Fine Grinding Project because of unmanageable circulating loads of composites.
- Relocation of some LGM circuit regrinding capacity into the zinc circuit (Figure 17). Together with the reintroduction of lime, this helped increase recovery to zinc concentrate by a further 5 per cent.

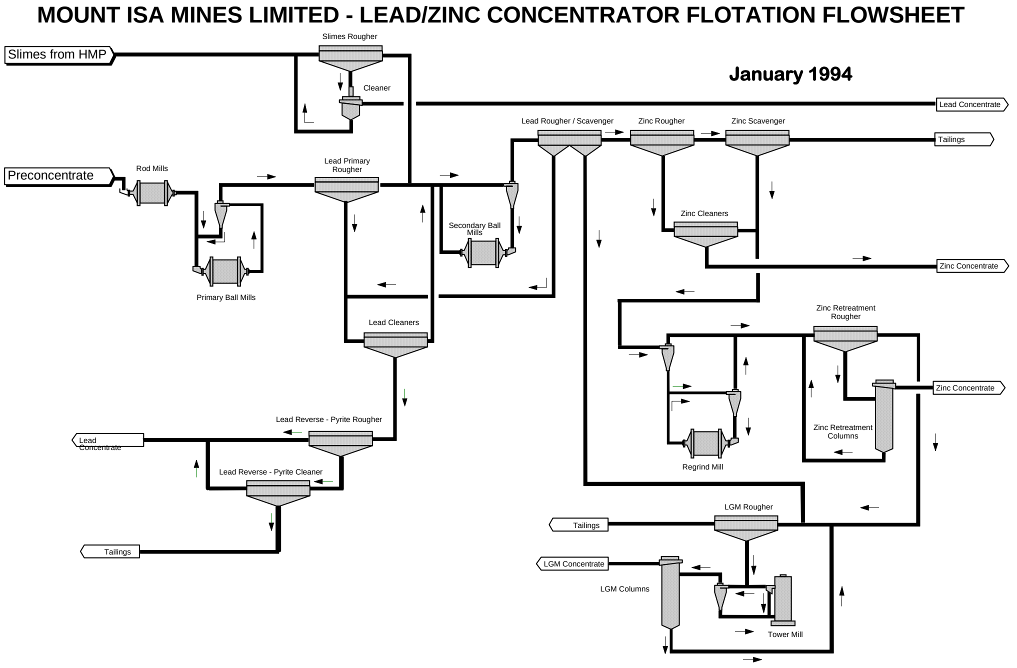


Figure 17. - Flowsheet after the relocation of regrinding capacity. (eight products).

- Use of some fresh water, instead of process (recycle) water, as dilution water in both zinc and lead cleaning. This reduced the impact of salt deposition (especially gypsum) on fine minerals surfaces. It also reduced the lime requirement in zinc cleaning as the slurry was previously supersaturated in calcium, and helped reduce frothing problems in both lead and zinc cleaning.
- Reintroduction of basic process control loops. Enormous efforts in advanced process control had previously yielded little gain, as the process was inherently unstable. Supervisory loops have been gradually introduced as the circuit has been simplified and stabilised. Tonnage/size/load grinding loops are used by operators over 85 per cent of the time on all grinding lines, and 18 flotation loops are used 70 per cent of the time. Tonnage based feed forward reagent ratio controllers are used for cyanide, copper sulphate and xanthate additions in the lead and zinc circuits. Adaptive controllers are used in the cleaners, adjusting both air and xanthate additions.

New Regrinding Technology

- Mount Isa Mines Limited developed revolutionary new ultrafine grinding technology for the McArthur River deposit, with prototypes developed in the Lead/Zinc Concentrator. The circuit has had two 1.1 MW mills regrinding lead concentrate, since 1995 (Figure 18) (Enderle et al, 1997). These mills have further increased liberation and recovery and simplified the circuit. The lead regrinding mills increased zinc recovery by 5 per cent by liberating sphalerite from composites that previously reported to lead concentrate. These mills regrind lead rougher concentrate to P80 = 15 um.
- The regrinding mills also eliminated the bleed stream of difficult lead middling particles to the LGM circuit, leaving the zinc treatment as the only remaining feed to the LGM circuit.

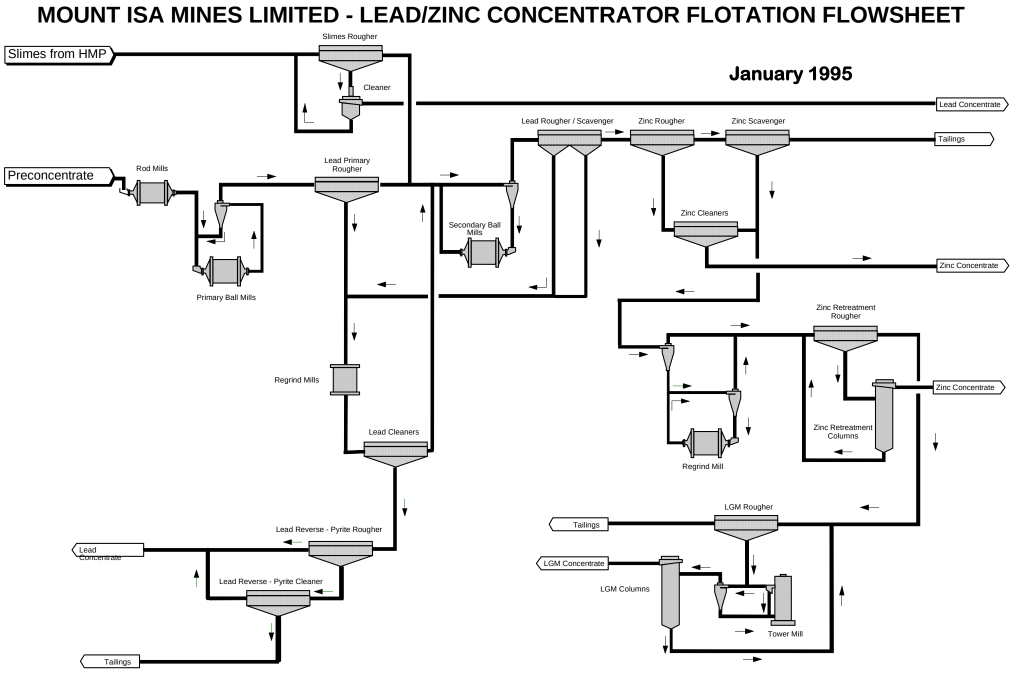


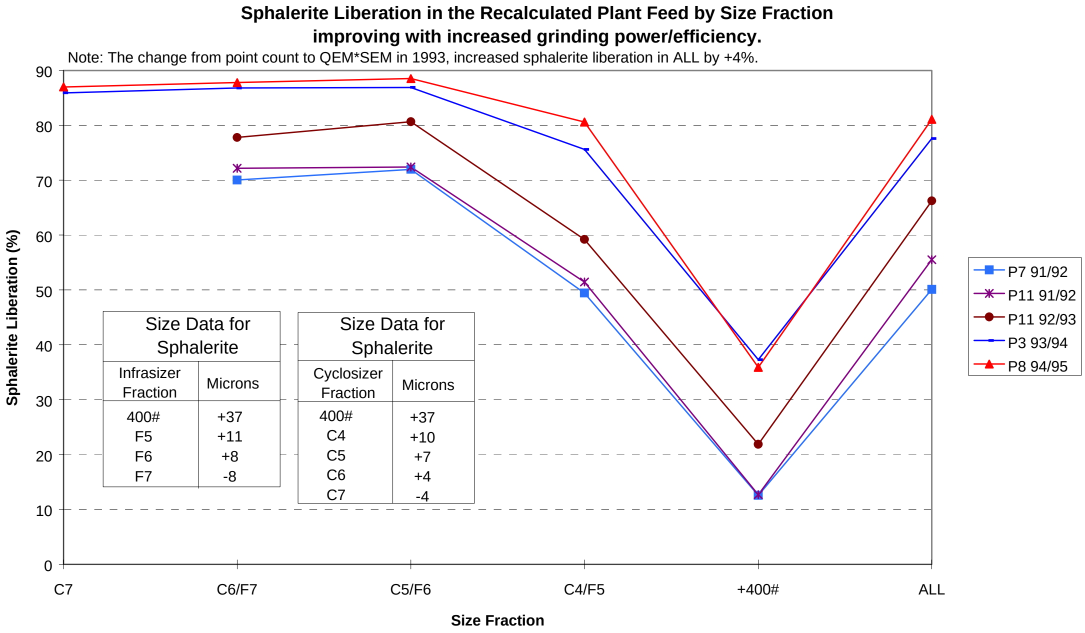
Figure 18 - Flowsheet after the installation of regrinding of lead rougher concentrate.

Size by size analysis

The increase in sphalerite liberation in the recalculated plant feed is shown on a sized basis in Figure 19 for selected months from 1991-1995. Figure 19 shows that liberation increased across all size fractions for each project which increased grinding power or grinding efficiency. The liberation increased in four main stages (Figure 14):

- Installation of Tower Mill in LGM circuit,
- installation of increased primary and secondary grinding power during the Fine Grinding Project,
- relocation of grinding power from LGM circuit to zinc circuit, and
- installation of lead rougher concentrate regrinding.

The zinc recovery by size data increased in a similar manner to the liberation by size data (Figure 20). The figure shows that all size fractions were more liberated with finer grinding and not just the coarse size fractions. The liberation was therefore improved by two methods; firstly by increasing the liberation of each size fraction and secondly, and more importantly, by moving particles from the coarse, less liberated size fraction to the finer, more liberated size fractions.



Size Data for Sphalerite	
Infrasizer Fraction	Microns
400#	+37
F5	+11
F6	+8
F7	-8

Size Data for Sphalerite	
Cyclosizer Fraction	Microns
400#	+37
C4	+10
C5	+7
C6	+4
C7	-4

Figure 19.

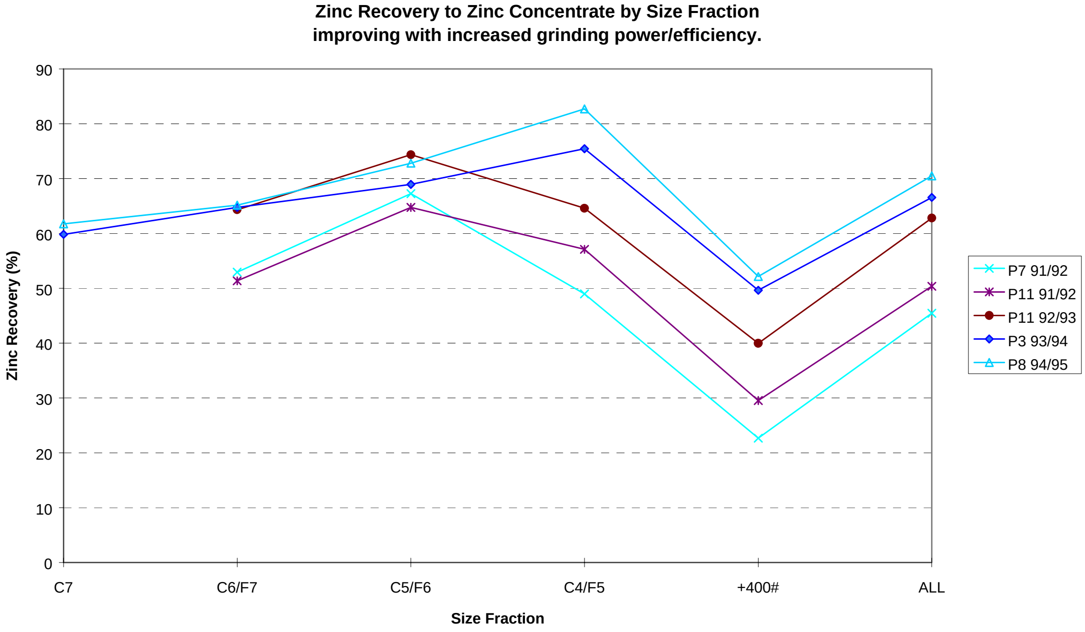


Figure 20.

CONCLUSION

Adoption of a rigorous, size-by-size mineralogical approach to plant operations was crucial to identify and solve the dramatic decline in ore quality and metallurgical performance.

The result was a 20 per cent increase in zinc recovery to zinc concentrate, 5 per cent increase in lead recovery to lead concentrate, improved quality for both lead and zinc concentrates, and 70 per cent reduction in the production of the low value LGM concentrate.

Important also is the simplification of the circuit. From 14 exit streams in 1992, the circuit had eight exit streams by 1995. This produced a dramatic improvement in circuit stability and increased ease of circuit operation. Three main indicators of circuit stability are:

- the willingness of operators to use simple process control loops to assist their decisions;
- the speed of achieving stability after plant start ups, ie. metallurgical results on start-up shifts are now indistinguishable from normal operating shifts; and
- plant spillage and hygiene. High side rubber boots are no longer issued, nor needed !

This case study is an excellent example of the benefits of applying a scientific approach to routine operations over a long period of time.

POSTSCRIPT - RECENT CHANGES

The metallurgical improvements described in this paper were driven by technology changes targeted at the fundamental nature of our fine grained, complex ore. The changes were highly successful and economically essential to business as ore quality declined. However, the improvements came at a price - high capital and operating cost. In 1996, the next improvement came from a comprehensive examination of the mine/mill/smelter business. This led to elimination of the LGM (bulk) concentrate and increased lead and zinc concentrate grades and recoveries, as well as prodviding considerable circuit simplification. These changes were achieved without capital and without extra operating cost. They will be the subject of future publications.

ACKNOWLEDGMENT

The consent of Mount Isa Mines Limited to the publication of the paper is gratefully acknowledged. The authors wish to thank the professional engineers, foreman, operators, maintainers and postgraduate and vacation students who were concerned with the research, development, commissioning and operation of the plant and with data collection and analysis for their contributions during the 1980’s and 1990’s.

REFERENCES

Bartrum, J, Dobrowolski, H J and Schache, I S, 1977. Developments in milling of silver-lead-zinc ores in the Mount Isa area since 1970. Lead-Zinc Update 1977, AIME, St Louis, pp157-182.

Challen, R H, Draper, N, Hoffmann, C W and Menzies, R B, 1968. The Design, Construction and Commissioning of No.2 Concentrator, Mount Isa Mines Limited. Proc AusIMM, No 226: 109-130.

Cunningham, K T, 1953. Milling at Mount Isa. Proc AusIMM, No 171: 307-340.

Davey, J M and Slaughter, P J, 1970. Changes in Lead-Zinc Flotation Practice at Mount Isa Mines Limited, 1955-1970. AIME World Symposium on Mining and Metallurgy of Lead and Zinc, St. Louis, Vol. 1, Chapter 22: 401-427.

Enderle, U, Woodall, P, Duffy, M and Johnson, N W, 1997. Stirred mill technology for regrinding McArthur River and Mount Isa zinc/lead ores. XX International Mineral Processing Congress, Aachen, Germany (In Press).

Espinosa-Gomez, R, Johnson, N W, Pease, J D and Munro, P D, 1989. The commissioning of the first flotation columns at Mount Isa Mines Limited. Proc Int Symp Process Complex Ore., Halifax., CIM, pp293-302.

Espinosa-Gomez, R and Johnson, N W, 1991. Technical Experiences with Conventional Columns at Mount Isa Mines Ltd. Proc Column ‘91., Vol. 2, pp511-523.

Fiedler, K F , Munro, P D and Pease, J D, 1984. Commissioning and operation of the 800 tph heavy medium cyclone plant at Mount Isa Mines Limited. The AusIMM Conference Darwin NT, pp259-271.

Gottlieb, P, Adair, B J I and Wilkie, G J, 1994. QEM*SEM Liberation Indices for grinding, Classification and Flotation. AusIMM Fifth Mill Operators Conference, Roxby Downs, pp5-13.

Grano, S R, Ralston, J and Johnson, N W, 1988. Characterisation and treatment of heavy medium plant slimes in the Mt. Isa Mines Lead/Zinc Concentrator. Min Eng, Vol 1, No 2, pp137-150.

Johnson, N W, Wickham, P and Watsford, R M S, 1982. Analysis of Lead/Zinc Concentrator Flotation Performance at Mount Isa Mines Limited. AusIMM Second Mill Operators Conference, Mount Isa, pp313-326.

Johnson, N W, 1992. Acquisition of Metallurgical Data for Particles Smaller than Ten Microns. 1992 AMIRA Annual Technical Meeting, Adelaide.

Kruttschnitt, J, Jacobsen, L K and Gross, K B, 1939. Milling Methods and Costs at the Mount Isa Mines, Ltd., Mount Isa, Queensland, Australia. Proc AusIMM, No 116: 413-458.

Munro, P D, 1993. Lead-Zinc-Silver ore concentration practice at the Lead/Zinc Concentrator, Mount Isa Mines Limited, Mount Isa, Queensland. Proc AusIMM, 208, pp17-21.

Riley, J F and McKay, W J, 1976, Textures in Lead-Zinc-Silver Ores of N.W. Queensland and the Northern Territory. MIM internal memo, Report No. 3859.

ALTERNATIVE PATHWAY TO LOWER EMISSIONS

'While current clean coal technologies are based on large-scale centralised plants with major CO_2 capture and storage – the alternative pathway is about smaller, more efficient plants that underpin renewables, and using niche capture as a last step.'

By Dr Louis Wibberley, CSIRO Energy Technology, Newcastle

At the Coal21 Annual Conference in 2007, the author proposed an alternative pathway to low CO_2 emissions electricity with coal as the key fuel. The proposed pathway avoids or minimises many of the issues facing current clean coal technologies focused on CO_2 capture and geological storage (CCS). These include supercritical pulverised coal with post-combustion capture (PCC), gasification with capture, and oxy-pulverised coal with flue gas liquefaction. The current focus on CCS is because improvements in efficiency for conventional coal technology will not give anything like the step reduction in greenhouse gas emissions required. This means that only limited reductions in CO_2 will occur (via efficiency improvements) until around 2025 when CCS is expected to be rolled out on a meaningful scale.

Presently, if a new coal plant were built in Queensland or NSW, there would be a strong preference for it to be a 700MW, dry-cooled, advanced supercritical unit, with provision for future staged CO_2 capture using PCC. With capture at the current stage of development, this technology would give a delivered cost of electricity (including capture, transmission and distribution penalties), of around $125/MWh, and with a fuel cycle thermal efficiency for delivered electricity of around 30%.

This article considers an alternative technology pathway to that of current clean coal technologies.

EFFICIENCY CHALLENGE FOR CLEAN COAL

The following graph (Figure 1) shows that the delivered thermal efficiency of coal-based generation (i.e. with transmission) has improved at a relatively constant rate over the last 115 years. Although there have been a number of disruptive technologies (the steam turbine, suspension or pulverised fuel (pf) firing, large unified designs, and supercritical steam conditions), when graphed over a longer time interval up to the present, there has been a steady increase in efficiency.

Based on best current estimates from the technology developers, efficiencies for dry-cooled plants under Australian conditions will increase by around 4% points by 2030. However, if CO_2 capture were adopted, there will be an efficiency penalty of around 8% points – which will effectively wipe out 50 years of efficiency improvements.

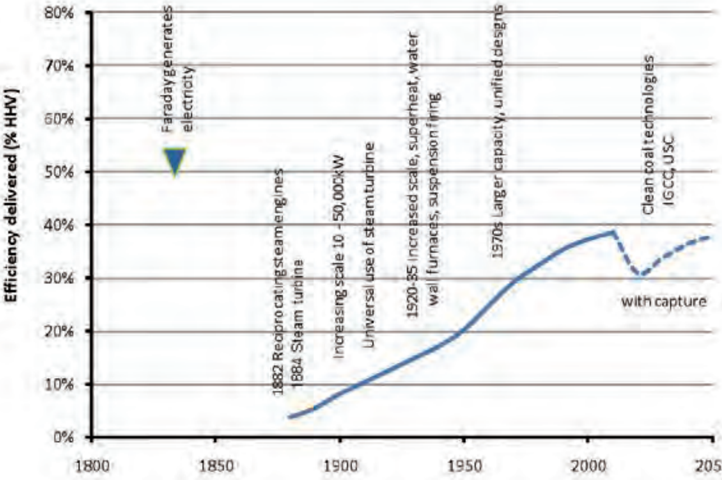


Figure 1 *Thermal efficiency of coal fired power generation over time*

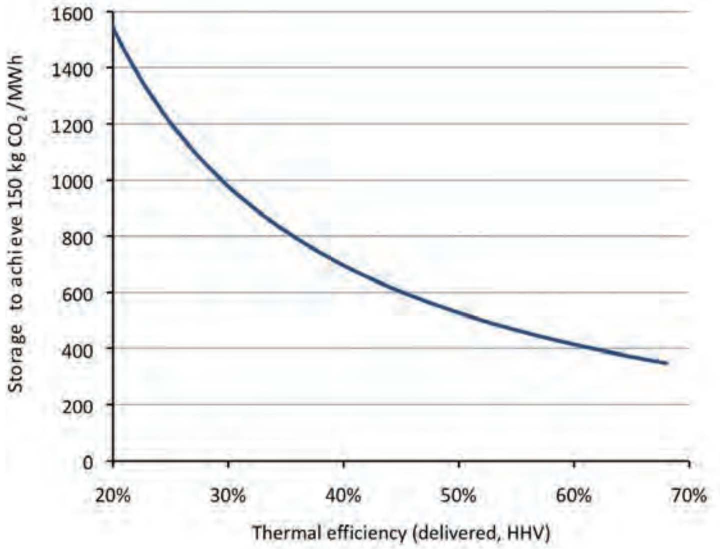


Figure 2 *CO_2 storage to achieve 150kg/MWh vs delivered thermal efficiency of the coal fired generation plant*

ENERGY PENALTY INCREASES STORAGE

The important point is not the efficiency number itself, but the proportional increase in the amount of CO_2 that will need to be stored as efficiency is reduced: if the overall efficiency of the system is reduced, then more CO_2 is produced and more must be captured, which increases the energy penalty for capture and further lowers the overall efficiency. This results in a disproportionate increase in the amount of CO_2 to be captured and stored to meet a given emissions intensity.

Figure 2 shows the amount of CO_2 to be stored to meet an emissions intensity of 150 kg/MWh versus the effective thermal efficiency (delivered electricity including capture).

With best current black coal plants, for each MWh delivered, around 1000 kg of CO_2 will need to be stored. This increases to around 1550–1600 kg/MWh for current brown coal plants.

Continued page 6 ▸

Continued from page 5

Making a current versus future technology comparison, if the delivered efficiency were increased to 45%, then the amount of CO_2 to be stored would decrease to 600 kg/MWh (a 40% reduction for black coal, and a 62% reduction for brown coal). If a delivered efficiency of 65% could be achieved, this would reduce the amount of storage required for black coal by 62%, and by around 76% for brown coal.

These higher delivered efficiencies would markedly reduce the aquifer volume required for CCS. As an example, to store the CO_2 from NSW's current coal plants would require injecting liquid CO_2 into around 8000 m^3 of reservoir pore volume every hour. Using nominal Department of Primary Industries values for aquifer porosity and the effectiveness factor, this means that around 1 000 000 m^3 of reservoir aquifer will be filled every hour. The higher efficiency value would reduce this to around 380 000 m^3 per hour for black coal.

As the cost of CO_2 mitigation per unit of power will also decrease in the same proportion, anything which can increase the system efficiency, and reduce the amount of CO_2 required to be stored, will be valued much more highly in the future.

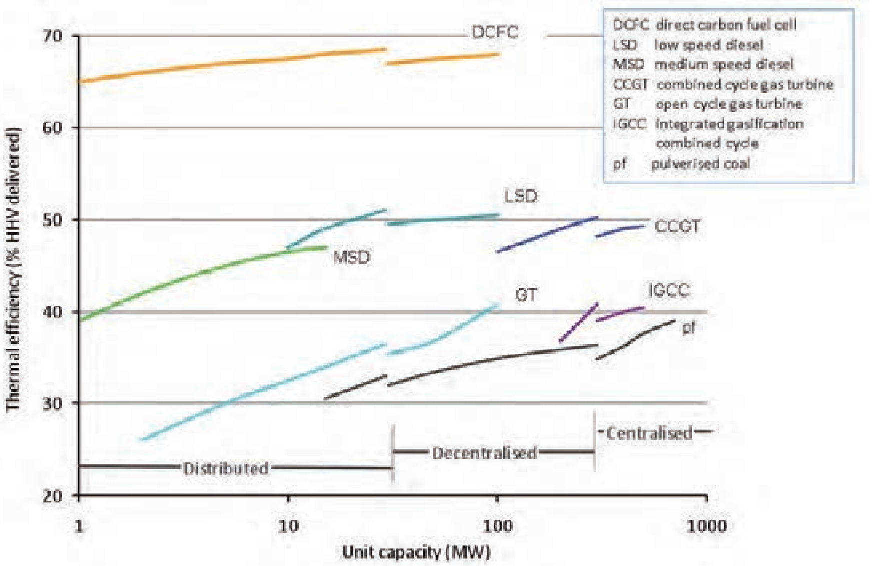


Figure 3 *Delivered efficiency vs unit capacity of a range of technologies*

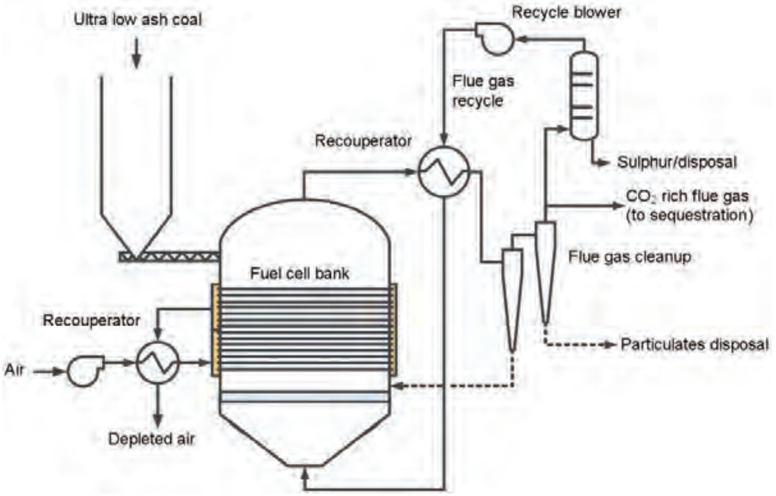


Figure 4 *Example schematic for a direct carbon fuel cell generator*

THE ALTERNATIVE PATHWAY

The alternative pathway being proposed by CSIRO has different priorities to those of the current clean coal approach:

- Achieving the highest possible efficiency for coal-based generation at smaller scale,
- Using the advantages of higher efficiency, smaller coal plants to underpin a high penetration of renewables, thereby giving more cost-effective renewables and a proportional reduction in CO_2 intensity of the system, and
- Using niche (rather than major) CO_2 capture and storage as the last step in meeting emissions targets.

So while current clean coal technologies are based on large-scale centralised plants with major CO_2 capture and storage – the alternative pathway is about smaller, more efficient plants that underpin renewables, and using niche capture as a last step.

MAXIMISING THERMAL EFFICIENCY

Figure 3 compares the thermal efficiency (from delivered fuel to delivered electricity – without capture) for a range of technologies. Size-for-size, the most efficient means of converting fuel energy to electricity is by using large diesel engines or direct carbon fuel cells (not yet commercially available, though the high efficiency has been demonstrated in the laboratory).

Coal could achieve a much higher efficiency if these technologies could somehow be used, enabling a delivered efficiency of around 50% from a direct injection coal engine (DICE) in the short-medium term, and at least 65% in the longer-term future from the direct carbon fuel cell (DCFC). Both could be achieved at a smaller scale than for conventional coal technologies – an attribute that could give a number of other important advantages associated with more decentralised electricity generation, including a reduction in transmission losses (electricity transmission losses are almost always significantly higher than those from the transport of fuel).

Although DICE and the DCFC require ultra low ash coals (say below 2% ash), technologies already exist to produce suitable fuels from coals (including from Victorian brown coals).

DIRECT INJECTION COAL ENGINE

Although never commercialised, the coal engine is not new, being the subject of a number of development programs over the last century (with key programs every 20 years approximately). The early work was led by Diesel and then co-worker Padowski, with engines running for many years on everything from lignite dust to coke. However, the most comprehensive test program was undertaken by the US Department of Energy (DOE) over the period 1978 to 1992, mostly for transportation applications. This included adapting and testing engines ranging from a 2MW single cylinder test engine (90 rpm), a 2MW locomotive (1000 rpm), and an 800 kW haul truck engine (1900 rpm). The most successful demonstrations used ultra low ash coal water fuels which were injected using modified conventional direct injection (i.e. solid injection) systems. Although the technical issues were overcome for the direct injection coal engine, with combustion efficiencies of over 99% being achieved at up to 1900 rpm, and with thermal efficiency equal to diesel fuel, the program was eventually terminated due to persistently low oil prices and before a commercial engine was developed.

WHAT'S CHANGED?

In revisiting this technology, it is important to consider a number of different drivers: the impending cost of CO_2 abatement, cooling water availability, the need to support a step increase in renewables, energy security and changes to the structure of the electricity supply industry.

DICE would have a number of advantages, including being implemented with smaller capital steps, and would be suitable for baseload, peaking and providing grid security or ancillary services. DICE would also be capable of cofiring of biomass fuels (char, crude bio-oils and algae soups), which could enable biomass to be utilised at double the efficiency of current biomass plants. Also, as waste heat is at sufficient temperature to provide the low temperature heat energy needed for CO_2 capture, this could be added with a smaller cost and energy penalty than for current clean coal technologies (instead of converting this energy to electricity with very small and inefficient turbo machinery).

In addition, there have been a number of key technology changes, including developments in coal cleaning, micronising, engine technology/materials/size and costs (about the same as for pf), which are all likely to further increase the viability of the coal engine over that in previous development programs, and reduce the time required for adoption of the technology. Currently, under the European HERCULES Project, new engine technologies are being developed which will increase the efficiency of large engines to over 60% (LHV, mech).

DIRECT CARBON FUEL CELL

The ultimate in thermal efficiency from coal could be achieved by using ultra low ash coals in direct carbon fuel cells. The DCFC works similarly to the hydrogen fuel cell, but is superior in terms of thermal efficiency, with over 75% being achieved in US laboratories for a range of carbon types, including a sample of ultra low ash coal produced by UCC Energy. There are a number of possible configurations being developed. An example schematic arrangement is shown in Figure 4, which comprises a solid oxide fuel cell bank, and a flue gas split (capture is not required) for CO_2 sequestration.

The DCFC is being commercialised for small applications (e.g. the US Army are funding the development of a unit to convert ration pack wrappers to electricity for field use), and current projections from US developers are that it will be available for large-scale commercial electricity production from around 2030. Development is also expected to piggy-back on technologies being developed for the hydrogen fuel cell.

From a system perspective, the DCFC would be highly advantageous, as the technology would be suitable for baseload and peaking duties, is highly suitable for biomass char, and as it produces essentially pure CO_2, it would have a smaller efficiency and cost penalty for CO_2 capture and storage. In the alternative pathway, it is envisaged that DICE power plants would ultimately be converted to DCFC, say post 2030.

In Australia, the DCFC is being researched for coal applications by the CSIRO, and the Universities of Newcastle and Queensland.

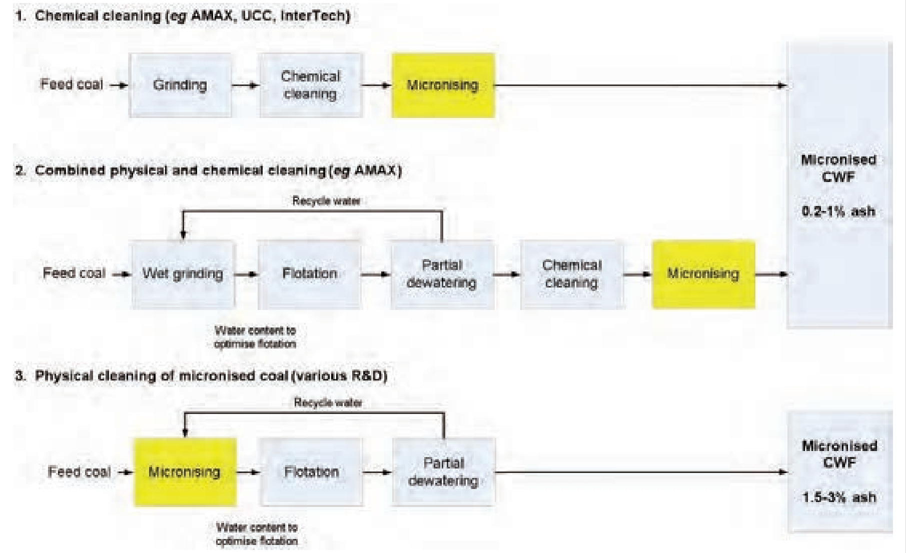


Figure 5 *Main coal cleaning processes (3) for production of engine grade CWF from bituminous coal*

ULTRA LOW ASH COAL

The alternative pathway depends entirely on cost-effective production of ultra low ash coals. Although the cleaner the better, detailed coal specifications for both large diesel engines and the DCFC remain unclear. The DOE work in the 1980s and early 1990s concluded that up to 2% ash is suitable for DICE, and this could be produced from chemical and physical coal cleaning processes.

The DOE studies showed that the most suitable method for fuel delivery was direct injection of micronised coal water fuel or CWF (so drying is not required after cleaning). Depending on engine size and operating speed, the micronised coal should have a top size of 20–30 μm, and a coal concentration of around 45%. This gives a CWF viscosity of around 300–500 mPa.s (at 100/s) which is sufficient to enable effective atomisation.

Although there are a number of methods of utilising CWF in the DCFC (including integrated flash drying and partial gasification combinations), it is expected that a similar quality coal would be required.

There are three main process types for cleaning coal and producing an engine grade CWF from bituminous coals, as shown in Figure 5. These involve either removing mineral matter from the coal, or the coal from the mineral matter. None of these processes are currently commercial, but have been operated at up to pilot scale.

CHEMICALLY CLEANING LOW ASH FEED COAL

The first process involves chemically cleaning a relatively low ash feed coal, producing a coal water slurry, and then micronising to produce an engine grade CWF. Examples are the AMAX and UCC processes which use a combination of caustic and acid treatments to dissolve away the ash and InterTech which does a similar thing using hydrofluoric acid.

FINE COAL PROCESSING

The second approach uses fine coal processing to remove as much of the ash as possible by conventional flotation, then uses chemical cleaning to remove the remaining ash, followed by slurrying and micronising to produce a CWF for an engine. This was developed by AMAX in the 1980s. However, as there are limitations on how much grinding can be tolerated without compromising the materials handling in the chemical cleaning process, the ability of the pre-cleaning step to reduce ash levels significantly (below that of normal coal preparation techniques) will be highly coal dependent.

ULTRA FINE COAL PROCESSING

The third approach involves slurrying and micronising the coal first (not last), and then using ultra fine flotation or selective agglomeration to remove the ash. Although several previous studies have shown that this process is technically feasible, the results obtained in the past have been highly variable, especially with respect to coal recovery rates. However, recent test work with a range of Australian coals (including tailings) has shown that if coal is ground finely enough, a consistent and low ash product is possible with very high coal recovery. As the micronised raw coal slurry has the appearance of crude oil prior to refining, this process has been termed micronised coal refining, and the product micronised refined coal (MRC).

As there is a trade-off between product ash and (nominal) processing costs (see Figure 6), the most suitable cleaning process will depend on the target application. In particular, there is likely to be a significant increase in production cost with ash contents below around 1.5–2%, because chemical cleaning is then required for most coals. For this reason, chemically cleaned coals at 0.2–1% ash are likely to be cost competitive with natural gas.

Continued page 8 ▶

Continued from page 7

Coal at around 2% ash would be cost competitive with thermal coals used in conventional coal power plants.

ULTRA LOW ASH COAL PROJECTS

Over the last few years there has been a steady increase in efforts to produce ultra low ash coals. By far the most advanced is the process being developed by UCC Energy, which is aiming to demonstrate the production of ultra clean coal (UCC) with ash contents down to 0.2%. This very low level is achieved by chemical cleaning. Although originally developed for gas turbines, UCC Energy is now aiming to optimise the process to produce a higher ash specification fuel for a wide range of diesel engines in applications that could replace natural gas/LNG turbines. It is projected that the revised process will give a step reduction in processing costs and CO_2 emissions from coal processing. A small-scale demonstration is presently being undertaken as part of the Asia Pacific Clean Fossil Fuel program (a component of the Asia-Pacific Partnership on Clean Development and Climate).

MICRONISED REFINED COAL

To compete with black coal pf plants, CSIRO, TUNRA Clean Coal, the University of Newcastle and Xstrata Technology are collaborating to investigate the use of new developments in milling and ultra fine coal cleaning to produce low-cost micronised refined coal (MRC) suitable for very large diesel engines. The process involves micronising run of mine coal to a d90 of <20 µm using an Isamill, and then physical cleaning and partial dewatering using advanced ultra fine coal flotation.

Cost estimates show that MRC could be produced at substantially lower costs than for processes using chemicals, which should enable MRC-DICE technology with the largest engines (mostly likely to be able to cope with higher residual mineral matter and other ash forming material in the fuel) to compete favourably with pf plants for CO_2 costs above $30/t.

To date, very promising results have been achieved using an Isamill to efficiently micronise the coal before flotation using TUNRA or Jameson cells. As an example, the Isamill-TUNRA Clean Coal technology combination has produced:

- 1.6% ash MRC from a 27% ash Queensland coal with a combustibles recovery of 94%,
- 2.8–3% ash MRC from a 14% Hunter Valley coal with a combustibles recovery of 90%, and
- 3% ash MRC from 52% ash Hunter Valley tailings with a combustibles recovery of 85%.

Similar results have been obtained by the University of Newcastle using alternative flotation techniques.

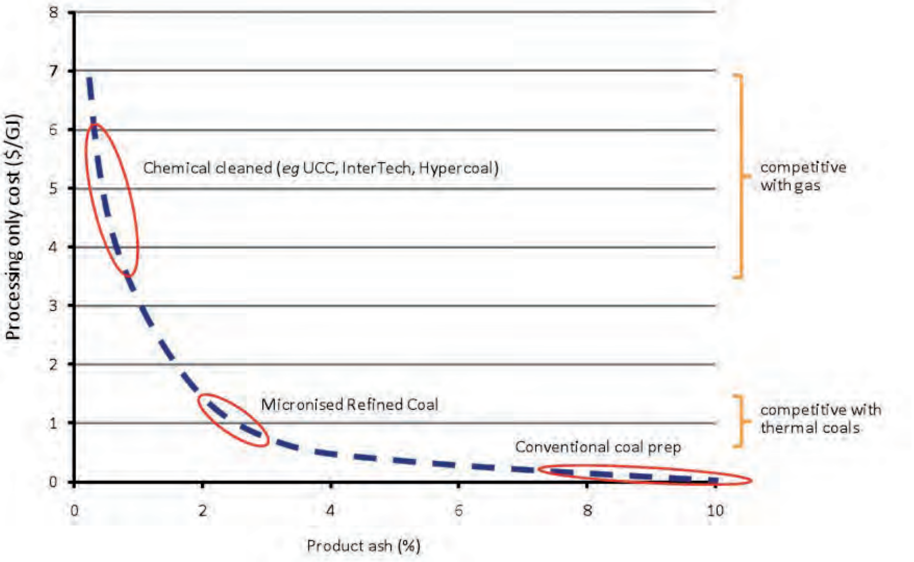


Figure 6 *Nominal processing only cost vs product ash*

Another interesting development is that TUNRA Clean Coal have dewatered MRC to 12% total moisture using selective agglomeration (at very low levels of organic additions). MRC is produced as a dry prill, which would give the option of transporting and storing MRC in solid form.

While most effort is on bituminous coals, Exergen are developing a diesel engine fuel from Victorian brown coal and lignites. As the ash content of these coals is already low, fuel production is focused on producing a suitable CWF by upgrading these coals by hydrothermal treatment.

MICRONISING

Efficient micronising of coal is essential for either coal engine or fuel cell applications. Conventional pf mills are totally unsuitable for producing the ultra fine particle size which is thought to be required for diesel engines (d50 of <10 µm and a top size of 30 µm). Conventional mills are even less suitable when treating higher ash coals. Since the extensive coal engine R&D for the DOE in the 1980s, new large mills have been developed for the minerals industry, routinely milling up to 10 000 tonnes of hard rock per day to sizes below 10 µm (e.g. see photograph of a typical Isamill in Figure 7). Preliminary studies by CSIRO have shown that coal can be micronised by an Isamill suitable for producing engine grade fuel with a projected energy consumption of around 60 kWh/t.

UNDERPINNING RENEWABLES

The second aspect of the alternative pathway is to use smaller high efficiency coal power plants to provide the grid security necessary to enable a higher penetration of renewables. Coal could have a more important role in underpinning the uptake and development of renewable energy, particularly those which are intermittent – wind and solar of course, but also seasonal biofuels. By doing this, coal

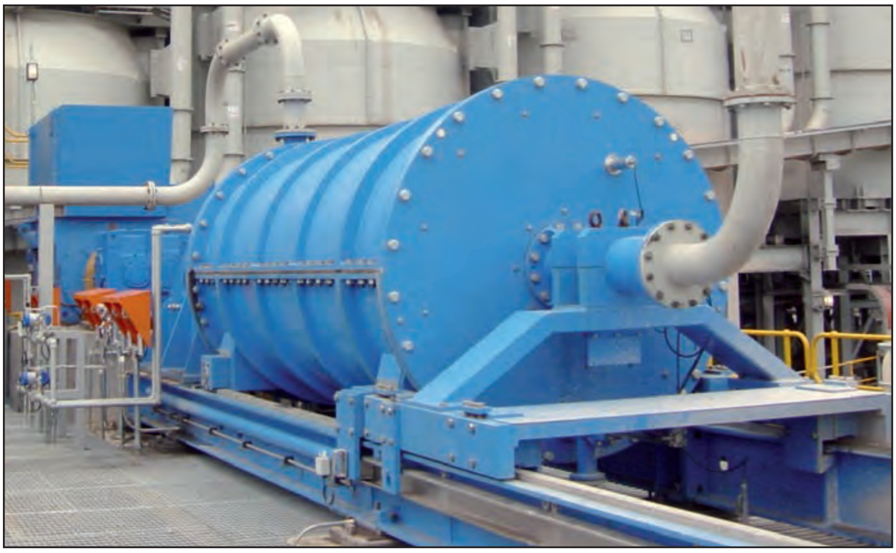


Figure 7 *A 3 MW M10,000 IsaMill™ operating in base metal regrind duty*

effectively reduces its own CO_2 intensity, and supports the transition to more sustainable generation.

Done effectively, this avoids the costs and inefficiency of energy storage or of providing the ancillary services required for a secure grid.

There are at least three ways that coal plants can do this – providing longer-term backup, shorter-term spinning reserve, and efficient utilisation of biofuels.

While the backup and spinning reserve roles are well known, to do this efficiently, without increased CO_2 emissions (especially with a higher penetration of renewables), requires technology which can give higher part-load efficiency and short start times. As large diesel engines have excellent part-load efficiency and also relatively short start times, DICE-based power plants should be better than pf and gas plants for this role.

The ability of DICE to efficiently utilise other biofuels is also likely to greatly assist the development of renewables. These could range from micronised chars, through a wide range of crude bio-oils, to soups of high lipid content algae. Relative to gas turbines, diesel engines are extremely tolerant to alkalis in the fuel, and compared to pf plants have only a marginal penalty for fuel containing up to 60% water. While these attributes are also shared with gasification-based technologies, DICE could do this with a much higher efficiency (8–15 percentage points depending on unit capacity) and remain effective at smaller scale.

CAPTURE AND STORAGE

Even with high efficiency and a high penetration of renewables, some CCS may eventually be required. As yet there appears to have been no R&D specifically undertaken for capture from a coal engine or a direct carbon fuel cell. However, as the coal engine exhaust will have properties between that of pf boilers and NG gas turbines, it should not pose any special difficulty for post-combustion capture. As engines reject their heat at a temperature which is usable for CO_2 stripping, this will greatly reduce the energy penalty for capture – especially if an integrated capture process is used to replace the small turbo machinery normally used to recoup extra exhaust energy.

Capture from a DCFC would be even easier, as the flue gas produced would be similar to that from oxy-pf. Capture would not be required, only dehumidification and liquefaction, giving an energy penalty only one-third that for oxy-pf.

It is likely that either technology has the potential to halve the energy penalty for capture, thereby further reducing the amount of CO_2 that has to be stored.

IMPLICATIONS FOR CO_2 INTENSITY

The diagram in Figure 8 shows the likely reduction in CO_2 intensity of delivered electricity for the current approach based on clean coal technologies with CCS, and that proposed for the alternative pathway. Although both pathways remain unproven, both could achieve a CO_2 intensity of around 100 kg/MWh into the future.

With clean coal technologies only a small reduction in intensity is likely until the rollout of major CCS – which by current projections is unlikely before 2025–2030.

As the alternative pathway can be implemented at smaller unit capacity (60–100 MW), it is proposed that this should result in smaller capital and risk hurdles which could facilitate earlier implementation. This would start with using ultra low ash coals into large diesel engines from 2015 giving a 20% reduction in CO_2 for new plants. These plants could be used to give both direct and indirect support to a high penetration of renewables (say 20%), providing a cumulative reduction in CO_2 intensity of around 40%. DICE plants would be replaced by DCFC/niche CCS as required post 2030 to meet future emission targets, and with a 400% reduction in the amount of CCS required. It is proposed that additional cumulative CO_2 savings could accrue from earlier implementation.

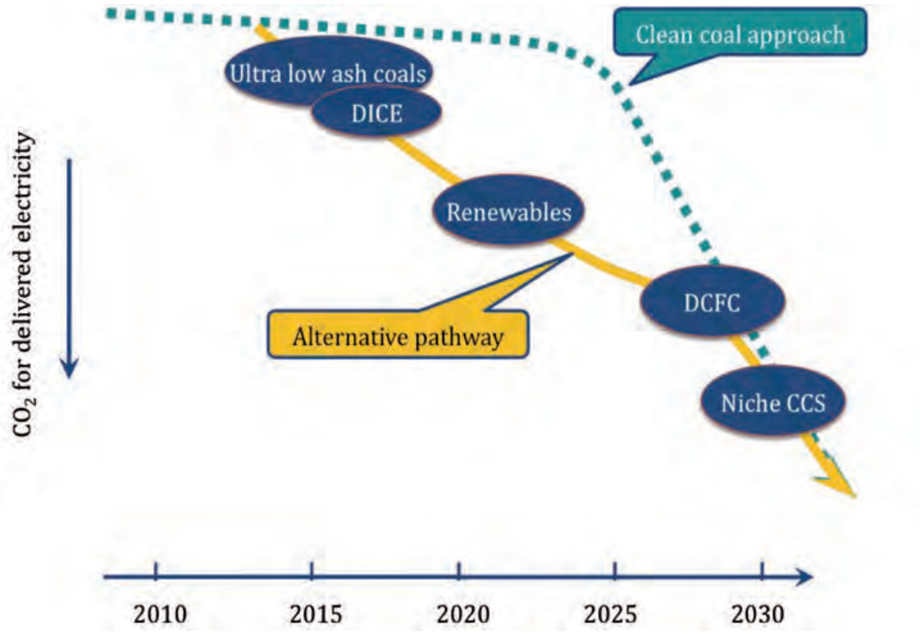


Figure 8 *Possible future CO_2 for delivered electricity*

WHERE TO FROM HERE

It is acknowledged that the proposed alternative pathway needs considerable development and demonstration to match the level of technical development of the clean coal technologies. Despite this, the proposed pathway has strong technical merit because of the ability to carry out a near-commercial scale demonstration at a relatively small scale. This greatly reduces development and commercialisation risk hurdles. In addition, while the overall approach is novel, most of the component technologies are based on adapting commercial and mature processes (and many are Australian). An additional feature is the high degree of flexibility and adaptability of all of the technologies.

A number of proposals are under development to establish small-scale demonstration projects in Australia. These mostly involve a two-stage program starting with pilot testing of fuel production at around 1 t/day for use in a 500 kW pilot engine (or several cylinders of a larger engine), together with engineering for a larger demonstration plant. This would be followed by a 10 000 hour demonstration plant with a 10 MW power plant. Even at this small scale, the demonstration plant would exceed the sent out thermal efficiency of Australia's best pf plants. In addition, this would be sufficient to scale up (through both multiples and scale) to a 100 MW power plant within 5 years.

As the most costly component of the demonstration is expected to be the power plant (around $25M for 10MW), there is considerable scope for cost-sharing this facility between a number of ultra low ash coal projects. Hopefully this could be achieved with strong interest from the large engine manufacturers.

In the meantime, CSIRO, partners and several industry groups continue to undertake research on how best to prepare coal water fuels from different coals/different cleaning technologies, and to better understand coal-engine interactions. Several Chinese groups are also working in this area.

JAMESON CELL 30 YEARS

The Use of the Jameson Cell to Improve Flotation Circuit Design

Rodrigo Araya
Glencore Technology Canada
Virginia Lawson
Glencore Technology Australia

Abstract

Mineral processing plants are facing an urgent need for more efficient technologies to produce saleable concentrate as the ore bodies are becoming more complex. This complexity typically translates in changes to the configuration of existing flotation circuits to be able to maintain productivity. The Jameson Cell is a high performance flotation cell able to recover mineral particles in a wide range of sizes due to a combination of small bubble size and intense mixing ideal for bubble-particle collisions. This paper presents case studies where the Jameson Cell has been successfully retrofitted in brownfield mineral processing plants transforming the existing flotation circuits into more efficient flowsheets.

Biography

Rodrigo Araya is a chemical engineer graduated from Universidad Católica del Norte, Antofagasta, Chile. He obtained a Master of Engineering degree from McGill University in Canada. Following graduation, he worked as a researcher at McGill University in the Mineral Processing Group, studying hydrodynamic properties of industrial flotation machines. He has worked on mineral processing operations including Taseko Gibraltar Mines, Teck Red Dog and Canadian Royalties Nunavik Nickel. Currently he works at Glencore Technology focused on flotation, grinding and atmospheric leaching technologies.

Virginia Lawson has 25 years experience in mineral processing in operations and technical management. Her operational experience covers precious and base metals in Australia and Canada. The last five years have been spent focused on applying her expertise to solving problems in clients' plants around the world. Virginia manages the global Jameson Cell business for Glencore Technology and enjoys using her engineering skills and interacting with people all over the world to help improve their mineral processing performance.

Introduction

The modern society is increasing its demands for metals to produce goods and infrastructure for an always growing population. It has been reported that the extraction of the main commodities has increased by a factor of 1.7, which accelerates the depletion of mineral resources (Calvo et al., 2016). Consequently, new ore bodies around the world are at deeper depths, present lower head grades, complex mineralogy and high variability often complemented with deleterious elements including arsenic, fluorine, and mercury (Huynh et al., 2014). Complex mineralogy often requires fine and ultrafine grinding to liberate valuable mineral particles. For example, Glencore's McArthur River Mine produces a concentrate at a P80 of 7 μm (Pease et al., 2010). Processing plants using conventional flotation technologies and fine/ultrafine particle size typically struggle due to poor hydrodynamics to recover fine liberated particles (Jameson, 2012). This paper proposes the Jameson Cell as a flotation technology to overcome these challenges. Case studies of industrial installations are presented.

Background

Flotation is a process that separates hydrophilic particles from hydrophobic particles by means of air bubbles that rise to the top of a flotation cell to produce a froth layer (Wills & Finch, 2016). Ideally, hydrophilic particles remain in the pulp phase and hydrophobic particles adhere to air bubbles and exit the flotation cell as concentrate; however, in practice there will always be water entrained as it is required to build the froth phase (Ireland et al., 2007). The froth layer is a system in which gas cells are enclosed by water (Weaire & Hutzler, 2001). The water that forms the froth layer is transported from the top of the pulp phase; therefore, the water in the froth has the same composition as the pulp. This transport of water from the pulp to the froth is the start of the transport of unwanted hydrophilic particles from the froth to the concentrate launder (Johnson, 2005). The negative effects of water entrainment depends on particle size. Relatively coarse hydrophilic particles settle more easily in the froth phase, whereas fine fractions (<10 μm) do not settle and are transported much more successfully through the froth (Johnson, 2005). In most cases, relatively deep froths are enough to produce clean concentrates when coarse hydrophilic gangue is present. The use of wash water is typically needed for fine hydrophilic gangue.

The Jameson Cell

The Jameson Cell is a high intensity flotation cell in which pulp and air are brought together in a co-current descending flow in a vertical tube called the downcomer. The pulp is introduced into the downcomer through a slurry lens orifice to produce a high velocity jet that generates violent mixing and fine bubble size. Due to these conditions, a dense foam is produced inside the downcomer where violent collisions between bubbles and particles occur, therefore allowing the collection of hydrophobic mineral particles (Evans et al., 1995). The pulp then discharges into a separation tank, which allows the air bubbles carrying the hydrophobic particles to raise to the top of the cell to form a froth layer, which is consequently collected in the concentrate launders. Figure 1 shows an illustration of the Jameson Cell. The main advantages of the Jameson Cell includes the very rapid collection leading to compact size; self-aspiration of air eliminating the need of a blower or compressor; froth washing to reduce entrainment of unwanted hydrophilic mineral particles. Another great advantage is that the Jameson Cell recirculates a portion of its tails so the volumetric flow rate to the downcomer is constant, therefore, mixing, bubble size, air flow rate and metallurgical performance are constant.

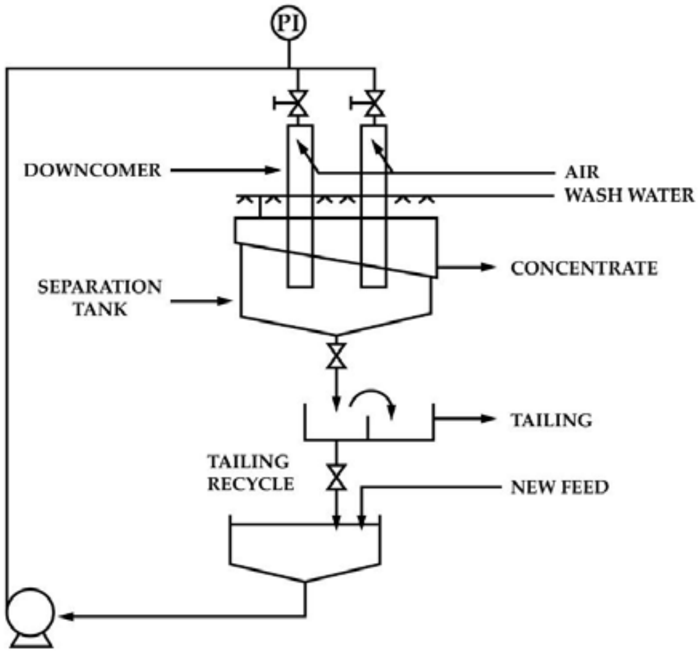


Figure 1 – Illustration of the Jameson Cell

Conventional Cell Cleaning Circuit

Cleaner circuits that use conventional mechanical cells are typically developed based on flotation tests to define the number of cleaning stages required to achieve the desired concentrate grade (Wills & Finch, 2016). There are a number of conventional cleaning circuits that have installed the Jameson Cell at the head of their cleaner circuits for several reasons including, increase cleaning capacity, removal of deleterious elements, increase of final concentrate grade (Bennet et al., 2012; Seaman et al., 2012). These installations serve as industrial demonstrations that the Jameson Cell is equivalent to three stages of cleaning in conventional mechanical cells, therefore implying that more efficient cleaner circuits can be designed with less machines, therefore less footprint and more power efficient.

Improved Cleaner Circuit Design

The proposed cleaner circuit design use the Jameson Cell in two separate duties: cleaner/scalper and recleaner (see Figure 2). The cleaner/scalper Jameson Cell produces a high grade concentrate by recovering the fast floating liberated mineral particles at the head of the cleaner circuit. The tails from the cleaner/scalper proceeds to a bank of mechanical cells as cleaner/scavenger to maximize recovery of a stream now with a lower degree of liberation. The concentrate of the cleaner/scavenger circuit feeds a re-cleaner Jameson Cell that also produces final concentrate. In flowsheets that include regrinding, the performance of the re-cleaner Jameson Cell can be used as a diagnostic tool to determine the grinding power required to liberate the value particles and consequently increase concentrate grade and recovery. Typically in base metals, the Jameson Cell when installed as cleaner/scalper is able to achieve final concentrate grades at unit recoveries ranging from 50 to 70%.

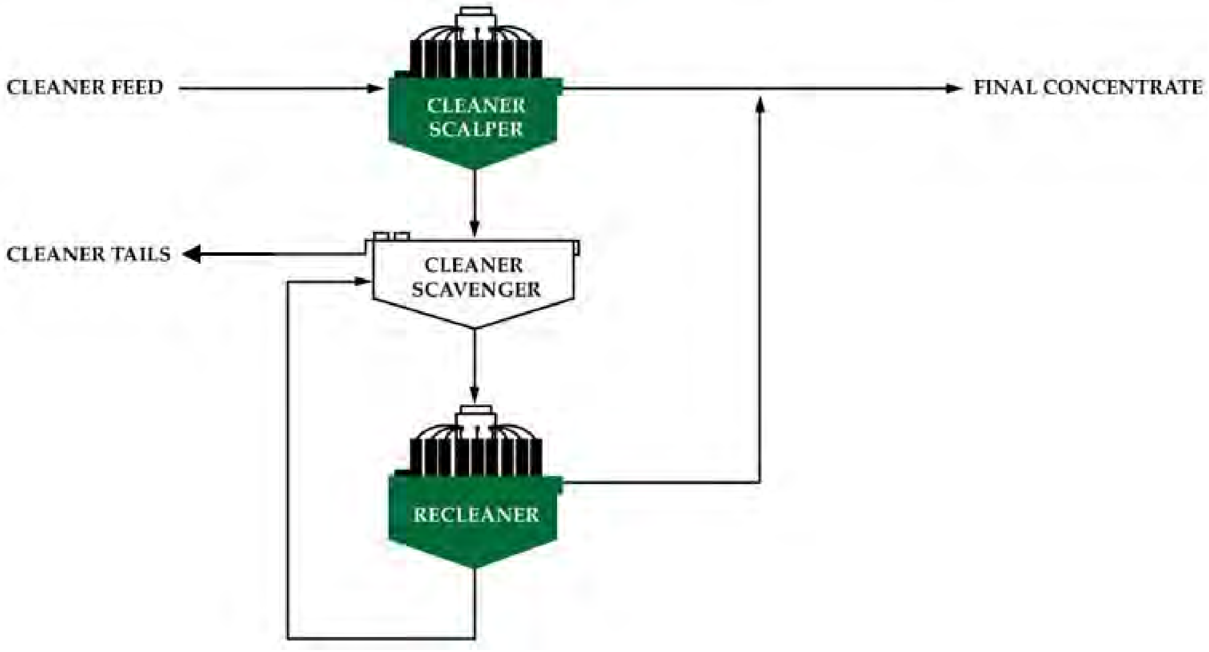


Figure 2 –Cleaner Circuit Design with Jameson Cell as Cleaner/Scalper and Recleaner

Case Study 1: Mount Isa Mines Copper Concentrator

Mount Isa Mines is one of the biggest mining operations in Australia that operates as a subsidiary of Glencore PLC, located near Mount Isa, Queensland. Mount Isa Mines operates two separate mining and processing streams: copper and silver-lead-zinc. The mining and smelting complex produces copper anodes, crude lead bullion (containing silver), and zinc concentrate. The copper operation mine chalcopyrite as the only significant copper mineral that occurs as a replacement deposit in a silica-dolomite host rock. The sulphide gangue consists of pyrite, minor amounts of pyrrhotite and cobaltite. The non-sulphide gangue (NSG) consists of dolomite, chlorite, quartz and talc.

The copper concentrator was built to process chalcopyrite ore and has throughout its history also processed converter slag when mine output allowed, to produce copper concentrate. The concentrator was commissioned in 1973 and by 2015 most of the flotation cells had reached their service life, therefore it was decided to perform an online refurbishment which gave an opportunity to re-design the cleaner circuit. Figure 3 shows the flowsheet before the refurbishment. The cleaner circuit consisted of 33 flotation cells: three 2.5m x 17m column cells, twelve 120 Agitair cleaners, eight 120 Agitair re-cleaners, eight 120 Agitair retreatment and two 100 m^3 Wemco cleaner/scavengers (Lawson et al., 2017).

Figure 4 shows the concentrator flowsheet after the cleaner circuit re-design where it can be seen a dramatic reduction from 33 to 5 flotation cells. The new cleaner circuit consists of one B5400/18 cleaner/scalper Jameson Cell, one B5400/18 recleaner Jameson Cell and three cleaner/scavengers (one B5400/18 Jameson Cell and the two 100 m^3 Wemco). The cleaner/scalper Jameson Cell takes rougher concentrate and produces final concentrate. The cleaner Jameson Cell receives the tails from the cleaner/scalper and the cleaner/scavenger concentrate (Jameson Cell and mechanical cells) to produce final concentrate. The cleaner/scavenger Jameson Cell takes the cleaner Jameson Cell tails plus the reground rougher/scavenger concentrate.

The Jameson Cell units were sized based on the surface area required to remove 5 $t/h/m^2$ and not residence time as is the conventional approach. The Jameson Cell can also be simulated from laboratory flotation tests. Figure 5 shows the comparison of the grade/recovery curve created with the laboratory flotation test and the curve produce with the industrial Jameson Cell after commissioning at Mount Isa Copper Concentrator.

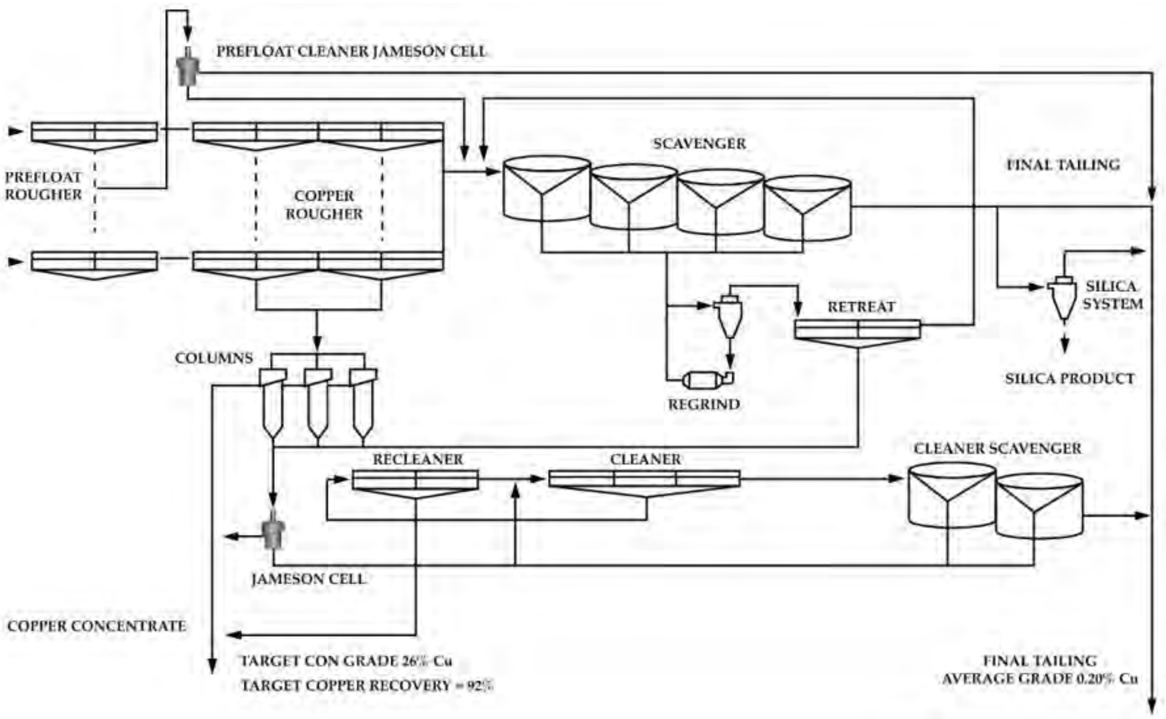


Figure 3 – Mount Isa Concentrator flowsheet before the refurbishment

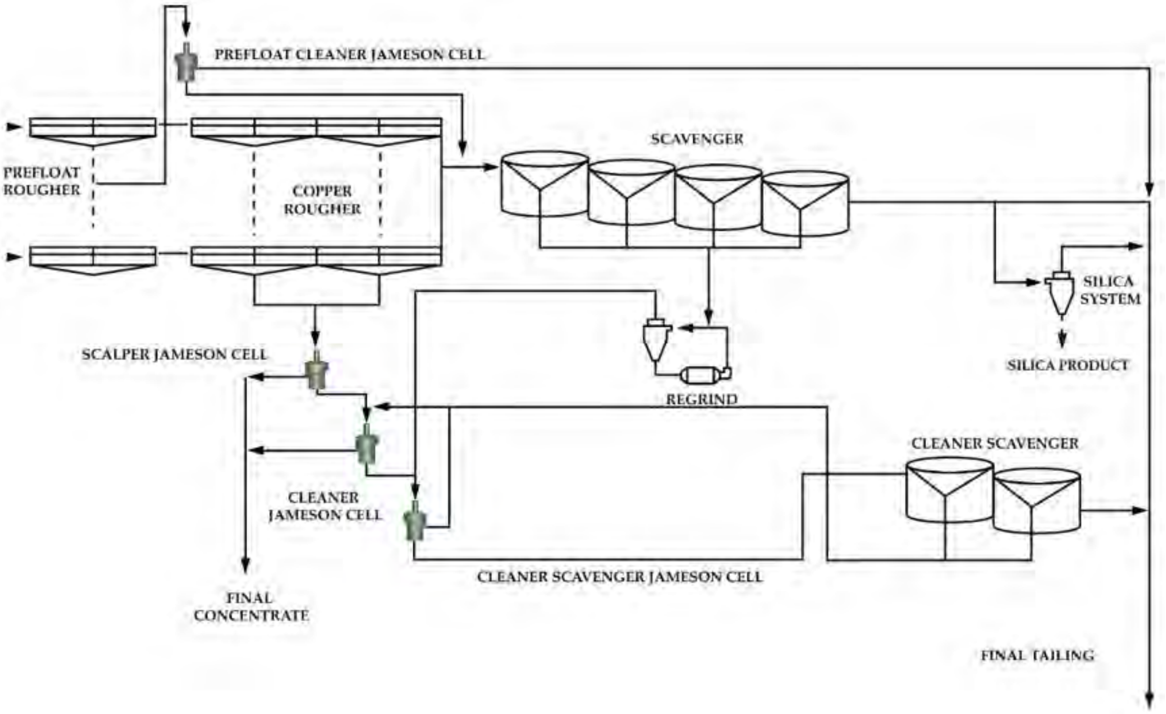


Figure 4 – Mount Isa concentrator flowsheet after the installation of Jameson Cells

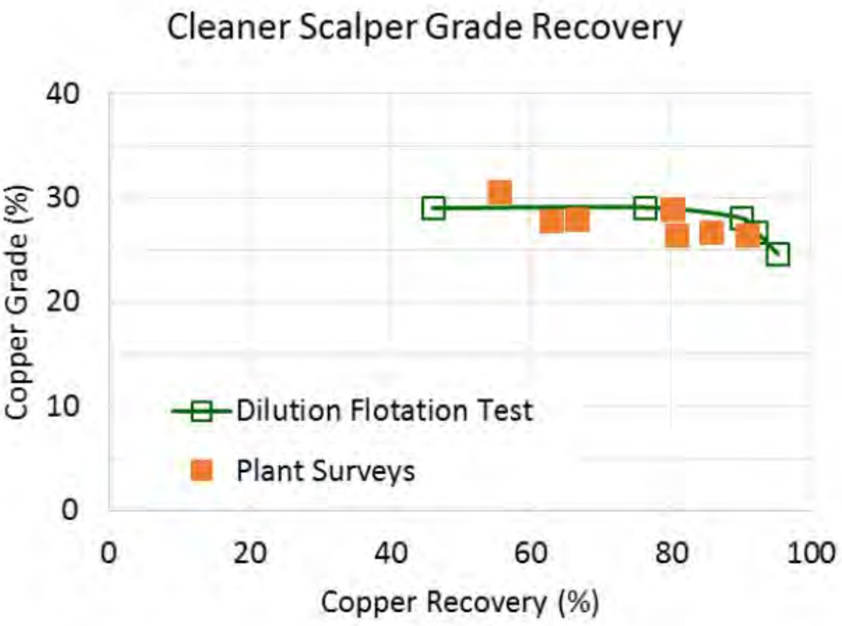


Figure 5 – Comparison between dilution cleaning test and industrial Jameson Cells at Mount Isa copper concentrator

Case Study 2: CSA Cobar

CSA Mine is an underground copper mine owned by Cobar Management Proprietary Limited (CMPL), a wholly owned subsidiary of Glencore PLC. It is located 14 km north of Cobar in New South Wales. The main copper species processed are chalcopyrite and cubanite. It is a high grade copper mine with head grades varying from 5 to 8%. The large variations in feed grade often overwhelmed the flotation process leading to the need to by-pass rougher concentrate directly to final concentrate. In addition to this challenge, the last stage of cleaning that consisted of Denver cells installed in 1965 that needed to be replaced. A plant upgrade project was requested with the ultimate goal of increasing the concentrate grade from 28% Cu at 96% overall recovery to 30% without compromising the overall copper recovery. The flotation circuit before the plant upgrade consisted of two banks of five 500 ft^3 Wemco Cells roughers, two banks of ten 1.8 m^3 Denver Cells as first cleaners, two banks of five 1.8 m^3 as second cleaners. The rougher and cleaner tails were sent to three 30 m^3 Outotec Cells.

Figure 6 shows the flowsheet of the new flotation circuit. The cleaner circuit was re-designed to increase the concentrate grade and to eliminate the need to by-pass the rougher concentrate to final concentrate, therefore making it easier to operate. A cleaner/scalper E4232/10 Jameson Cell was added to produce final concentrate. The second cleaners (ten 1.8 m3 Denver Cells) were replaced with only one E1732/4 Jameson Cell. Figure 7 shows the characteristic grade vs. recovery curves for the cleaner/scalper and re-cleaner Jameson Cell at CSA Mines. It can be seen that the cleaner/scalper is removing most of the mass of copper as final concentrate with unit recoveries greater than 80%, due to collecting the fast-floating-liberated mineral particles. On the other hand, the more challenging grade/recovery curve of the re-cleaner Jameson Cell clearly indicates that it is recovering a greater number of composite particles (Huynh et al., 2014). This new cleaner circuit was capable of producing the target grade and recovery, 30% Cu and 96%, respectively.

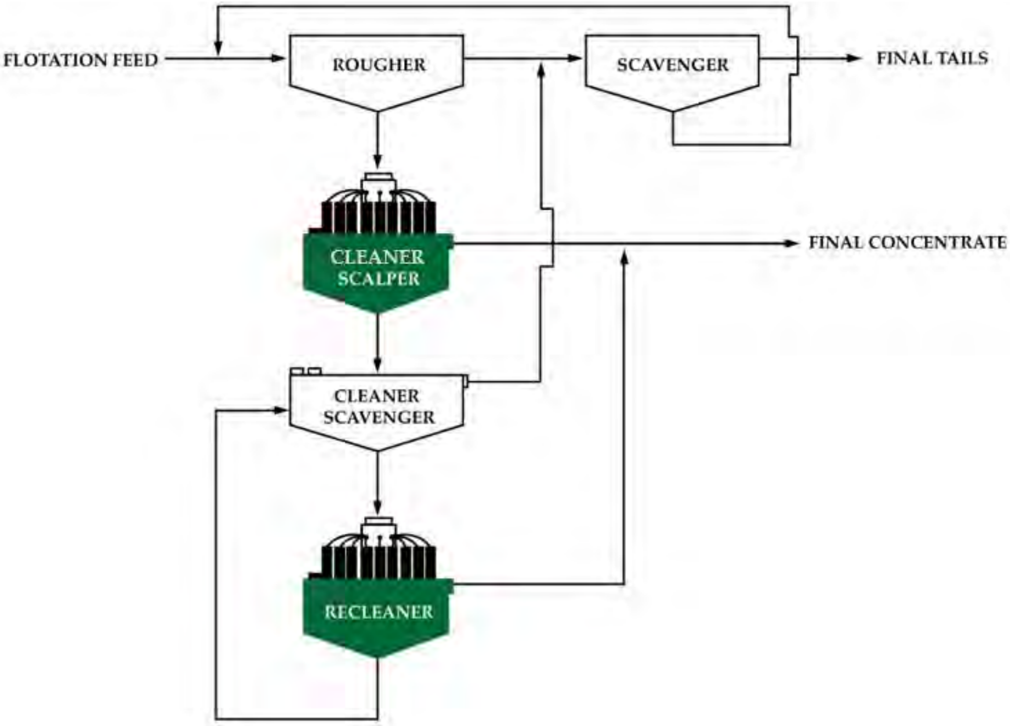


Figure 6 – Flowsheet at CSA Mines with the New Cleaner Circuit Design

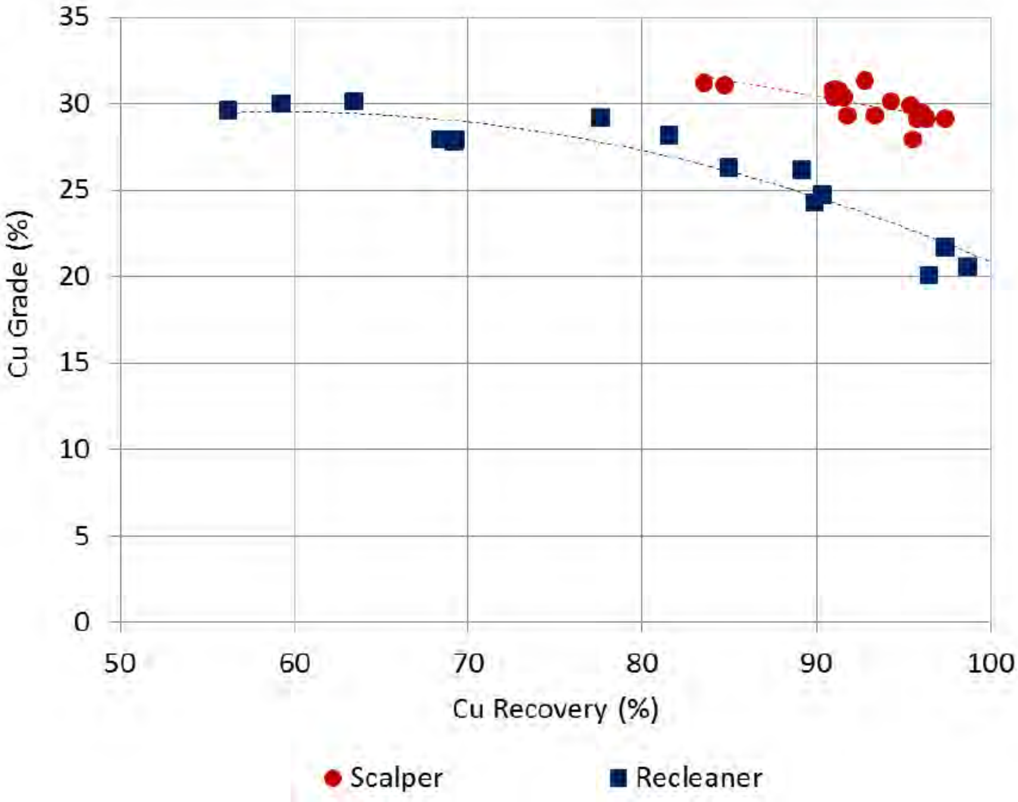


Figure 7 – Grade vs. Recovery Curve of the Two Jameson Cells at CSA Mines

Conclusions

Depletion of world's mineral resources is causing the need to process more complex ore bodies. This challenge is forcing mineral processing plants to adopt more efficient flotation circuits and flotation technologies. An improved cleaner circuit designed was proposed. This circuit is able to achieve ultimate metallurgical targets, great stability in the face of feed variations, reduced plant footprint, and a more robust operation. The Jameson Cell represents a key element in this improved cleaner circuit due to its exceptional metallurgical performance. Case studies were presented in which conventional cleaner circuits were re-designed to implement in industrial scale the improved cleaner circuit with exceptional results. The Mount Isa Mines Copper Concentrator was upgraded resulting in a reduction from 33 flotation cells in the cleaner circuit to only 5 flotation cells without compromising overall concentrate grades and recoveries (Lawson et al., 2017). CSA Mines was able to improve its already spectacular targets from 28% Cu grade at 96% overall recovery to 30% Cu grade at 96% recovery (Huynh et al., 2014).

References

Bennet, D., Crnkovic, I., & Walker, P. (2012). Recent process developments at the Phu Kham copper-gold concentrator, Laos. 11th Mill Operators Conference, (pp. 257-272). Hobart, Tasmania, 29-31 October.

Calvo, G., Mudd, G., Valero, A., & Valero, A. (2016). Decreasing ore grades in global metallic mining: A theoretical issue or a global reality? Resources 5 (4) 36.

Huynh, L., Araya, R., Seaman, D., Harbort, G., & Munro, P. (2014). Improved cleaner circuit design for better performance using the Jameson Cell. 12th AUSIMM Mill Operators' Conference, (pp. 141-152). Townsville.

Ireland, P., Cunningham, R., & Jameson, G. (2007). The behaviour of wash water injected into a froth. International Journal of Mineral Processing, 99-107.

Jameson, G. (2012). The effect of surface liberation and particle size on flotation rate constants. Minerals Engineering 36-38, 132-137.

Johnson, N. W. (2005). A review of the entrainment mechanism and its modelling in industrial flotation processes. Centenary of Flotation Symposyum, (pp. 487-496). Brisbane.

Lawson, V., Muller, M., Radulovic, P., & Wallace, J. (2017). Mt Isa copper concentrator cleaner circuit redesign. Metplant , (pp. 279-287). Perth, WA, 11-12 September.

Pease, J., Young, M., Curry, D., & Johnson, N. (2010). Improving fines recovery by grinding finer. Mineral Processing and Extractive Metallurgy, 216-220.

Seaman, D., Burns, F., Adamson, B., Seaman, B., & Manton, P. (2012). Telfer processing plant upgrade - The implementation of additional cleaning capacity and the regrinding of copper and pyrite concentrates. 11th Mill Operators Conference, (pp. 373-381). Hobart, Tasmania, 29-31 October.

Weaire, D., & Hutzler, S. (2001). The physics of foams. Oxford University Press.

Wills, B. A., & Finch, J. (2016). Will's Mineral Processing Technology. Elsevier.

MOUNT ISA MINES NECESSITY DRIVING INNOVATION

*V. Lawson, H. DeWaal, G. Heferen, N. Aslin, P. Voigt, and M. Hourn

Glencore Technology
Level 10, 160 Ann St
Brisbane, Australia, 4000
*(*Corresponding author: virginia.lawson@glencore.com.au)*

ABSTRACT

Mount Isa Mines (MIM) acquired a reputation for the successful application of R&D to develop break-through technologies for the mining industry starting in the 1978's through until the early 2000's. The ISAPROCESS^{TM} tank-house technology has been licensed to copper refineries throughout the world, and a significant per cent of the world's copper is refined using this technology. Since development in the late 1980's more than 20 ISASMELT^{TM} copper and lead smelting furnaces are now installed in countries around the world. Jameson Cell flotation technology developed jointly by Mount Isa Mines and Professor Graeme Jameson is widely used in the Australian coal mining industry and increasingly in the base-metal and gold industry. The IsaMill^{TM}'s developed at Mount Isa and McArthur River made it possible to develop the McArthur River and George Fisher orebodies and has been successfully implemented into base metal fine grinding applications around the world. The most recent commercialised innovation is the atmospheric leach Albion Process^{TM} with its supersonic HyperSparge^{TM} gas sparger, is being adopted as a solution to the increasing complexity of orebodies.

MIM's contribution to the industry was significant given the size and the remote location of its operations with Townsville Copper Refineries more than 1350 km and Mount Isa 1800 km from the nearest state capital of Brisbane. This paper will briefly discuss the development of each of these technologies and why MIM – now owned by Glencore - was so successful innovating and developing such technologies over a period of nearly 40 years.

KEYWORDS

Innovation, Mount Isa Mines, ISAPROCESS^{TM}, IsaKidd^{TM}, ISASMELT^{TM}, IsaMill^{TM}, Jameson cell, Albion Process^{TM}, HyperSparge^{TM}, ZipaTank^{TM}

INTRODUCTION

Mount Isa is located in the Gulf Country region of Queensland about 1800 kilometers North West of Brisbane (see Figure 1). It came into existence because of the world class mineral deposits found in the area. In 1923 the orebody containing lead, zinc and silver was discovered by the miner John Campbell Miles. Mount Isa Mines Limited (MIM) was founded in 1924 to develop the minerals discovered by Miles, but production did not begin until May 1931. It paid its first dividend in 1947 after 16 years of troubled production. In 1954 the 1100 copper orebody was discovered and with rapidly rising reserves during the 1950's and 1960's led to the construction of new concentrators to treat lead/zinc/silver ores in 1966 (#2 concentrator) and copper ore's in 1973 (#4 concentrator). The difficult nature of the Mount Isa lead-zinc orebodies has meant that the company had always needed to be at the forefront of mining technology. In the 1970's through to the 1990's, it became a world leader in developing new mining techniques and processing technologies as a response to declining metal prices and rising costs. Mount Isa has been smelting copper since 1953 and lead since the early 1930's. Copper Refining at Mount Isa's fully owned subsidiary of Copper Refineries Proprietary Limited (CRL) had commenced operations in 1959.

Figure 1 – Location of Mount Isa and Townsville relative to Brisbane the nearest Capital City

Technologies to come out of Mount Isa include the ISAPROCESS$^{\text{TM}}$ copper refining technology, the ISASMELT$^{\text{TM}}$, The Jameson Cell, the IsaMill$^{\text{TM}}$, the Albion Process$^{\text{TM}}$ and the Hypersparge$^{\text{TM}}$. Mount Isa Mines Ltd was acquired by Xstrata in 2003 and Xstrata was then merged with Glencore in 2015. The level of innovation achieved at Mount Isa Mines is unsurpassed and was the result of the difficult nature of the Mount Isa ore bodies and its response to declining metal prices and rising operational costs in the 1970's and 1980's. By the 1990's, Mount Isa had become a world leader in innovative mining techniques and state of the art processing technologies. The processing technologies are discussed below.

INNOVATIONS

Each of the innovations developed at Mount Isa Mines had a driver but the overarching desire was to make technology more efficient and cost effective. Each of these process developments will be discussed separately.

ISAPROCESS^{TM}

The development of the ISAPROCESS^{TM} tank house technology had its beginning in the zinc industry. During the mid-1970s, MIM was considering building a zinc refinery in Townsville to treat the zinc concentrate produced by its Mount Isa operations. As a result, MIM staff visited the zinc smelters using the best-practice technology and found that modern electrolytic zinc smelters had adopted permanent cathode plate and mechanised stripping technology. MIM realised that the copper refineries performance was constrained by the conventional practice of copper starter sheets. The preparation of these copper starter sheets was labour intensive and the overall cycle was several weeks in duration.

MIM initiated a research program aimed at developing similar permanent cathode technology for copper refining. CRL, a subsidiary of MIM, had been operating in Townsville since 1959, using conventional starter-sheet technology and treating blister copper produced in the copper smelter at Mount Isa. Permanent cathode technology was developed and adapted over many years of in-plant experimental work and successfully introduced to the Townsville refinery in 1978. . The fundamental difference between the new ISAPROCESS^{TM} and the conventional starter sheet technology is the use of a permanent reusable cathode blank instead of a non-reuseable copper starter sheet and the introduction of mechanised and automated electrode handling machines replacing labour-intensive manual operations. The vertical edges had plastic strips and the bottom cased in wax to prevent copper cathode from growing around the edges of the cathode plate during, stripping and allowing two separate copper sheets from each cathode plate. This technology led to major advances in the electrode handling systems and automation in copper tank houses. The improved geometry of the cathode plates and the significantly shorter cathode cycle times allowed for increased intensity and efficiency of the refining process. Introduction of permanent cathode technology resulted in higher capacity, better copper cathode quality with less defects, safer operation and a four-fold improvement in productivity. Considerable development work was required to modify the original stripping machines from their zinc cathode origins due to the heavier cathodes. The stripping capacity of the machines has increased from 250 plates per hour to 600 plates per hour in the latest designs. More recent developments include the elimination of wax masking from the cathode plate, robotic electrode handling machines, and the introduction Duplex Stainless Steel cathode plates giving greater durability and corrosion resistance. Through the use of ISAPROCESS^{TM} user forums, to exchange ideas and developments in the technology and to share operational experiences, the technology has enjoyed continued improvement with higher productivity and improved quality at low cost.

Figure 2 – The IsaKidd process

In mid 1981 Falconbridge Limited commissioned a copper smelter near Timmins to treat concentrate from its Kidd Mine. The original copper cathode produced at Kidd suffered from the presence of higher concentrations of lead and selenium and could not meet customer specifications. It was determined that the use of copper starter sheets was preventing the Kidd refinery from meeting its cathode quality targets. Testwork began with the use of permanent stainless steel cathodes after preliminary tests showed a significant reduction in deleterious elements. The Kidd Process cathode used a solid copper header bar welded onto stainless steel resulting in a lower voltage drop than the ISAPROCESS™. Falconbridge began marketing the Kidd Process technology in 1992 providing competition between the two suppliers of permanent cathode technology. Between 1992 and 2006, 25 Kidd technology licenses were sold and 52 ISAPROCESS™ licenses.

The development of the ISAPROCESS™ and Kidd Process set the scene for a run of technology developments that continued until the mid 2000's. Xstrata took over MIM in 2003 and then Falconbridge in 2006. The Kidd Process technology consequently became part of the tank house package and together they have since been marketed as IsaKidd™ representing the dual heritage of the technology. The current robotic stripping machine (Figure 2) is based on over 30 years of copper refining and winning technology. Today over 100 licensees are using IsaKidd™ technology.

ISASMELT™

The sinter plant/blast furnace combination was the dominant technology for lead smelting throughout the 20th century. In the early 1970's companies using this technology came under sustained political and economic pressure as tighter environmental regulations were introduced, and energy costs increased, leading to higher capital and operating costs (Fewings 1988). It was in this environment that Mount Isa Mines sought a process that would improve the performance of the operations at their lead smelter in Mount Isa. After investigating the various processes under development, researchers turned their attention to the Sirosmelt lance. It had recently been developed on a laboratory scale at the Commonwealth Scientific and Industrial Research Organisation (CSIRO) in Melbourne. Following initial investigations Mount Isa Mines recognised the potential of the novel concept for smelting of lead concentrates and embarked on an extensive development program.

In 1978 a joint project was initiated between Mount Isa Mines and CSIRO to investigate the application of the Sirosmelt submerged-combustion technology to the smelting of Mount Isa lead concentrates. The ISASMELT™ process, as it became known, was developed to maturity for smelting copper, nickel, lead and zinc feeds by Mount Isa Mines through the 1980's and 1990's using incremental scale up. Commercialization only occurred once the process had been proven on laboratory, pilot and demonstration scale over many years. Approximately ten years were required for development of the lead and copper ISASMELT™ from crucible to demonstration scale (refer to Figure 4). During this decade the core know-how that was accumulated enabled the development team to reach the point where they were much better equipped to design and construct a full scale commercial plant – the final stage of the scale up process. Key aspects in this process were the selection of the scale up factors and the systematic design, development and re-engineering of several components of the technology. Figure 3 shows a comparison for the scale up stages for the lead and copper ISASMELT™ processes. Pilot scale was defined as unity for scale up comparison.

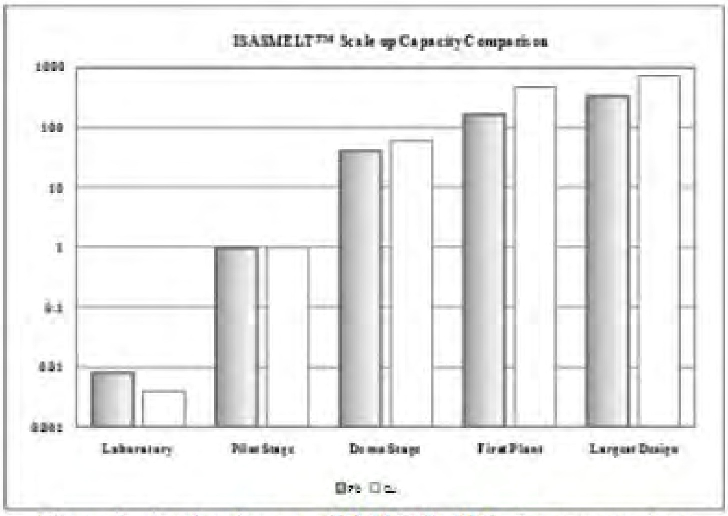


Figure 3 – Lead and copper ISASMELT™ Scale up comparison

During the scale up process, refer to Table 1, several aspects of the technology were developed to a high standard that allowed the ISASMELT™ technology to become a commercial success. As a result, ISASMELT™ technology now operates successfully at numerous plants around the world. The methodical approach to development of the technology has allowed owners to modernise their existing operations or create new businesses with significantly reduced technical risk.

An important parameter in the evolution of the ISASMELT™ technology has been the refractory campaign life. Figure 5 shows the history of the refractory campaigns at the commercial copper ISASMELT™ plant at Mount Isa since commissioning. At the time Mount Isa Mines management considered the installation of water cooling on the furnace refractories undesirable because of the potential for fatal incidents and increased operating costs. As a result the commercial scale furnaces were constructed with minimal water cooling. Although this led to shorter campaign lives initially, a development program was begun that focussed on optimising refractory materials selection and installation methodology. When coupled with process control strategies and continuous on-line monitoring of the bath temperature using systems developed over more than 10 years of operation, it allowed Mount Isa Mines to achieve campaign lives of more than 3 years without using any water cooling of the furnace refractories.

Figure 4 – Tapping matte from the copper ISASMELT at Kazzinc

Table 1 – Key Indicators of ISASMELT™ Plants from pilot to commercial scale

Topic	Unit	Pilot Scale		Demo Scale		First Full Scale		Current Design[1]	
		Pb	Cu	Pb	Cu	Pb[3]	Cu	Pb	Cu
Furnace ID	m	0.4	0.4	1.8	2.3	2.5	3.75	3.6	4.4
Lance Diameter	mm	38	38	150	250	250	350	250	500
Lance Control	-	Manual		Semi Automatic		Semi Automatic		Automatic	
Oxygen Enrichment	%	21	21	21	28	35	45	70	90
Nominal Feed Rate	tph	0.12	0.25	5	15	20	101	40	183
Offgas Treatment	-	Flue System / Baghouse		Gas cooler/ Baghouse		WHB		WHB[2]	

Notes:

ID: Internal Diameter; WHB: Waste Heat Boiler

[1] Refers to maximum throughput

[2] Some of the plants use a combination of radiation section and evaporative cooler for offgas treatment

[3] Refers to the smelting furnace from the two stage lead ISASMELT™ process

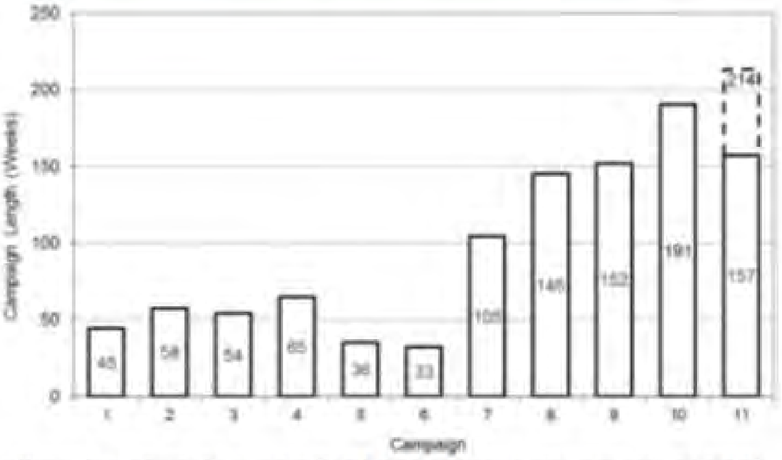


Figure 5 – Mount Isa copper ISASMELT™ plant campaigns (as of 2013)

Jameson Cell

The Jameson Cell (Figure 6) was jointly developed by Mount Isa Mines and Laureate Professor Graeme J Jameson (AO) of the University of Newcastle. Mt Isa had commenced operations with conventional flotation cells but was installing columns in cleaning duties in the mid 1980's. The columns had the benefit of froth washing that was likely to allow significant grade benefits in the very fine lead-zinc circuit. The first observations of the columns was that the collection process was slow necessitating long residence times and large volumes which remains a limitation of columns even today. In 1985 Professor Jameson was commissioned to undertake a project to improve the column sparger design.

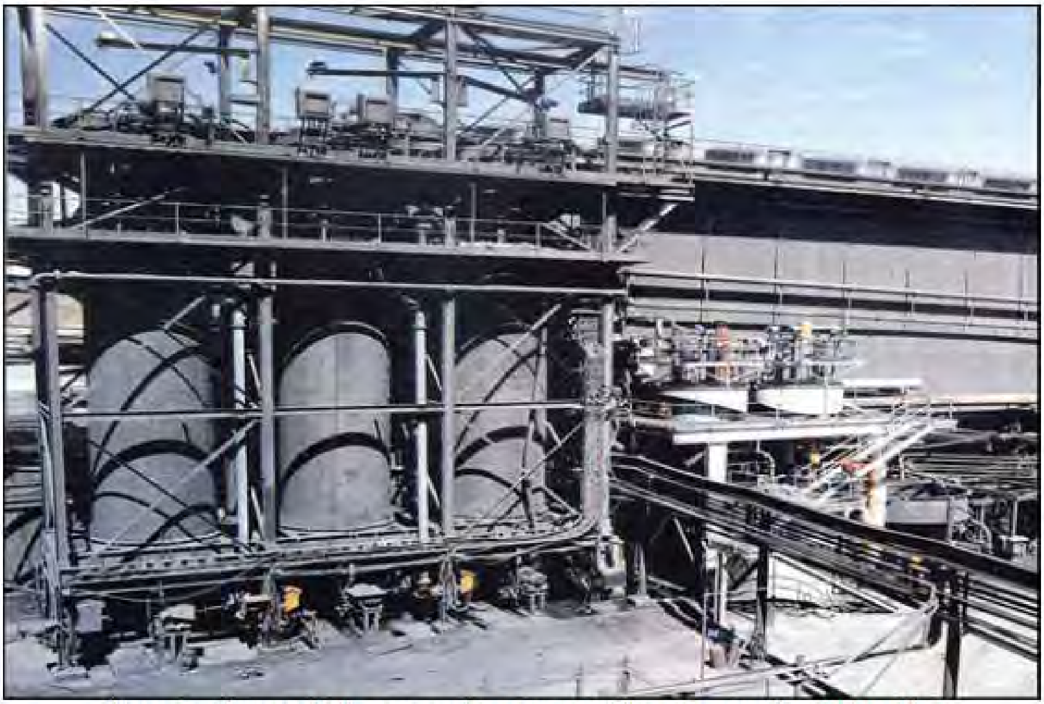


Figure 6 – Jameson Cells compared to columns of the same capacity at Mount Isa

Following initial work to provide an alternate method to bring together bubbles and particles, the downcomer was created. In the downcomer the air and the slurry are co-current with the air being entrained into the plunging jet under vacuum. Investigation showed that all of the bubble particle contact took place in the downcomer and thus the flotation tank could be much smaller. The first application at an industrial scale was in the lead zinc concentrator on the heavy media plant (HMP) lead slimes circuit. The initial improvement in performance were attributed to the very short residence time that allowed the minimisation of oxidation of galena fines. The cells were significantly smaller than the columns and there is no doubt the performance was superior as shown in Figure 3.

The testwork and trials in the early applications showed improved metallurgical performance when operated correctly. The challenge was operating them correctly. The technology hadn't been sufficiently developed to be successfully adopted into plant operations. The cell fell out of favour in base metals and in the 1990's was adopted into the Australia Coal industry and into niche SXEW applications where the main design challenges were resolved. The operability was improved by the introduction of a partial recycle to maintain constant flow and the maintainability of the cell was improved through various design modifications in operating plants. It was a period of continuous improvement. The result was a robust, low maintenance, easy to operate cell with the original features of excellent bubble particle contact.

The final obstacle was overcome when its adaption into the flowsheet was recognised to enable successful installations at the head of cleaner circuits and as low cost brownfield expansions. It is clear that the fast failures have had a significant effect on the success of the cell limiting its adoption into the industry. It is interesting that a significant proportion of sales are to return customers. Once you get over the hurdle of getting a Jameson Cell into your plant then seeing is believing. 2016 was the best year for Jameson cells into base metals and include the first sales back into South America where the cell had been abandoned after the difficulties of operations and maintenance of the Alumbrera installation. The metallurgical performance in Alumbrera was never the issue but the operators and maintainers hated the cells and they failed fast and hard.

The Jameson Cell celebrates its 30th birthday this year and has finally been adopted into mainstream base metals concentrators mainly as cleaner scalper at the head of the cleaner circuit. The cells generally recover up to 80% of the cleaner feed at high grades enabling much lower capital expenditure on the entire circuit. Process performance can be predicted from laboratory and pilot plant testing with demonstrated direct scale-up. It may have taken 30 years but the Jameson Cell is finally a success story. There are many lessons that can be learned from the implementation of innovation into industry from this case study.

IsaMill™

Unlike the developments of some of the other technologies at Mount Isa where efficiency was the main driver, the IsaMill was developed based on necessity. Figure 7 shows photomicrographs with the same scale of 40 micron demonstrating the increased complexity of Mount Isa ore over Broken Hill ore and the very difficult McArthur River ore. Although McArthur River was discovered in 1955 it was not able to be economically processed until the successful development of ultrafine grinding. McArthur River processing began in 1995 – 40 years after discovery when the IsaMill™ made it technically and economically feasible to grind all of the rougher concentrate to 7 micron to facilitate the rejection of non-sulphide gangue. Even at 7 micron galena liberation is not possible and a bulk zinc-lead concentrate is produced.

Figure 7 – Photomicrograph of a) Broken Hill ore b) Mount Isa ore c) McArthur River ore

Investigations into fine-grinding started at Mount Isa started in the 1970s using conventional grinding technology to increase mineral liberation by grinding to fine sizes. These technologies were not only found to have high power consumption but also proved to be detrimental to flotation performance as a result of pulp chemistry and iron contamination from steel media. These poor results were revisited during pilot plant and tower mill testwork in the 1980s which also showed an inability of tower mills to economically achieve the required sizes. When it became clear that the solution to efficient fine-grinding did not exist in the minerals industry, MIM looked for ideas to "crossover" from other industries that also ground fine particles – pigments, pharmaceuticals, foodstuffs (e.g. chocolate). While these mills operated at a much lower scale and treated high value products they demonstrated the principle that stirring fine media at high speed was highly efficient. The challenge was transferring this concept to continuous, high tonnage and lower-value streams in the minerals industry.

In 1991 the introduction of a Netzsch laboratory stirred mill to the Mount Isa site was a turning point in fine-grinding and ultrafine grinding. The ½ litre bench scale mill resembled a milk shake maker and used fine copper smelter slag as grinding media. Testwork on McArthur River ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the Mount Isa pilot plant. The testwork showed that high speed, inert, horizontal mills could efficiently grind to 7 microns at laboratory scale providing major improvements in metallurgical performance. To make ultrafine grinding applicable to full-scale production a program of development was undertaken between Mount Isa Mines Limited and NETZSCH-Feinmahltechnik GmbH.

After 7 years of development and testing of prototypes in the Mount Isa operations, the IsaMill™ evolved. It was large scale, continuous, and most importantly robust because it was developed by operators. The crucial breakthrough was the perfection of the internal product separator – this allowed the mill to use cheap natural media (sand, smelter slag, ore particles) and to operate in open circuit. These are significant advantages for operating cost and circuit simplicity. Scale-up was tested using trial installations at the Hilton and Mount Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill™ (1.1MW) was installed in the Mount Isa concentrator. Improvements to the technology were continually made by the operators, maintainers and engineers working with the technology.

In 1998 the rights for commercialisation of the IsaMill™ where transferred from Mount Isa Mines Limited to MIM Process Technologies (now Glencore Technology) and under an exclusive agreement with Netzsch. In December 1998, the IsaMill™ technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. The IsaMill™ is now a mainstream fine grinding machine with over 130 installations around the world.

The Albion Process™

In the 1990's, MIM were studying options for the development of the large Frieda River/Nena project in PNG through its subsidiary Highlands Pacific. The Nena ores were not amenable to smelting, due to the elevated arsenic content, and several hydrometallurgical options were examined. Out of this work, MIM developed the Albion Process™, named after the suburb in Brisbane where MIM's development laboratory was located. The Albion Process™ is a combination of ultrafine grinding using Glencore Technology's IsaMill™, followed by oxidative leaching at atmospheric pressure in a series of reactors designed to achieve high oxygen mass transfer efficiency. The HyperSparge™ was also developed to deliver oxygen to the reactors efficiently.

Various small scale continuous pilot plant campaigns were conducted in 1994 and 1995. A larger pilot plant (120kg zinc cathode/day) was constructed in 1997 to conduct testwork as part of a feasibility study on the zinc/gold resources of Pueblo Viejo in the Dominican Republic. Extensive piloting was also conducted on lower grade chalcopyrite concentrates for Cyprus Amax in 1998, and for Mount Isa Mines in 2000. Pre-feasibility and feasibility pilot testing was conducted on the zinc/lead bulk concentrates from McArthur River and Mount Isa in Australia between 2001 and 2005. During this time the Albion Process™ was successfully tested on over 70 different ores and concentrates. The process is designed to recover gold and base metals from refractory ores. The key to the process is the ultrafine grinding stage followed by a hot oxidative leach at atmospheric pressure.

In the period from 1994 until 2004, the Albion Process™ (see Figure 8) was seen as strategic to the MIM/Xstrata group, and was not marketed externally. In 2005, a decision was made to offer the technology to external clients under licence, and a marketing agent – Core Resources, was appointed to market the technology globally. Interest in the technology has been very strong in the subsequent period, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Certej Project. The technology moved into commercial production in 2010 with the commissioning of Glencore's Albion Process™ plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Glencore of a second plant in Germany (16,000 tpa zinc metal). The Las Lagunas refractory gold project commissioned in 2012, and the GPM Gold refractory gold project commissioned in 2013.

Figure 8 – The Albion Process oxidative leach plant in Armenia

The major scale up risk with any oxidative leaching technology is oxygen mass transfer. High agitator power demands are common to achieve the shear rates in the vessel required for effective mass transfer at a commercial scale. A different approach was taken in the design of the Albion Leach Reactor to lower the agitator power demand. Glencore developed the HyperSparge supersonic gas injection lance to provide gas injection velocities of the order of 500 $m.s^{-1}$ within the leaching vessel, compared to the 4 – 8 $m.s^{-1}$ achieved with a typical agitator. Supersonic oxygen injection is a far more efficient method of generating shear than conventional agitation, allowing the total power input into the vessel to be significantly reduced, and greatly reducing the scale up risk for the oxidative leach.

The Albion Process™ was enabled by the fine grinding of the IsaMill™ and the process was designed to deliver a lower cost processing option for treating refractory mineral resources. There are now six operating Albion Process™ plants and the process has now an extensive database of potential applications.

CONCLUSIONS

MIM developed a significant number of processing innovations that are technical and economic successes. The ability to innovate at MIM was enabled by very challenging orebodies and the need to process efficiently to remain economically viable. The success has been attributed to the development of these technologies on an operating site with the R&D group solving the technical issues on small scale. Each subsequent scale up was completed in the operating plants where the operators, maintainers, engineers and metallurgists were required to achieve production goals at each step of the scale up to ensure funding for the next step.

The number of innovations, at MIM, was disproportionate to the scale of operations and may have been enabled by the remoteness of the site and the researchers and operators working collaboratively to solve economic and technical problems. The research group were not capital city based but worked on the same site and were required to assist with installation, commissioning and operation of the various stages. This co-operation led to adoption into the plant and a fast feedback loop for improvements. The ultimate success of the innovations has been their widespread adoption into the mainstream industry where feedback from operating sites based on a user group model has enabled continuous improvement of each of the technologies.

ACKNOWLEDGEMENTS

The authors would like to acknowledge Glencore Copper and Glencore Technology for permission to publish and to all the research and production personnel who enabled and improved the

technologies in their plants. The success of these developments continues with the input from end users in the ongoing development.

REFERENCES

Armstrong, W. "The Isa Process and its contribution to electrolytic copper," paper presented at the Rautomead Conference, Scotland, August 1999.

Fewings J.H., Management of Innovation – The IsaSmelt Process. Presented at AMIRA "Innovation in Metal Production" Technical Meeting at Mount Isa, October 3-4, 1988.

Fountain C. Isasmelt and IsaMills – Models of successful R&D – AusIMM Young Leaders conference, 19-21 March 2002, Kalgoorlie, W.A.

Nihill D.N., Stewart C M. and Bowen P. the McArthur River Mine – The First Years of Operation. The AusIMM '98 Mining Cycle. Mount Isa 19-23 April 1998.

Pease, Joe (August 2005), "Complex leaching becomes much simpler" (PDF), Australian Mining, pp. 26–32, archived from the original (PDF) on 12 September 2009, retrieved 6 January 2010

Pease, Joe September 2016, "Crossing the innovation valley of death" Presented to JKMRC.

Premium coal fuels with advanced coal beneficiation

Louis J Wibberley[1] and Dave Osborne[2]
[1]Leader DICE Development Program
CSIRO Energy Technology, Mayfield West 2304, Australia
Email: louis.wibberley@csiro.au
[2] Coal Technology Consultant and Managing Director,
Somerset International Australia, Brisbane, 4000, Australia
Email: dosborne@somersetpty.com

Abstract

Research and pilot plant trials over the last 5 years have shown that a range of premium quality coal fuels, with ash contents as low as 1%, can be produced economically from a wide range of coals (including coal tailings). This capability can improve both the economic and environmental performance of coal through increased grade recovery, recovery of saleable coal from existing tailings emplacements, and new higher value products such as a fuel oil replacement in boilers, enhanced coking blends, and for higher efficiency power generation using the direct injection carbon engine (DICE). While a conventional coal preparation plant can be used to produce suitable ultra-low ash coal for these fuels, changes to both the philosophy of coal preparation and operating strategy are needed for achieving the best results. Further improvements are possible via the application of the latest milling and dewatering technologies. These fuels also require a rethink of the supply logistics. The paper discusses findings from both laboratory and pilot scale trials in Australia, in the context of new export products for DICE and boiler fuels.

Key words: ultra-low ash coal, coal grain analysis, coal slurry fuel, advanced beneficiation, DICE, high efficiency engine

INTRODUCTION

Producing very low ash coal products has for many years been a challenge for coal treatment specialists and researchers. Much of the early work in the 1970s and 1980s used chemical cleaning processes involving leaching with acidic or caustic solutions, most of which proved uneconomic when scaled up to commercial application. The most successful of these include the AMAX 2-stage leach process developed with US-DOE funds in the mid-1980s[1] and the Australian UCC process[2].

The use of physical cleaning to ultra-low ash levels, using more conventional coal processing technologies, has proven more challenging. Old school thinking typically regards ash contents below 2-3%, as both technically and economically unviable. This is due to several factors/misnomers:

- The so-called "inherent ash" content of coal is usually regarded as the lowest ash possible
- Lower ash contents require finer grinding to increase liberation which is costly and produces a product for which there are few markets
- Flotation of ultra-fine coal can be problematic, and requires high reagent consumption

---

[1] Berggren, M. H., Chemical coal cleaning and cost refinement for coal-water slurry manufacture. Report DOE/MC/20700-T1, February 1985.

[2] http://www.yancoal.com.au/page/key-assets/technology/UCC

- Fine coal products are inevitably high moisture (> 35%) which means costly dewatering and/or drying to produce saleable products
- Ultra low ash coal products are uneconomic to produce

All of these factors are incorrect, or at best very misleading, as recent research/pilot plant tests have shown.

Fine coal cleaning is generally regarded as being the treatment of coal sized below 0.5mm (500µm) but the challenges have tended to descend further down the size distribution to 0.1mm (100µm) and lower. Most recently, with the widespread use of micro-flotation using the Jameson and Microcell technologies, the elusive size fraction is now <0.05mm (50µm) and this is probably where the current research and development focus lies. Such a size range usually means that coal particles are extensively liberated from the mineral content and are recoverable, providing that the concentrate can be recovered and dewatered to a commercially acceptable level.

Not surprisingly, much of the research and development that has occurred in ultra-fine coal cleaning has taken place for higher value products, often metallurgical, and almost all techniques for both cleaning and dewatering have undergone some form of improvement or development during the past decade in this context. These improvements have been in all areas - optimised design, improved materials of construction and improved wear resistance[3], process control and monitoring, integrated automation and sampling/analysis.

This paper describes recent work carried out at laboratory and pilot scale in Australia, that has successfully produced coal concentrates with <2% ash content from a wide range of coals and tailings, using a combination of fine grinding and froth flotation. The potential to achieve even lower ash levels is also discussed. The main objective of this work has been to produce a stable, coal-slurry fuel (i.e. micronised refined coal or MRC) for use in DICE, an application which demands the lowest possible ash content in the slurry fuel (preferrably <1%) and very fine particle size distribution (typically d_{80} of 20-40µm). This is much finer than the conventional coal water slurry fuel (CWSF) for boilers or gasifiers.

CHEMICALLY CLEANED COAL

Ultra-clean coal can be produced by chemically cleaning coal, and a 2- and 3-stage caustic leach process has been successfully demonstrated in Australia by UCC Energy Pty Ltd. A significant feature of the process is that the coal needs only to be coarsely milled (typically to -2mm) to provide sufficient contacting: all other processes for producing ultra-low ash coals require a higher degree of milling. The main process consumables are sulphuric acid, and lime; caustic soda is regenerated in the process. The waste materials are environmentally benign (gypsum and calcium alumina silicates) and the final ultra-clean coal product has 0.5-0.7% ash (with less than 0.2% possible for some coals). UCC Energy Pty Ltd (a wholly owned subsidiary Yancoal Australia Pty Ltd) owns and has operated a pilot plant at Cessnock in the Hunter Valley, New South Wales. The nominal 350kg/h facility (see Figure 1) was commissioned in 2002[4]. The main focus has been in the production of low ash coal fuels for gas turbines and DICE, as well as for electrode carbon and other high value applications – several using briquettes.

---

[3] Osborne, D. G.; 2012. Milestones in Fine Coal Development, *Challenges in Fine Coal Processing, Dewatering and Disposal*, editors M.S. Klima, B. J. Arnold and P.J. Bethell, SME, TN816.C375 2012, pp3-32

[4] http://www.uccenergy.com.au/the-process/

Fig 1 Ultra-Clean Coal pilot-plant, Cessnock, NSW (courtesy UCC Energy Pty Ltd)

Fuel from this plant has been used for longer duration tests in an adapted diesel engine (*ie* DICE) at the CSIRO, and scale up of the process is now under consideration by Yancoal and partners.

PHYSICAL CLEANING

Although chemical cleaning has been proven to produce very low ash coal products for the highest value applications (e.g. electrode carbon and MRC for DICE – especially smaller engines), the process is more costly than physical cleaning, is less tolerant of coal rank, and requires a coarser lower ash feed coal for optimum results. Physical cleaning is therefore preferred for the production of coals with ash contents <2% from a variety of coals and tailings, and is the focus of the present paper.

Liberation – removing the "inherent ash" constraint

Efficient analysis of the occurrence of mineral matter within the feed coal is key to economically producing premium coal products by physical cleaning. CSIRO in Australia has developed an optical reflected light microscopy system for assessing coal petrography samples. This system collects and creates mosaic images so that quantitative information can be obtained on individual coal/mineral grains. Size and compositional information (the amount of vitrinite, inertinite, liptinite and mineral type) is obtained for each particle, and this information can then also be used to estimate the density and ash value of each particle (both macerals and minerals) and that of the sample.

Although coal grain analysis (CGA) is generally conducted on a small representative sub-sample of -1 mm material, the technique has also been successfully used to characterise particles up to 4 mm in size (a size which perhaps report to a middlings circuit), and for samples which have been micronised to give particle size of less than 20 µm (e.g. for producing ultra-low ash concentrate by flotation).

For the ultra-fine material, CGA is particularly useful to assess the degree of liberation and therefore assist in optimising the beneficiation process to produce the lowest possible ash products (for uses such as DICE, or even direct carbon fuel cells[5]).

---

[5] O'Brien G., Firth B., Adair B. The Application of the Coal Grain Analysis Method to Coal Liberation Studies, 2011, *International Journal of Coal Preparation and Utilization 31*: pp 96–111

For advanced beneficiation, CGA enables liberation to be assessed and the expected yield at different target ash levels to be estimated. As the CGA information also provides detail on the intrinsic and entrained minerals in a sample, it benchmarks the lowest ash value which could be obtained if all entrained minerals are removed from the sample at that top-size.

An example of its application in the production of MRC for DICE is given in Figure 2 below. The characterised CGA images shown illustrate the extent of liberation for raw tailings (-250µm) and the corresponding micronised concentrate with a d_{80} of 15µm. The results show that milling has enhanced significantly the particle liberation, and has produced essentially single component grains. This indicates that it is physically possible to reduce the intrinsic mineral being carried over to the product. This implies that an appropriate circuit of grinding and flotation operations should be capable of producing a very low ash product concentrate. In this case sufficient liberation has occurred to enable a product coal with less than 1% ash to be obtained, provided entrainment within the froth can be eliminated. Although it is anticipated that the fineness of the entrained mineral particles will make separation more difficult, there are a number of process optimisations available to achieve this (e.g. use of reagents which enhance the hydrophobicity of the coal and hence allow more aggressive application of wash water to be used, optimisation of the pull between flotation stages). Both have been shown to be effective in practice at the pilot scale.

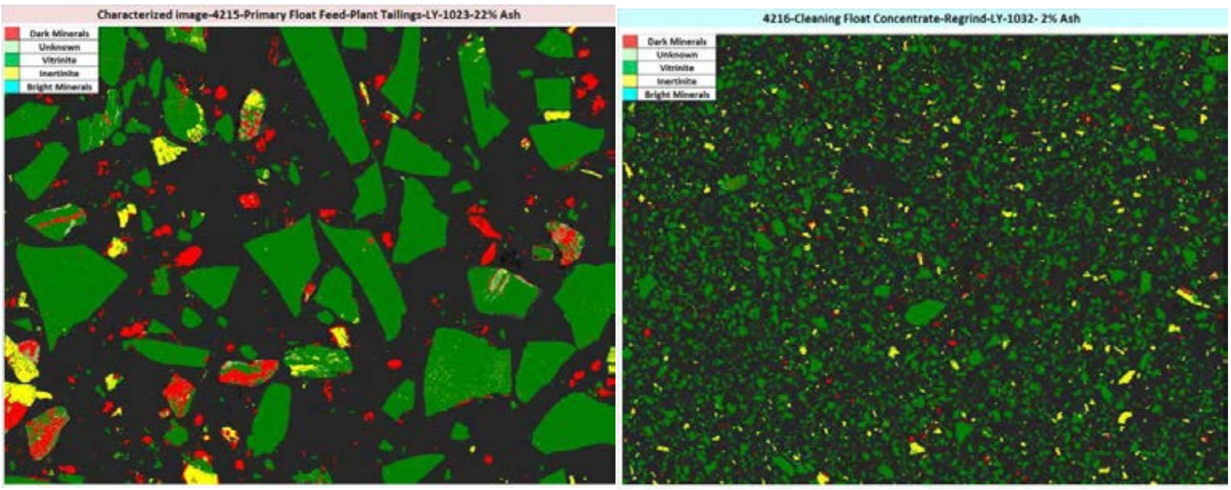


Fig 2 Characterised images for raw coal tailings feed and final concentrate (courtesy QCAT-CSIRO)

The results obtained from several CGA studies with a pilot plant, and other minerals tracking studies by the CSIRO with 9 Australian bituminous coals/tailings, have provided a valuable insight into the separation and recovery performance of the current beneficiation work, by tracking the selective recovery of individual grain types. In particular, this has enabled optimising the milling and flotation steps for processing raw tailings samples.

Overall, CGA has proved a valuable method for applying research in a commercial environment.

Milling/flotation process approach

In order to achieve lower ash levels, a liberation step needs to be introduced by fine grinding. Typically fine grinding for coal water slurry fuels is by ball mills; -for example, ball mills are commonly used in China where over 30Mt of slurry fuel is currently produced annually with ash levels below 4%. However, to enable even lower ash contents and maximum liberation (and with a reduced energy consumption), the present work has used bead mills (IsaMills) for both laboratory and pilot scale tests.

Using this mill, a 3 tonne/day pilot-plant, see Figure 3, owned and operated by Glencore Technology (formerly Xstrata) has been used for proving the feasibility of producing ultra-low ash coal from freshly generated raw coal tailings. The plant is sited at Bulga mine, a large thermal coal operation in the Hunter Valley. The pilot plant comprises two Jameson flotation cells, a small IsaMill and a membrane filter press, which can be configured to simulate a number of circuit designs.

Fig. 3 The 3 t/day coal tailings treatment pilot-plant (Courtesy Glencore)

A range of different coal feed types have been tested with raw coal ash contents ranging from ~ 30% feed ash (ad) from easily treated coal seams, to much higher ash coal seams containing difficult-to-separate colloidal clays. The plant is usually configured in a rougher-scavenger circuit where the tailings from the rougher (primary cell) feeds to the scavenger (secondary cell).

The cleaning efficiency of this circuit is clearly shown by the difference in colour of the two product streams in Figure 4 below; the black concentrate is high in vitrinite and the white tailings stream is predominantly kaolinite clay.

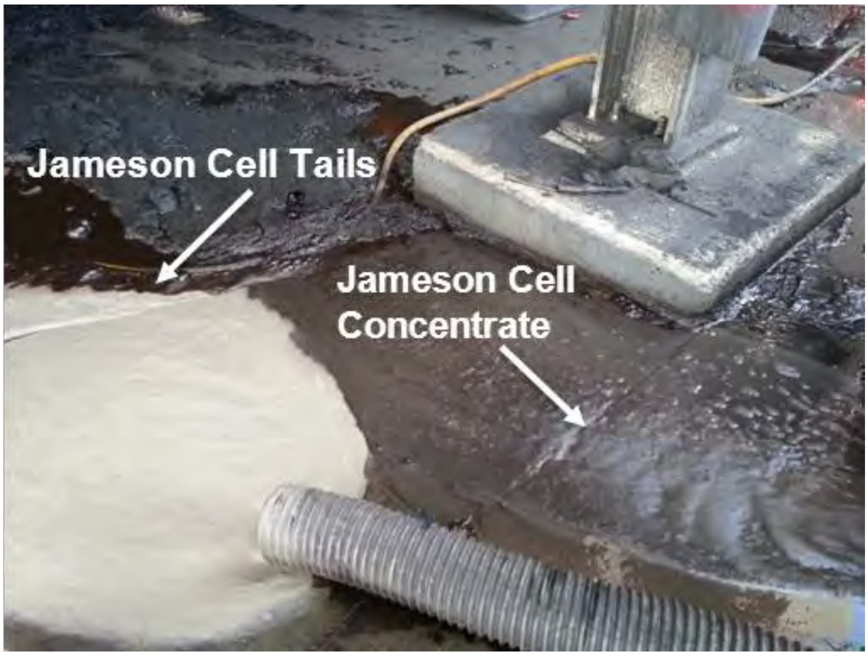


Fig. 4 Concentrate and tailings streams from pilot plant test-work (courtesy Glencore)

A secondary effect of cleaning is a marked improvement in dewatering: Fine mineral matter and clay in a coal concentrate significantly reduces filtration rate and increases product moisture. By removing much of this material, a significant reduction in product moisture can be achieved – essential for metallurgical and briquette products.

Cost effective and efficient dewatering is thus essential to producing premium coal products, for treating both the concentrate and barren tailings streams. Dewatering options available are listed in Table 1. Recent advances in "by zero" fines dewatering has enabled total moistures of flotation product to be reduced to a target whereby it has become commercially viable to include flotation concentrates into the final thermal coal product at a greater number of mine sites. Tailings dewatering still has some challenges, but high-G solid-bowl centrifuges offer promise for on-line dewatering.

Table 1 Equipment Used for Dewatering Coal (adapted from Table 13-1, *The coal handbook*[6])

Equipment	Footprint size	Throughput (t/h dry solids)	Product moisture (% w/w)	Feed preparation	Application
High frequency screen	0.6-2.4 x 3m	10-100 t/h	15-25	cyclone underflow	fine coal
Screen scroll centrifuge	0.5-1.5m dia.	45-100 t/h	11-18	cyclone underflow	fine coal
Horizontal vacuum belt filter	75-150m^2	50-130 t/h	20-30	flocculation	ultrafine coal
Screen bowl centrifuge	1.1m dia. x 3.3 m long	20-60 t/h	16-27	thickening	ultrafine coal
Solid bowl centrifuge	1.1 m dia. x 3.3 m long	15-20 t/h	15-20	thickening	ultrafine coal slimes

---

[6] Bickert, G. 2013. Solid-separation Technologies for Coal, Chapter 13, The coal handbook; Volume 1; Woodhead Publishing Series in Energy, pp 422 – 444, edited by D. G. Osborne, 2013

Equipment	Footprint size	Throughput (t/h dry solids)	Product moisture (% w/w)	Feed preparation	Application
Disc filter	120-200 m^2	50-150 t/h	20-32	thickening /flocculation	ultrafine coal
Hyperbaric disc filter	70-200 m^2	30-150 t/h	17-25	thickening /flocculation	ultrafine coal
Paste thickening	25 m dia x 6-12 m high	100 t/h	45-55	flocculation	barren tailings
Solid bowl centrifuge	1.1 m dia	20-60 t/h	30-45	thickening	barren tailings
Belt press filter	3-3.5 m wide	10-20 t/h	25-45	thickening /flocculation	barren tailings
Filter press	200-800 m^2	15-30 t/h	14-32	thickening	ultrafine coal & barren tailings

The economic argument for pursuing fines recovery in this application is now very compelling. Technology advances, particularly in dewatering, allow ultrafine product to be included in final product streams without penalising product quality (or introducing handling difficulties) as shown by the simplistic revenue scenario below.

Based on the following assumptions:

- Raw plant feed containing 10% passing 0.1 mm
- Thermal coal operation of 16 Mt/y ROM (1.6 Mt/y of raw feed currently sent to waste)
- Assuming a nominal 50% yield which equates to 0.8 Mt/y of potential saleable product
- At a benchmark price of US$85/product tonne (equivalent to US$69M revenue loss per annum, excluding freight, port, tonnage adjustments, etc.)

Including conservative capital and operating costs produces a reasonably attractive investment opportunity as given in Table 2 below.

Table 2 Economic evaluation of brownfields flotation installation

Capex $M	Operating $/t feed	Coal rate t/h	Direct costs $/t	Tax rate %	Discount rate %	NPV $M	IRR %	Payback years
50	15	230	30	30%	10%	78	51	2.5

Future direction – reducing product ash and moisture

The challenge of cost-effectively recovering a saleable fines component from tailings has been with us for many years, and periodically an apparent solution emerges. Glencore Technology has been operating the pilot plant described above for over 4 years, and this plant incorporates the combination of ultra-fine grinding by an Isamill and Jameson Cell flotation technology[7]. The inclusion of a fine grinding stage enables slurries to be prepared whereby mineral components are almost completely liberated from the carbonaceous material, thereby facilitating recovery of a highly concentrated ultrafine, low ash coal product, with ash levels well below traditional inherent ash limits – even from tailings.

This combination has already been proven capable of achieving very good combustible recoveries (material dependent, but normally over 90%) for coal derived from the raw tailings stream. The milling step produces a feed with a d_{80} of 15-30µm, enabling enhanced flotation recovery by a combination of increased liberation and the formation of fresh surfaces on the ultra-fine coal particles. This also reduces reagent consumption to levels significantly below traditional fine coal flotation.

Further enhancement by the addition of improved ultrafine dewatering of the flotation concentrate using a membrane filter press, or equivalent, and appropriate mixing system has resulted in the preparation of ultra-low ash, highly stable slurries with solids concentrations over 60% (w/w). This product should be very suitable for DICE, a potential large new market for a range of power generation markets:

- To replace fuel oil and natural gas,
- To provide highly efficient, highly flexible and modular coal-based power to backup increased renewables (giving an extremely low CO_2 and secure electricity system), and
- For incremental, low CO_2, coal-based generation capacity[8].

In the shorter term, slurries with a slightly coarser particle size distribution (typically with a d_{80} of 75µm) can be prepared in a similar way to create conventional coal-water slurry fuels, at over 70% solids, for direct firing into boilers as a replacement for heavy fuel oil (HFO) and for gasifiers.

Figure 5 below shows a nominal flowsheet based around an Isamill for micronising, and the Jameson Cell for separation. Adding the milling step can be optional for boiler grade product, but is essential for producing MRC for DICE. Various dewatering options, as described earlier, need to be carefully evaluated to achieve the desired solid-liquid outcome for each coal source/product combination. Alternatively, a briquetted product can be included in the normal product stream, thus avoiding slurry or fines related problems in product handling and transportation.

The integrated plant design thus has the functionality of the dual product offering, i.e. coal briquettes that can be added to the conventional product and/or coal-water slurry fuel for heavy fuel oil replacement for boilers, or for more innovative, value added applications such as DICE.

To facilitate logistics, and to enable early adoption of the technology, the concept of coal water slurry supply chains is being promoted to industry, whereby slurry fuels employ existing heavy fuel oil infrastructure to transport and store the fuel at the customer's facility. The use of these systems was been extensively demonstrated in Japan in the 1990s, and more recently in China.

---

[7] Mercuri, F., Osborne, D.G. and Young, M; 2014. The Future of Thermal Coal Flotation. Australian Coal Preparation Society conference 2014, Gold Coast, Qld.

[8] Wibberley, L.J. DICE – The best option for coal? Publication on the DICEnet web-site www.dice-net.org http://www.dice-net.org/images/pdf/Media/Papers

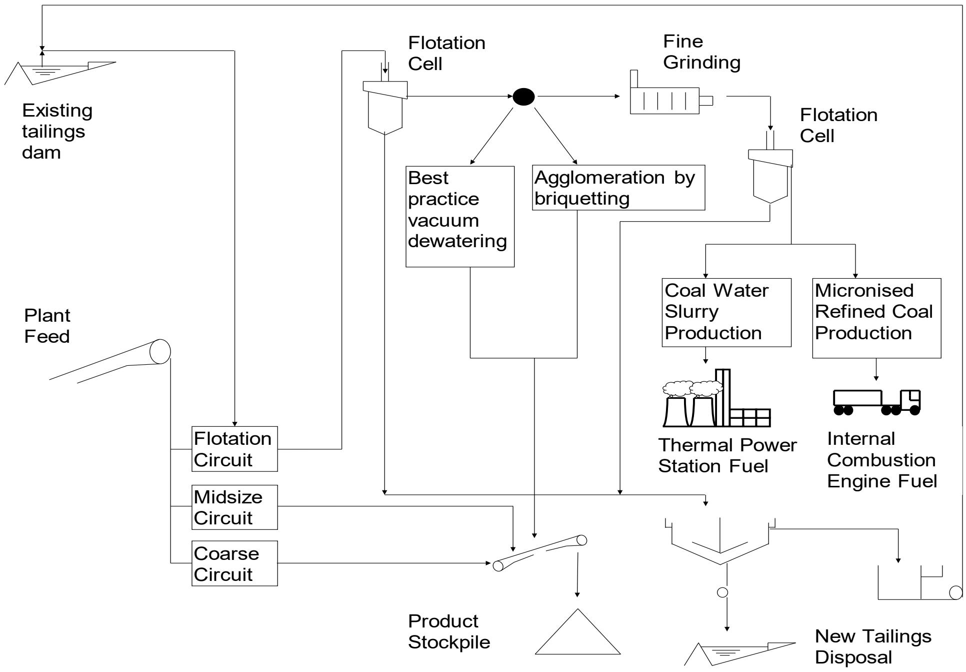


Fig. 5 Alternative fines treatment for thermal coals

Conclusions

For more than two decades the appeal of so-called "deep cleaning" of coal via liberation and subsequent beneficiation has been recognised in terms of the significant downstream improvements that would result - maximised resource recovery, minimised transport and handling costs, numerous end-user process improvements, reduced maintenance and wear, lower environmental impacts and sustainable improvements.

While the capability of ultrafine particle separation has matured via progressive improvements in froth flotation, the capability of dewatering the concentrate has continued to be the major barrier until recently. The emergence of larger capacity membrane filter presses, hyperbaric dewatering via decanter centrifuges or large disc filters now offer commercial solutions.

Binderless briquetting has also progressed to machines with capacities of up to ~40t/h for fine coal applications, and providing an alternative product for conventional transportation infrastructure.

The manufacture of stable coal-water slurries with over 65% solids content, and MRC slurries with over 60% solids content, have also reached commercial adoption; such products can also be produced from tailings.

With these technical barriers now overcome, the scene is set for these new applications to progress to become the new generation of clean coal technologies for a wide range of applications, with key economic drivers being a much higher fuel cycle efficiency (i.e. lower carbon intensity), highest grade recovery, and lowest solid waste disposal.

References

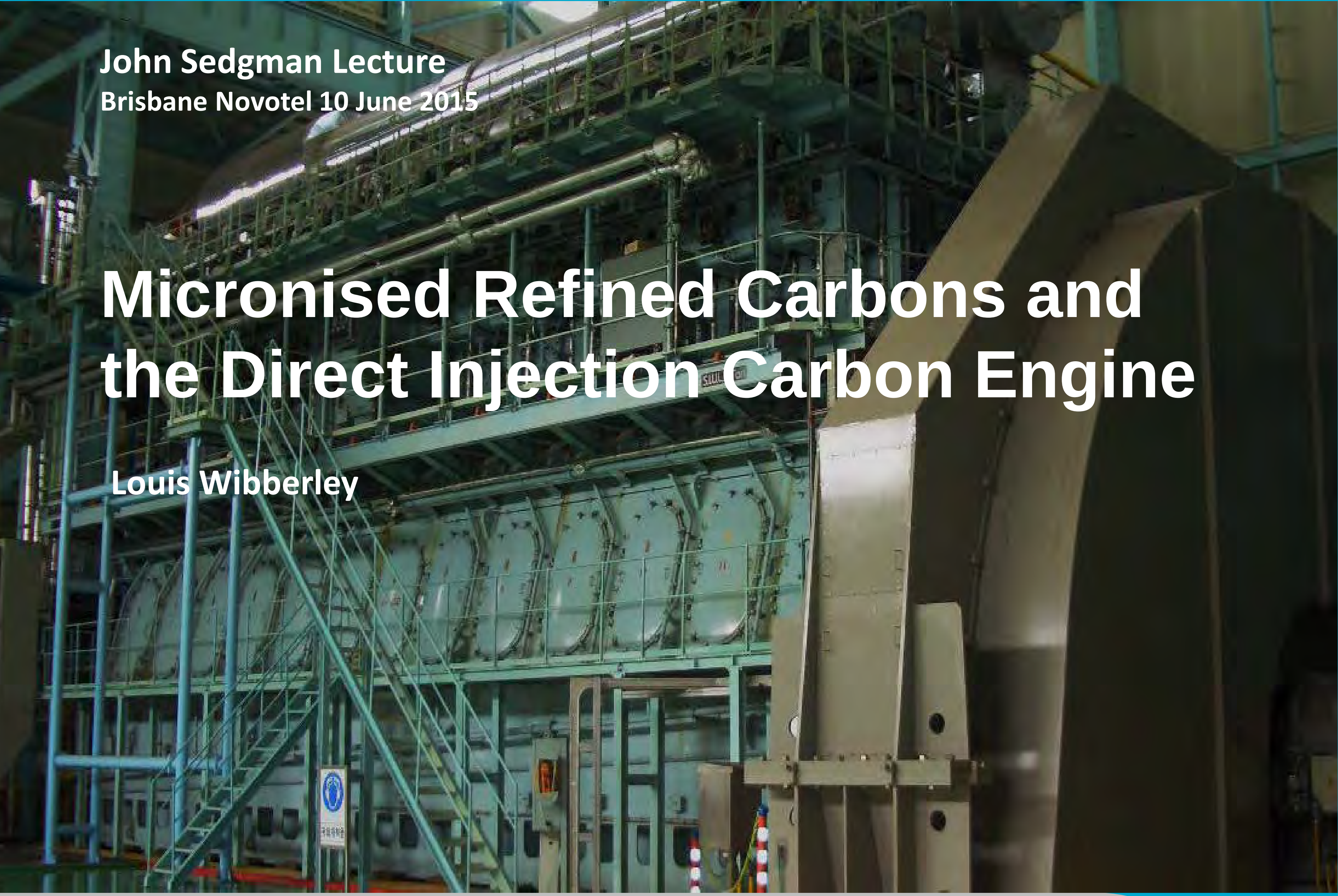
1. Osborne, D. G.; 2012. Milestones in Fine Coal Development, *Challenges in Fine Coal Processing, Dewatering and Disposal*, editors M.S. Klima, B. J. Arnold and P.J. Bethell, SME, TN816.C375 2012, pp3-32.

2. O'Brien G., Firth B., Adair B. The Application of the Coal Grain Analysis Method to Coal Liberation Studies, 2011, *International Journal of Coal Preparation and Utilization 31*: pp 96–111.
3. O'Brien G., Firth B., Warren, K and Hapugoda, P; 2015. Principles of Coal Selection for Advanced Beneficiation. 2015 Clearwater Clean Coal Conference; *40th International Technical Conference on Clean Coal & Fuel Systems*
4. Mercuri, R., Osborne, D.G. and Young, M; 2014. The Future of Thermal Coal Flotation. *15th Australian Coal Preparation Society Conference*, Gold Coast. QLD.
5. Bickert, G; 2013. Solid-separation Technologies for Coal, Chapter 13, The coal handbook; Volume 1; *Woodhead Publishing Series in Energy*, pp 422 – 444
6. Wibberley, L.J. 2014: DICE – The best option for coal? Publication on the DICE-net web-site http://www.dice-net.org/images/pdf/Media/Papers

Acknowledgements

The authors acknowledge the contributions provided by Glencore Technology, Glencore Coal, Yancoal and the CSIRO towards the preparation of this paper and for permission to include figures and data from other recently published papers and articles on this subject.

Dr Osborne thanks Somerset Coal International for encouragement and support in attending and participating in the 2015 Clearwater Clean Coal conference.



John Sedgman Lecture

Brisbane Novotel 10 June 2015

Micronised Refined Carbons and the Direct Injection Carbon Engine

Louis Wibberley

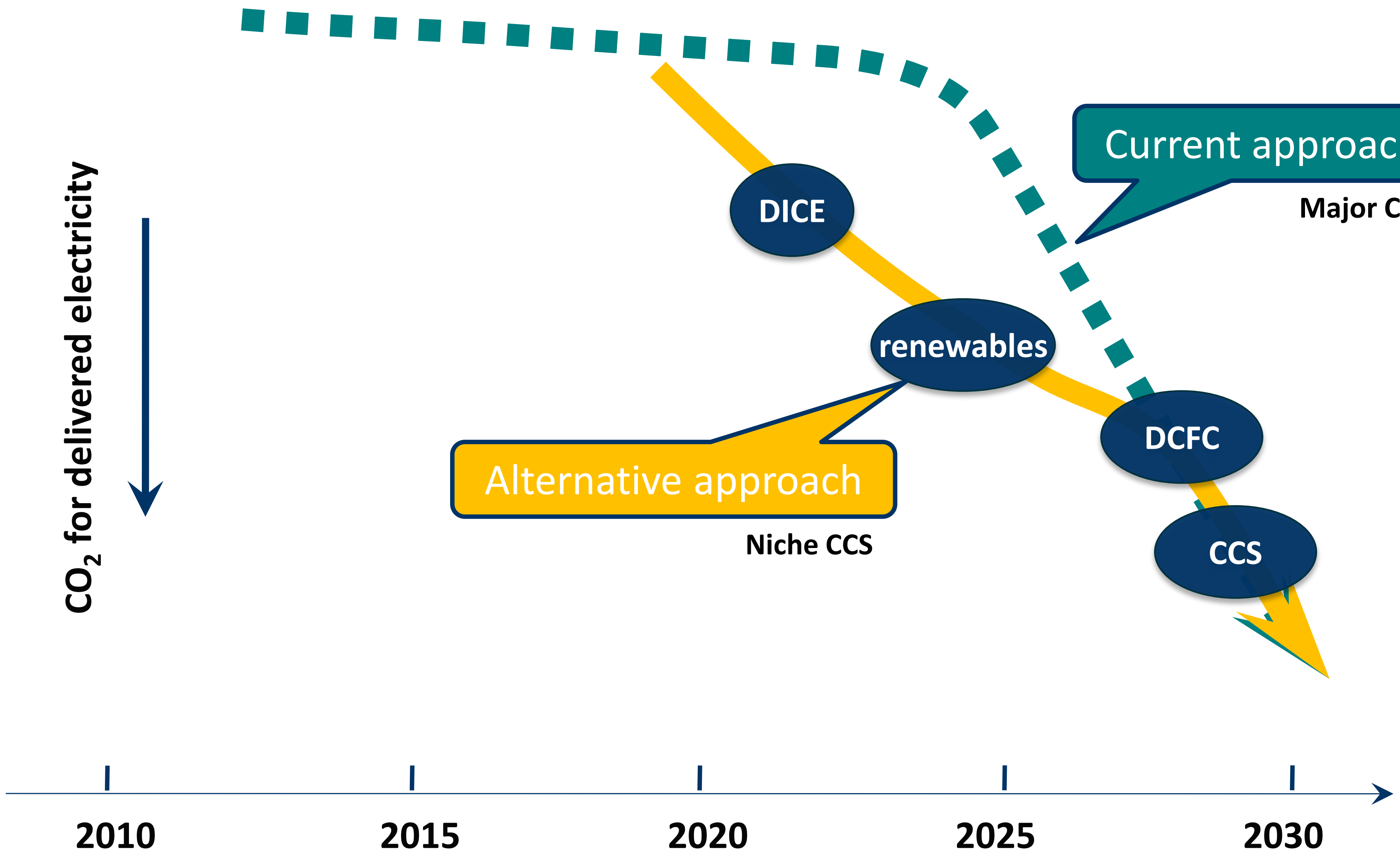
Key messages

1. Combining the low cost and availability of coal with the superior thermal efficiency, flexibility and lower capital cost of the diesel engine provides a <u>step change technology</u> for coal
   – this requires <u>ultra low ash coal</u>

2. Ultra low ash (and other premium coals) can be economically produced from a wide range of coal sources, including tailings - using <u>conventional equipment</u>
   … but this requires a <u>change in philosophy</u>

Why? … provides an alternative LE pathway

Philosophy: higher efficiency + underpinning renewables + niche CCS



MRC-DICE fuel cycle

Carbons

Premium water-based slurry fuel

micronised refined carbons (MRC)

Ultra efficient diesel engine generation

direct injection carbon engine (DICE)

MRC – the most efficient way of converting carbons into liquid fuels

Micronised refined carbon (MRC) has been produced from a range of sources

- desanded and hydrothermally treated low rank coals
- deashed black coals (including tailings)
- chars and algal matter (blended)

Fuel choice determines carbon footprint

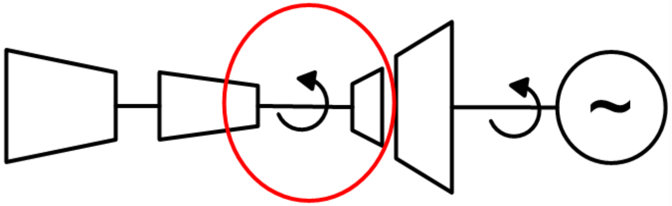
Process has very high energy conversion efficiency >97% (LCA basis)

Diesel engine – efficient, flexible and fuel tolerant

(but some adaptation required for coal)

Gas turbine

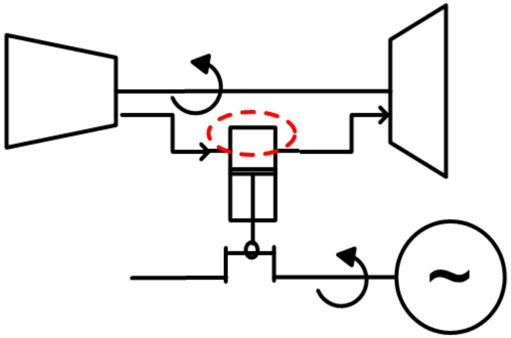
Continuous combustion - hot section is exposed continuously to 1450-1500°C gas at 2-4 MPa
Exotic alloys, hot strength, oxidation
Fouling issues for impure fuels



Turbocharged diesel engine

Higher "Carnot" efficiency
Cyclic hot space allows dirtier fuel without fouling or the need for exotic alloys
Large expansion ratio = smaller waste heat recovery

Batch combustion - hot section hot for <10% of the time, cyclic @ 1-5 Hz (larger engines)
Higher T & P possible and without fouling, >1500°C, 15-25 MPa



Cycle comparisons – too much water?

	External cycle				Power cycle				
Cycle-fuel $P/T_1/T_2$[1]	Fuel (dry t)	Water (t)	Air (t)	η (%HHV)	Fuel (dry t)	Water (t)	Air (t)	η (%HHV)	η sent out (%HHV)
Steam-black 250/650/650	1	0.1	11	88%		12		48%	42%
Steam-brown 250/650/650	1	2.2	10	73%		11		48%	35%
Steam-CWM 250/650/650	1	0.7	11	84%		11.5		48%	40%
Diesel-HFO 200/1500					1	0-1	15	54%	52%
DICE-MRC 200/1450					**1**	**1**	**15**	**51%**	**49%**

[1] *bar/°C/°C*

DICE offers game-changing attributes in 5-6 years

1. Match and compete with natural gas; rapid start/stop and load following capability
   - excellent match to electricity grid with high intermittent renewables
2. Step reduction in carbon emissions for electricity generation without CCS
   - 20-35% reduction in carbon emissions versus current black coal
   - 35-50% reduction in carbon emissions versus current brown coal (in Victoria)
3. Cost competitive with new conventional coal

… not possible with other coal technology

4. Small capital investment steps
   - can achieve large power plant size incrementally using 20-100 MW units
   - shorter construction time
5. No cooling water
6. Can be used for various biomasses
7. Capture ready/capture efficient
   - 30-40% lower cost of CO_2 abatement over conventional coal
8. Short path to commercialisation
   - adaptation of current large engines, short cycle time to implement changes, relatively low development cost

ability for commercial-scale demonstration at a small scale (say 30 MW)

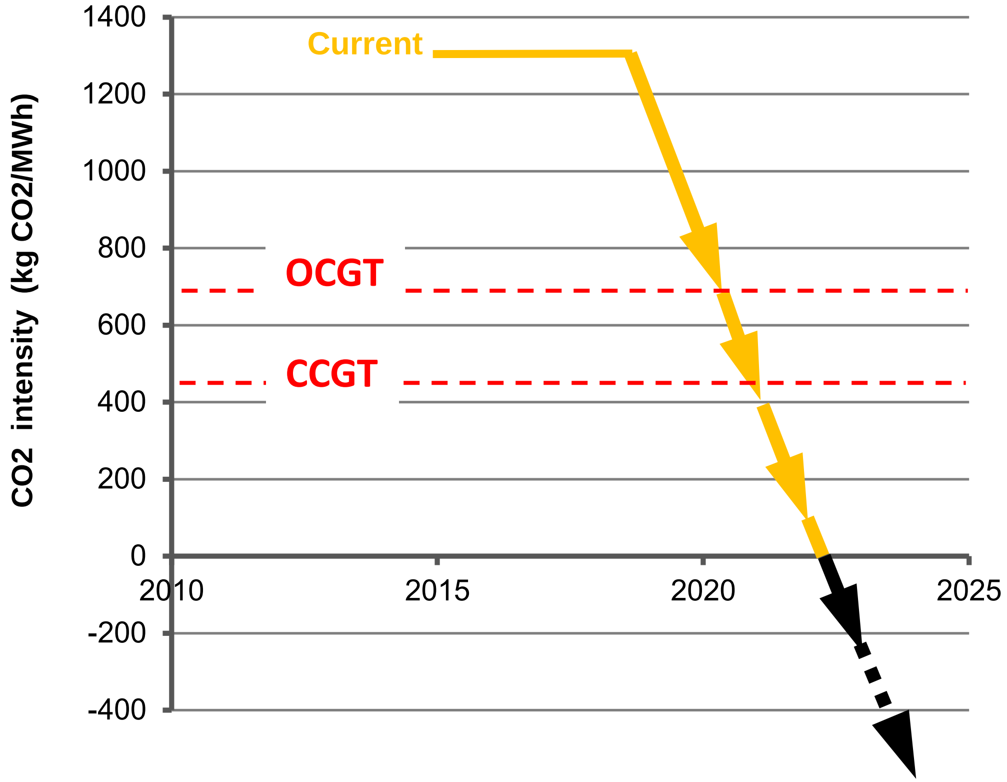
What about USC? … development opportunities restricted and costly

- Thermodynamic efficiency of pf generation is severely limited by the availability of materials that can operate at these conditions for practical service lifetimes
  - EU, USA, Japan, India and China all have extensive material research programs aiming for steam temperatures of 700°C (advanced ultra-supercritical)
  - development cost of billions of $ and long lead times (creep testing)
  - anticipated that a commercial unit could be brought on-line in 2031 (IEACCC/229)
  - high capital cost of advanced ultra-supercritical is of particular concern (high pressure steam pipes currently 80% of the boiler cost)
- While the combustion conditions in the diesel engine are more extreme, the diesel cycle is a batch process
  - high temperature conditions are present for less than 10% of the time, which avoids the need for major exotic alloys

A pathway to net negative CO_2 emissions?

(DICE efficiency first, then high penetration renewables with bio-CCS and lastly partial CCS)

Example: Victorian generation



** landscape & soil carbon sequestration credits

Carbon management sequence

- DICE efficiency
- Biomass co-firing
- Underpinning major renewables
- Bio-CCS**
- Partial CCS

If successful DICE could address many aspects of the coal dilemma

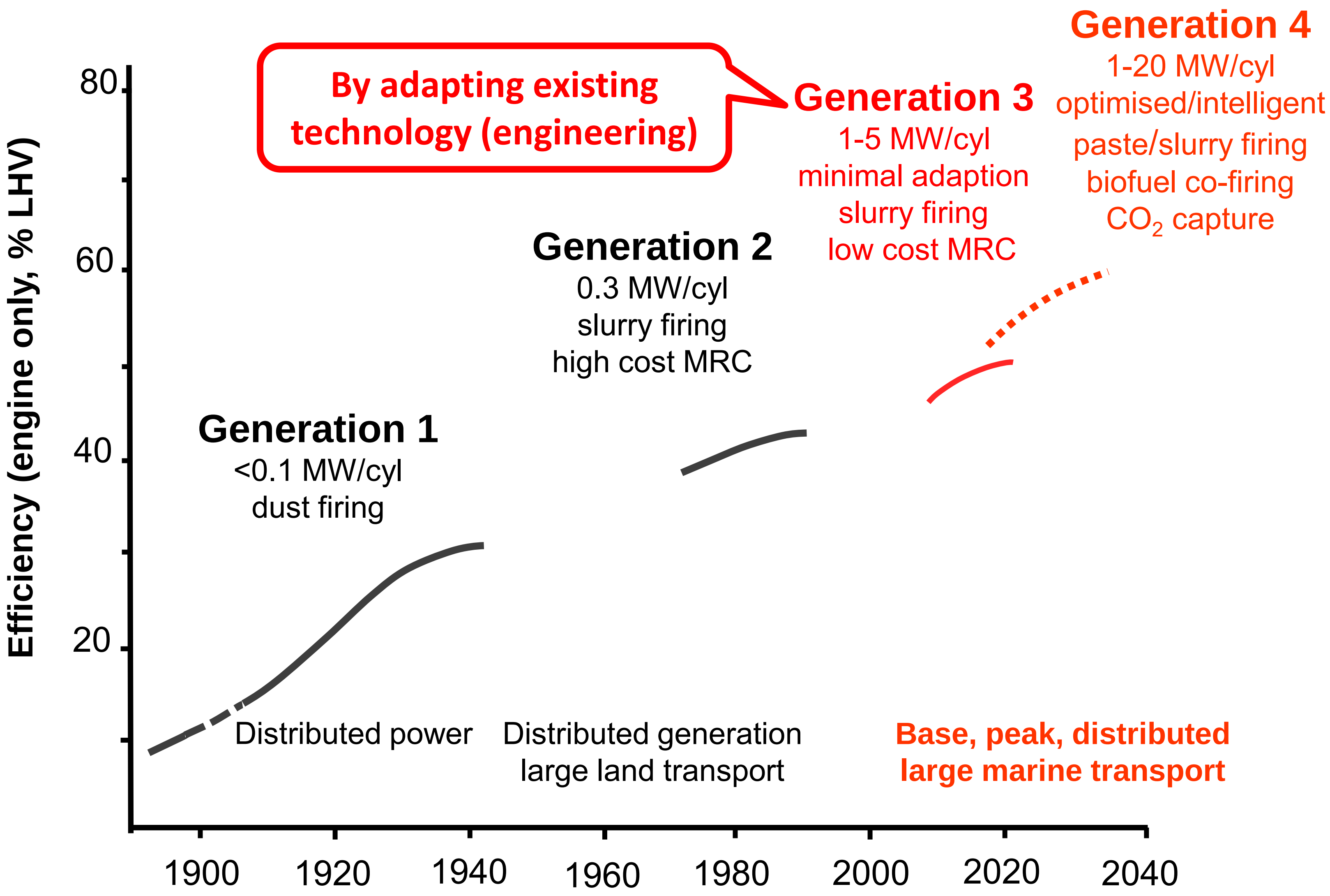
Including:

- that large centralised plants are needed for efficiency
- the nexus between water/CO_2 and cost (dry cooling)
- technology development by a fragment of generation industry
- inefficient (even if cheap) is no longer acceptable
- very poor image of low rank coals
- poor project economics from long development times
- the higher flexibility needed for current and future electricity markets

Could DICE become the benchmark coal generation technology?

Not new … commercialisation in G3?

Considerable upside



By adapting existing technology (engineering)

Generation 1
<0.1 MW/cyl
dust firing

Generation 2
0.3 MW/cyl
slurry firing
high cost MRC

Generation 3
1-5 MW/cyl
minimal adaption
slurry firing
low cost MRC

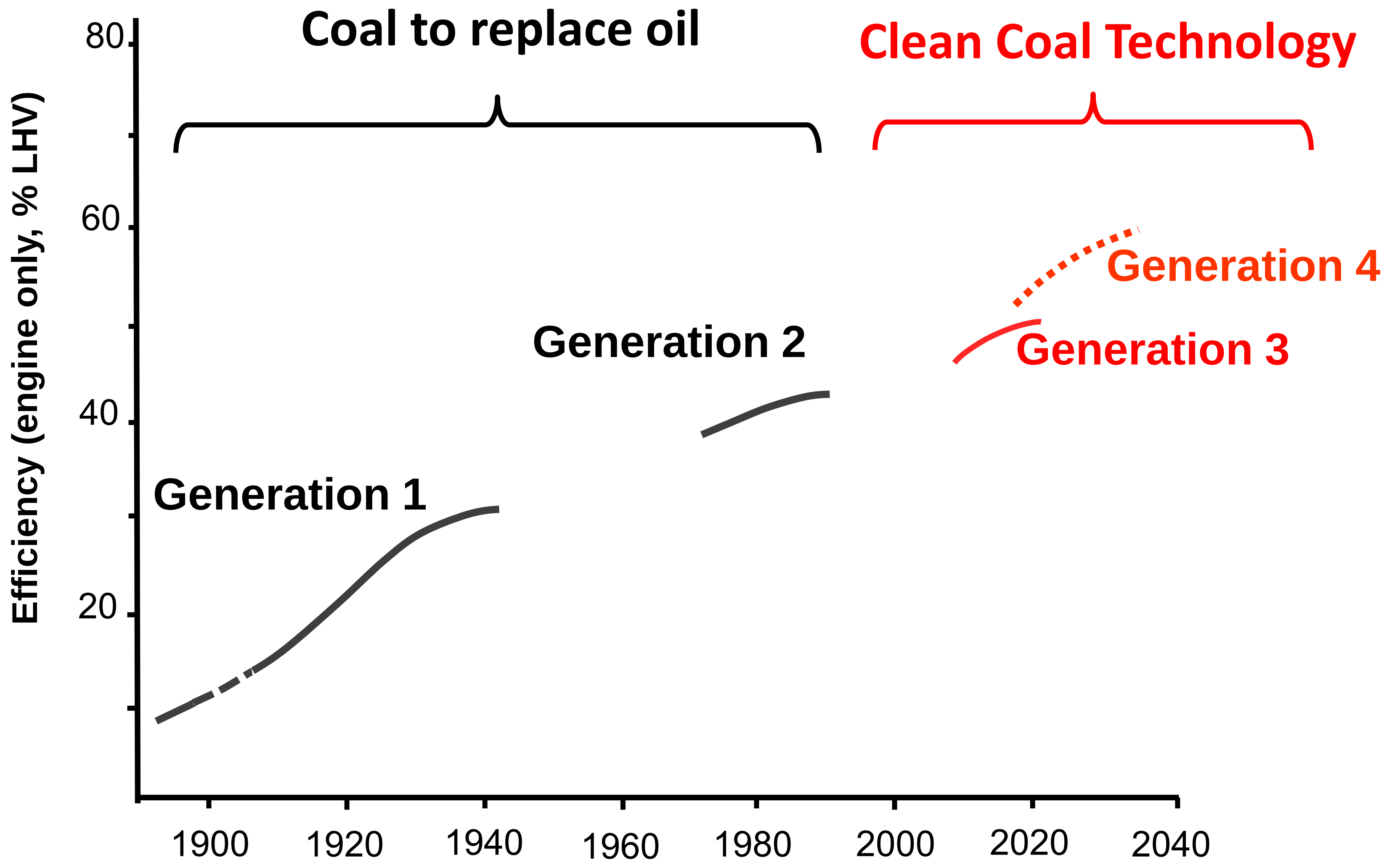
Generation 4
1-20 MW/cyl
optimised/intelligent
paste/slurry firing
biofuel co-firing
CO_2 capture

Distributed power

Distributed generation
large land transport

Base, peak, distributed
large marine transport

… with a changed philosophy



The fuel side … towards commercial production of MRC

Coal water fuel in China – stepping stone to MRC?

Coal water mixtures for boilers

~40 Mtpa in China
5-8% ash, d90 ~150µm
typically 70% coal
2000 mPa.s @ 100/s

Micronise and float

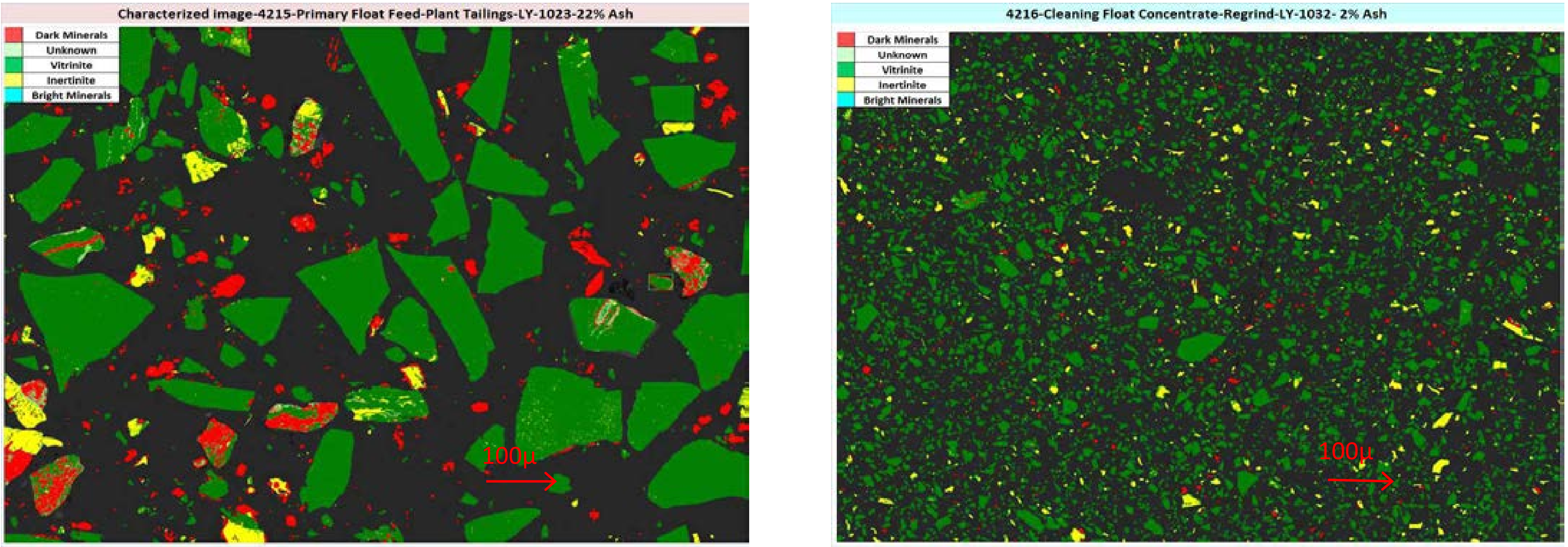
MRC for DICE

nominal specifications
1-2% ash, d90 ~40µm
typically 55% coal
<300 mPa.s @ >200,000/s

MRC is based on ultra-fine coal beneficiation

- Physical cleaning – coal structure not changed, but comminution is needed for optimal liberation/improved flotation response
  - potential product <1% ash, including from tailings
- “Old school” thinking typically regards ash contents below 2-3%, as both technically and economically unviable because
  - “inherent ash“ of coal is usually regarded as the lowest achievable ash content
  - lower ash requires milling to impractical ultra-fine sizes for liberation
  - … but flotation of ultra-fine coal is problematic requiring higher reagent dosages
  - fine coal concentrates are inevitably high in moisture (> 35%) which means costly dewatering and/or drying to produce saleable products
- All of these factors are incorrect, or at best very misleading, as recent research/pilot plant tests have shown

Excellent liberation by micronising



Feed

After liberation by milling

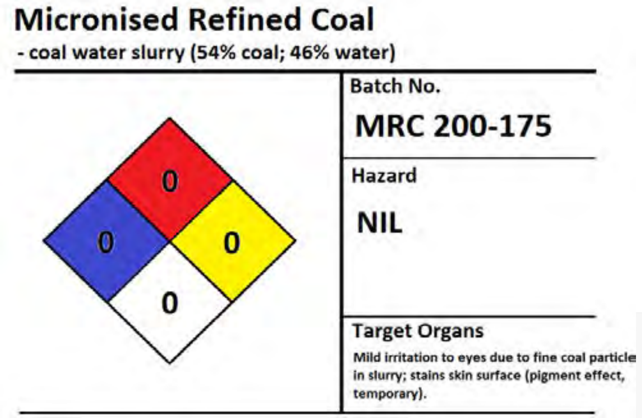
Characterised images for raw coal tailings feed compared with final concentrate (QCAT-CSIRO)

Micronising and sub-50µm coal flotation – excellent separation and recovery

Courtesy Glencore Technology

Pilot scale production of MRC for MAN Diesel & Turbo

1. MRC cake production at Bulga Pilot Plant *(Glencore)*
2. Formulation & rheology trim *(CSIRO)*
3. Certification *(ALS)*



Micronised Refined Coal

- coal water slurry (54% coal; 46% water)

Batch No.

MRC 200-175

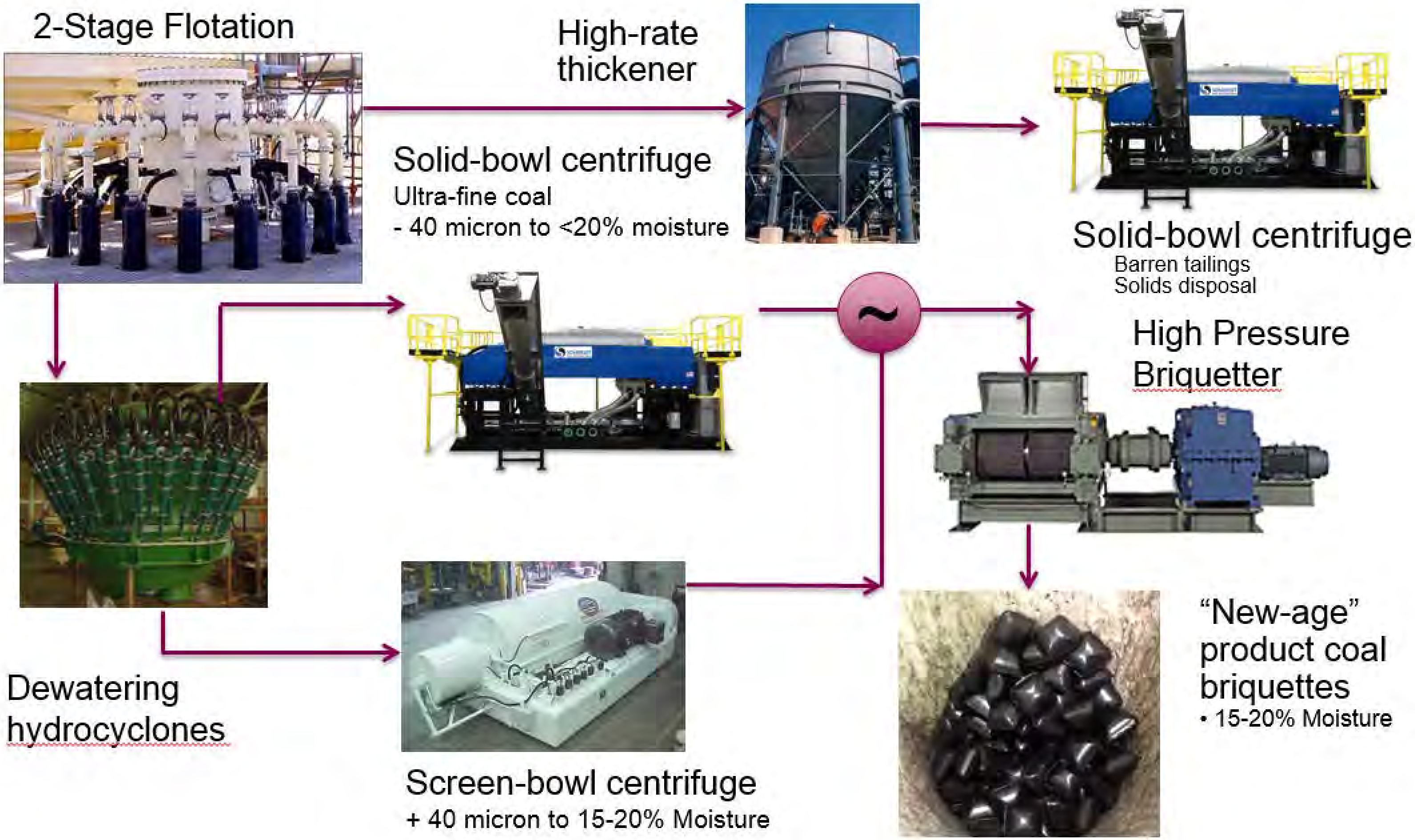
Hazard

NIL

Target Organs

Mild irritation to eyes due to fine coal particle in slurry; stains skin surface (pigment effect, temporary).

MRC … strategic part of a bigger picture of “Premium Coal Products”



2-Stage Flotation

High-rate thickener

Solid-bowl centrifuge
Ultra-fine coal
- 40 micron to <20% moisture

Solid-bowl centrifuge
Barren tailings
Solids disposal

High Pressure Briquetter

Dewatering hydrocyclones

Screen-bowl centrifuge
+ 40 micron to 15-20% Moisture

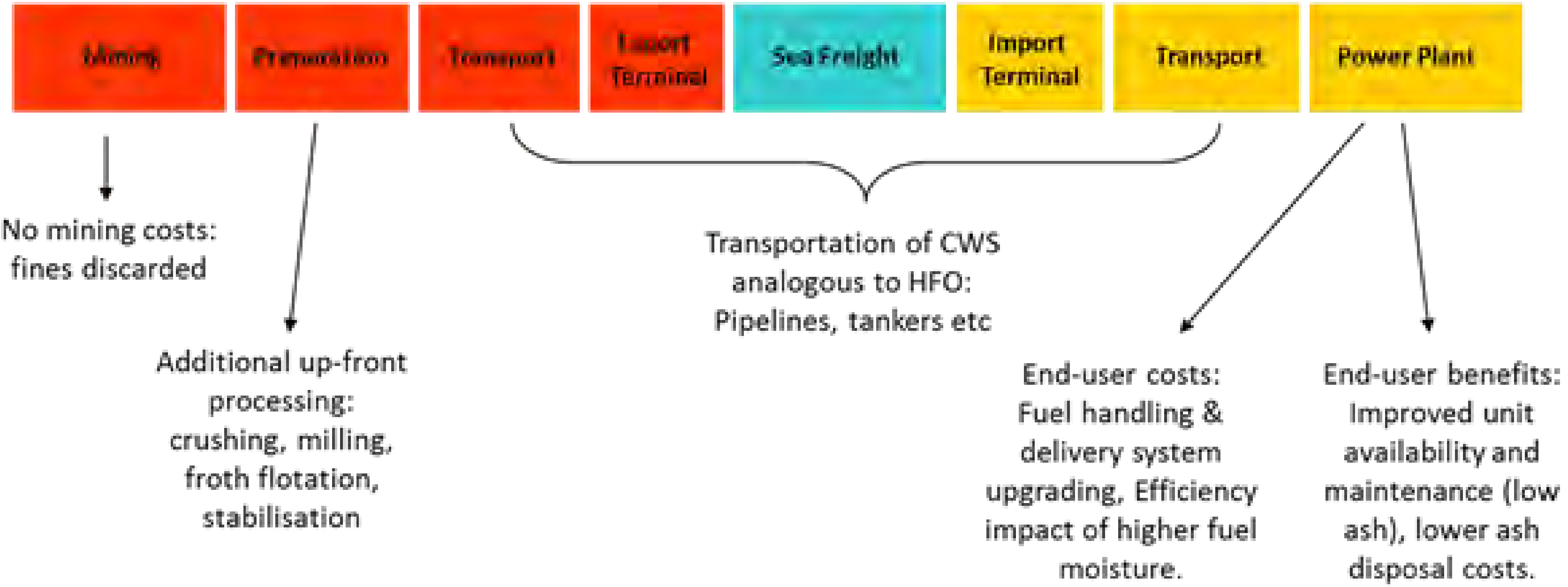
“New-age” product coal briquettes
- 15-20% Moisture

Context

Coal products	Conventional	Steaming
		Coking
“Premium” (<3% ash)		Coal water fuel
		Electrode carbon
		Micronised refined carbons for DICE (MRC)
Electricity generation	Conventional	Pf
	Alternative	IGCC
	Emerging	**Direct injection carbon engine (DICE)**
	Novel	Direct carbon fuel cell (DCFC)

… and an alternative Coal Supply Chain

- Current Coal Supply Chain hampered by an inability to dewater and efficiently transport fine coal
- Innovative approach - recover and use all the ultra-fines as coal water slurry thereby recovering potential “lost coal” creating higher yield and lower cost/tonne



Mining | Preparation | Transport | Export Terminal | Sea Freight | Import Terminal | Transport | Power Plant

No mining costs: fines discarded

Additional up-front processing: crushing, milling, froth flotation, stabilisation

Transportation of CWS analogous to HFO: Pipelines, tankers etc

End-user costs: Fuel handling & delivery system upgrading, Efficiency impact of higher fuel moisture.

End-user benefits: Improved unit availability and maintenance (low ash), lower ash disposal costs.

The engine side … towards commercial DICE

Recent developments

www.dice-net.org

CSIRO DICE program since 2008

- **de-risking based R&D** program (Yancoal, Exergen, Newcrest/JGC, BCIA, Ignite Energy Resources, Xstrata)

MAN Diesel & Turbo have taken a lead position in DICE development

MAN Diesel

Umbrella organisation established to facilitate DICE development internationally

- 17 participants includes MAN, RWE (Germany), JGC (Japan), Sinarmas (Indonesia), Exergen, Ignite Energy Resources, BCIA, Energy Aust, AGL, Newcrest, Yancoal, Worley Parsons, GHD, ACALET and CSIRO

Recent interest from groups in Korea and China

Stage-gated development

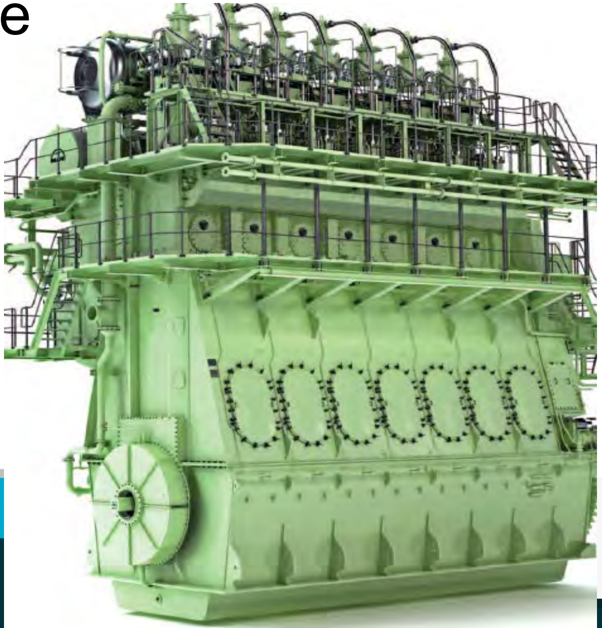
2014-16 Small scale demonstration, initial demonstration/validation DICE,1MW single cylinder (brown and black coals)

Joint funding with BCIA, ANLEC, CSIRO and industry

2016-17* Development/design of components for prototype engine

2017-19* Full scale demonstration MRC production with a 12-30 MW prototype engine for 8000h campaign

2020*- First commercial DICE power plant [$1.4-2 M/MW] possible given appropriate funding support



**** based on MAN D&T estimates of 3-5 years for engine dev***

DICE deployment strategy

DICE favoured when natural gas price >$6-7/GJ forecast to occur by 2020 for most countries

- Australia; $5-6/GJ; forecast >$8.50/GJ in 2020
- China (import); $13.70/GJ; forecast >$10/GJ in 2020
- Europe; $10.80-12.20/GJ; forecast >$8.80/GJ in 2020
- UK; $10-14.60/GJ; forecast >$10/GJ in 2020
- Japan/Korea; $14.20-16/GJ; forecast >$13/GJ in 2020
- Limited incentives in USA (low cost gas,1100 lb CO_2/MWh regulations)

DICE suitable for new coal capacity, and to replace old capacity nearing the end of its economic life (or as it becomes socially unacceptable)

Final comments

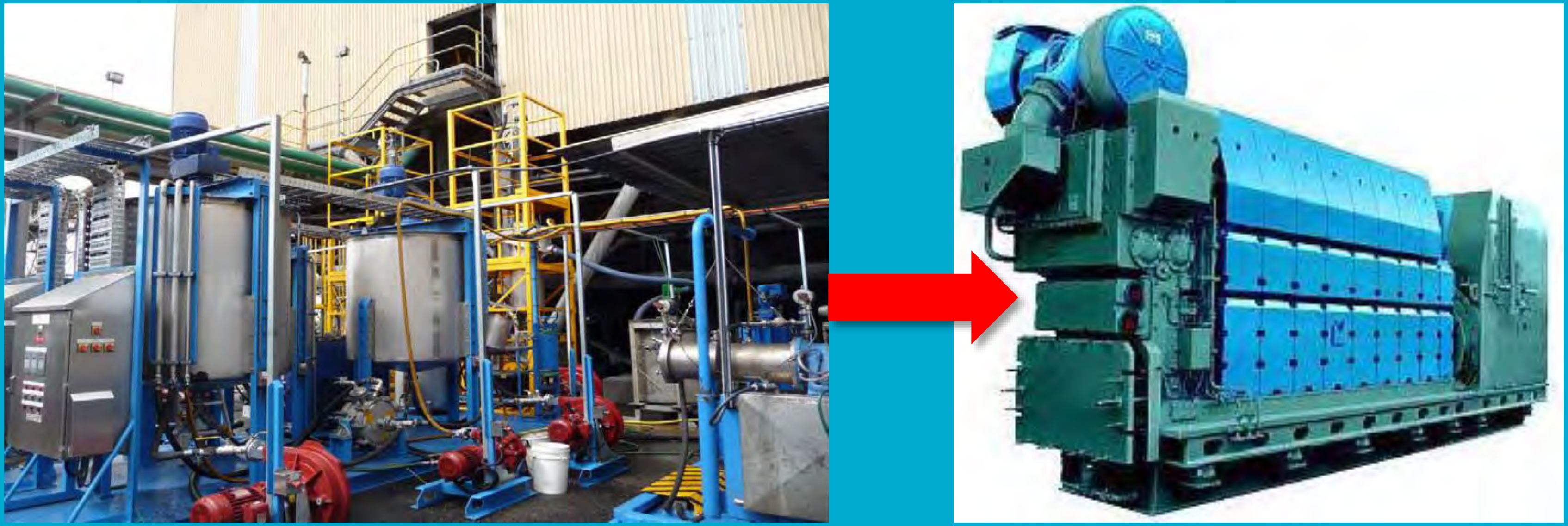
1. DICE could provide coals with a innovative step technology to increase its cost competitiveness and environmental acceptance
2. Barriers to commercialisation are mostly engineering
    - adaptation of commercial process & engine technologies
3. Rapid development possible - can be demonstrated at commercial scale at a relatively small cost
    - short lead time between technology development & implement
4. Logistical barriers to commercialisation of the fuel cycle needs broad intra-industry support
    - as part of premium coal products for maximum benefit

Energy Technology
Louis Wibberley
Leader DICE Development Program

t +61 2 4960 6050
e louis.wibberley@csiro.au
w www.csiro.au/lorem

Premium coal fuels with advanced coal beneficiation

Louis Wibberley - CSIRO
Dave Osborne – Somerset International



Premium coal fuels with advanced coal beneficiation

Contents

Introduction

- Greater resource recoveries are being sought by mine operators to maximize investment returns.
- Current industry trend - “by zero” recovery of fine coal to maximize resource yield and minimize environmental footprint.
- Fine coal size fraction faces the greatest barriers towards qualification as a product component.
- However, advances in liberation, flotation and dewatering create new opportunities for thermal coal operations. So…………

What are the historical barriers and how are they being overcome?

Why have advanced processing options now become viable?

Ultra-fine Coal Beneficiation

Two distinct approaches:

1. Chemical cleaning – coal structure is changed via chemical decomposition – potential is <0.2% ash residue.
2. Physical cleaning – coal structure not changed, but comminution may be applied for liberation – potential is <1% ash residue.

“Old school” thinking typically regards ash contents below 2-3%, as both technically and economically unviable because:

- “Inherent ash "of coal is usually regarded as the lowest achievable ash content
- Lower ash requires milling to finer particle size to increase liberation
- Flotation of ultra-fine coal can be problematic often requiring higher reagent dosages
- Fine coal concentrates are inevitably high in moisture (> 35%) which means costly dewatering and/or drying to produce saleable products.

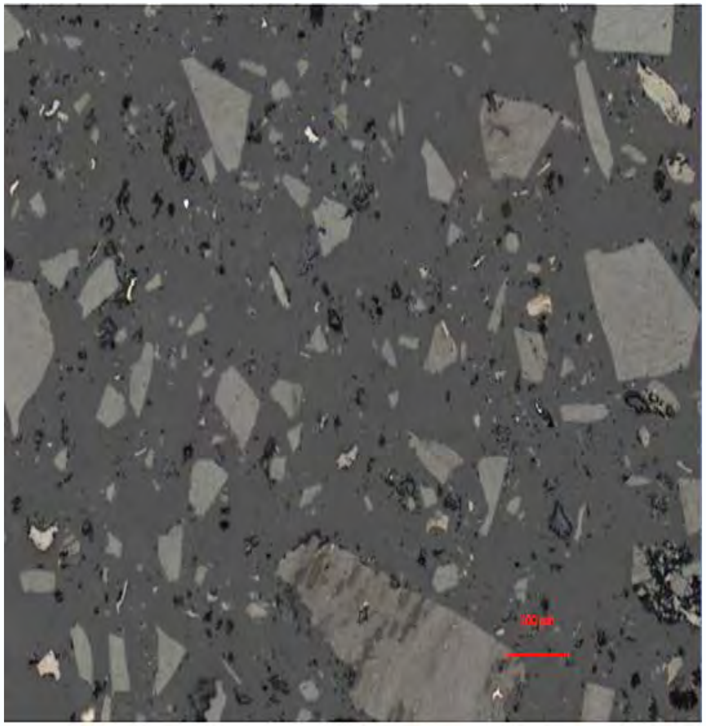
Chemically Cleaned Coal

- Caustic leach process has been successfully demonstrated in Australia.
- Similar to the well-proven Bayer alumina process and also the AMAX 2-stage leach process developed with US-DOE funds in the mid-1980s.
- Ultra-low ash residue <0.2%
- Uses include slurry fuel or briquettes
- Costly option difficult to justify in current climate.

Ultra-clean Coal pilot-plant, Cessnock, NSW (Courtesy UCC Energy Pty Ltd.)

Liberation - Inherent ash constraint

CSIRO has developed an optical reflected light microscopy system for assessing coal petrography samples.



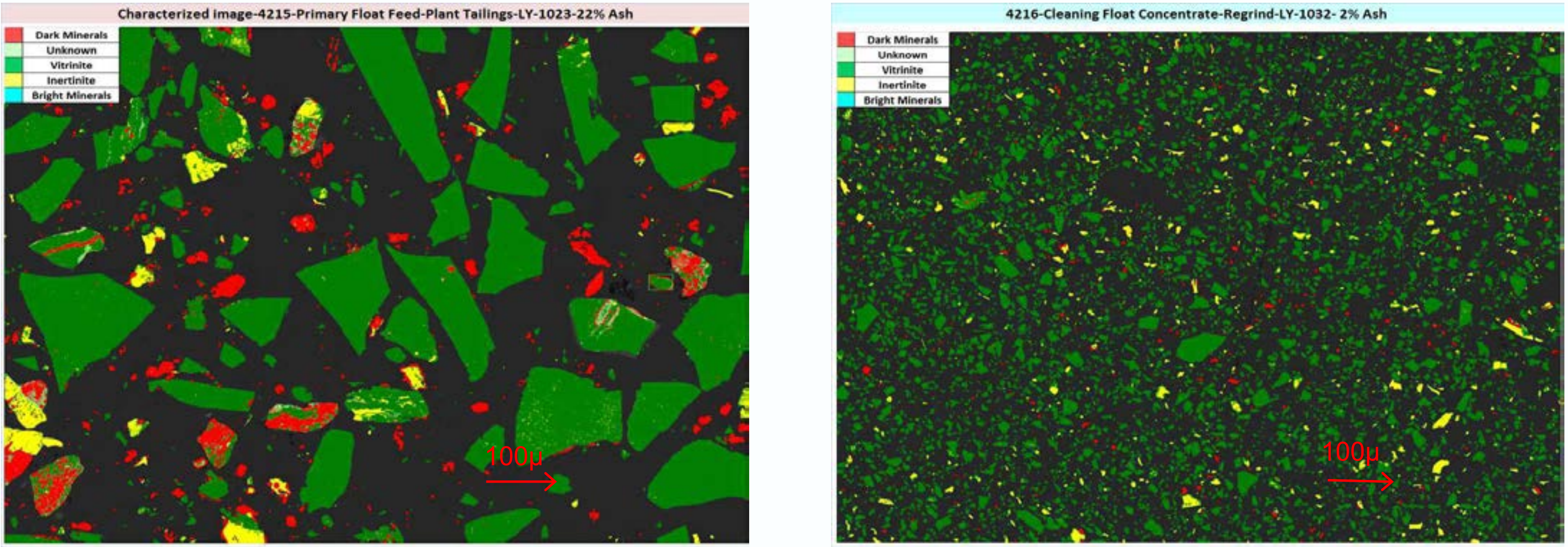
System collects and creates mosaic images so that quantitative information can be obtained on individual coal grains, i.e. **Coal Grain Analysis** (CGA).

CGA generally requires only small representative sub-samples of <1mm material.

Size and compositional information, i.e., **macerals** vitrinite, inertinite, liptinite and **minerals** can be determined for each particle.

Information can then also be used to estimate % mass, density & “ash” value of each particle.

Liberation – CGA Images Confirm Status



Characterised images for Raw Coal Tailings Feed compared with Final Concentrate. (Courtesy QCAT-CSIRO)

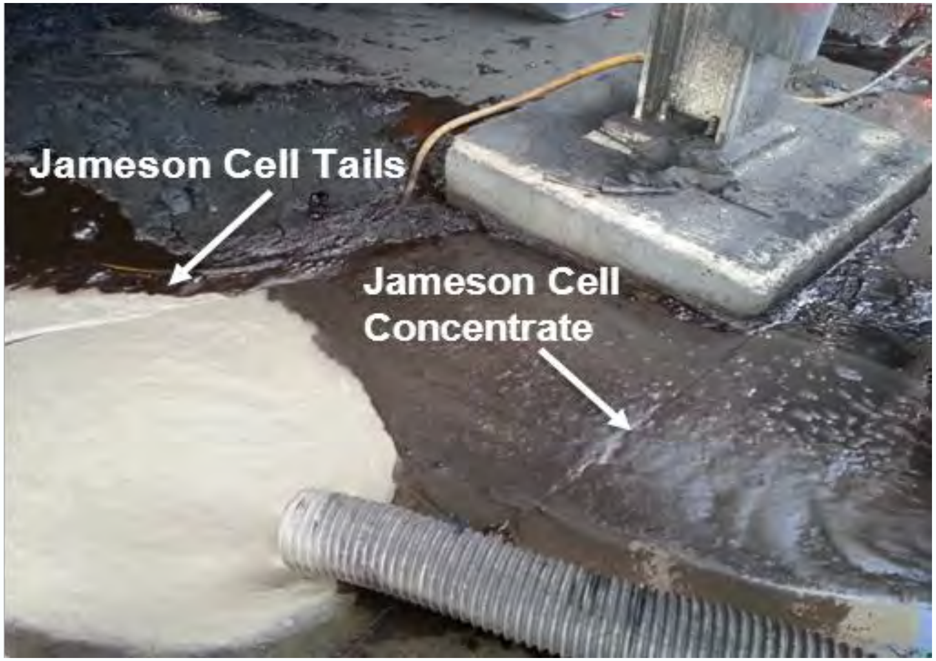
Flotation/Milling process approach

A 1tonne/h pilot-plant, owned and operated by Glencore Technology (formerly Xstrata) comprising

- two Jameson flotation cells,
- a small IsaMill, and a
- membrane filter press, etc.,
- Located at a large thermal coal operation in the Hunter Valley.

Currently testing freshly generated raw coal tailings to produce coal water slurry fuels

Milling and Sub-50µ Coal Flotation



NSW Coal Water Slurry Fuel (CWSF) Pilot Plant

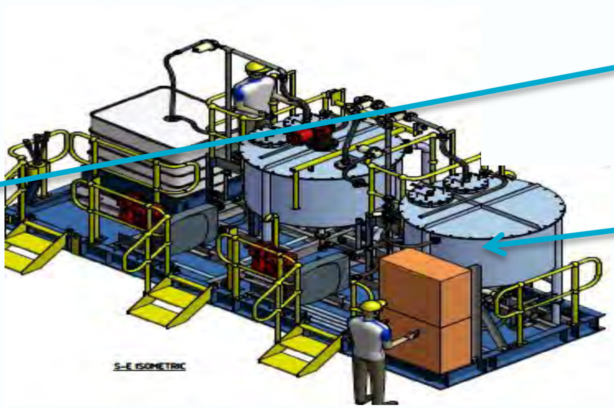
- Successfully produces stable Coal Water Slurry Fuel (CWSF) from coal tailings
- CWSF can then be further refined to produce very low ash (<1% ash) Micronized Refined Coal (MRC)
- MRC produced from 2011 - 2015 for diesel engine tests
- Process information obtained also used for design of CWSF modules including a package plant and fuel handling rig.

Pilot Plant Fuel Production

Jameson cells

Flotation rig

Isa Mill



Fuel Preparation

Fuel Delivery

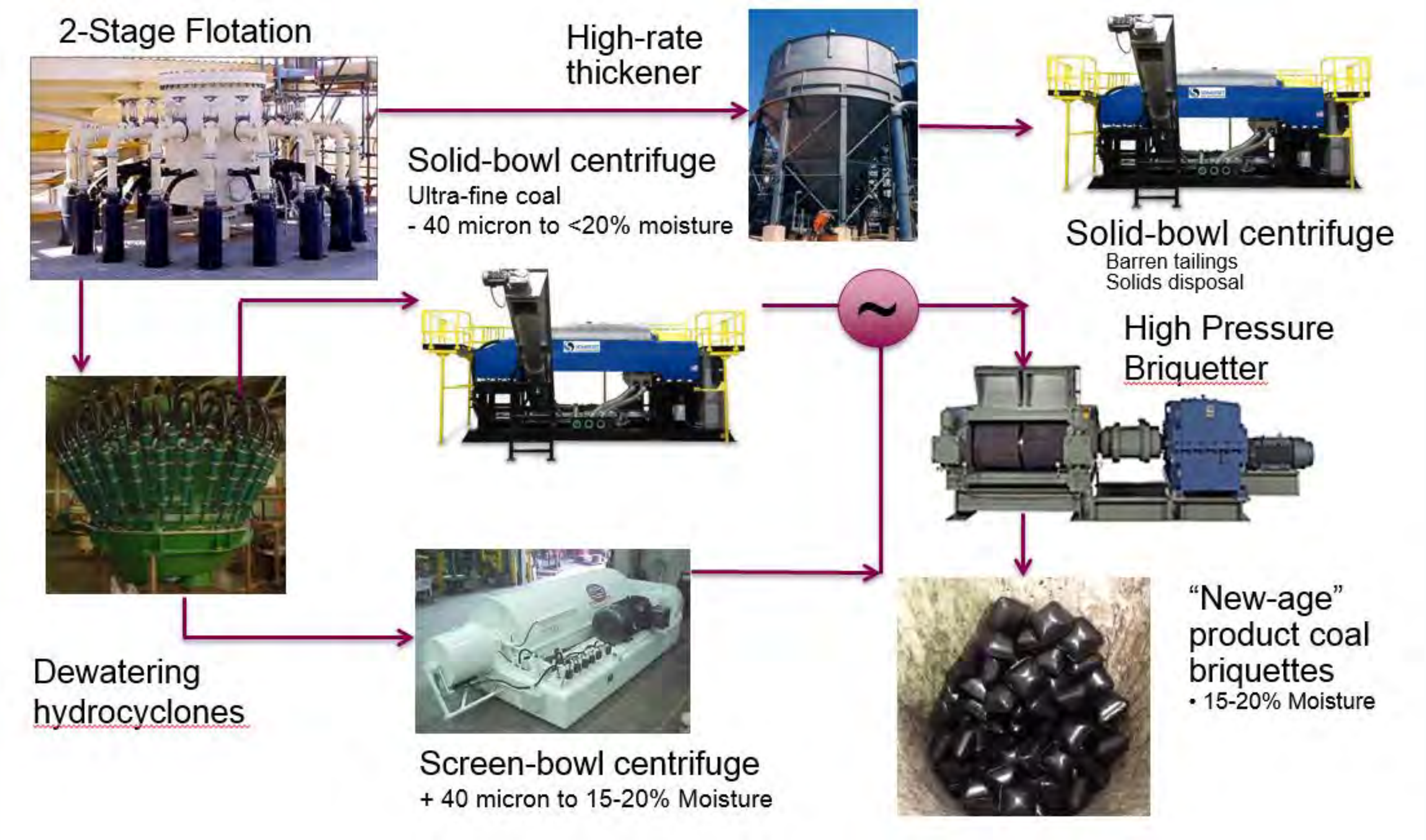
Dewatering Technology

Equipment	Throughput (dry solids)	Product Moisture (% w/w)
Horizontal belt filter	50-130 t/h	20-30
Screen bowl centrifuge	20-60 t/h	16-27
Centribaric centrifuge	15-20 t/h	15-20
Vacuum disc filter	50-150 t/h	20-32
Hyperbaric disc filter	30-150 t/h	17-25
Solid bowl centrifuge	10-18 t/h	18-25
Membrane Filter press	15-30 t/h	14-32

Ultrafines

Key objective is to achieve a cake moisture of 20% or lower for ultrafines

Emerging Coal Fines Treatment Circuit



2-Stage Flotation

High-rate thickener

Solid-bowl centrifuge
Ultra-fine coal
- 40 micron to <20% moisture

Solid-bowl centrifuge
Barren tailings
Solids disposal

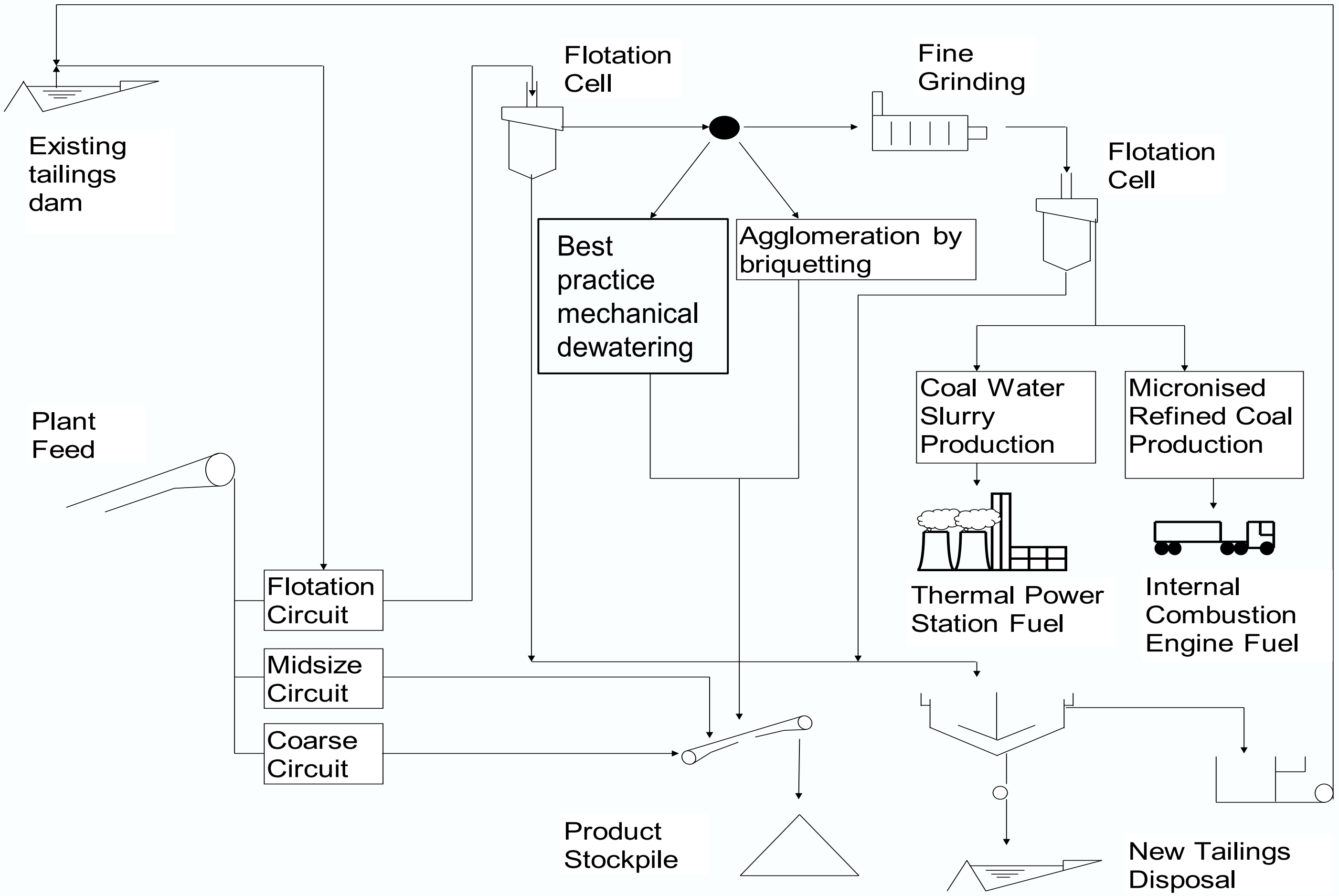
High Pressure Briquetter

Dewatering hydrocyclones

Screen-bowl centrifuge
+ 40 micron to 15-20% Moisture

"New-age" product coal briquettes
- 15-20% Moisture

Wider Fines Treatment Options



Existing tailings dam

Plant Feed

Flotation Circuit

Midsize Circuit

Coarse Circuit

Flotation Cell

Best practice mechanical dewatering

Agglomeration by briquetting

Fine Grinding

Flotation Cell

Coal Water Slurry Production

Micronised Refined Coal Production

Thermal Power Station Fuel

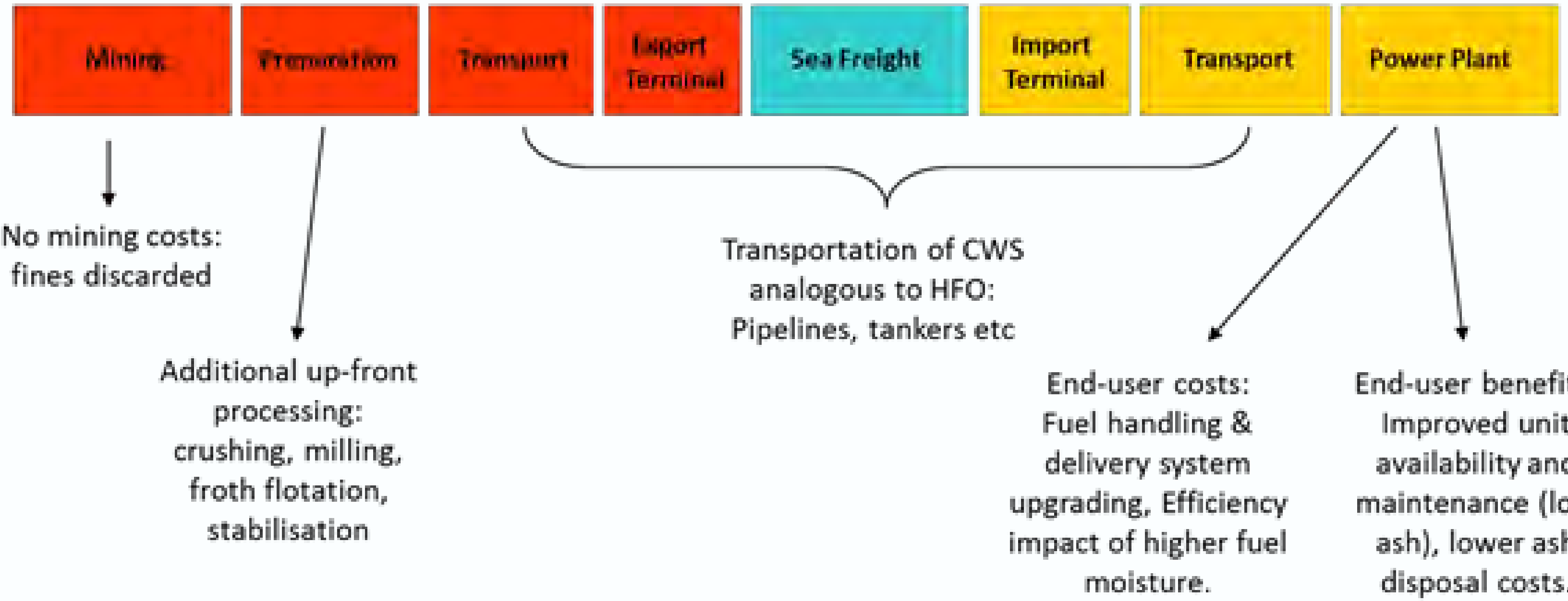
Internal Combustion Engine Fuel

Product Stockpile

New Tailings Disposal

Coal Supply Chain

- Current Coal Supply Chain (CSC) has been hampered by an inability to dewater and efficiently transport fine coal.
- Innovative approach - recover and use all the ultrafines via coal-water slurry thereby recovering potential "lost coal" creating higher yield and lower cost/tonne.



Mining

Preparation

Transport

Export Terminal

Sea Freight

Import Terminal

Transport

Power Plant

No mining costs: fines discarded

Additional up-front processing: crushing, milling, froth flotation, stabilisation

Transportation of CWS analogous to HFO: Pipelines, tankers etc

End-user costs: Fuel handling & delivery system upgrading, Efficiency impact of higher fuel moisture.

End-user benefits: Improved unit availability and maintenance (low ash), lower ash disposal costs.

Innovative Coal Supply Chain

New Low-cost Fuel

- Coal-water slurry fuel (CWSF) at ~70% solids prepared from coarser (bi-modal) particle size distribution (p80 of 0.075 mm)
- Use for direct firing to boilers as a potential replacement for heavy fuel oil (HFO), or partial replacement for Pulverized Fuel (PF)
- Transport as slurry fuel - avoids sticky, wet or dusting coal problems
- Lower tailings disposal cost - paste-thickening, further dewatering for co-disposal with coarse plant discards and mining waste.

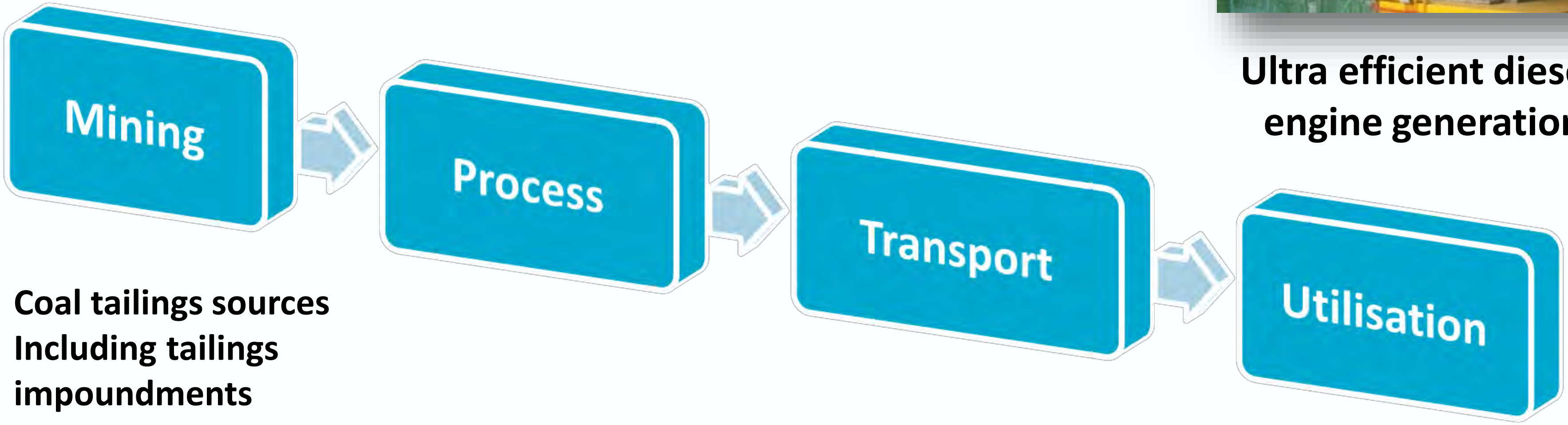
User Benefits

- No further grinding needed, significantly lowering cost
- Major O&M savings and lower ash disposal cost
- Reduced thermal efficiency offset by cost reductions from recovering lost coal from tailings.
- Potential to replace > 30% of the pulverised coal capacity.
- **Value Proposition:** a 1.0 to 1.5 c/kWh saving once the boiler has been converted for CWSF.

Micronized Coal Water Slurry

Optimization of the fuel cycle (DICE)

Ultra efficient diesel engine generation



- Coal tailings sources
- Including tailings impoundments

- Increased grade recovery
- Recovery of ultra fines
- Minimal dewatering

- Road/rail/ship – cake or slurry
- Pipeline coal water fuels
- Higher solids paste for longer distance

- Mine-mouth or centralized
- Distributed generation
- Support of renewables

Conclusions

- “Deep cleaning” via liberation and subsequent beneficiation has offered significant potential downstream improvements, i.e.,
  - maximised resource recovery,
  - minimised transport and handling costs,
  - numerous end-user process improvements,
  - reduced maintenance and wear,
  - lower environmental impacts and
  - other sustainable improvements.
- Ultrafine coal beneficiation has matured via progressive froth flotation improvements
- Dewatering the concentrate was a barrier, but emergence of membrane filter presses, hyperbaric disc filters or high-g decanter centrifuges now offers commercial solutions.
- Briquetting and agglomeration has progressed to machine capacities of up to ~40 ton/h for fine coal applications to improve product handling.
- Manufacture of stable coal-water slurries with > 65% solids and stable micronized slurries with > 60% solids have reached commercial adoption.
- **Scene is now set** for new generation clean coal technologies with minimal wasted energy, lowest ash disposal costs and reduced SO_x, NO_x and CO_2 emission costs.

Acknowledgements

The authors acknowledge the contributions provided by

- Glencore Technology,
- Glencore Coal,
- Yancoal, and the
- CSIRO

towards the preparation of this paper and for permission to include figures and data from other recently published papers and articles on this subject.

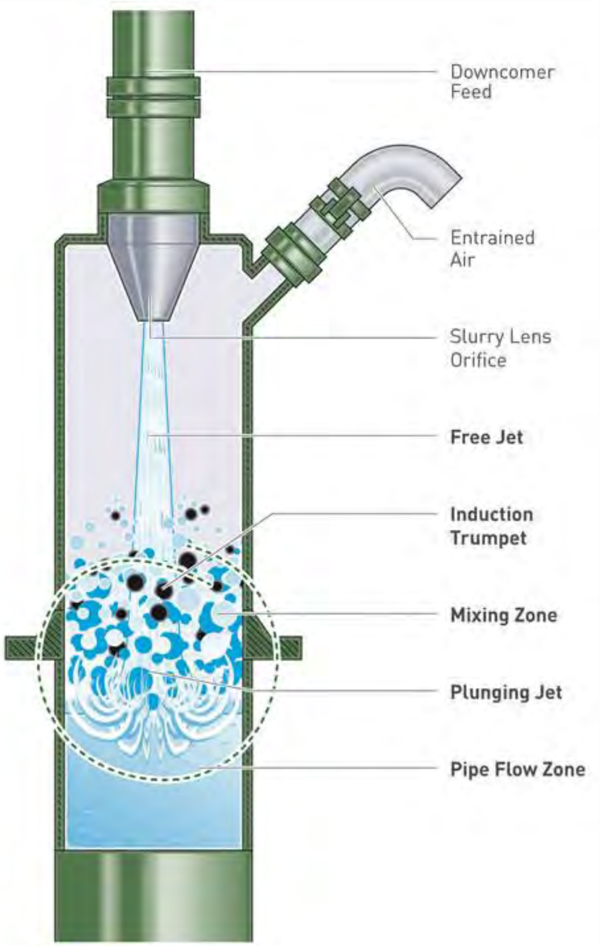
Dave Osborne thanks Somerset Coal International for encouragement and support in attending and participating in the 2015 Clearwater Clean Coal conference.

Thank you for your time

Questions?

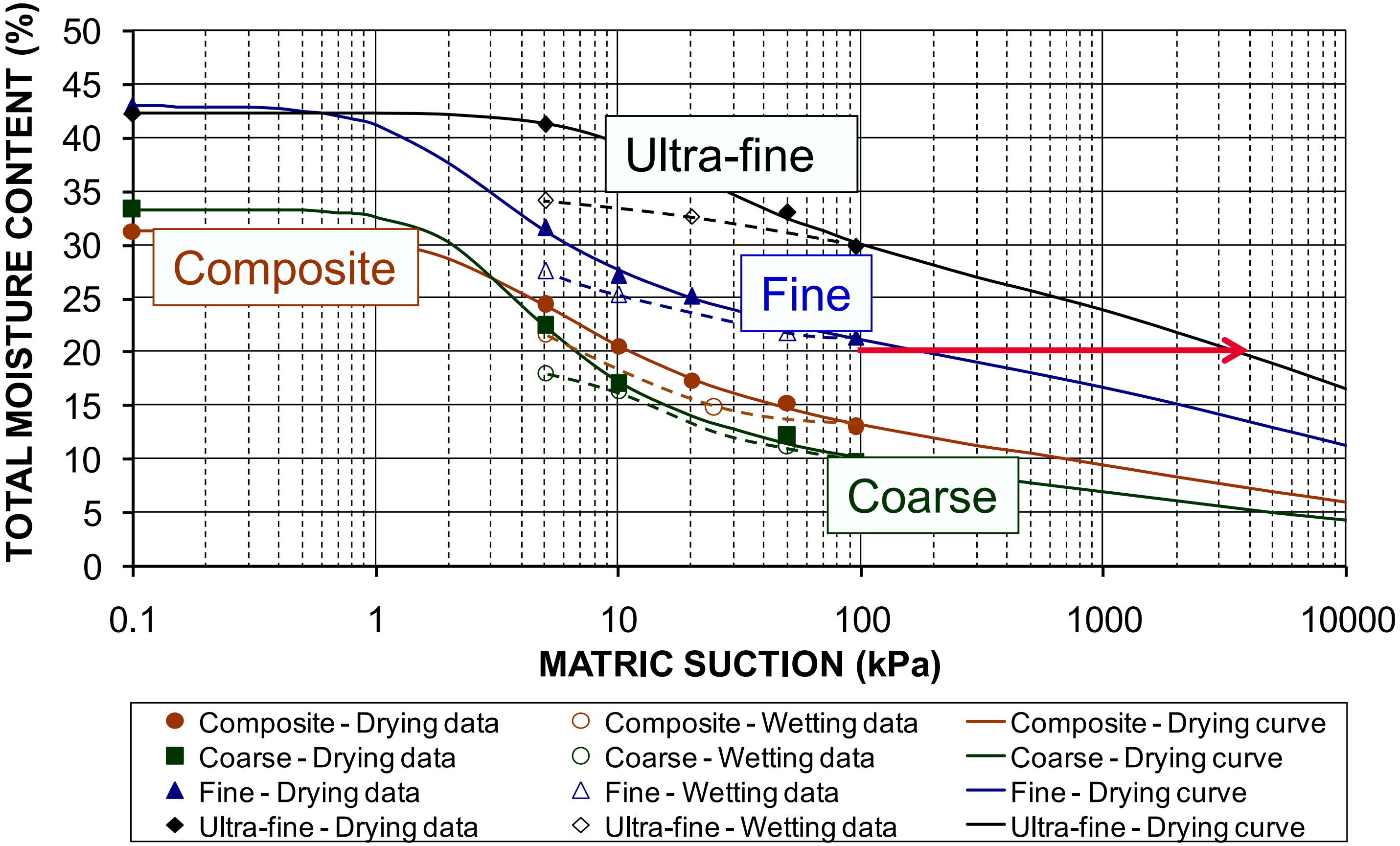
Louis Wibberley Phone: +61 7 3327 4457 Email: Louis.wibberley@csiro.au
Dave Osborne Phone: +61 7 3010 9443 Email: dosborne@somersetpty.com

The Jameson Cell - Downcomer



Product Coal Moisture Relationships

- Soil Water Characteristic Curve (SWCC), related to pore size distribution, in turn related to Particle Size Distribution[1].
- Matric Suction is related to Applied Pressure



---

[1] Source: Prof David Williams; Univ Queensland - D.Williams@uq.edu.au

Coal Water Slurry Fuel

Comparison of Coal Supply Chain Costs for Electricity Generation

Mining	Preparation	Transport	Export Terminal	Sea Freight	Import Terminal	Transport	Power Plant
Conventional Coal			**FOB 80 $/t**	**CIF 90 $/t**			**Total Costs = 144 $/t**
60 $/t	5 $/t	10 $/t	5 $/t	10 $/t	5 $/t	5 $/t	44 $/t
							Total Costs = 5.5 c/kWh
2.0 c/kWh	0.2 c/kWh	0.5 c/kWh	0.2 c/kWh	0.5 c/kWh	0.2 c/kWh	0.2 c/kWh	1.6 c/kWh
Coal Water Slurry			**FOB 36 $/t**	**CIF 47 $/t**			**Total Costs = 105 $/t**
	20 $/t	11 $/t	5 $/t	11 $/t	5 $/t	6 $/t	47 $/t
							Total Costs = 4.0 c/kWh
	0.8 c/kWh	0.5 c/kWh	0.2 c/kWh	0.5 c/kWh	0.2 c/kWh	0.2 c/kWh	1.7 c/kWh

paper 7A

The Future of Thermal Coal Flotation

F. Mercuri[1], D.G. Osborne[2] and M.F. Young[2]
[1] Benefish Consulting
[2] XT

ABSTRACT

The drive toward greater resource utilisation, and increased yields, has placed the issue of thermal coal flotation in focus of the industry. Historically, the opportunities for the flotation of non-coking coal have been limited based on the net energy gain achieved by the beneficiation and dewatering technologies available at the time. Recent advances in coal liberation, flotation and dewatering have created an opportunity to increase resource yield by the addition of a clean, fine coal component whilst maintaining or potentially increasing the energy content of the final product.

Increased focus on plant efficiencies, and the sustainability of operations in general, has identified opportunities once deemed uneconomic or inherently not viable as plausible improvement options. Cost pressures are heightening the need to maximise energy recovery from existing operations. The industry as a whole is now realising that the most economical source of adding more value is via the recovery of the fine coal currently being lost to tailings. Such losses have in the past been largely due to inefficient plant designs and/or unsuitable flowsheets that are not capable of providing for optimisation of the resource.

This paper investigates the industry trend towards thermal coal flotation and the drivers and technology responsible for this shift. The techniques used to exploit this opportunity will be investigated in detail and case studies will be presented.

THERMAL COAL PROCESSING

History of Thermal Coal Processing

Thermal coal beneficiation is generally governed by 3 sets of major parameters:

- coal washability
- particle size distribution
- marketing specification.

Thermal coal is generally washed to an energy target (or ash value, which is directly related), and this can be affected by changes in the above parameters throughout a mine's life. The aim of thermal coal beneficiation is to maintain as large a particle size as possible in the beneficiation phase, whilst achieving adequate liberation and optimal yield. This has typically been achieved historically with dense medium processes including drums, baths and cyclones or in an earlier generation of plants, via jigs alone. Fine material (nominally -1 mm) is typically processed through water-based gravity separation devices predominantly spirals, teetered bed separators (TBS), or more recently, reflux classifiers (RC). Material -0.1 mm is typically discarded to tailings due to low energy contribution to the final product and cost of processing.

The mine operator has certain controls at his disposal to optimise output whilst meeting the required product quality. These are:

- density cut-point (D_{50}) control
- blending
- bypass of seam types or particle size fractions directly to product or waste.

Due to the unit process efficiencies available at the time, moisture content has generally precluded the introduction of the flotation process from being viable due to a net product energy loss created by the addition of the concentrates with higher clay and moisture contents, as evident in a typical Hunter Valley thermal coal. Technology advances have allowed yield and resource optimisation to improve over the years. These have included improvements in both processing and dewatering via the avenues of process control, plant stability, maintenance programs and new wear materials.

Historically Accepted Losses

The traditional thermal coal flowsheet typically routes desliming cyclone overflow directly to the tailings circuit. No attempted recovery is undertaken and the material is simply thickened and then pumped to a site tailings storage facility. A typical size distribution in the Hunter Valley can contain 5–10% of ROM material in this stream. This material routinely contains a high amount of clay but also contains recoverable and saleable coal values. The viability of this stream can no longer be ignored and must be considered for recovery in new and existing operations.

MOISTURE IMPACTS

The level of total moisture contained within a thermal coal is critical to how much energy within the coal is available for customers to generate useable energy. Any energy used to dewater the coal becomes essentially wasted and does not contribute to kWh generation. The presence of fines in a thermal coal product also plays a crucial role in the overall total moisture holding capacity and this is primarily a function of surface area.

To optimise the effect of particle surface area, "bottom size analysis" (Osborne et al, 2014) investigates via detailed simulation, the optimum bottom size for a particular resource to maximise saleable product volumes and quality resulting in maximum revenue. This simulation estimates the impact of fine particle surface moisture on the saleable product quality. The simulation shown below (Figure 1) illustrates the interaction between increasing product volumes and increased product total moisture as the bottom size decreases. A critical input to this process is the use of Non-Centrifugal Moisture (NCM) testing (Firth et al, 1998).

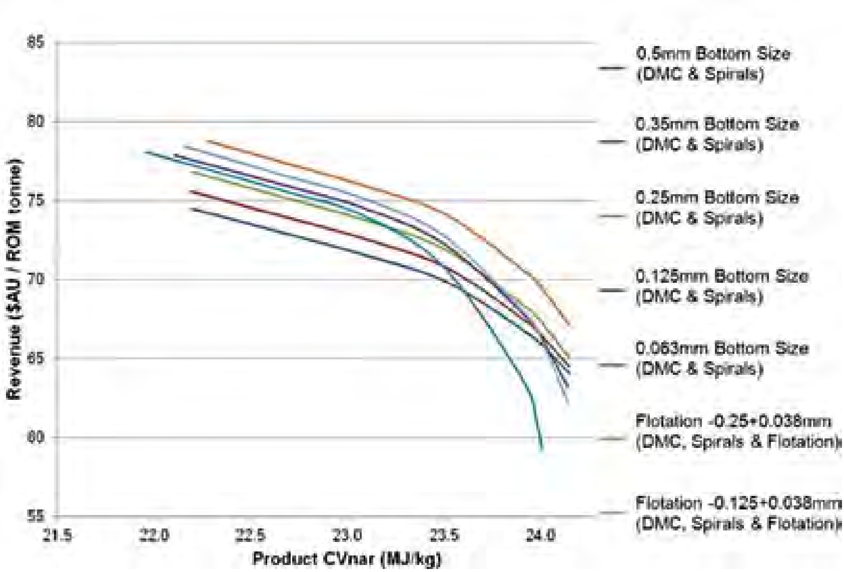


Figure 1 Example Bottom Size Analysis Results (Osborne et al, 2014)

ADVANCES IN FLOTATION TECHNOLOGY – CURRENT INDUSTRY BEST PRACTICE

Pneumatic Flotation

In order to achieve low product ash levels in flotation, pneumatic style flotation technologies are generally employed. Technologies including the Jameson Cell, Micro Cell and Pneuflot exhibit features that target low concentrate ashes. The fundamentals of pneumatic flotation are well explained by the operation of the Jameson Cell (Evans, Atkinson and Jameson, 1993). It involves feed slurry being pumped through a restriction to create a high pressure jet which then enters a cylindrical contact chamber where particle contact occurs. The jet of liquid first shears and then entrains air from the atmosphere. Removal of air into the jet causes a vacuum to be generated inside the contact chamber. When a hydraulic seal is formed at the bottom of the chamber, the vacuum causes a column of slurry to be drawn up inside the chamber. The jet of slurry plunges onto the liquid surface and the high kinetic energy of the jet will then disseminate the entrained air into very fine bubbles. In this zone, the high intensity of the system creates a very favourable environment for the bubbles and particles to collide and attach. The air bubbles, coal and mineral particles move continuously down the chamber before exiting into the tank. The particle laden bubbles then float to the top to form the froth whilst the hydrophilic rock and mineral particles remain in the pulp phase to be removed as tailings. To ensure consistent operation, tailings recycle can be employed. This dampens feed fluctuations to the cell allowing the downcomer to operate at a constant feed pressure and flowrate.

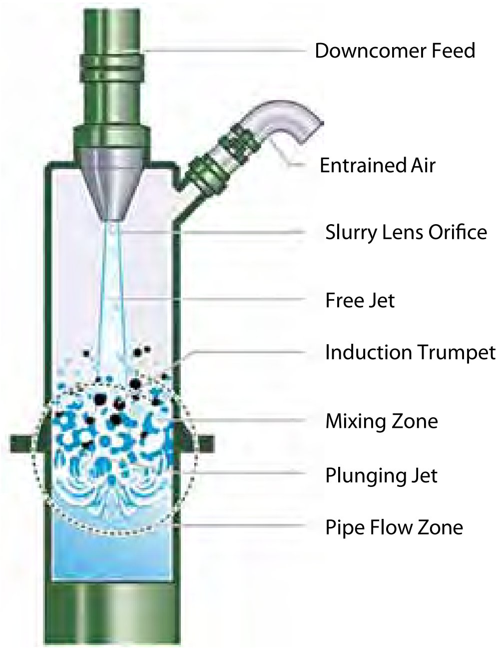


Figure 2 Jameson Cell Downcomer (Courtesy XT)

This technology employs rapid kinetics so the flotation vessel volumes tend to be very much smaller than that for mechanical and column cells of equivalent capacity leading to economic advantages.

Froth Washing

A flotation technology that has froth washing capabilities has advantages over those that do not. Froth washing is one of the most effective methods employed to reduce entrainment thereby enabling concentrates with the lowest ash (and highest energy) value to be produced. The amount of wash-water used is therefore an important process variable, i.e., the system must be able to be operated over the desired flow-rate range for a flotation cell of a particular size. This must be based on the designed concentrate tonnage range and solids content of the concentrate.

In practice, the wash-water addition is dependent on the process operation and in particular factors such as the structure and stability of the froth and the required ash value target. This in turn is influenced by factors such as particle size and hydrophobicity of the particles recovered in the froth. A stable froth allows froth washing to be effective in producing a clean concentrate without having a detrimental effect on combustibles recovery.

Bubble Size

Bubble size is one of the most important factors in any flotation system as it has a strong influence over flotation kinetics. Fine bubbles increase the flotation kinetics across all particle sizes (Diaz-Penafiel and Dobby 1994, Ahmed and Jameson 1985), and not just recovery of fine particles as has often been hypothesised. In coal flotation, fine bubbles also improve separation as they intensify the difference in the kinetics of the coal from non-coal particles, allowing concentrates with lower ash value to be produced without loss in yield.

Carrying Capacity

Bubble size dictates the carrying capacity of a flotation machine. Finer bubbles increase the carrying capacity (often measured as the mass flow rate (t/h) of concentrate per m^2 of surface area of the flotation machine) as there is more bubble surface area per volume of air added for particles to attach. Essentially, this means that if two different types of flotation machine are used to float the same coal using the same amount of air, the one generating air bubbles (measured as a distribution) that are half the mean-size of the other machine will have four times the bubble surface area available for flotation. This is therefore a very important consideration, especially for coal feeds offering very high potential yields. The more bubble surface area that is available from the machine, the lower will be the cross-sectional area required and fewer cells will be required to recover all the coal.

The air bubbles generated by the Jameson Cell are in the range of 0.3 to 0.7 mm (Sauter mean diameter, D32) (Evans, Atkinson and Jameson, 1993). Figure 3 compares the bubble size of a range of industrial mechanical and columns cells (Nesset, Finch and Gomez, 2007) to that of the Jameson Cell (Osborne et al, 2013). All results collated in Figure 3 were determined by the same bubble size measurement technique as developed by McGill University and described by Chen, Gomez and Finch (2001) and Gomez and Finch (2007).

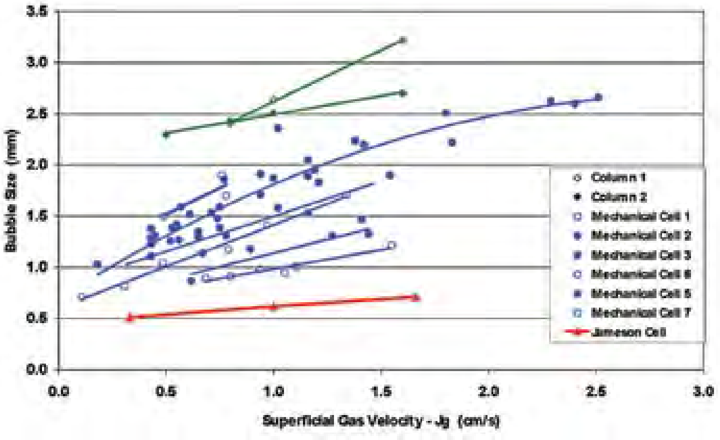


Figure 3 Bubble Size Measurement (Osborne et al, 2013)

PILOT PLANT PERFORMANCE ON A HUNTER VALLEY THERMAL COAL TAILINGS STREAM

A range of different thermal coal types have been tested in a two stage pilot scale Jameson Cell flotation circuit.

These have had qualities ranging from a 30% ash (ad) tailings sample from easily treated coal seams to very high >60% ash (ad) coal seams containing difficult fine clays.

Table 1 Pilot Plant Coal Quality Results by Seam

Seam	Feed Ash % (ad)	Concentrate Ash % (ad)
A	30–35	7–8
B	35–45	8–12
C	>60%	15–16

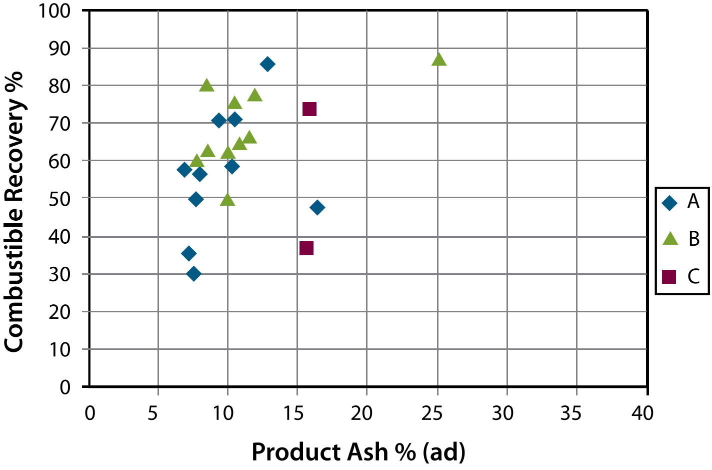


Figure 4 Pilot Plant Results

It can be seen that the Jameson Cell flotation circuit was able to achieve a low ash concentrate of 7–12% (ad) (Figure 4) at up to 70–80% combustible recovery for some feed types with the lower clay content, and 15–17% (ad) ash for coal seams with much higher clay content.

The fine mineral matter and clay in the coal concentrate reduces filtration rate and increases product moisture, by removing more of these particles improved product moisture is achieved. The plant is configured in a rougher-scavenger flowsheet where the tailings from the rougher (primary cell) feeds to the scavenger (secondary cell).The cleaning effect can be seen by the colour of the two product streams (concentrate which is high in vitrinite and tailings which is predominantly kaolinite clay) from the Jameson Cell circuit, shown in Figure 5.

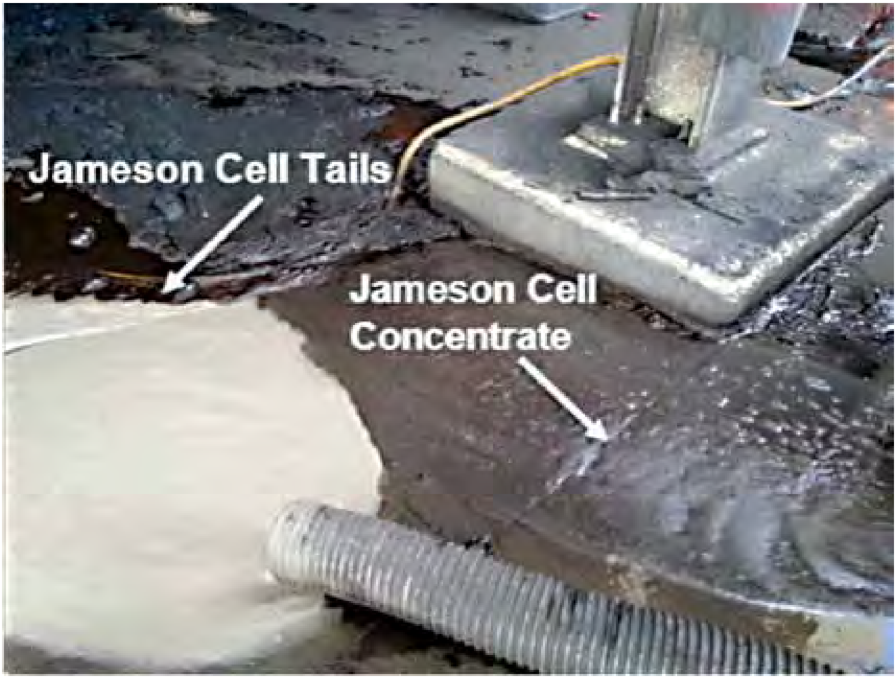


Figure 5 Concentrate and Tailings Streams from Pilot Plant Testwork

DEWATERING TECHNOLOGY

Dewatering options available for fine coal are listed in Table 2. Recent advances in "by zero" fines dewatering has enabled total moistures of flotation product to be reduced to a target whereby it can become viable, from an energy balance perspective, to include flotation concentrates into the final thermal coal product at a greater number of mine sites.

Table 2 Equipment Used for Dewatering Coal (adapted from Bickert's chapter in Osborne 2013)

Equipment	Size	Throughput (dry solids)	Product Moisture (% w/w)	Feed Preparation	Application
High frequency screen	0.6–2.4 x 3 m	10–100 t/h	15–25	Cyclone underflow	Fine coal
Screen scroll centrifuge	0.5–1.5 m dia	45–100 t/h	11–18	Cyclone underflow	Fine coal
Horizontal vacuum belt filter	75–150 m^2	50–130 t/h	20–30	Flocculation	Ultrafine coal
Screen bowl centrifuge	1.1 m dia x 3.3 m long	20–60 t/h	16–27	Thickening	Ultrafine coal
Centribaric centrifuge	1.1 m dia x 3.3 m long	15–20 t/h	15–20	Thickening	Coal slimes
Disc filter	120–200 m^2	50–150 t/h	20–32	Thickening / flocculation	Ultrafine coal
Hyperbaric disc filter	70–200 m^2	30–150 t/h	17–25	Thickening / flocculation	Ultrafine coal
Paste thickening	25 m dia x 6–12 m high	100 t/h	45–55	Flocculation	Coal tailings
Solid bowl centrifuge	1.1 m dia	20–60 t/h	30–45	Thickening	Coal tailings
Belt press filter	3–3.5 m wide	10–20 t/h	25–45	Thickening / flocculation	Coal tailings
Filter press	200–800 m^2	15–30 t/h	14–32	Thickening	Ultrafine coal and tailings

ECONOMICS

The economic argument for pursuing fines recovery is compelling. Technology advances allowing ultrafine product to be included in final product streams without penalising product quality is supported by the simplistic revenue scenario below.

Raw plant feed containing 10% passing 0.1 mm.

Yield of 50%.

For a thermal coal operation of 16 Mt/y ROM = 1.6 Mt/y of raw feed currently sent to waste.

Assuming a nominal 50% yield equates to 0.8 Mt/y of potential saleable product.

At a Newcastle benchmark price of US$86/product tonne **= US$69M revenue loss per annum** (not accounting for freight, port, tonnage adjustments, etc).

Including conservative capital and operating costs produces a very attractive investment opportunity.

Table 3 Economic Evaluation of Brownfields Flotation Installation

Capex $	Opex $/feed t	Rate t/h	Direct Costs $/t	Tax rate %	Discount Rate %	NPV $	IRR %	Pay back yrs
50M	15	230	30	30%	10%	78M	51	2.5

FUTURE DIRECTION – REDUCING PRODUCT ASH AND MOISTURE

The challenge of cost-effectively recovering a saleable fines component from tailings has been with us for many years and periodically an apparent solution emerges. XT has been operating a pilot-plant as described earlier and this plant incorporates the combination of fine grinding and Jameson Cell flotation technology for the preparation of coal-water slurry fuel (CWSF). The inclusion of a fine grinding stage enables slurries to be prepared whereby the non-value components are liberated from the carbon material thereby facilitating recovery of a highly concentrated ultrafine, low ash coal product.

Figure 6 1.0 t/h Coal Tailings Treatment Pilot-plant

This combination has already been proven capable of achieving very good combustible recoveries (material dependent, but normally over 90%) for coal from the raw tailings stream. The milling step reduces particles down to a p80 of 0.015 mm enabling enhanced flotation recovery by increased liberation and the formation of fresh surfaces on the ultra-fine coal particles. Further enhancement of the addition of improved ultrafine dewatering of the flotation concentrate using a membrane filter press or equivalent and a "fit-for-purpose" mixing system has resulted in the preparation of highly stable slurries with solids concentrations over 60% (w/w). This fuel has been specifically prepared for use in direct coal injection engines (DICE) to replace diesel fuel.

Other coal-water slurries with a slightly coarser particle size distribution (p80 of 0.075 mm) can be prepared in a similar way to create a coal-water slurry fuel (CWSF) at over 70% solids that can be used for direct firing to boilers as a potential replacement for heavy fuel oil (HFO).

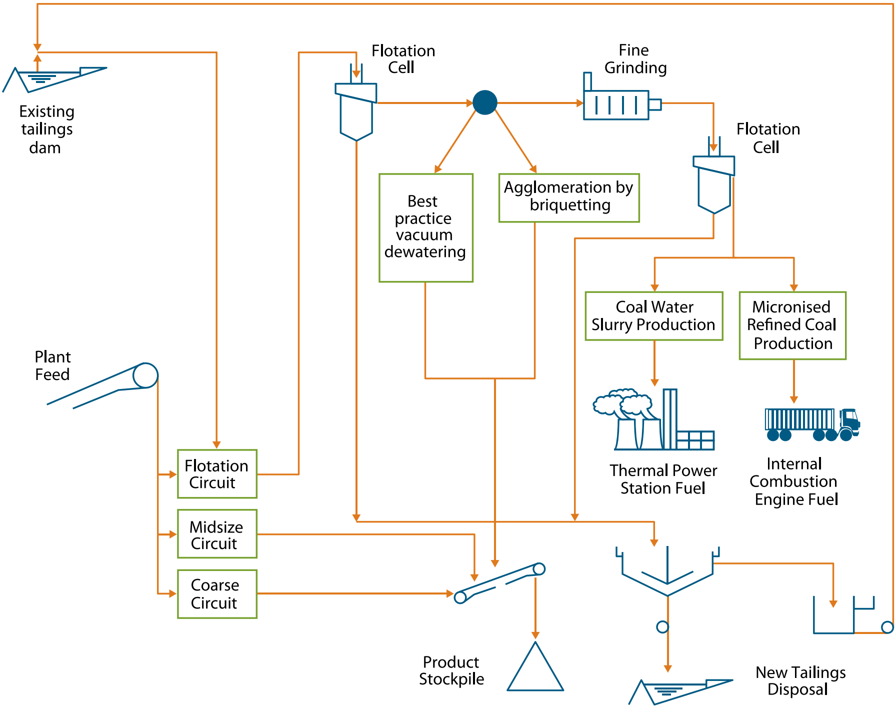


Figure 7 Alternative Fines Treatment of Thermal Coals Flowsheet

Figure 7 shows the flowsheet for these alternatives with the inclusion of the Jameson Cell and fine grinding for the purposes described earlier. Adding the milling step will be optional for CWSF but essential for micronised refined coal (MRC). Various filter options, also described earlier will need to be evaluated to achieve the required solid-liquid outcome for each coal source and each product component. Alternatively, a briquetted product that can be included in the normal product stream is another way of dealing with increasing fines related problems in coal product handling and transportation.

The integrated plant design has the functionality of the dual product offering, i.e., coal briquettes that can be added to the conventional product and/or coal-water slurry fuel (CWSF) for either heavy fuel oil replacement of more novel applications such as the MRC and DICE combination The concept of CWSF supply chains is being promoted to industry whereby slurry fuels employ existing heavy fuel oil infrastructure to transport and store the fuel at the customer facility.

COAL SUPPLY CHAIN CONSIDERATIONS

The current Coal Supply Chain (CSC) is hampered by an inability to dewater and transport fine coal. As a result, significant loss of coal value to tailings occurs also often leading to a significant loss of yield, which in turn impacts the coal mining price/tonne, adding further cost due to the need for tailings handling. The irony is that it is often necessary to exclude fines from product coal in order to transport it to the power station because of reduced revenue due to moisture, dust or sticky coal handling issues, etc. However, on arrival at the user's power plant a significant amount of money is then spent grinding it down to the conventional specification of ~70% minus 0.075mm.

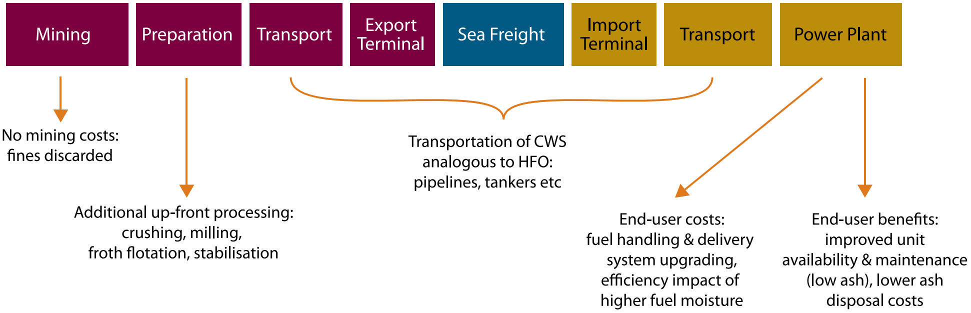


Figure 8 Coal Water Slurry Fuel Supply Chain (Osborne, 2013)

An innovative way to address this is to recover and use all the fines via coal-water slurry thereby recovering potential lost coal creating higher yield and lower cost/tonne. This could then be handled and transported as liquid fuel, avoiding the ever increasing problems associated with sticky, wet or dusting solids. There would then be less tailings material to dispose of, further lowering cost especially if the residual tailings can be paste-thickened and further dewatered so as to be disposable with coarse plant and mine waste.

No further grinding would be required at the power station, significantly lowering their cost and despite thermal efficiency being reduced, this would be offset to a large extent by cost reductions on coal that otherwise would not have been sold. Figure 8 shows a simplistic example of the chain. Some very simplistic cost-in-use analysis was conducted for this model which suggested a very definite benefit of about 1.0 to 1.5 c/kWh saving, would be generated for the user once the power plant boiler had been converted to accept CWSF to replace about 30% or more of the pulverised coal capacity (Osborne, 2013).

SUMMARY

This paper has provided insight as to the potential offered from thermal coal flotation. Waste in coal operations has for many years been "justified" by the impact that the added moisture has on the net value of the coal products. This may not be tolerated indefinitely because of increased awareness of sustainability issues and environmental impacts of adopting lowest cost disposal practices. With the concomitant emergence of better flotation technologies and improved solid-liquid separation equipment, arguments against recovery of wasted coal can be constructively challenged.

ACKNOWLEDGEMENTS

The authors would like to express their gratitude to the ACPS for accepting this paper and the associated presentation at this conference and to the senior management of XT for support and encouragement, in particular Mr Lindsay Clark, GM Mineral Processing.

REFERENCES

Ahmed, N. and Jameson, G.J., 1985, **"The Effect of Bubble Size on the Rate of Flotation of Fine Particles"**, *International Journal of Mineral Processing*, V 14, pp 195–215.

Bickert, G., 2013, **"Solid-liquid Separation Technologies for Coal"**, *The Coal Handbook*, V 1, Ch 13, pp 422–444.

Chen, F., Gomez, C.O. and Finch, J.A., 2001, **"Bubble Size Measurement in Flotation Machines"**, *Minerals Engineering*, V 14, N 4, pp 427–432.

Diaz-Penafiel, P. and Dobby, G.S., 1994, **"Kinetic Studies in Flotation Columns: Bubble Size Effect"**, *Minerals Engineering*, V 7, N 4, pp 465–478.

Evans, G.M., Atkinson, B.W. and Jameson, G.J, 1993, **"The Jameson Cell"**, *Flotation Science and Technology*, Marcel Dekker, New York, pp 331–363.

Firth, B., White, T., Stanmore, B., Hoskin, A., O'Brien, M. and Hu, S., 1998, "**Product Moisture After Centrifuging Coarse Coal – Stage 2 and Development and Demonstration of Standard Test"**, *ACARP Report* C4049.

Gomez, C.O. and Finch, J.A., 2007, "**Gas Dispersion Measurements in Flotation Machines"**, *International Journal of Mineral Processing*, V 84, pp 51–58.

Huynh, L., 2013, **"Design and Performance Aspects of Coal Flotation"**, *Proceedings: Coal Processing – Increasing the Value of Coal, Southern African Coal Processing Society International Coal Conference*, Graceland, South Africa, July 23rd to 25th, 2013.

Nesset, J.E., Finch, J.A. and Gomez, C.O., 2007, **"Operating Variables Affecting Bubble Size in Forced-air Mechanical Flotation Machines"**, *Proceedings of the Ninth Mill Operators' Conference*, AusIMM, Fremantle, Western Australia, pp 55–65.

Osborne, D., Huynh, L., Kohli, I., Young, M. and Mercuri, F., 2013, **"Two Decades of Jameson Cell Installations in Coal"**, *Proceedings of the 17th International Coal Preparation Congress*, Istanbul, Turkey, pp 353–358.

Osborne, D.G., 2013, **"Adding Value to the Coal Delivery Chain Via Integrated Science and Technology"**, *10th Australian Coal Science Conference*, Brisbane, Australia, November 19th and 20th.

Osborne, D.G., Huynh, L., Young, M.F., Sherritt, G., Perrin, M.J. and Collins A.R., 2014, **"Factors Affecting Design Optimisaton of Fine Coal Cleaning Circuits and Impacts on Value-in-use"**, *Proceedings: 21st Century Challenges to the Southern African Coal Sector; The Southern African Inst. Min. Met.*, South Africa, Symposium Series S79, pp 137–156.

BUSTING THE MYTHS OF FLOTATION IN THE AUSTRALIAN COAL INDUSTRY

L Huynh, I Kohli, D G Osborne
Xstrata Technology

ABSTRACT

The early 1990s saw the emergence of advanced flotation technologies such as the Jameson Cell to the Australian coal industry. Results from laboratory and pilot plant campaigns followed by full scale installations consistently proved distinct performance improvements over existing mechanical cells. These new technologies were able to produce lower ash products and high yield and recoveries in single cells (rather than banks) and a single stage of flotation. An ACARP report from the mid-1990s predicted that the industry would soon be dominated by these technologies (Sanders & Williamson, 1996).

Since its first installation in 1990 at Xstrata Coal's (now Glencore's) Newlands operation in the Bowen Basin, the Australian Coal Industry has proven to be an important breeding ground for the Jameson Cell technology allowing it to be continuously developed and improved over two decades. Now there are over 120 Jameson Cells installed in coal washing plants in Australia. Although the modern day Jameson Cells are much more robust, easier to operate, fully automated and require very little maintenance, flotation continues to be a major challenge in many operations with performance of the Jameson Cells less than optimal. However, the exact reasons for this have never been fully examined. This has led to many myths surrounding flotation, the technology used and the solutions required to solve the issues currently experienced in the industry.

This paper aims to address the common myths in the Australian coal industry which has perhaps negatively influenced the real effort that is required to address and rectify the current problems. It will explain processes that relate fundamentally to the flotation process, rather than the technology used, in order to describe the role of frothers in controlling flotation performance. It will also discuss how the term 'residual frother' has become a convenient excuse for a plant's sub-optimal flotation performance which can lead to masking of the real plant bottlenecks, and finally, the paper will examine the commonly held belief that flotation cells should always operate at the 'knee' of the ash-yield curve when in reality flotation needs to be optimised in relation to the entire washing plant and hence, consideration of gravity unit separation processes too.

Coal flotation circuits will be compared to those employed in modern base metal flowsheet designs to address the perception that the current designs used in the industry are inadequate and more sophisticated circuits should be developed. Plant data gathered from different operations will be used to demonstrate how diagnosing and addressing the real bottlenecks, rather than changing flotation technology used, or cell arrangement, is the best approach for any plant to achieve the desired flotation performance. Also of vital importance is education and training which is necessary to help plant personnel understand the fundamental flotation process and learn how to troubleshoot and respond to unwanted changes. Finally, fines circuit designs must incorporate an inherent procedure to institutionalise learnings from the errors in the past.

INTRODUCTION

By 1990 flotation was well established as an integral part of Australian coal preparation practice, although there was growing dissatisfaction among coal washing plant operations with the performance of conventional mechanical cells, the standard technology used at that time. The early 1990s saw the emergence of advanced flotation technologies such as the

Jameson Cell, which represented a major step change in process performance. The Jameson Cell was first tested and commercially installed in a coal washing plant at Xstrata Coal's (now Glencore's) Newlands mine (Jameson et al., 1991). The fines stream was cyclone overflow material which was previously discarded after thickening (minus 20-25 microns in particle size normally with ash content ranging from 15 to 40%). Pilot plant testing showed it was possible to achieve greater than 90% combustibles recovery with a product target of 10% ash and a similar moisture content value of ~10% via vacuum filtration. This led to the installation of the first generation full-scale, (so-called Mark I) Jameson Cells in 1990. Following the initial Newlands installation, many sites tested the technology which was shown to consistently produce low ash concentrates and achieve high combustibles recovery whilst being forgiving to variations in feed ash (Harbort et al., 1992; Atkinson et al., 1993; Manlapig et al., 1993). BHP Coal's (now BHP Billiton Mitsubishi Alliance - BMA) Goonyella 1,800 t/h coking coal operation in Central Queensland tested the advanced flotation technologies and subsequently replaced the entire 32 mechanical (Wemco) cell circuit with 8 Jameson Cells operating in a 2-stage configuration (Caretta et al., 1997). Figure 1 compares the performance of the Jameson Cells at this plant after commissioning to the old mechanical cell circuit. The ability of the Jameson Cell to consistently deliver a low ash product at high combustibles recoveries contributed to an overall plant yield increase of ~3.5% and led to production records at that time.

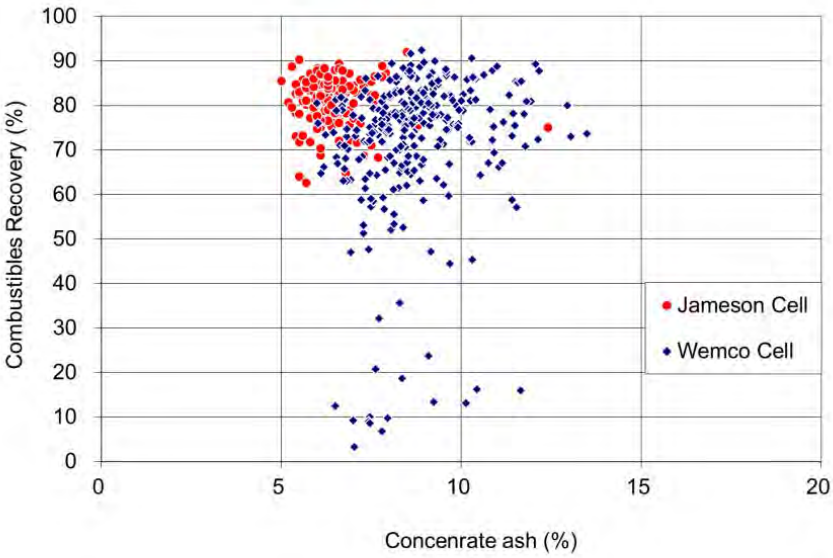


Figure 1
Full scale Jameson Cell performance at Goonyella mine compared to the original mechanical (Wemco) cell circuit

Amongst the key benefits was froth washing and the simplicity afforded by the Jameson Cell, it being easy to operate and maintain; with no moving parts, and needing no auxiliary equipment except for the feed pump. Since its first commercial installation, the Jameson Cell technology has been continuously developed and improved making it more robust and easier to use. The development of the technology can be divided into four phases, designated Mark I to IV as shown in Figure 2. There are now over 120 Jameson Cells operating in coal applications in Australia, with the current largest installation being at Wesfarmers' Curragh Mine in Central Queensland which treats over 5 million tonnes of coal fines per year using twelve cells. Some of the early installations have gone through upgrading to Mark III and IV designs targeting the benefits from such an upgrade.

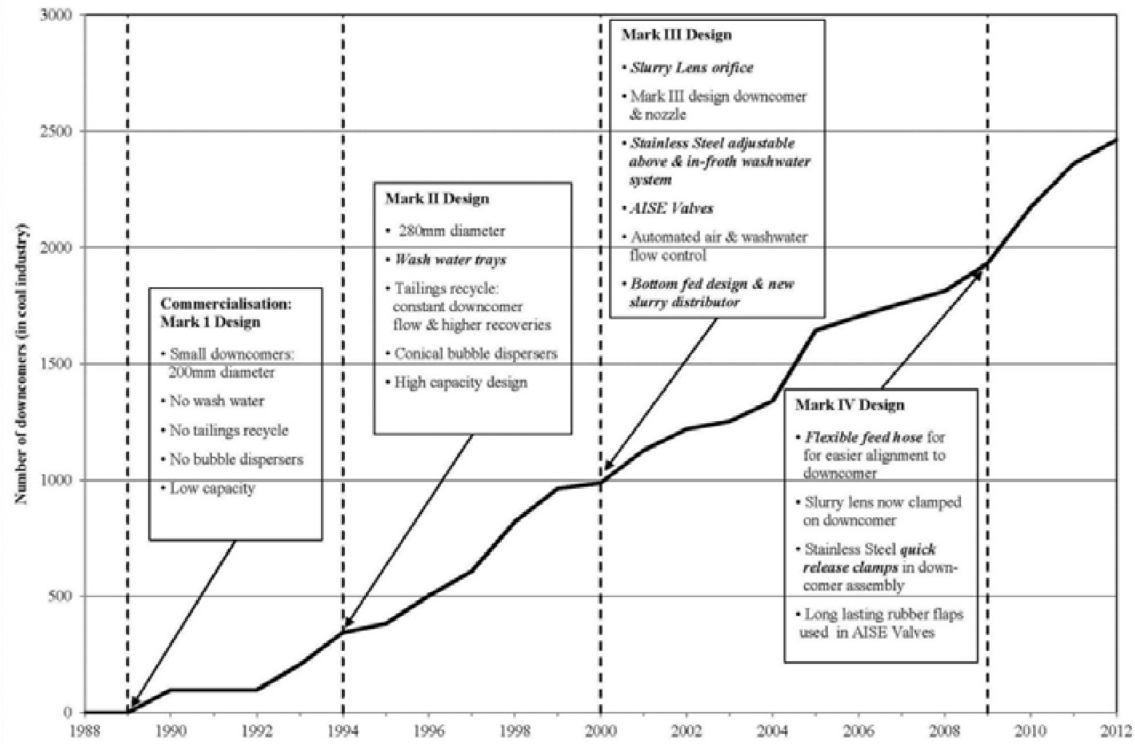


Figure 2
Jameson Cell development path

Although the modern day Jameson Cells are now much easier to operate, fully automated and requires very little maintenance, the fines circuit continues to be a major challenge, for example, with performance of the Jameson Cells being operated at less than optimal at many operations. However, the exact reasons for this have never been fully examined. This conundrum appears to have generated many myths surrounding flotation, i.e., the technology used and the solutions required to solve the real issues currently being experienced in the industry.

This paper will highlight and address five common myths in the Australian coal industry that relate to flotation. It will provide fundamental and logical explanations to hopefully dispel myths that primarily blame the technology and thereby create a more beneficial outcome by directing the focus on the real issues facing the industry today. Flotation performance data from existing sites using the Jameson Cells will be used to illustrate how plants working with the usual design constraints can rise above the 'mediocre' levels in combustibles recovery to consistently achieve high performance. Removing froth handling and downstream dewatering bottlenecks will be a key element for many plants to achieve in order for them to realise the full potential of their installed Jameson Cells. Education and training of industry personnel is also an essential need for optimising performance and creating an appreciation of the fact that chemistry factors controlled by feed characteristics and reagents are just as vital in the flotation process as the machine itself.

MYTH 1: FLOTATION IS A SEPARATE "ADD-ON" UNIT OPERATION IN THE 'BACK-END' OF A WASHPLANT

Flotation in all too many cases has been the last process to appear in a washing plant, either conceptually or as a retrofit in plants already built and operating. This has been mainly due to the "installed cost vs. apparent benefits" argument, and the resultant operation has invariably underperformed against expectations. The main reasons for this can be simply categorised as having been one or more of the following; poor design, inadequate design data, poor equipment selection and “cutting corners” or cost saving measures prior to the final sign-off and installation.

A washing plant consists of three main circuits which treat coal of different a particle size range. For example, the modern day washing plants in Australia usually have a coarse (50x1.5mm), intermediate (1.5x0.25mm) and fines (0.25x0mm) circuits. The coarse and the intermediate circuits which typically account for 70-80% of the throughput, use gravity-based separation techniques to produce their product components. Gravity methods become increasingly inefficient as the feed gets finer, and the magnitude and nature of this deterioration is well known and often predictable. However, flotation, an entirely different separation technique which is surface chemistry based is a very different proposition. There are no Ep’s for flotation equipment and their performance must be determined by test work. Although the fines circuits treat a much smaller proportion of the raw coal feed compared to the gravity circuits, it can still play a vital role in controlling the overall product quality from the plant. It can usually be manipulated to enable optimisation of the overall yield of the overall plant for the specified quality of the combined product.

In many plants, operators tend to optimise the gravity and flotation circuits separately often overlooking the fact that there is only a single product produced from the plant. In actual fact, this product is composed of perhaps 3 or 4 components which emanate from circuits having differing control capabilities. So a flotation component needs to be integrated into an overall operating strategy to achieve the desired plant yield and product quality.

MYTH 2: FOR OPTIMUM FLOTATION PERFORMANCE, FROTHER CONCENTRATION MUST BE ABOVE 15 PPM (THE ‘MAGIC NUMBER’) AS IMPOSED BY THE FLOTATION TECHNOLOGY VENDOR

Due to the very high yields (mass pull) often encountered in fine coal flotation, there must be sufficient bubble surface area to enable the capture and recovery all the required coal particles. This is particularly so in the treatment of coking coals where up to 80-90% of the feed mass reports to the concentrate, machines that can produce a large number of very small air bubbles (small mean size) are therefore advantageous because for the same volume of air, more individual bubbles are generated representing vast bubble surface area, i.e. carrying capacity. Large carrying capacity means higher productivity than other machines which generate larger bubbles (large mean size). The Jameson Cell is able to produce ultra-fine bubbles via by the shear action of a plunging jet (Evans, Jameson & Atkinson, 1992). Air bubble mass generated lies in the range of 300 to 700 μm (Sauter mean diameter, D_{32}) (Evans, Atkinson & Jameson, 1993).

Frothers are required in the flotation process to prevent air bubbles from coalescing. Regardless of the technology used, frothers should be used at dosages above the critical concentration of coalescence (CCC), as this is the minimum concentration required to prevent coalescence (Cho & Laskowski, 2002; Finch, Nesset & Acuna, 2008). The graph in Figure 3 shows bubble size as a function of frother concentration for the commonly used MIBC (methyl isobutyl carbinol) using the Jameson Cell and a mechanical cell. The CCC is a property of the frother and independent on the machine used as clearly shown in Figure 3. The CCC is measured to be 15 ppm for MIBC. However, the machine dictates the minimum bubble size generated and the figure clearly illustrates that the Jameson Cell produces significantly smaller bubbles than the mechanical cell. This dispels the myth that a ‘magic number’ is imposed by a flotation machine vendor to make their machines ‘work’, as often implied. The inability to add sufficient frother due to any inherent design problems, or plant bottlenecks will affect ALL flotation machines without prejudice, and hence also similarly affect carrying capacity. So even at concentrations below the CCC, the Jameson Cell will generate smaller bubbles and hence, have greater potential carrying capacity than most other machines. If a plant operator wishes to change the type of frother they are using, they would need to take into account the CCC characteristic of the potential replacement and review dosages on a fair and comparable basis. The frother reagent supplier should be able to provide this information for customers.

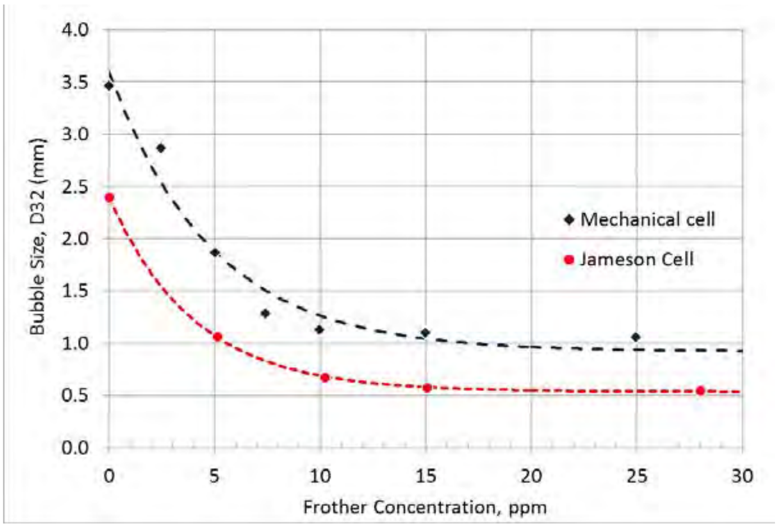


Figure 3
Bubble size as a function of frother concentration

MYTH 3: 'RESIDUAL' FROTHER PROBLEMS IS LIMITING FROTHER DOSAGE

Low yield of low combustibles recovery from the Jameson Cell can be attributed to insufficient frother addition (below that required for the CCC), which decreases the carrying capacity of the machines. Without truly understanding the real issues at individual sites, operators nowadays appear to have developed an exaggerated fear of 'frothing out the plant' which all-too-often causes them to automatically reduce the frother dosage. 'Residual' frother is defined in this paper to mean frother remaining in the water exiting the fines circuit either via the flotation concentrate, or tailings streams. This water then gets recirculated and used in other areas of the washing plant. This can compromise the performance of other unit operations such as the dense medium cyclone circuit, as 'residual' frother can cause frothing in sumps, pumping issues and affect density gauge readings etc. Its effect on a plant is largely dependent on the design of the water circuit. A site where the water recirculates around the plant quickly tends to have more issues because the build-up of frother concentration is faster than its decomposition or break-down. However, it has been found that in reality, 'residual' frother is not always the first, nor the primary, culprit for operators to 'turn-down' the frother dosage. Instead, it may be caused by a number of other issues within the fines circuit itself. Common contributors are a number of dewatering issues occurring downstream of the flotation stage. Examples include; using frother dosage as a way of deliberately decreasing the tonnage and volume produced from the Jameson Cells, e.g., to try to avoid a 'sloppy concentrate' in the case of horizontal belt filters (HBF); or motor trips, in the case of screen bowl centrifuges (SBC). Essentially, the dewatering process is controlling the separation process which is far from an ideal situation.

In recent years, SBCs seem to have gained in popularity compared to HBFs, probably due to their smaller footprint and lower installed capital cost. However, their suitability for dewatering high quality Bowen Basin type coking coals is debatable as SBCs cannot capture material less than 45 micron, which is consequently discarded with the tailings. Unlike some Australian thermal coals which may have higher ash in the finest fractions due to clay material, coking coals usually has high vitrinite in this size fraction as coal macerals tend to be friable and therefore highly concentrated in the raw coal fines. But plants employing SBC's are often left with what is installed due to the anticipated cost of retrofitting a more effective solution. Disposal of the effluent is normally via the tailings thickener but as this stream is extremely frothy (stabilised by fine hydrophobic coal particles) there are always pumping issues due to inadequate design of sump and pumps. At too many plants with SBC's, effluent spillage is a primary reason to 'de-rate' the flotation recovery. As expected, the effluent stream is high in frother concentration and yet, it is designed to exit the fines circuit whereas for plants where HBFs are used, the filtrate is returned to the flotation feed as a means of minimising 'residual' frother effects. Although it is far from ideal not to capture the minus 45 micron coal-rich fines fraction, it is still advisable to return the centrate to the SBC feed rather than the flotation feed. Some plants have introduced this practise thereby avoiding the inevitable build-up of a circulating load of fines in the flotation circuit. It clearly makes no sense to continually return coal fines already recovered back to the separation process despite the original poor choice of dewatering technology for the feed type. Clearly, the selection procedure for the dewatering technology needs to carefully consider the particle size and type of coal treated and not use capital and/or operating cost as the main decision driver.

Generally, poor design of the overall fines circuit appears to be the main culprit for sub-optimal flotation recovery at many sites. Improvement must be made in the design of sumps to handle large fluctuations in froth volumes, and appropriate pumps installed to transfer frothy streams. Dewatering units like SBCs, which are sensitive to flows, may benefit from a buffer tank upstream or more appropriately a coal thickener, which not only reduces the volume but increases the solids content that is to be treated.

The challenges experienced at individual sites are not always the same and proper diagnosis is the key to addressing the right issues at each operation. A review of all the Jameson Cell installations in Australia shows the bottlenecks can be, in general, placed into one of 3 categories. These are education, measuring and/or addressing minor design issues (30%), concentrate dewatering capacity and/or issues (60%) and flotation capacity issues (10%). This shows that the majority of sites do not actually have a problem with the Jameson Cells, highlighting the need to focus on addressing the fundamental issues rather than just the flotation machine. That is why any debate directed towards solving current industry issues by swapping the flotation technology used (as discussed later in Myth 5), is perhaps premature if not mistaken. Only Category 3 relates to the flotation machine, but this is due to inadequate flotation capacity. It does not imply that the Jameson Cell will perform poorly when treating the feed for which it was originally designed.

MYTH 4: THE ADVANCED FLOTATION TECHNOLOGIES SUCH AS THE JAMESON CELL 'ARE NOT WHAT THEY ARE CRACKED UP TO BE' AND 'PERFORM POORLY' AS DESIGNED

The Australian Coal Industry was rapid in its adoption of the Jameson Cell soon after its commercialisation in 1989, thereby significantly contributing to its eventual wider adoption and growth over a period of only two decades. The latest model (Mark IV design) of Jameson Cell is therefore a great improvement over the first generation of Jameson Cells. Currently, typical duties are far less demanding as the raw coal fines feed tends to be coarser (than the original Newlands circuit) and together with the improvements made to the technology, it would reasonable to expect the Jameson Cell would be much more robust and perform at least equal, if not better, than in earlier installations. However, there is a perception in the industry, particularly in recent years, that once installed the Jameson Cell technology often 'performs poorly'. This is most often undefined and does not specify or quantify why the Jameson Cell cannot achieve expected values of concentrate ash, yield/combustibles recovery or either. In fact, Xstrata Technology believes that the exact reasons for 'sub-optimal' flotation performance in many operating plants have never been properly investigated.

The key to optimising flotation is firstly, to understand that it is a complex multifaceted separation process that is driven by surface chemistry and not by gravity like the other unit operations in a coal washing plant. It is controlled by three main factors that in turn can be categorised into three areas: the coal, the chemistry and the machine as shown below in Figure 4.

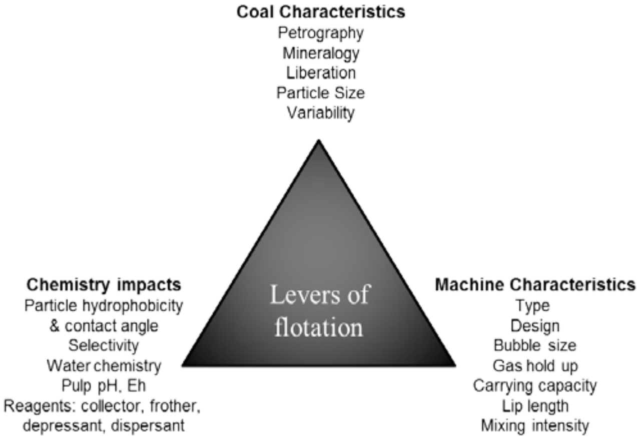


Figure 4
Flotation ‘triangle’ showing the key factors affecting performance

The flotation machine is one of three facets important to the overall process, but seems to get the most attention and is often blamed when flotation performance is poor. However, the variability of the coal and flotation reagent control are two equally significant factors, relating to the other facets of the triangle, and both are perhaps surprisingly, often overlooked. The greater the number of different coal seams and sources that are treated, the more challenging is the task of achieving effective flotation and meeting the targeted qualities and recoveries. Plant designers must therefore take into account all factors that are going to be influential in ensuring expectations are met. Plant operators must be properly trained to fully understand how the circuit is intended to perform and then be able to respond appropriately. This includes the correct response to changes in tonnage, particle size distribution and flotation behaviour of the different coal types by making the necessary adjustments to reagent dosages and process variables to ensure optimised performance.

In many plants, monitoring of flotation performance is irregular and often a ‘knee-jerk’ or spontaneous change is made when performance has clearly deteriorated. Furthermore, it may be impossible to conduct sample surveys via the feed, concentrate and tailings because sample points do not exist. Even in the more modern (recently built) plants, the flotation feed often cannot be easily collected as it usually consists of more than one stream which gravity flows into a large collecting sump. In many Jameson Cell installations, operations personnel unknowingly collect the downcomer feed and use the ash result from this stream in the two-product formula to calculate yield and combustibles recovery. This is then erroneous as the downcomer feed is in fact an internal stream. The correct and incorrect streams to collect around the flotation circuit are clearly illustrated below in Figure 5.

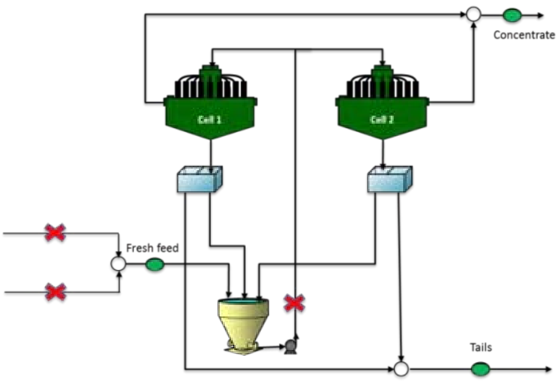


Figure 5
The correct (green circles) and incorrect (red crosses) sample collection streams for flotation surveys

Regular flotation surveys are necessary for measuring and gauging performance. Results need to be plotted as an ash-yield curve and benchmarked against a characterisation curve for each coal type treated. Figure 6 shows six months of plant data from an existing operation (Plant A). A wide range of combustibles recovery values is shown and at ash values which generally lie on or around the characterisation curve. This is typical for a Jameson Cell circuit, in either single or two-stage configurations. But this result may not be achieved using other flotation technologies machines. For example, mechanical cells are likely to produce higher ash concentrates, when used in the same configuration because this type of machine is less selective and commonly does not employ froth washing. As a consequence the data points obtained will lie to the right-hand side of the characterisation curve. The large variation in the combustibles recovery shown in Figure 6 means that high recoveries are possible (i.e., the Jameson Cell itself is not the problem) but the process has not been properly optimised. Knowing this, Plant A needs to diagnose then address the real issues with the aim of reducing the variation to consistently achieve the desired target for combustibles recovery.

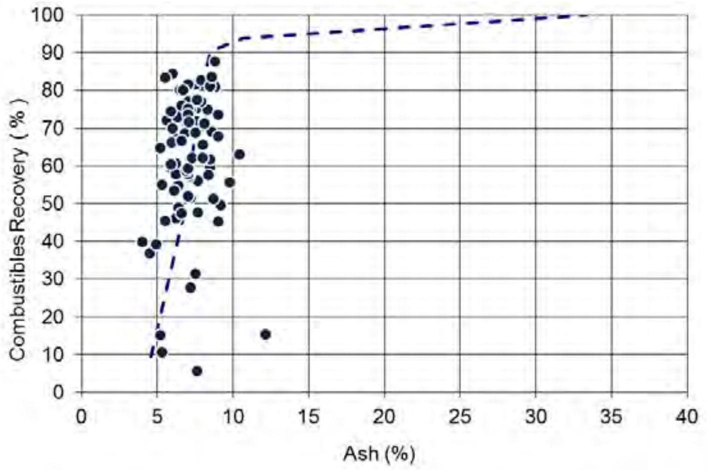


Figure 6
Plant data from Plant A benchmarked against the characterisation curve

Figure 7 shows performance data from another operating coal preparation plant (Plant B). This plant when compared to Plant A, has reached the next stage of flotation optimisation. Clearly, the data points are centred around 5 to 7% ash (i.e., achieving the set target) at combustibles recovery values of around 70 to 90%. Plant B operates with many design constraints just like other plants, but this operation has obviously implemented effective strategies for dealing with these issues and is operating within these limitations.

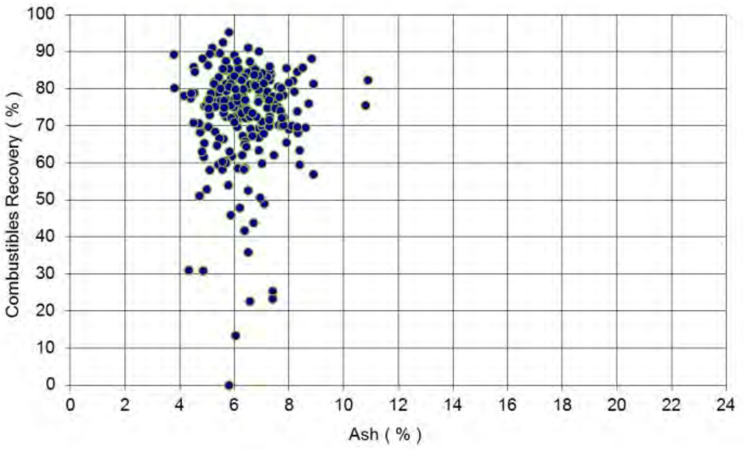


Figure 7
Plant data from Plant B

Figure 8 shows the performance data from a third preparation plant, Plant C. This operation has focussed on eliminating the major bottleneck which has been downstream concentrate dewatering. This is a good example whereby the Jameson Cells are unaffected by extraneous factors and can be operated as designed to realise the full potential.

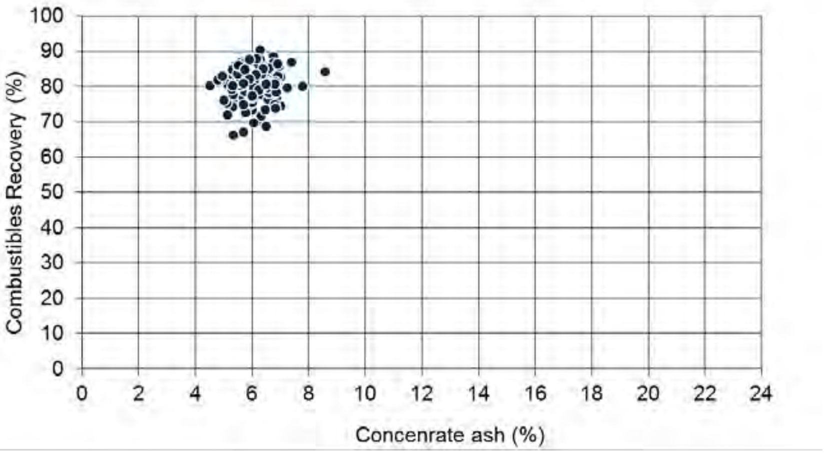


Figure 8
Plant data from Plant C.

The flotation performance of all three different sites were graphed using data collected from six months of shift or daily samples, all during 2013. The results obtained should clearly dispel the myth that the flotation machine is the problem and counter speculation that the Jameson Cells perform badly ‘as designed’. Instead, this highlights the importance of sampling and regular monitoring, understanding the flotation principles (the flotation triangle) and diagnosing the real plant issues. The Jameson Cell is not a stand-alone piece of equipment. To be fully utilised it has to be integrated into fines circuit design that allows it to be used to its full potential for all types of feed.

MYTH 5: THE AUSTRALIAN COAL INDUSTRY SHOULD CONSIDER A RETURN TO MECHANICAL CELLS AND DESIGN COAL CIRCUITS SIMILAR TO BASE METALS INDUSTRY

In recent times, a body of opinion has emerged that is suggesting that perhaps the Australian Coal Industry may be better served by reverting to using mechanical cells to treat raw coal fines. This is possibly driven by a misconception that Jameson Cells, and columns, have in some cases been the root cause of below par performance and resultant low yields/combustibles recovery, i.e., loss of coal to tailings. It may be easier to blame the technology, but this is clearly not the primary cause. Re-adopting a conventional technology that has been superseded by more advanced ones will not help the industry to address the underlying issues or identifying the true cause. To achieve this requires a rigorous diagnostic approach followed by progressive debottlenecking of the fines circuit that have been shown to be preventing the installed cells from being fully utilised. With a clean sheet of paper the design of new fines circuits must overcome current and/or old issues and be better integrated into the overall plant. In this regard, properly structured teaching and training of operations and maintenance personnel will ensure a better understanding of the basic principles of flotation and the capabilities of the flotation circuit as a whole. The key reward will be the confident performance of operators, and valuable know-how generated for plant designers and consultants.

The current design and operational issues, all too frequently encountered in the fines circuits have perhaps detracted from the realised benefits that the advanced flotation technologies have provided the Australian Coal Industry. For the Jameson Cell, the most noticeable of these are the high quality concentrates that can be produced, the small footprint and very low maintenance requirements. It is difficult to understand how a return to mechanical cells would be at all advantageous as conventional technologies are simply less effective in treating coking coal deposits in Australia. For example, a mechanical cell circuit will not be able to achieve a comparable high yield and combustible recovery at the lowest ash content. Add to this the footprint and the high maintenance due to moving parts (rotors), motors and blowers and the benefits swing even more towards the more advanced technology. However, this swing has not yet been as strong in other coal producing countries around the world. Countries like India and Russia have been slower to change and are only beginning to realise the benefits of adopting advanced technologies over the conventional cells which they have been using as standard for decades. In contrast, new coal producing regions such as Mongolia and Mozambique already use the advanced flotation technologies as standard in design of their new washing plants.

Across the many differing types of industries utilising flotation, the momentum has shifted towards advanced technologies in favour of conventional cells. In particular, the Jameson Cell technology is successfully used in a number of different industries to address plant issues and/or improve plant performance. In the metals industry where flotation is the dominant separation process, the mining of more complex ore-bodies and decreasing metal head grades means the standard flotation circuit traditionally utilising all mechanical cells, can no longer achieve the desired performance. Inclusion of Jameson Cells into conventional cleaner circuits is necessary to produce clean saleable concentrates that are sufficiently low in non-sulfide and silicate gangue, penalty elements (such as fluorine and mercury) or deleterious elements such as Uranium (Araya et al., 2013). Other examples include the phosphate industry where a method was developed to recover ultrafine phosphate that had until that time been long considered as unrecoverable in this industry (Teague & Lollback, 2012) and another good example is the use of the Jameson Cell to enhance the recovery of bitumen in the oils sands industry (Neiman et al, 2012).

Regardless of industry and application, flotation circuits should always be of the simplest and most robust design to achieve the desired product quality and recovery. Feed characteristics and the degree of upgrade desired will dictate the number of stages required to achieve the desired performance. For example, base metal flotation often requires multiple stages to sequentially upgrade the very low grade feed to an acceptable final product quality. The mass passing to the final concentrate (equivalent of yield) is always very low in comparison to coal and typically no higher than 10%.

Figure 9 shows an example of a flow sheet designed to treat copper ores which consists of rougher and scavenger stages followed by three stages of cleaning. In contrast, the upgrade required in coal applications is very low and can be easily achieved in a single step so the circuits can be much simpler, as shown in Figure 10. Circuits for coal treatment only typically require single or two-stages.

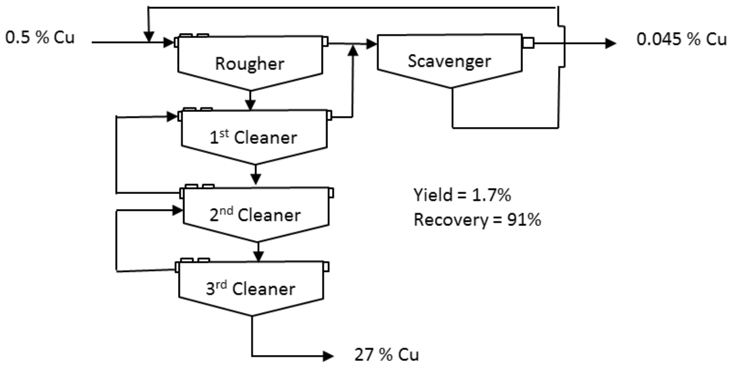


Figure 9
Example of a flotation circuit to recover copper

Another major difference between base metal and coal processing is the so-called mass pull, or yield. In base metals circuits, the yield is low and the multiple processing stages and circulating loads are necessary to achieve high recoveries. Most tailings streams cannot be operated in open circuit. In contrast, coal applications have very high mass reporting to the concentrate. For example to achieve the same recovery (91%) as the base metal example shown in Figure 9, the yield in coal flotation is 73% versus only 1.7% for copper.

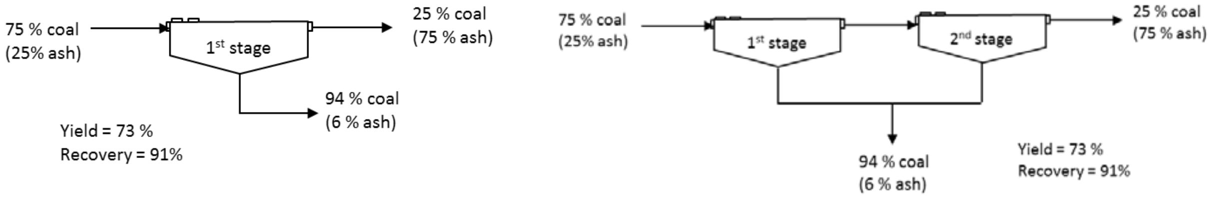


Figure 10
Typical design of coal circuits consisting of one or two-stage configurations

Therefore, it is highly advantageous to use flotation machines which have very high carrying capacity to minimise number of machines required for the duty. Coal circuits should avoid circulating loads as this will only serve to increase the carrying rates of the machine even more. Residence time is of lesser importance, particularly when treating high quality coking coal, as it's normally naturally hydrophobic and does not rely on this factor to achieve maximum recovery. Another difference between coal and base metals flotation is the much larger volumes that are treated which is mainly due to the difference in feed solid content. Feed treated in coal flotation is typically 5 to 10 wt% solids compared to 20-40 wt% solids for base metals applications.

In the Australian Coal Industry, there is ongoing debate about the merits of Jameson Cells arranged in single versus two-stage configurations. Figure 11 compares the performance of two operating plants treating the same feed in these two configurations. The Jameson Cell produces a low ash concentrate at very high (70 to 90%) combustibles in both cases. This highlights the robustness of the Jameson Cell in either configuration.

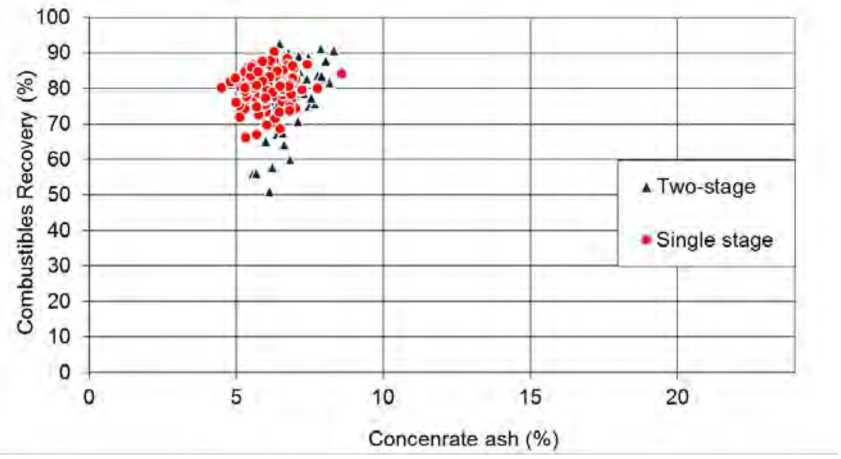


Figure 11
Comparison of Jameson Cell performance in single and two-stage configuration; using data from two operating plants

Bearing all this in mind, the design of any flotation circuit should never rely on a "one-size-fits-all" approach, as a critical driver for the design is the coal(s) treated. Every source of coal is different and is also likely to change over time as a result of changes in the deposit and in the mining conditions, etc. Factors relating to surface chemistry on the flotation process become more important as coal quality decreases. Therefore, circuits designed for poorer quality coking coals, oxidised or weathered coal and thermal coals of varying rank, need to carefully consider other aspects such as reagent addition control (i.e., some may need staged addition), if reagent emulsification is required and conditioning for proper adsorption of collector on coal particle surfaces. For example, it is the variable nature of these various types of coal that lends itself to adopting two-stage circuit designs.

It is also important to highlight that two-stage Jameson Cell circuits do not need double the amount of cells compared to a single-stage circuit if treating the same amount of feed. Single stage circuits are designed at higher tailings recycle, typically around 40 to 50% because the higher recycle is required to ensure the cells can have sufficient carrying capacity to produce the desired product tonnage. In two-stage circuits, the recycle is reduced to 10 to 20% and designed only to dampen fresh feed fluctuations and thereby ensure that the Jameson Cells can be operated at a constant feed pressure and volumetric flow rate. In most cases, the total number of cells will be the same, although the two-stage stage option will probably require the next cell size larger than the cells chosen for single-stage.

CONCLUDING REMARKS

The Australian Coal industry was a leader in adopting advanced flotation technologies such as the Jameson Cell, into its operations in the early 1990s. However, in many subsequent installations, inadequate design and improper integration of the fines circuit into the overall washing plant have prevented operators from realising the full benefits of this improved technology. Myths perpetuated in the industry relating to flotation, from fundamentals aspects to the machines themselves, have not helped operating plants to identify and address the real issues. It is hoped that the explanations and discussions in this paper will help to address these myths allowing sites and the Australian Coal Industry in general to focus their efforts towards overcoming the current challenges. Proper education and training is necessary for operators to better understand and appreciate the complexity of the flotation process, and to learn that the surface chemistry aspects are just as important as the machine used. In our view it should be mandatory for all plants to measure the flotation performance in order to be able to manage it. Performance must be quantified to allow each site to gauge the current performance and then implement solutions to debottleneck the fines circuit as necessary. Only then will the Jameson Cell circuits be used to their full potential. It is hoped that the proven performance of the flotation circuit demonstrated from three sites reported in this paper will help other sites realise that such improvements are possible and therefore motivate them to do the same. There are many lessons to be learnt from overcoming the current challenges and these must be institutionalised by every operating company to ensure that mistakes are not repeated in new their plant designs and projects.

ACKNOWLEDGEMENTS

The authors would like to thank the operating sites that have allowed us to use their data for this paper. We would like to acknowledge the help of everyone in the Australian Coal Industry who have provided valuable discussions on the operation of their fines circuit and flotation in general. This paper could have not been prepared without your valuable feedback and input. Finally, we would like to express our gratitude to the senior management of XT for permission to offer this paper to the ACPS.

REFERENCES

Araya, R. Cordingley, G., Mwanza, A. & Huynh, L., 2014, Necessity driving change and improvement to the cleaner circuit at Lumwana concentrator, *Proceedings of the 46th Annual Canadian Mineral Processors Operators Conference*, CIM, Ottawa, Ontario, pp. 329-341.

Atkinson, B.W., Conway, C.J. & Jameson, G.J., 1993, Fundamentals of Jameson Cell operation including size-yield response, *6th Australian Coal Preparation Society Conference*, Mackay, Queensland.

Caretta, M.F., Graham, J.N., Dawson, W.J., 1997, Jameson Cell scale-up experiences at BHP Coal's Goonyella coal preparation plant, *Coal Prep '97*, Lexington, Kentucky.

Cho, Y.S. & Laskowski, J.S., 2002, Effect of flotation frothers on bubble size and foam stability, *International Journal of Mineral Processing*, vol. 64, pp. 69-80.

Evans, G.M., Atkinson, B.W. & Jameson, G.J., 1993, The Jameson Cell, *Flotation Science and Technology*, Marcel Dekker, New York, pp. 331-363.

Finch, J.A., Nesset, J.E. & Acuna, C., 2008, Role of frother on bubble production and behaviour in flotation, *Minerals Engineering*, vol. 21, pp. 949-957.

Harbort, G.J., Manlapig, E.V., Jackson, G.J., 1992, Jameson Cell use in fine coal flotation, *ASEAN-Pacific Coal Conference '92*, Cebu, Philippines.

Jameson, G.J., Goffinet, M., Hughes, D., Operating experiences with Jameson Cell at Newland Coal Pty Ltd, 5th Australian Coal Preparation Society Conference, Newcastle NSW, May, 1991.

Manlapig, E.V., Jackson, B.R., Harbort, Cheng, C.Y., 1993, Jameson Cell coal flotation, *10th International Coal Preparation Exhibition Conference*, Lexington Kentucky.

Neiman, O., Hilscher, B. & Siy, R., 2012, Secondary recovery of bitumen using Jameson downcomers, *Proceedings of the 44th Annual Canadian Mineral Processors Operators Conference*, CIM, Ottawa, Ontario, pp. 115-124.

Sanders, G.J, & Williamson, M.M., 1996, Coal flotation technical review, ACARP Report C4047.

Teague, A.J. & Lollback, M.C., 2012, The beneficiation of ultrafine phosphate, *Minerals Engineering*, vol. 27-28, pp. 52-59.

Improved Cleaner Circuit Design for Better Performance Using the Jameson Cell

L Huynh[1], R Araya[2], D R Seaman[3], G Harbort[4] and P D Munro[5]

ABSTRACT

In today's operating environment many conventional cleaner circuits struggle to produce final grade concentrate with acceptable levels of non-sulfide gangue and/or penalty elements. This is primarily due to poor selectivity and the entrainment of gangue, and is further exacerbated by the processing of lower grade ores with more complex mineralogy that require finer grinding for liberation. In new concentrators, the standard approach to cleaner circuit design is to define the number of stages of mechanical cells required to achieve the desired concentrate grade. Residence time is the primary factor to size cells and to determine the number of cells required to attain target recovery. While it is required for recovery, a bank of mechanical cells is not the ideal solution to maximise concentrate grade. Relatively coarse bubble size means slow flotation of fines, requiring long residence times. The combination of long residence times and no froth washing means higher entrainment as the pulp becomes more barren and froth stability decreases down a bank.

In recent years, the Jameson Cell technology has been retrofitted into a number of concentrators around the world to solve both cleaner circuit capacity and concentrate grade issues. Jameson Cells have been installed at the head of conventional cleaner circuits to produce final grade concentrate in a single step of flotation using a single cell (referred to here as 'cleaner scalping' duty). The experience gained from these installations has provided the platform for the design of improved and simpler cleaner circuits that will perform better and be more robust to operate.

This paper analyses traditional cleaner circuit designs and explains the philosophy behind new hybrid cleaner circuits, which use Jameson Cells to produce final concentrate and mechanical cells for a 'cleaner scavenger' duty. The advantages of the new cleaner circuit design are demonstrated by a case study from an operating plant. Engineering studies show that the better metallurgical performance can be achieved with less equipment than conventional cleaning designs, reducing capital cost, operating cost, and with more than 30 per cent energy savings.

Finally, the paper describes simple modifications to the standard laboratory flotation test procedures which accurately simulate this new cleaner circuit using routine equipment.

INTRODUCTION

Depletion of the world's mineral resources imposes great pressure on the efficiency of modern concentrators. The combination of lower head grades, more complex mineralogy and environmental requirements places new demands on concentrate quality, both in terms of concentrate grade and the need to remove deleterious elements such as arsenic, fluorine, mercury and uranium. Regrinding of the rougher concentrate is increasingly required to liberate the valuable minerals sufficiently to produce a saleable concentrate. The finer the grind, the more difficult is subsequent flotation separation in the cleaning circuit to produce the desired concentrate quality. Despite these increasing challenges, there does not appear to have been significant design changes in cleaner circuits in recent decades. Most new plants today, particularly in Australia, are still designed predominantly with conventional mechanical cells for cleaning.

The primary purpose of a flotation cleaning circuit is the separation of valuable (hydrophobic) minerals from non-valuable minerals. Non-valuables include hydrophilic gangue, mildly hydrophobic minerals such as iron sulfides, and minerals locked in composite particles. In industry, cleaning is commonly achieved with multiple stages of counter-current banks usually consisting of mechanical flotation cells. However, in some recent brownfield projects, cleaning circuit modifications have been required to either increase capacity,

---

[1] MAusIMM, Flotation Manager, XT Pty Ltd, Level 10, 160 Ann Street, Brisbane Qld 4000. Email: le.huynh@xt-t.com

[2] Process Engineer, XT Canada, 10th Floor, 700 West Pender Street, Vancouver BC V6C 1G8, Canada. Email: rodrigo.araya@xt-t.com

[3] MAusIMM, Principal Metallurgist, Newcrest Mining Ltd, 193 Great Eastern Highway, Belmont WA 6104. Email: david.seaman@newcrest.com.au

[4] FAusIMM(CP), Process Manager, AMEC Mining and Metals, Level 14, 144 Edward Street, Brisbane Qld 4000. Email: greg.harbort@amec.com

[5] FAusIMM, Senior Principal Consulting Engineer, Mineralurgy Pty Ltd, Unit 2, 42 Morrow Street, Taringa Qld 4068. Email: pdmunro@bigpond.com.au

to improve recovery, or to increase concentrate grade and remove penalty elements. These upgrade projects have been an opportunity to develop and demonstrate alternative approaches to cleaning circuit design. This includes using froth washed cells such as columns or Jameson Cells, and the use of small bubble size and very high intensity mixing, such as the Jameson Cell, to increase the flotation rate of fine particles. Combining these advantages of high flotation rate and froth washing with long residence times possible in mechanical cells has developed hybrid circuits that combine the benefits of both technologies in more efficient, more effective, lower energy circuits. Having been developed to improve existing circuits, it is now clear these circuit concepts offer a step change improvement for the design of new circuits.

In conventional mechanical flotation cells, the recovery to concentrate by entrainment can be described by a size dependent parameter (ENT_i) that relates the recovery by entrainment of particles in size fraction i ($R_{ent,i}$) to water recovery across the cell (R_w) as follows (Johnson, McKee and Lynch, 1974; Savassi *et al*, 1998):

$$R_{ent,i} = ENT_i \cdot R_w$$

In mechanical cells, it is necessary to reduce water recovery in order to lower entrainment recovery – this can be achieved by reducing water recovery through feed dilution while maintaining valuable mineral recovery if there is sufficient residence time in the flotation cell or bank. In froth washed systems, such as columns and Jameson Cells, the entrainment factor (*ENT*) is reduced by froth washing with a positive downward water bias.

The Jameson Cell is a high intensity flotation technology jointly developed in the mid-1980s between Mount Isa Mines (MIM, now Glencore) and Professor Graeme Jameson from the University of Newcastle. What started as a research project to improve the sparger design in the column cells installed in MIM's zinc cleaner circuit, culminated in the development of a completely different bubble-generation device called a downcomer. When it was discovered that, in addition to being a bubble generation device, bubble-particle collisions also occurred inside the downcomer (Harbort, Manlapig and DeBono, 2002), it became apparent that the large residence time and hence, large volumes required for the collection zone in column cells was not needed. This meant that the downcomers could be placed in much smaller tanks, to achieve the same or better metallurgical results. This is shown graphically in Figure 1 (Jameson, 1988).

Since its commercialisation in 1989, the Jameson Cell has undergone two decades of development in operating plants, with the latest Mark IV designs being highly reliable and operable. Figure 2 shows the development path of the technology in that time frame. Modern designs are fully instrumented and require very little maintenance due to significant design improvements to the downcomer, the unique feature of the technology. The technology has been 'tried and tested', and installed in many different duties but in recent years it has been almost exclusively used in cleaner circuits. The ability of the Jameson Cell to produce final grade concentrates in a single stage of flotation, through fine bubble generation, intense bubble-particle contact and the use of froth washing to minimise entrainment, has seen it being successfully retrofitted into a number of concentrators around the world to add cleaner capacity and/or solve concentrate grade issues (Araya *et al*, 2013).

This paper applies the developments from these brownfield installations to provide a new design basis for improved 'greenfield' cleaning circuits. The new concept is for hybrid cleaning circuits that are simpler, cheaper, and more robust, and that provide better grade/recovery performance. The feature of the improved 'hybrid' circuit is the use of Jameson Cells to control grade and produce the entire plant final concentrate, and mechanical cells for cleaner scavenger duty to provide overall cleaner circuit recovery.

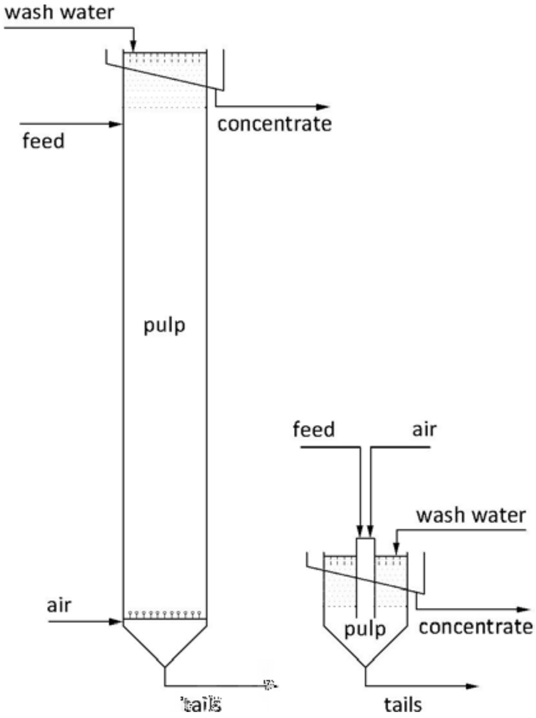


FIG 1 – Comparison of the size of a column cell (left) to a Jameson Cell (right). Reproduced from Jameson (1988).

The operation of such a circuit will be demonstrated in a case study from a recent installation. Engineering considerations such as circuit residence time, footprint and installed power of this new cleaner circuit compared with traditional 'standard' designs, to show that the better performance is accompanied by 30–40 per cent reductions in equipment footprint and energy. The methodology to accurately simulate this circuit design using standard laboratory equipment is described.

DESIGN AND PERFORMANCE OF CONVENTIONAL CELL CLEANING CIRCUITS

The conventional approach to development of cleaner circuit flow sheets is to conduct batch flotation tests to define flotation kinetics and the number of cleaning stages required to achieve the desired concentrate quality. This work should be carried out in conjunction with mineral liberation and regrinding studies. Usually, locked cycle tests are performed to simulate the entire concentrator design in closed circuit with circulating loads. This determines stream flow rates and scale-up of cell capacities. Copper and lead cleaning circuits typically employ two to three stages of counter-current cleaning. Generally, the finer the feed and the lower the head grade, the greater are the number of cleaning stages required to achieve product grade, since separation becomes less effective and entrainment more prevalent. For example, at Glencore's McArthur River and MMG's Century Mine operation, the feed size to the cleaning circuit is around D_{80} of 6–8 microns, requiring five or more cleaning stages to produce the desired zinc concentrate quality. Generally, residence time is the primary factor used to determine the number and size of flotation cells needed for target recovery. But in the last cleaning stage carrying capacity

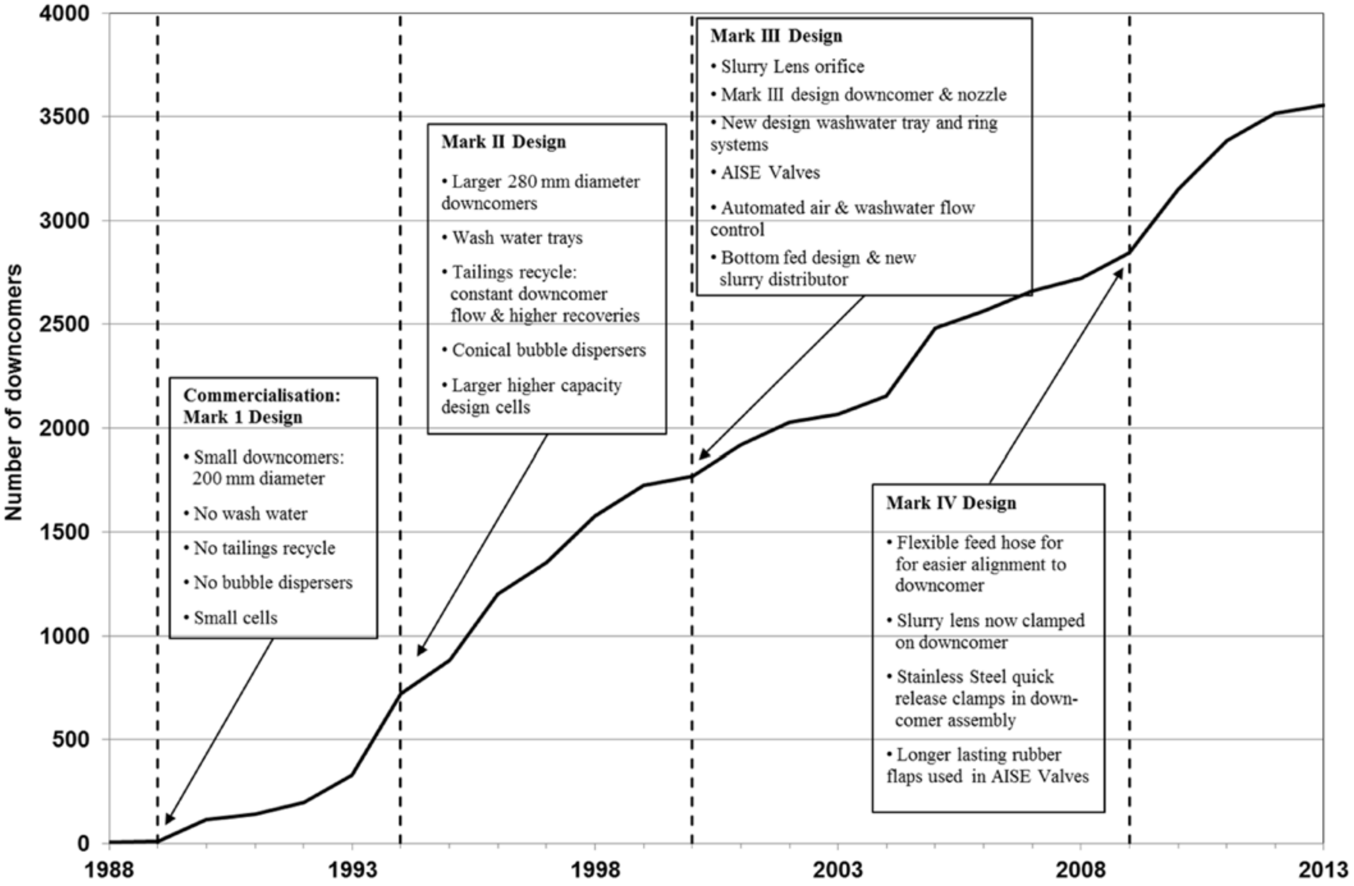


FIG 2 – Development path of the Jameson Cell technology over two decades.

(bubble surface area available) and lip loading (mass flow rate of concentrate per concentrate lip length) may be governing design parameters. In turn, lip loading and carrying capacity are primarily determined by particle size and density, so 'rules of thumb' developed for coarse flotation will be incorrect for finer flotation.

In many plants, issues with conventional cleaner circuits usually relate to either insufficient capacity (and hence recovery), or inability to consistently achieve final grade quality. The former may be due to under-design of the original circuit and/or increased loading as a result of the common practice to 'push more tonnes'. In these cases, performance is very sensitive to feed grade and solids content. Upgrade ratios of valuable minerals generally decrease in each successive stage, hence the need for multiple stages of cleaning (although this also depends on other flotation factors such as ore characteristics, chemistry and particle size). Performance also depends on how effectively operators can make use of the normal flotation controls, ie dart valves for froth depth and valves for air flow rate, to control froth drainage and mass pull, particularly at the last stage of cleaning. Ideally, controls would be available for every cell but this is seldom the case in multistage cleaner circuits. Even in plants with conventional trough cells, level control is across an entire bank and air addition is controlled by a single automated air valve with manual butterfly valves to each cell, which are typically difficult for an operator to access. In any case, banks of conventional cells are not the most efficient way to produce both high-grade and high recovery. The relatively large bubble size and low mixing intensity means relatively slow flotation rates, especially for finer particles. High recoveries can be still be achieved by long residence times, but this is at the expense of grade, since the long flotation times without froth washing means high entrainment. Columns offer the benefit of froth washing to improve grade, but their low mixing intensity (and often the large bubble size) means slow flotation rates and slow recovery of fines.

AN IMPROVEMENT ON CONVENTIONAL CELL CLEANER CIRCUITS – JAMESON CELL CLEANER SCALPING

The above discussion suggests that the best solution may require a combination of technologies. Producing high-grade concentrates (particularly from finer streams) needs concentrate to be froth washed. To be practical, froth washing needs to be applied in a small area – so it needs high flotation rates and shorter residence times. Achieving high flotation rates (particularly for fines) needs small bubbles and intense mixing. But achieving high recovery of the slowest floating particles needs long residence times. No technology has demonstrated the ability to achieve all of these simultaneously. However, by combining the best features of Jameson Cells with mechanical cells, it seems possible to design circuits to achieve higher grade (from fully froth washed concentrate) with high recovery (from adequate residence time in mechanical cells), in an overall smaller installation (from fast flotation rates of small bubbles and intense mixing in the Jameson Cell downcomer).

These design concepts were tested on plant cleaner feed streams in several existing operations. It has been consistently shown that Jameson Cells can produce a higher grade in a single stage compared with multiple (typically two or three) stages of conventional cleaning. An example is shown in Figure 3. The data for the Jameson Cell and the existing plant circuit was collected at the same time during a recent pilot

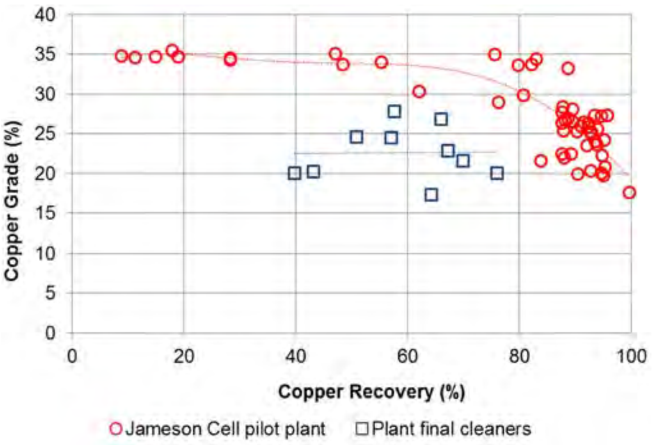


FIG 3 – Jameson Cell performance in a scalping duty compared to a plant with two stages of conventional cleaning.

plant campaign. The Jameson Cell is clearly more selective and operates on a superior grade/recovery curve. It produces a copper concentrate at 35 per cent Cu grade (the ore contains chalcopyrite and secondary copper minerals) in a single stage of flotation, whilst the operating plant cannot consistently produce a final concentrate above 28 per cent Cu using two stages of cleaning in conventional mechanical cells.

Since the increasing application in base metals around 2005 (Young *et al*, 2006), Jameson Cells have been successfully retrofitted to solve capacity and grade issues in a number of copper operations around the world (Araya *et al*, 2013). The referenced brownfield projects and the reasons for Jameson Cell inclusion are as follows:

- PanAust's *Phu Kham* operation in Laos – to increase cleaner circuit capacity (Bennett, Crnkovic and Walker, 2012)
- Newcrest's *Telfer* operation in Western Australia – to reject non-sulfide gangue (NSG) and improve cleaner circuit recovery (Seaman *et al*, 2012)
- Barrick's *Lumwana* operation in Zambia – to reduce uranium mineral entrainment allowing the plant to consistently produce a saleable concentrate below acceptable limits (Araya *et al*, 2014)
- OZ Mineral's *Prominent Hill* operation in South Australia was a new plant in a greenfield project where fluorine was identified to be an issue during the development phase of the project (Barns, Colbert and Munro, 2009).

At these sites, Jameson Cells were installed at the head of the existing conventional cleaner circuit. The Jameson Cell treats the cleaner feed stream producing a very clean high-grade final concentrate. The tailings from the Jameson Cell then go to a cleaner section with conventional mechanical cells, which will now treat a lower grade feed. Because the mineral load to these cells is now significantly lower they can be 'pulled' much more slowly allowing better froth drainage. The feed to the mechanical cells is now at a lower density which aids separation efficiency and reduces entrainment. The concentrate from the final stage of mechanical cell cleaning is combined with the concentrate from the Jameson Cell to produce the overall plant final concentrate. A review of the Jameson Cell performance in these plants showed that:

- The actual unit recovery of the Jameson Cell was generally higher (typically 60 to 80 per cent) than the original design (typically 50 per cent). Maximum recovery was found to be controlled by mineralogy and the amount of liberated mineral together with the quantity of fast-floating particles in the feed, rather than any machine limitations.
- The 'shape' of the selectivity curve between valuable mineral and gangue is dictated by liberation. When treating liberated streams (ie after regrinding), high rejection of gangue (>90 per cent) can be achieved at a valuable mineral recovery as high as 80–90 per cent.
- Carrying capacity was never an issue due to the low mass pull requirement as the upgrade ratios are generally high in this duty (typically greater than five). The tailings recycle designed into the Jameson Cell allows it to handle any large fluctuations in feed and the short 'residence time' in the cell (less than three minutes) allows it to respond quickly to both plant disturbances or changes in process variable settings by operators.
- The Jameson Cell is very forgiving to changes in feed grade and (slurry) density. The operation is robust and the cells require little maintenance.

The performance of the conventional cleaner circuits downstream of the Jameson Cell was found to vary widely across these different sites. A summary at each operation is as follows:

- *Phu Kham* – the Jameson Cell achieved its role in adding sufficient cleaner circuit capacity. After the Jameson Cell was installed, the improved performance of the cleaner circuit increased the overall plant copper recovery by 0.8 per cent (Bennett, Crnkovic and Walker, 2012).
- *Telfer* – the addition of the Jameson Cell increased the copper recovery in the overall cleaner circuit from approximately 85–95 per cent. There was also an overall net benefit in concentrate quality. Figure 4 compares the NSG in the final plant concentrate before and after the Jameson Cell installation. Clearly, the plant is able to produce a cleaner concentrate lower in NSG after Jameson Cell installation (also shown for comparison is the cumulative frequency plot for NSG in the Jameson Cell concentrate). However there was no noticeable difference in the overall copper grade in the plant final concentrate. This is because the Jameson Cell at the head of the circuit recovers most liberated copper at a higher grade, allowing the recleaner cells to recover a larger fraction of the composite particles. That is, the plant is converting the increased cleaning power to higher recovery rather than higher grade.

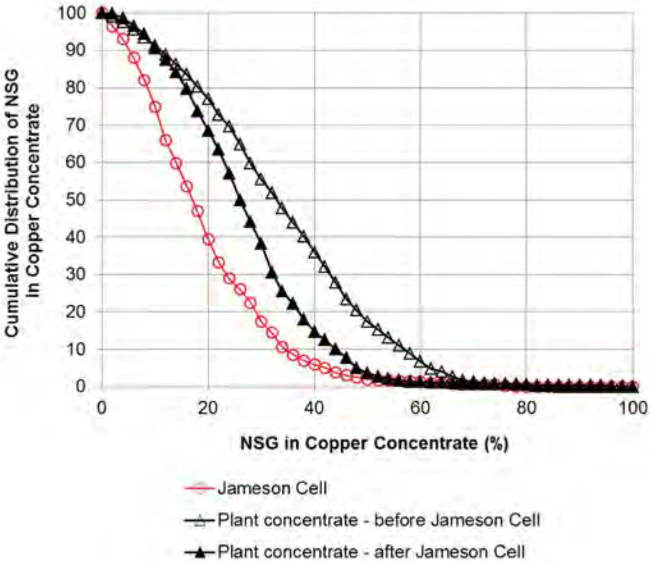


FIG 4 – Cleaner circuit performance at Telfer showing non-sulfide gangue in Jameson Cell concentrate and final plant concentrate before and after installation of the Jameson Cell.

- *Lumwana* – soon after commissioning, the scalper Jameson Cell was switched from a scalping duty to instead replace the plant conventional recleaner bank. This chance was due to equipment constraints and bottlenecks elsewhere in the circuit. The change was successful and the plant now consistently produces on-specification concentrate, eliminating the previous need for concentrate blending. While the Jameson Cell was originally employed to solve a concentrate quality issue at Lumwana, its installation and other initiatives undertaken by site personnel over the past two years has seen recovery increase at the plant of 1.3 per cent (Araya *et al*, 2014).
- *Prominent Hill* – the Jameson Cell produces a high-grade copper concentrate and is very effective in rejecting entrained gangue thereby minimising fluorine levels. The three stages of mechanical cell cleaning produces a lower grade copper grade with higher levels of fluorine (as expected). However, the Jameson Cell contributes a greater proportion (more than the original design) to the final concentrate than the final stage of the conventional cleaners, so the overall final concentrate is still within acceptable fluorine level.

PROGRESSING TO AN OVERALL NEW CLEANER CIRCUIT DESIGN

It is well known that froth washing is the most effective method for reducing the recovery of gangue by entrainment, particularly in cleaning circuits. It is therefore logical that all streams reporting to final concentrate should employ a froth washing type of machine, and the circuit should be designed to maximise overall cleaner circuit recovery. Froth washing is not usually used in mechanical cells. This is partly due to practical constraints – it is difficult to design effective froth washing systems to fit on top of the cells; and the cells do not generally provide a deep and stable enough froth bed to sustain froth washing. Furthermore, the slow flotation rates (due to large bubble size and less intense mixing) means long flotation times and a large surface area required to achieve recovery. To froth wash such a large area requires an impractical amount of water that would overload the rest of the circuit.

The examples described clearly indicate that the addition of cleaner scalper Jameson Cells to these previously all-mechanical cell cleaner circuits was beneficial to the metallurgical performance. The cleaner scalper cells were able to produce a very clean concentrate at relatively high stage recoveries (typically 50–70 per cent), reducing the valuable mineral load on the subsequent mechanical cleaner cells. Final stage concentrate from mechanical cells was higher in NSG content than before the installation of the scalper cells due to the mechanical cleaning circuit receiving a lower grade feed from the Jameson Cell tails. However, the net effect of the new circuit design is an overall reduction in entrainment to the final concentrate and an increase in the overall cleaner circuit recovery.

There is room to further improve these circuit designs to achieve maximum entrainment rejection. At Lumwana, after the installation of the cleaner scalper, further rejection of fine liberated non-floating gangue particles was sought. Since the majority of this gangue was reaching final concentrate from the mechanical recleaner cells, the decision was made to switch the duty of the Jameson Cell from a scalper to a recleaner. This improved overall gangue rejection as the Jameson Cell now produces the entire plant concentrate. In circuits where the first (and/or second stages of cleaners) are capacity limited (eg launder, froth surface area or residence time), this alteration would be unsuitable as it would impact (reduce) the resulting overall cleaner circuit recovery.

Figure 5 proposes an improved cleaner circuit design to combine and maximise the advantages of both Jameson Cells and mechanical cells. It comprises a Jameson Cell cleaner scalper, a single bank of mechanical cells for scavenging cleaner tailings for high recovery, and a second Jameson Cell for final cleaning. The first Jameson Cell recovers the fast floating liberated minerals and produces a high-grade concentrate. It also acts as a buffer in the overall cleaner circuit as it is designed with sufficient capacity to handle varying mass and mineral loads from the rougher/scavenger cells. A Jameson Cell will typically recover 50–70 per cent in a single pass. This recovery range is typical in the final cleaning stage, supported by a sufficient circulating load to provide a high overall circuit recovery. In this design, the circulating load is supported by the mechanical cleaner scavenger cells, which can provide the long residence time for ultimate recovery, while sending the (relatively low-grade) concentrate to another Jameson Cell rather than directly to final concentrate.

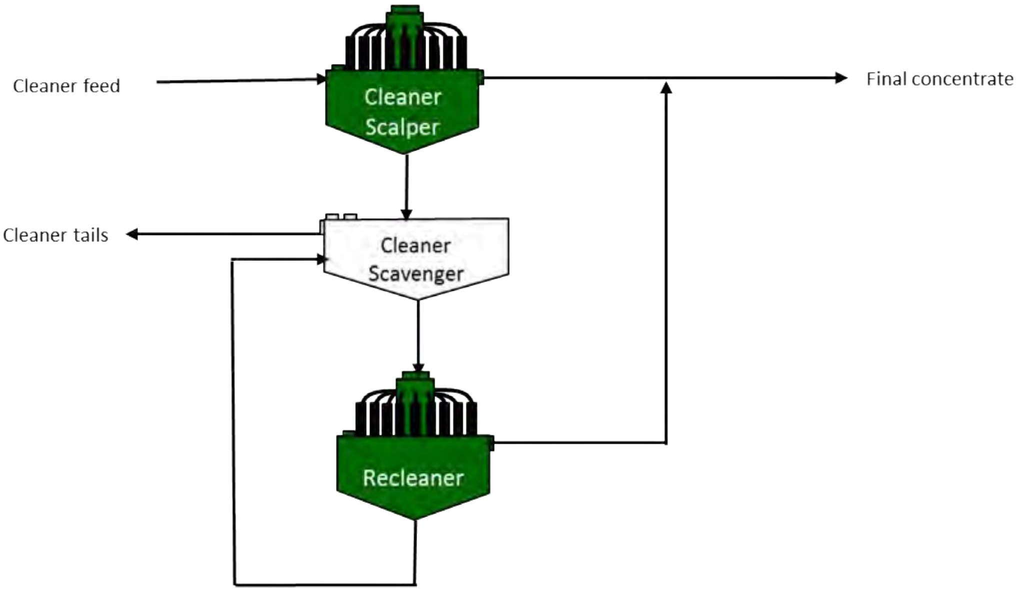


FIG 5 – Proposed new cleaner circuit design with Jameson Cell in cleaner scalper and recleaner duties.

This circuit achieves the grade benefits of fully froth washed concentrate with the long flotation times needed for ultimate recovery, in an overall smaller circuit than either technology could achieve alone.

In comparison to multistage mechanical cleaning circuits, this new circuit can replace two to three stages of traditional mechanical cell cleaning circuits. If further cleaning is required in the case of ultra-fine regrind circuits, the concept can be extended to include an additional third stage Jameson Cell, or an additional mechanical stage between the cleaner scavengers and the recleaner Jameson cell. If only two stages of mechanical cell cleaning are required, it may be possible to dispense with the second Jameson Cell and instead have the cleaner scavenger concentrate report to the cleaner scalper Jameson Cell. If regrinding is required in the circuit, the cleaner scalper cell could either be installed before or after the regrind mill. The optimum location is case-specific and depends on the liberation characteristics of the rougher/scavenger concentrate. The correct water addition for froth washing is determined once the mass flows around the circuit have been estimated.

The overarching principle of the proposed new circuit is to ensure that froth washed flotation machines produce the entire concentrate reporting to final product. The use of a scalper cell minimises the impact of feed variation on circuit performance and mechanical cleaner scavenger cells ensure a high overall cleaner circuit recovery. The circuit should be able to achieve a high enough and dynamically stable cleaner circuit recovery to allow the open-circuit disposal of cleaner tailings. Recently, CSA Mine has refurbished its old concentrator. The old two stage mechanical cell circuit was replaced with the new circuit as described here. The performance of the new circuit is presented as a case study later in this paper.

Choice of flotation technology

Although this paper describes the use of the Jameson Cell technology, the improved cleaner circuit design principles can be applied to other flotation machines incorporating froth washing, such as column cells. Column technology was invented in the 1960s but did not gain wide acceptance in base metals operations until the early 1980s. It is still a popular technology in the cleaning circuits of many operations in the Americas. Schena and Casali (1994) reported the use of this technology in various cleaner circuit configurations at different South American copper operations. However, column cells have not featured strongly in Australian operations. A review of these installations is provided by Lane and Dunne (1987).

The Jameson Cell has often erroneously been classified as a column cell probably due more to the similarity of applications rather than the separation technology. However, it operates on completely different principles and a more accurate description for the Jameson Cell is a reactor/separator type of flotation technology as stated by Finch (1995). Table 1 provides a comparison of the main characteristics of column and Jameson Cell technologies. Technology selection will always be a topic of debate and will be affected by individual experience. Ultimately, the best indicator of the strength of a flotation technology is the analysis of its performance in operating plants. This will account not for the 'design' or laboratory performance, but the actual performance over time in real plant conditions. This will include the ease of operation and maintenance, and the consistency of critical performance variables such as bubble size. Feedback from users and rigorous analysis of actual operating performance should prevail over laboratory or pilot results or 'head-to-head' testing. This is only meaningful if each technology can be accurately scaled up and can maintain the small-scale

TABLE 1

Characteristics of Jameson Cell and column flotation technology.

Column cell	Jameson cell
• Spargers (jetting or shearing types) required for bubble generation • Compressors required	• Air for flotation entrained from atmosphere • No compressors or blowers required
• Long residence time • Need large tank volumes	• Very short residence time • Small tank volumes (approximately five times less than 'equivalent' column cell) • Collection/contact in downcomer, tank used for separation only
• Low intensity – need large collection/contact zone (typically >15 minutes) • Disadvantage for fine particle recovery (Finch, 1998)	• High intensity for efficient bubble-particle contact (five to ten seconds in downcomer) • Total residence in tank (for separation only) is two to three minutes • Advantage for fine particle recovery
• Low/medium gas hold-up – ten to 20 per cent (Hernandez, Gomez and Finch, 2003)	• Very high gas hold-up (in collection zone) – 40 to 60 per cent (Jameson, 1988)
• Bubble size is typically 1500 to 3000 microns – jetting and shearing spargers (Pyecha *et al*, 2006; Lizama, Carrion and Estrella, 2008)	• Bubble size – typically 300 to 700 microns (Evans, Atkinson and Jameson, 1995) • Fine bubbles increase kinetics of *all* particles, not just fine ones • Fine bubbles mean higher carrying capacity and hence, productivity (production tonnage per square metre cell area)
• Need scale-up factor to account for short-circuiting (bypass) and differences in froth recovery/froth drop-back between pilot and full-scale cells (Dobby, 2002)	• Direct scale-up. Hydrodynamic conditions (jet velocity and mixing intensity) are the same between pilot plant and full-scale cells (Young, Barns and Pease, 2006) • No chance of short circuiting (so no bypass) – all feed has to pass through the downcomer at least once
• Medium/high maintenance – spargers and compressors • Sparger replacement typically six to 18 months	• Low maintenance – centrifugal fixed speed pump • Slurry lens orifice (to create high-pressure jet) replacement typically every +5 years
• Power required for compressors and recirculation pumps (Microcels and Cavitation tube-type columns only) • A portion of *tailings* is recirculated through cell, equal to many multiple times the feed flow	• Power required for downcomer *feed* pump • Tailings exit cell, a portion (less than or equal to feed) is mixed with fresh feed and passed through the downcomer again (giving another chance for particles to be collected) • Tailings are not simply recirculated through cell

performance (eg bubble size, operability) over the long run in a real plant.

Both columns and Jameson Cells use froth washing to control entrainment. However to achieve the higher grade in a smaller circuit needs fast flotation rates. In this respect column technology has a disadvantage because of its low flotation rates for fine particles due to the low mixing intensity (Finch, 1988; Dobby, 2002). In contrast, high flotation rates are a distinguishing feature of Jameson Cells, since they consistently provide very small bubble size and highly intense mixing. This is particularly important for the cleaning of finer streams, which are becoming more prevalent. This results in fast flotation rates necessary for the smaller more efficient circuits described here as ores become more complex and need finer regrinding for mineral liberation.

CSA Mine case study

CSA Mine is a copper operation 14 km north of Cobar in New South Wales, owned by Cobar Management Proprietary Limited (CMPL), a wholly owned subsidiary of Glencore. Its processing plant treats mainly chalcopyrite ore from its underground mine. Feed to flotation has a head grade around five per cent Cu varying from three to eight per cent Cu. The old flotation circuit consisted of roughing, scavenging and two stages of mechanical cell cleaning to produce final grade concentrate of 27.5 per cent Cu at over 96 per cent recovery. The large variations in the copper head grade often overloaded the circuits so operators had to use bypass plates in the launders of the first cells of both the roughers and the first cleaner bank to divert some concentrate directly to the final concentrate. The flotation circuit required considerable operator intervention and 'bypassing' material to final concentrate was never a preferred solution as it reduced the final concentrate grade.

In 2011, a project was initiated to increase throughput from 160 t/h to 205 t/h. This project was to be implemented over several years and included major changes to the grinding circuit as well as a complete overhaul of the flotation circuit which then consisted entirely of Denver cells installed in 1965 (Erepan, Rajiwate and Beehan, 2014). There was an initiative to increase concentrate grade and to reduce concentrate rail transport costs to the port of Newcastle. CSA Mine management requested a new cleaner circuit to increase concentrate grade to 30 per cent Cu without sacrificing plant recovery. The cleaner circuit now consists of an E4232/10 model (10 downcomer) Jameson Cell for cleaner scalping, three new Outotec TK40 tank cells for cleaner scavenging, and an E1732/4 model (4 downcomer) Jameson Cell for recleaning the conventional cell concentrate. The CSA flow sheet with the new cleaner circuit is shown in Figure 6. This cleaner circuit upgrade project was managed in-house by CMPL and was completed in October 2013.

Figure 7 shows the performance of the two Jameson Cells from surveys conducted during commissioning. The grade/recovery (Figure 7a) and selectivity curves (Figure 7b) for the two Jameson Cells, although quite different are in-line with expectations: the cleaner scalper cell produces a higher grade concentrate than the recleaner cell. The recovery in the cleaner scalper was quite high owing to the very large quantity of fast floating copper mineral particles in the rougher concentrate stream. The recleaner appears to be recovering a lot more composite particles as evidenced by the different copper selectivity response with silica compared to the cleaner scalper cell. The wide variation in data points is due to deliberately operating the Jameson Cells with different parameters (air flow rate, froth depth and wash water) to vary performance demonstrating that the Jameson Cells can be 'tuned' to operate at any point on the grade/recovery curve.

During the Jameson Cell surveys cleaner scavenger tailings were also collected to allow calculation of recovery across the entire cleaner circuit. Eight surveys over two days showed the overall cleaner circuit recovery to be over 98 per cent while producing a final plant concentrate grade averaging 29.5 per cent Cu. The overall plant recovery was maintained above the target of 96 per cent. Commissioning of Jameson Cells and the new cleaner circuit took place at a plant throughput of 160 t/h, and not the design criteria of 205 t/h for sizing the Jameson Cells as the plant is yet to ramp up to the higher tonnage. At the time of writing, the CSA Mine had been operating the new cleaner circuit for around five months. Figure 8 shows the plant concentrate grade approximately two years prior to the installation and since the new circuit was installed. In the months after the new circuit was installed, the plant treated more challenging ore, which has unusually

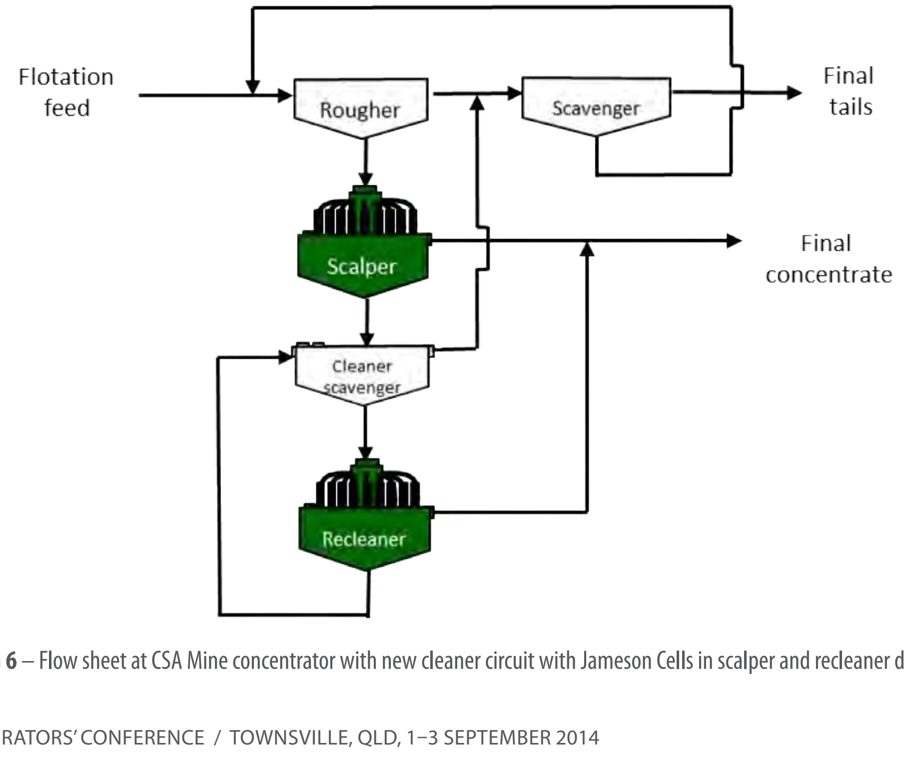


FIG 6 – Flow sheet at CSA Mine concentrator with new cleaner circuit with Jameson Cells in scalper and recleaner duties.

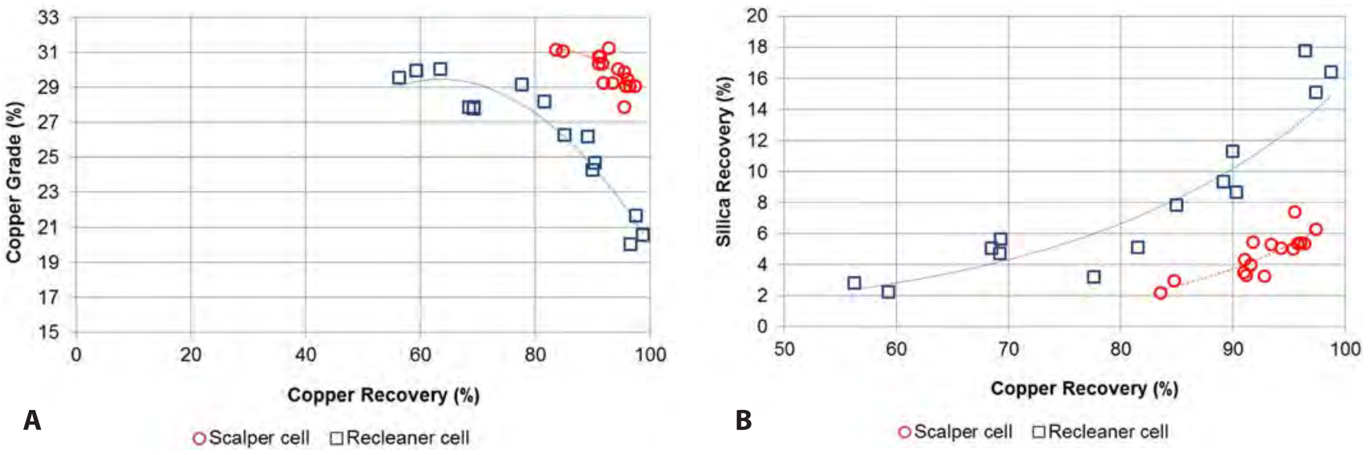


FIG 7 – (A) Copper grade/recovery and (B) copper/silica selectivity curves for the Jameson Cells at CSA Mine.

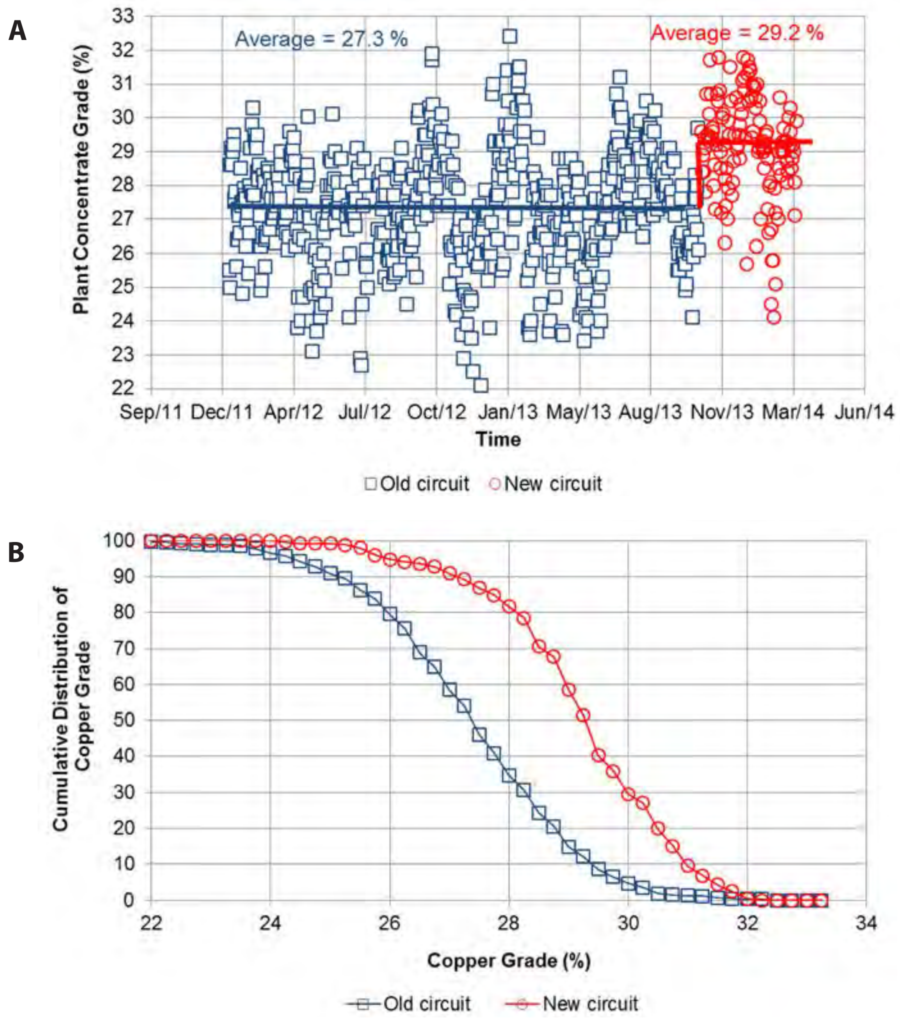


FIG 8 – CSA Mine final concentrate grade before and after installation of new cleaning circuit. (A) Plant final concentrate grade as timeline and (B) same data shown as cumulative distribution.

high quantities of readily floating pyrrhotite gangue and cubanite ($CuFe_2S_3$, 23.4 per cent Cu), which both tend to lower the final concentrate grade. However, the shift assays have showed that the new circuit is still capable of producing a higher average grade final concentrate (29.2 per cent Cu) compared with the old circuit (averaging 27.3 per cent Cu). The difference in final concentrate grade of the two circuits is more evident when the same data is plotted as a cumulative distribution curve as shown in Figure 8b. The ongoing aim is to elevate the final concentrate to at least 30 per cent Cu which is expected to happen when the plant returns to treating the 'normal' ore type.

As well as higher concentrate grade, the plant reports significant operational improvements. The new cleaner circuit has proven to be much more robust and better able to handle the fluctuations in feed grade without operator intervention or the need to bypass material. The new operating strategy for the CSA Mine concentrator is to operate the cleaner scalper

Jameson Cell on the 'flat' part of the grade/recovery shown in Figure 6a and then 'tune' the recleaner Jameson Cell to achieve the overall desired cleaner circuit performance. For example, for maximum plant final concentrate grade, the recleaner should be operated below 80 per cent unit recovery as silica recovery increases significantly above this point as shown in Figure 6b. However, if a circulating load builds up around the cleaner circuit and starts to affect (decrease) overall plant recovery, the recleaner Jameson Cell can be 'pulled' harder to reduce the load to the cleaner scavenger bank. The short residence time of the new circuit ensures it responds quickly to such changes.

Interestingly, another copper operation in Australia has been operating Jameson Cells in the new cleaner circuit configuration since 1995, although that was not the initial intent. Jameson Cells were originally designed to operate in a two-stage cleaning configuration, but during commissioning it was quickly realised that the first cleaning stage was already producing final grade concentrate, making a second stage of cleaning redundant. As a result these cells were altered to treat the concentrate from the conventional cleaner scavenger bank instead. So in fact the improved cleaner circuit is not even new. The only difference is that the Jameson Cells installed at CSA Mine are much more 'user friendly' being the latest Mark IV design compared with those at this copper operation which are the much earlier Mark II design.

Other proposed benefits

So far discussions on the new cleaner circuit design have focused purely on the benefits in metallurgical performance. To explore other potential benefits for greenfield projects, the new cleaner circuit design was compared to commonly designed 'standard' flow sheet. The size and number of mechanical (or column) cells required for the standard cleaner circuit was selected by an independent engineering design house while design of the new circuit was undertaken by the authors. It is assumed the three concentrators are treating copper ore (chalcopyrite as the economic mineral) with head grades between 0.5 to 1.0 per cent Cu. The three projects are described as follows:

- Project A is a small/medium sized plant (<1000 t/h) where the cleaner circuit has three stages of conventional cleaning, designed in the usual counter-current closed circuit configuration
- Project B is a medium/large sized plant (>1000 and <3000 t/h) where the cleaner circuit has three stages of conventional cleaning, again designed in the usual counter-current closed circuit configuration
- Project C is a very large copper plant (>5000 t/h similar to those typically seen in porphyry copper operations) where there are also two stages of cleaning, but mechanical cells are only used for the first cleaner stage whilst column cells are chosen for recleaning.

Table 2 compares the most important design aspects of the two cleaner circuit designs:

- number of cells installed – less cells means lower installation costs, less ongoing operating, maintenance and spare parts costs
- circuit total residence time – shorter residence time makes for a more robust and responsive circuit that is faster to 'tune' and optimise
- footprint – smaller footprint has obvious advantages in foundation and structures, building and installation costs
- installed motor power – for consideration of ongoing energy usage and cost.

Table 3 compares these parameters for the two different circuit designs. It shows significant benefits in the new cleaner circuit, mainly as a result of the reduction in number of cells and total cell volumes. In summary, it shows that for the three scenarios, the improved circuit design reduces residence time by around 70 per cent, footprint by 30–50 per cent and uses 30–40 per cent less power. This design comparison is a simple exercise that can be independently estimated by potential users for their application.

The development of a lower cost circuit that is also more effective is a significant step. As is often the case, the 'step change' is not really new; rather it has been 'assembled' by combining existing knowledge and developments. It has been developed in brownfields applications with specific constraints and objectives. This provided a low risk demonstration and development of the concepts, equipment and techniques. Several other brownfield projects are now following this success and implementing circuits to produce all final concentrate from Jameson Cells. In the common pattern, successful application at brownfield sites leads to adoption for greenfield projects. At the time of writing several projects at definitive feasibility stage (DFS) have the new cleaner circuit locked into the flow sheet design.

Laboratory tests to simulate 'new' cleaner circuit design

Flow sheet development occurs early in the metallurgical development of an orebody and is typically conducted on a small number of blended drill core intervals to represent the ore. At this stage, required reagent schemes, pulp chemistry

TABLE 3

Potential savings of the new cleaner circuit design over standard designs.

Project	Circuit residence time (%)	Footprint (%)	Installed motor power (%)
A	71	54	37
B	73	33	30
C	74	43	30

TABLE 2

Comparison of standard and new cleaner circuit designs.

Project	Standard circuit design				New cleaner circuit design			
	Number of cells	Residence time (minutes)	Footprint (m^2)	Installed motor power (kW)	Number of cells	Residence time (minutes)	Footprint (m^2)	Installed motor power (kW)
A	15	319	88	385	7	92	41	243
B	19	210	188	1413	7	58	126	995
C	22	245	732	5498	12	64	417	3839

conditions, grind size targets and overall circuit configuration are developed by conducting a series of 'standard' tests. The overall flow sheet is settled before applying the same set of conditions over a much greater range of ore-variability samples to develop flow sheet design criteria and metallurgical models for economic evaluation. Flow sheet development test work is conducted in laboratory batch experiments, with locked cycle tests generally being conducted on a select number of samples to confirm closed-circuit performance and flow rates for design purposes. Depending on the complexity of the orebody and the owner's risk appetite, a continuous pilot plant evaluation of a large bulk sample may be carried out, or mini-continuous pilot plant runs may be completed on multiple samples.

To evaluate the improved cleaner circuit for a new orebody, a practical and accurate laboratory evaluation procedure is needed during this stage of design. Such a procedure, using standard laboratory equipment, has been developed and is described here.

The Jameson Cell scales up very accurately from continuous laboratory and pilot scale rigs. However this requires volumes that can be impractical for laboratory testing. Fortunately, a procedure to accurately simulate Jameson Cell performance using standard laboratory batch cells has been developed and verified. Several of the brownfields applications described in this paper successfully used this approach. For example, Seaman *et al* (2012) described laboratory tests used to justify installation of two Jameson Cells at Telfer without the need for pilot plant testing. The full-scale performance of these cells was predicted by dilution batch tests. This procedure can be applied with reasonable confidence for the new cleaner circuit design when developing a flow sheet for a greenfield project.

Figure 9 compares dilute batch flotation test selectivity of copper versus penalty element recovery with parallel pilot Jameson Cell performance over a wide range of operating conditions. In each case, the batch dilute flotation test was carried out by collecting a sample from the operating plant, then diluting the slurry in process water to ensure the slurry solids content was less than ten per cent w/w. Following this a batch flotation test was conducted with frother as the only reagent added (and only as required) to ensure generation of a stable froth during the test. Unlike conventional batch tests, these dilution tests were conducted with very low scraping (froth removal) rates, eg ten to 20 second intervals (with froth paddles slowly pulled across the froth) to allow as much drainage from the froth as possible.

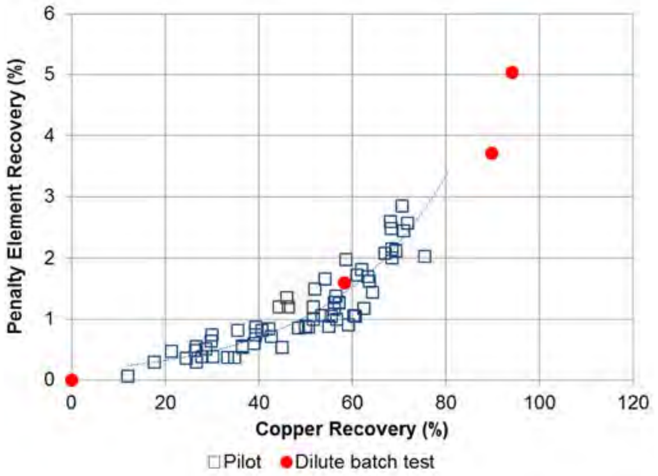


FIG 9 – Pilot Jameson Cell versus dilute batch flotation test selectivity.

Figure 10 shows an example of a complete testing regime that can be conducted for a new project to simulate the new cleaner circuit design. The recommended procedure is as follows:

- After regrinding (if required), place slurry (rougher concentrate) in a laboratory batch flotation cell and dilute to less than 10 wt per cent solids.
- Condition with relevant collectors, depressants, pH modifiers.
- Add frother throughout the test as required to maintain a stable froth phase and maintain the level of the cell with addition of water to give approximately 15–30 mm froth depth. Adjust airflow so that the froth is almost free-flowing over the launder.
- Collect concentrates by using a very slow scraping rate of the entire froth phase into a container. Suggested scrape rates are once every 20 seconds, with the paddle pulled very slowly (taking approximately 15 seconds for one scrape) across the froth in order to maximise concentration of valuables and to minimise entrainment of gangue.

The above procedure is applied to simulate both Jameson Cell stages (cleaner scalper and recleaner). To test the cleaner flow sheet shown in Figure 10b, a suitable flotation time or concentrate mass pull for the cleaner scalper must first be identified. This involves conducting a test where four to five concentrates are recovered from the rougher concentrate (Figure 10a) and analysing the resulting selectivity curve of the

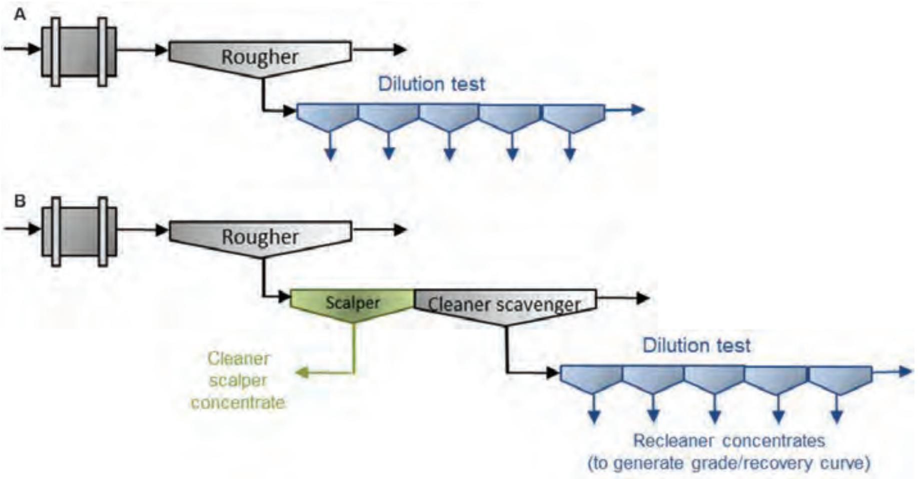


FIG 10 – Laboratory flotation test procedure to simulate the new cleaner circuit design. (A) Cleaner scalper batch test, (B) complete cleaner circuit batch tests.

cleaner scalper. The optimum point would be located left of the 'knee' of the selectivity curve. For example, in Figure 9 this would target copper recovery of approximately 50 per cent. In this case, the concentrate was collected only after one minute (three scrapes). If flow sheet development time or ore sample quantity is limited, an arbitrary time of one minute (or three scrapes) should be used as standard — experience has shown that this is close enough to the targeted recovery/grade for a copper circuit for a cleaner scalper duty.

The cleaner scavenger stage batch flotation test is carried out as a regular batch test following the removal of the cleaner scalper concentrate using the normal ten second scraping rates to remove the remainder of the floatable mineral particles. Then, a final dilution cleaner test will need to be conducted on the cleaner scavenger concentrate using a similar flotation time as that used to generate the rougher concentrate, with multiple concentrates collected in order to generate a grade/recovery curve to represent the entire cleaning circuit. The procedure described above for the new cleaner circuit design and shown in Figure 10b is also suitable for locked cycle tests as per typical practice for any other cleaner circuit design.

The actual residence times in the cleaner scalper and recleaner stage are not relevant for scale-up determination as residence time plays no role in the sizing of Jameson Cells. This information is only required for the sizing of the cleaner scavenger bank. The results from these tests should then allow the design engineers to use the grade and recovery numbers generated for equipment selection. The sizing of the Jameson Cell for each duty is provided by the flotation technology vendor. Rather than looking at each duty in isolation, the criteria for the design of the entire cleaner circuit is:

- to ensure there is sufficient capacity in the Jameson Cells to produce the desired range of plant final concentrate tonnages
- to ensure that there is sufficient capacity in the cleaner scavenger bank to allow the tailings from these cells to be sent to final tailings.

CONCLUSIONS

Case studies have been presented showing that the introduction of Jameson Cells into cleaning circuits has enhanced selectivity by increasing fine gangue rejection whilst maintaining or improving overall cleaner circuit recovery. A new cleaner circuit design has been presented, along with operating data from CSA Mine where the circuit was implemented as part of an expansion to the operating plant. The circuit design promises not only improved metallurgical performance, but also greater stability in the face of feed variation, a more robust operation and is more operator-friendly with a lower maintenance requirement.

The improved cleaner circuit design is relatively new for greenfield projects, although it has been already well proven in several brownfield cases. A laboratory procedure for development of this new circuit has been presented using dilution batch tests which have been proven to scale-up to both Jameson Cell pilot plant and full-scale plant data. This allows accurate prediction of the performance of this circuit in routine laboratory flow sheet development. This new approach will position the industry to process the more challenging ore types that are expected in the future where the emphasis will be on consistent operation and achieving high recoveries at target grades from poorer quality ores. The relatively small number of current installations has already demonstrated the potential for both new and existing plants and the opportunities will continue to emerge.

As well as the significant improvement in concentrate grade/recovery performance, the new cleaner circuit significantly reduces the required number of cells and equipment compared to current standard cleaning circuit designs. This means that better performance can be achieved in less space, for lower capital cost, lower operating cost, and significantly less energy use.

ACKNOWLEDGEMENTS

The authors would like to thank the management of CMPL for permission to publish their data. Thank you also to Dave Osborne and Joe Pease from XT for their review and input to this paper.

REFERENCES

Araya, R, Cordingley, G, Mwanza, A and Huynh, L, 2014. Necessity driving change and improvement to the cleaner circuit at Lumwana copper concentrator, in *Proceedings 46th Canadian Mineral Processors Conference*, pp 329–341 (Canadian Institute Mining, Metallurgy and Petroleum: Montreal).

Araya, R, Huynh, L, Young, M and Arburo, K, 2013. Solving challenges in copper cleaning circuits with the Jameson Cell, in *Proceedings Tenth International Mineral Processing Conference,* pp 261–271 (GECAMIN: Santiago).

Barns, K E, Colbert, P J and Munro, P D, 2009. Designing the optimal flotation circuit – the Prominent Hill case, in *Proceedings Tenth Mill Operators' Conference,* pp 173–182 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Bennett, D, Crnkovic, I and Walker, P, 2012. Recent process developments at the Phu Kham copper-gold concentrator, in *Proceedings 11th AusIMM Mill Operators' Conference*, pp 257–272 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Dobby, G, 2002. Column flotation, in *Proceedings Mineral Processing Plant Design, Practice, and Control Symposium* (eds: A Mular, D Halbe and D Barratt), pp 1239–1252 (Society for Mining, Metallurgy and Exploration: Littleton).

Erepan, P, Rajiwate, F and Beehan, D, 2014. Flotation circuit and dewatering improvements at CSA concentrator, *12th AusIMM Mill Operators' Conference,* pp 00-00 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Evans, G M, Atkinson, B W and Jameson, G J, 1995. The Jameson Cell, in *Flotation Science and Technology*, pp 331–363 (Marcel Dekker: New York).

Finch, J A, 1995. Column flotation: a selected review – part IV: novel flotation devices, *Minerals Engineering*, 8(6):587–602.

Finch, J A, 1998. 1998 Gaudin Lecture: Fundamental fallout from column flotation, *Mining Engineering*, 50(12):49–56.

Harbort, G J, Manlapig, E V and DeBono, S K, 2002. Particle collection within the Jameson Cell downcomer, *Transactions of the Institutions of Mining and Metallurgy, Mineral Processing and Extractive Metallurgy*, 111:C1–C10.

Hernandez, H, Gomez, C O and Finch, J A, 2003. Gas dispersion and de-inking in a flotation column, *Minerals Engineering*, 16:739–744.

Jameson, G J, 1988. A new concept in flotation column design, in *Proceedings Column Flotation '88*, pp 281–285 (Society of Mining Engineers: Littleton).

Johnson, N W, McKee, D J and Lynch, A J, 1974. Flotation rates of non-sulphide minerals in chalcopyrite flotation processes, *Transactions of Society of Mining Engineers*, 256:204–209.

Lane, G S and Dunne, R C, 1987. Column flotation – an Australian perspective, in *The AusIMM Kalgoorlie Branch Equipment in the Minerals Industry: Exploration, Mining and Processing Conference*, pp 81–94 (The Australasian Institute of Mining and Metallurgy: Kalgoorlie Branch).

Lizama, H, Carrion, J and Estella, D, 2008. Improvements in column flotation through the use of Microcrel spargers at Antamina, in *Proceedings 46th Canadian Mineral Processors Conference*, pp 363–375 (Canadian Institute Mining, Metallurgy and Petroleum: Montreal).

Pyecha, J, Lacouture, B, Sims, S, Hope, G and Stradling, A, 2006. Evaluation of a Microcel sparger in the Red Dog column flotation cells, *Minerals Engineering*, 19:748–757.

Savassi, O N, Alexander, D J, Franzidis, J P and Manlapig, E V, 1998. An empirical model for entrainment in industrial flotation plants, *Minerals Engineering*, 11(3):243–256.

Schena, G and Casali, 1994. Column flotation circuits in Chilean copper concentrators, *Minerals Engineering*, 7(12):1473–1486.

Seaman, D R, Burns, F, Adamson, B, Seaman, B A and Manton, P, 2012. Telfer processing plant upgrade – the implementation of additional cleaning capacity and the regrinding of copper and pyrite concentrates, in *Proceedings 11th AusIMM Mill Operators' Conference*, pp 373–381 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Young, M F, Barnes, K E, Anderson, G S and Pease, J D, 2006. Jameson Cell: the comeback in base metals applications using improved designed flow sheets, in *Proceedings 38th Annual Meeting of the Canadian Mineral Processors*, pp 311–332 (Canadian Institute Mining, Metallurgy and Petroleum: Montreal).

Материалы 42-ой Ежегодной конференции Канадского общества переработки минерального сырья - 2010

ДОКЛАД 14

“Нужное оборудование в нужном месте”: увеличение производительности по никелю на комбинате Cosmos компании Xstrata Nickel Australasia

Дэн Керри[1], главный управляющий комбината

Майкл Купер[1], управляющий комбината

Джош Рубенстайн[2], старший инженер-технолог

Том Шоулдис[3], президент

Майкл Янг[2], главный металлург

---

[1] Xstrata Nickel Australasia, Level 3, 24 Outram Street, West Perth, WA 6006
Тел. +61 (8) 9213 1588
dcurry@xstratanickel.com.au
mcooper@xstratanickel.com.au
[2] Xstrata Technology, Level 5, 509 Richards Street, Vancouver, BC V6B 2Z6
Тел. (604) 699 6402
jrubenstein@xstratatech.com.au
myoung@xstratatech.com.au
[3] G & T Metallurgical Services Ltd., 2957 Bowers Place, Kamloops, BC V1S 1W5
Тел. (250) 828 6157
tshouldice@gtmet.com

Ключевые слова: IsaMill™, флотомашина Джеймсон, Wemco® SmartCell™, никель, технологическая минералогия

42-ое Ежегодное собрание Канадского общества переработки минерального сырья

19-21 января 2010 года, Оттава, Онтарио, Канада

АВТОРЕФЕРАТ

Комбинат Cosmos компании Xstrata Nickel Austrasia (шт. Западная Австралия) в настоящее время осуществляет проект устранения узких мест в технологической схеме с целью увеличения производства никеля при сохранении металлургических показателей и низкой удельной себестоимости продукции. Результаты минералогического анализа показали, что узкие места в технологической схеме можно устранить, изменив отдельные участки технологического цикла, а также увеличив мощность производства с использованием новых технологий, таких как IsaMill™, флотомашина Джеймсон, Wemco® SmartCell™ и фильтр Larox.

Флотомашина Wemco® SmartCell™, идеально подходящая для флотации грубого минерального сырья, и флотомашина Джеймсон, признанная оптимальной для быстро флотируемого и мелкого минерального сырья, могут быть установлены на основной технологической линии цикла флотации в качестве флотомашин скальпирования основной флотации. Установка двух дополнительных флотомашин позволит получить достаточную мощность флотации с учетом ожидаемого повышения содержания никеля в питании и тоннажа.

Установка мельницы IsaMill™ для доизмельчения позволит сделать более грубым первичный продукт мельницы полусамоизмельчения, тем самым повысив производительность. Среди преимуществ, предлагаемых мельницей IsaMill™, – возможность работы в открытом цикле (внутренняя классификация), инертная (керамическая) среда измельчения, облегчающая последующую флотацию (более чистые новые поверхности), компактная конструкция (удобство установки на действующем производстве) и энергоэффективность. Одновременно с увеличением мощности основной флотации, схема установки остальных существующих флотомашин была изменена с учетом новой мельницы доизмельчения и требований цикла перечистки. Существующий фильтр-пресс Larox 24м^2, используемый для обезвоживания концентрата, был заменен фильтром 32 м^2 с учетом увеличенного выхода концентрата.

ВВЕДЕНИЕ

Горно-обогатительный комбинат Cosmos расположен в 680 км к северо-востоку от Перта, шт. Западная Австралия. Разработка месторождения была начата в 1999 году компанией Jubilee Mines NL, предыдущим владельцем комбината Cosmos, а первый никелевый концентрат был получен в апреле 2000 года. Комбинат перерабатывает высокосортную никелевую сульфидную руду в коматиитовой вмещающей породе; сульфидные минералы преимущественно представлены пентландитом и пирротитом, в меньшей степени пиритом и халькопиритом.

Компании Jubilee принадлежали одни из самых перспективных и относительно малоизученных никелевых месторождений в мире, что обеспечивало отличный потенциал расширения производства. Это хорошо сочеталось со стратегией развития компании Xstrata Nickel, и в октябре 2007 года Xstrata сделала предложение о покупке Jubilee Mines NL. В феврале 2008 года Xstrata приняла на себя функции управления активами Jubilee и учредила компанию Xstrata Nickel Australasia (XNA), новое производственное подразделение Xstrata Nickel.

Стратегия интенсивного роста XNA потребовала внесения изменений в перерабатывающее производство с учетом увеличения производства никеля из руды, добываемой на руднике Просперо комбината Cosmos (рис. 1). Увеличение производства никеля было связано с увеличением не только тоннажа, но и содержания никеля в руде. Обогатительная фабрика комбината Cosmos был спроектирован более 10 лет назад и был рассчитан примерно на 3 года отработки одноименного месторождения. Учитывая возраст и проектный срок службы фабрики, а также его компактную планировку, нельзя было не ожидать определенных трудностей, связанных с его модернизацией для увеличения мощности. В ноябре 2008 года обогатительная фабрика Cosmos имела массовую производительность 25 т/ч и производительность по никелю в питании 1,3 т/ч. Чтобы получить запланированный на 2009 год выход никеля, необходимо было увеличить производительность по никелю в питании 2,8 т/ч (+215%) и массовую производительность мельницы полусамоизмельчения 45 т/ч (+180%).

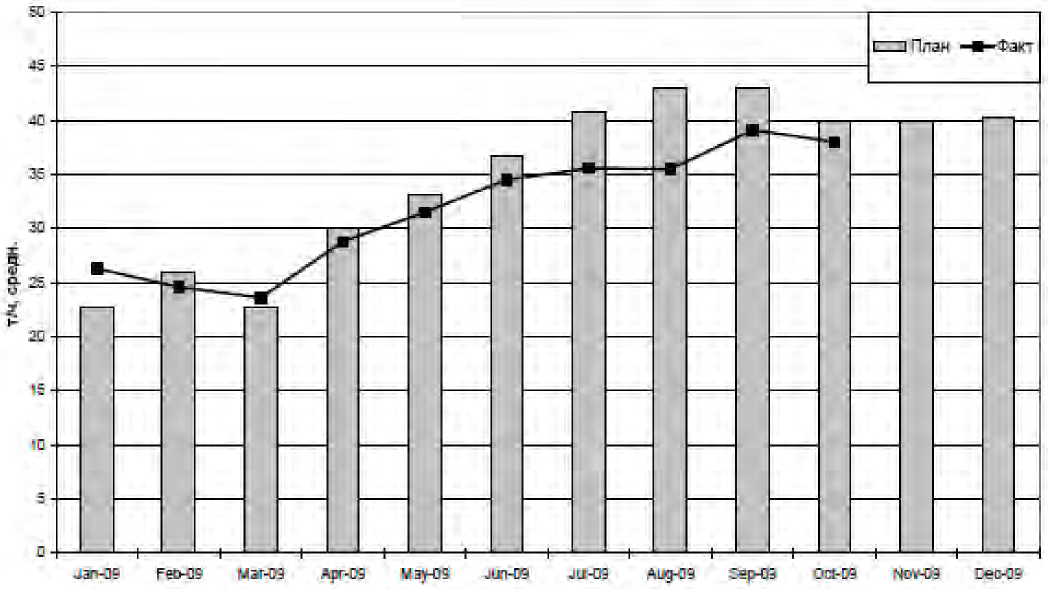


Рисунок 1. Производительность мельницы полусамоизмельчения на комбинате Cosmos (фактическая в 2009 году на указанную дату против плановой)

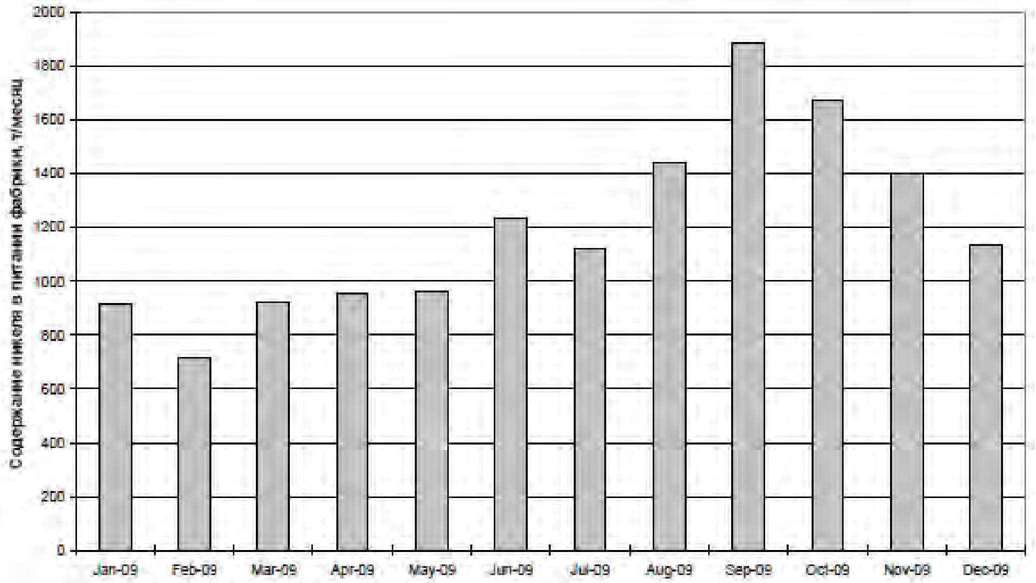


Рисунок 2. Производительность обогатительной фабрики Cosmos по никелю в питании, т/месяц согласно плану на 2009 год

Были выполнены минералогические и металлургические исследования, результаты которых показали, что выйти на высокие плановые показатели можно при условии установки и реконфигурации нескольких единиц ключевого оборудования. На основании полученных результатов и с использованием принципа применения оптимального оборудования была начата программа устранения узких мест в технологической схеме. Хотя она еще не завершена, программа уже доказала свою эффективность и будет выполнена согласно плану (за 12 месяцев) и с минимальными (по меркам отрасли) затратами.

Решения, используемые при устранении узких мест в технологической схеме, были и остаются основаны на минералогическом подходе. Вносимые в технологию изменения основаны на данных анализа извлечения по фракциям крупности, высвобождения и минералогических подсчетах. Для оценки эффективности вносимых изменений ключевое значение имели определения извлечения целевых минералов по фракциям в основных технологических потоках. Принцип “нужного оборудования в нужном месте” подразумевал реконфигурацию части оборудования для получения измененной технологической схемы и применение новых технологий, таких как IsaMill™, флотомашина Джеймсон и Wemco® SmartCell™, для получения требуемых металлургических показателей.

Ниже представлено краткое описание исходной технологической схемы и соответствующей технологической минералогии; далее следует описание изменений, внесенных при устранении узких мест, и их влияние на технологические показатели обогатительной фабрики по фракциям крупности.

ИСХОДНАЯ ТЕХНОЛОГИЧЕСКАЯ СХЕМА И ЕЕ ЭФФЕКТИВНОСТЬ (РЕЗУЛЬТАТЫ ИССЛЕДОВАНИЯ, ВЫПОЛНЕННОГО В ИЮЛЕ 2008 ГОДА)

В этой части будет рассмотрена схема фабрики, характеристики питания флотации, показатели флотации, состав конечного концентрата и показатели уноса с хвостами до программы устранения узких мест, причем последние два аспекта имеют особое значение. Иными словами, что присутствует в концентрате из того, чего в нем не должно быть, и наоборот, - что присутствует в хвостах из того, чего в них не должно быть. На рис. 3 показана исходная технологическая схема обогатительной фабрики Cosmos.

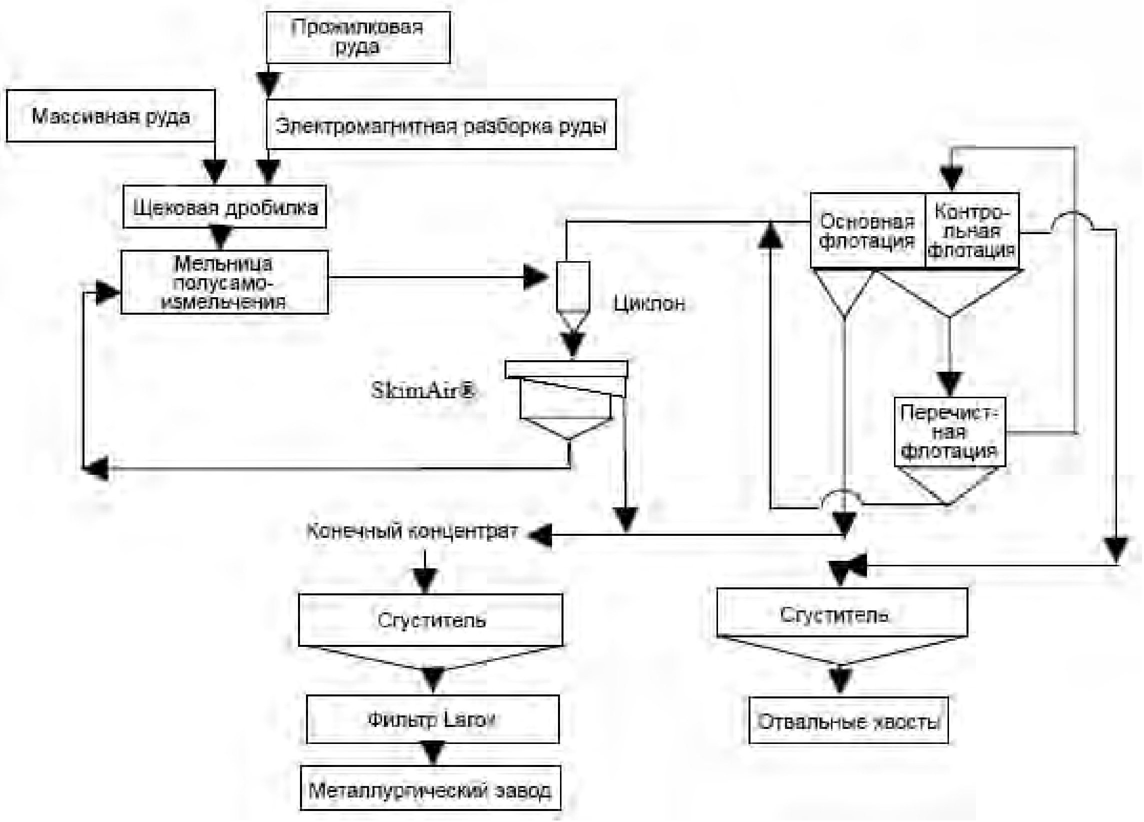


Рисунок 3. Исходная технологическая схема

Цикл измельчения состоял из одной щековой дробилки, продукт которой поступал непосредственно в одностадийную мельницу полусамоизмельчения закрытого цикла с одной парой циклонов Cavex 400 мм. Схема классификации не отличалась гибкостью и эффективностью, в результате чего значительное количество высвобожденного пентландита попадало в нижний продукт циклона. Поэтому флотомашина SkimAir® демонстрировала высокие показатели эффективности, однако это указывало на неудовлетворительную эффективность цикла измельчения, поскольку эти высвобожденные частицы должны были поступать в верхний продукт циклона. SkimAir® извлекала большую часть никеля (61%) при содержании в концентрате 20%. Остальная часть цикла флотации производила концентрат с содержанием 16%, обеспечивая извлечение еще 30% никеля. Общее извлечение никеля в цикле достигало 92% при содержании в концентрате 19% (таблица 1).

Цикл флотации состоял из двух стадий основной флотации и трех стадий контрольной флотации в основном технологическом потоке питания флотации (таблица 2 и 3). Концентрат первой основной флотации поступал в конечный концентрат вместе с продуктом SkimAir®. Примечательно, что концентрат и хвосты перечистной флотации находились в закрытом цикле с основной/контрольной флотацией. В результате во всех потоках флотации присутствовала смесь различных флотореагентов, которые были вынуждены одновременно выполнять функции активации, сбора и депрессии. Это делало цикл проще в управлении, однако большие рециркулирующие нагрузки, сложности, возникающие при дозировании реагентов, и непрерывное разубоживание концентрата перечистной флотации питанием основной флотации указывали на возможность несложного и значимого усовершенствования цикла.

Верхний продукт циклона содержал высвобожденный на 86% пентландит и высвобожденные на 97% несульфидные жильные минералы, что также указывало на переизмельчение. Столь высокие показатели высвобождения более подходили питанию перечистной флотации, чем циклу основной флотации, и позволяли сделать вывод, что можно увеличить производительность мельницы полусамоизмельчения и получать более грубый верхний продукт циклона без ущерба для извлечения в цикле основной флотации.

Таблица 1. Общий металлургический баланс, июль 2008 года

Продукт	Масса, %	Содержание, %					Распределение, %				
		Ni	Cu	Fe	S	As	Ni	Cu	Fe	S	As
Питание, пересчет	100	4,61	0,17	12,2	8,69	0,08	100	100	100	100	100
Концентрат SkimAir®	13,8	20,6	0,68	34,5	30,6	0,24	61	54	39	49	43
Концентрат 1-й основной флотации	8,5	16,4	0,77	27,7	26,1	0,13	30	37	19	25	15
Конечный концентрат	22,2	19,0	0,70	31,9	28,9	0,21	92	91	58	74	58
Отвальные хвосты	77,8	0,5	0,02	6,5	2,9	0,04	8	9	42	26	42

Таблица 2. Питание флотации, июль 2008 года

Минерал	Символ	Содержание, %
Пентландит	Pe	13,5
Халькопирит	Cp	0,5
Пирротит	Po	9,7
Пирит	Py	1,0
Несульфидные жильные минералы	Gn	75,3

Таблица 3. Состав минералов в питании флотации, июль 2008 года

Состояние минерала	Питание флотации, 104 мкм, K_{80}		
	Pe	Po	Gn
Высвобожденный	85,9	77,4	97,0
Бинарные частицы с Pe	—	6,0	1,1
Бинарные частицы с Po	3,2	—	0,6
Бинарные частицы с Gn	6,9	9,4	—
Бинарные частицы с проч.	0,7	0,6	0,3
Мультифазные частицы	3,3	6,6	1,0

На рис. 4 показано извлечение в зависимости от крупности в цикле SkimAir®. Извлечение из крупной фракции имеет отличные показатели при незначительном вовлечении жильной породы (особенно учитывая относительно высокую плотность пульпы, необходимую в этом цикле для управления водным балансом в мельнице полусамоизмельчения). Извлечение из мелких фракций (<50 мкм) хуже, однако извлечение происходит позже в цикле основной флотации, как показано на рис. 5. Примерно 30% высвобожденного пирротита в верхнем продукте циклона извлекается в концентрат первой основной флотации (максимальное извлечение наблюдается при диаметре частиц от 10 до 50 мкм). Вовлечение несульфидных жильных минералов в концентрат первой основной флотации выше, чем в концентрат SkimAir®.

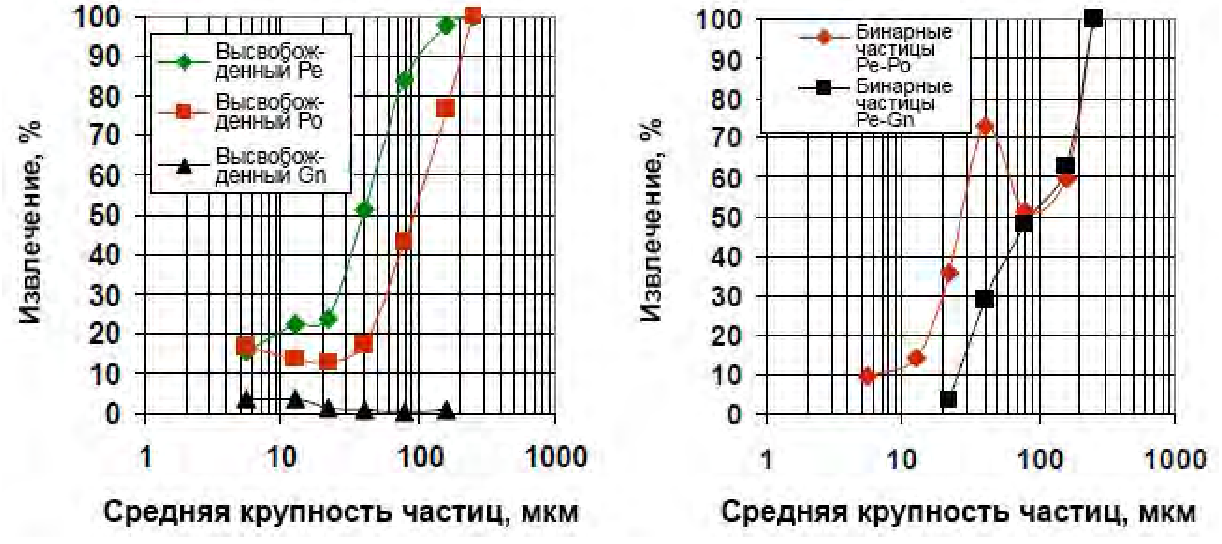


Рисунок 4. Извлечение в концентрат SkimAir® в зависимости от крупности и класса

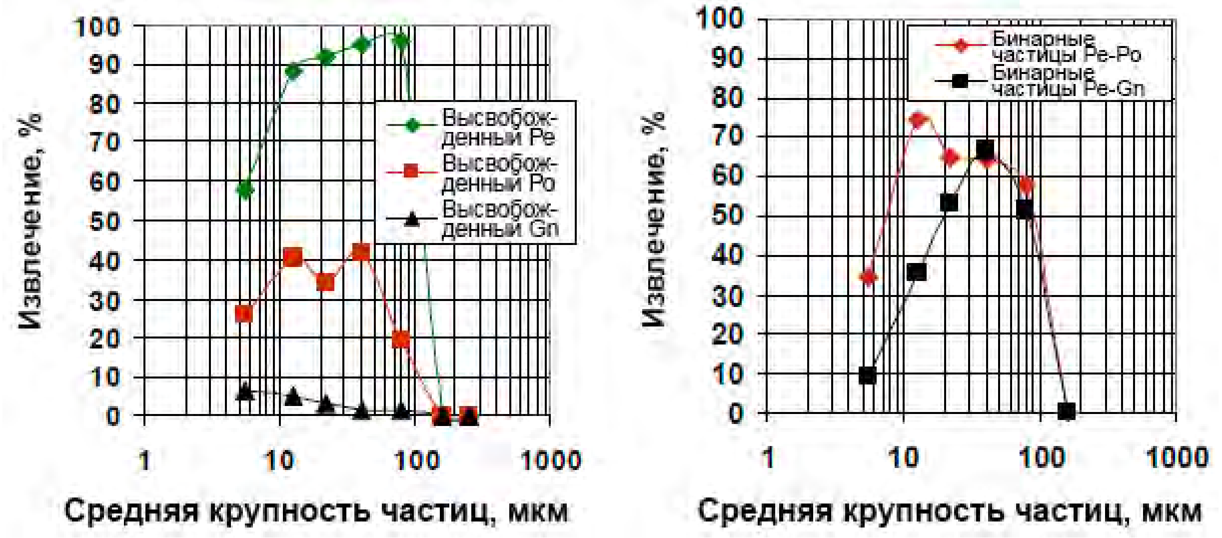


Рисунок 5. Извлечение в концентрат первой основной флотации в зависимости от крупности и класса

Как показано на рис. 6, основными разубоживающими минералами в концентрате SkimAir® являются пирротит и несульфидные жильные минералы (21% и 12% от массы концентрата, соответственно). Концентрат первой основной флотации содержал примерно в два раза больше несульфидных жильных минералов и примерно столько же пирротита, сколько содержал концентрат SkimAir® (рис. 7). Основные разубоживающие минералы преимущественно присутствуют в виде высвобожденных зерен. Теоретическая кривая зависимости содержания от извлечения на рис. 8 показывает, что при условии удаления большей части высвобожденных разубоживающих минералов можно получить концентрат более высокого сорта.

В то же время, увеличение крупности измельчения (приводящее к меньшему высвобождению) может не повлиять на содержание в конечном концентрате. Если применить традиционный цикл перечистки и/или промывку пены для удаления высвобожденных вовлеченных несульфидных жильных минералов, существует возможность получения концентрата с более высоким содержанием.

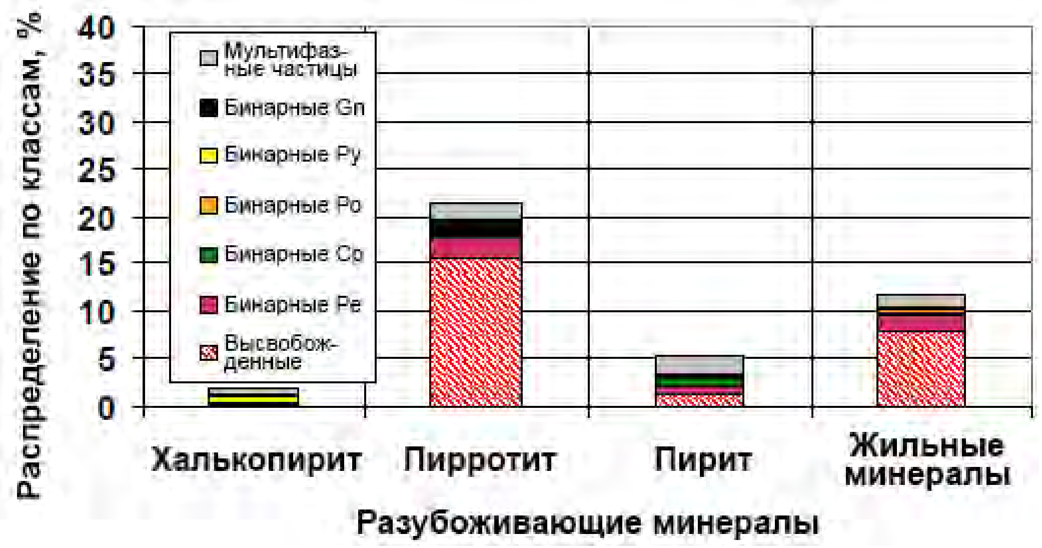


Рисунок 6. Распределение разубоживающих минералов в концентрате SkimAir®

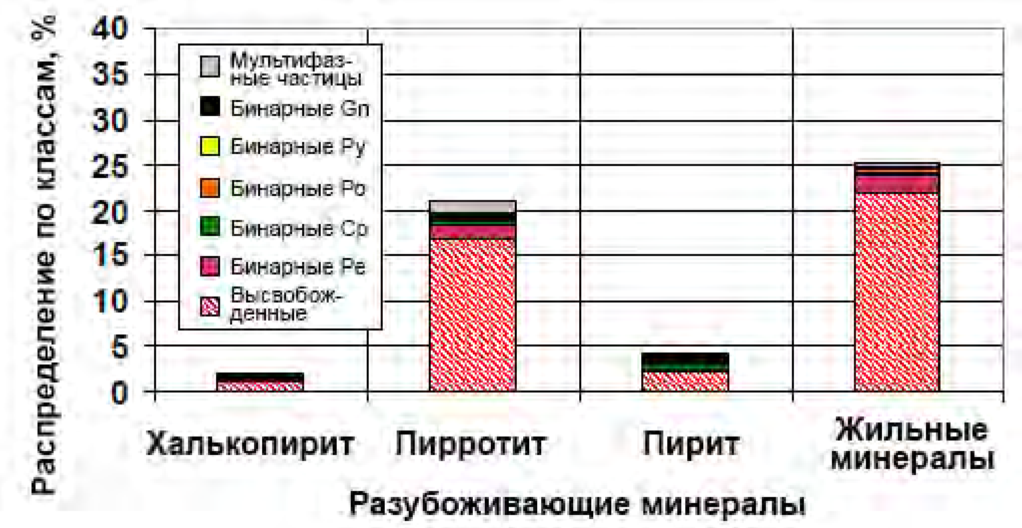


Рисунок 7. Распределение разубоживающих минералов по массе в концентрате основной флотации

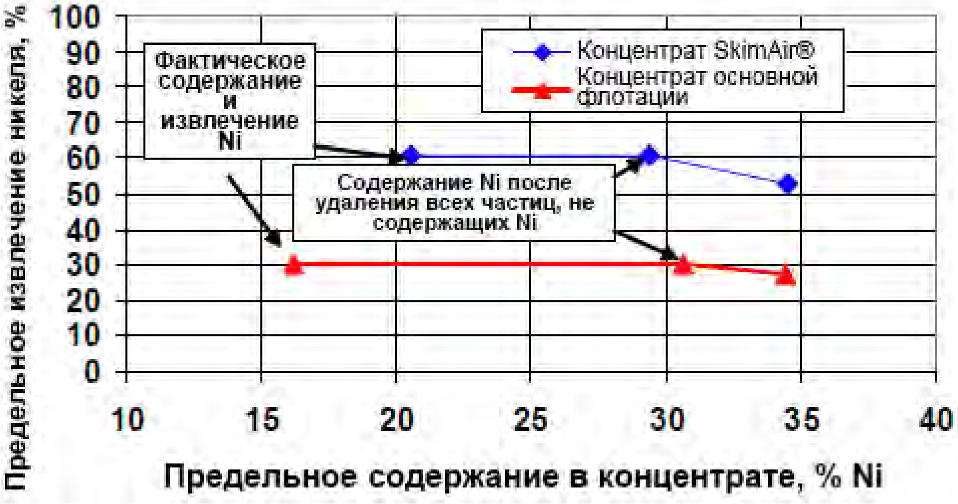


Рисунок 8. Предельные минералогические характеристики содержания и извлечения

8% никеля в руде теряется с отвальными хвостами. 60% потерь приходится на унос высвобожденных зерен пентландита (рис. 9), значительная доля которых имеет крупность менее 11 мкм. Это закономерно, учитывая отсутствие полноценного цикла перечистки и сложный химический состав пульпы.

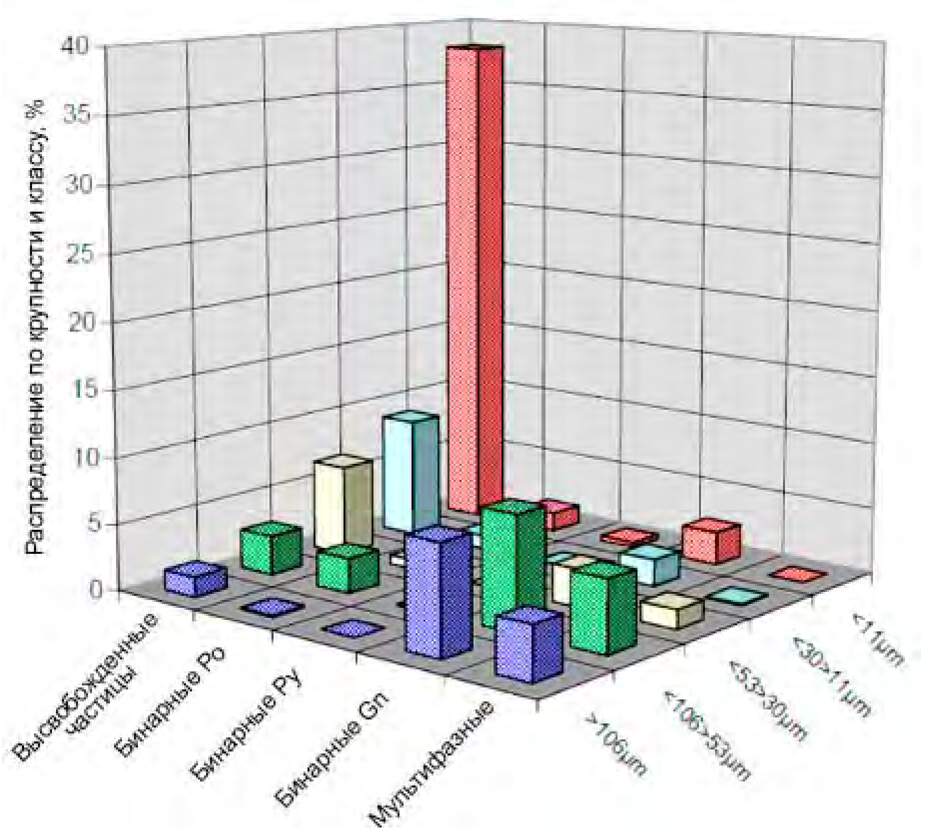


Рисунок 9. Распределение уноса никеля с хвостами по крупности и классу, % от содержания пентландита в хвостах

ТЕХНОЛОГИЧЕСКАЯ СХЕМА ПОСЛЕ УСТРАНЕНИЯ УЗКИХ МЕСТ И ЕЕ ЭФФЕКТИВНОСТЬ (РЕЗУЛЬТАТЫ ИССЛЕДОВАНИЯ, ВЫПОЛНЕННОГО В АПРЕЛЕ 2009 ГОДА)

Производительность мельницы полусамоизмельчения была увеличена с учетом большего тоннажа добычи с получением более крупного питания флотации. Увеличение производительности мельницы полусамоизмельчения показано на рис. 10. Среди существенных доработок цикла полусамоизмельчения можно назвать переход со смеси шаров 80/100/125 мм на шары 125 мм, увеличение угла лифтера корпуса мельницы полусамоизмельчения с 15 до 25 градусов с учетом повышенной скорости мельницы, увеличение размера отверстий и живого сечения сит на стороне разгрузки. Из-за невозможности добиться сбалансированного режима работы двух циклонов Cavex 400 мм при больших колебаниях расхода они были заменены четырьмя циклонами Cavex 250 мм. Результаты минералогического анализа показали, что уровень высвобождения пентландита и несульфидных жильных минералов в питании флотации может быть снижен при большей крупности измельчения без ущерба для показателей извлечения основной и контрольной флотации. Это позволило бы извлекать большую долю композитных частиц. Чтобы добиться постоянства содержания в конечном концентрате, композитные частицы необходимо доизмельчать до перечистной флотации. Мельница IsaMill™ была установлена в цикле для доизмельчения.

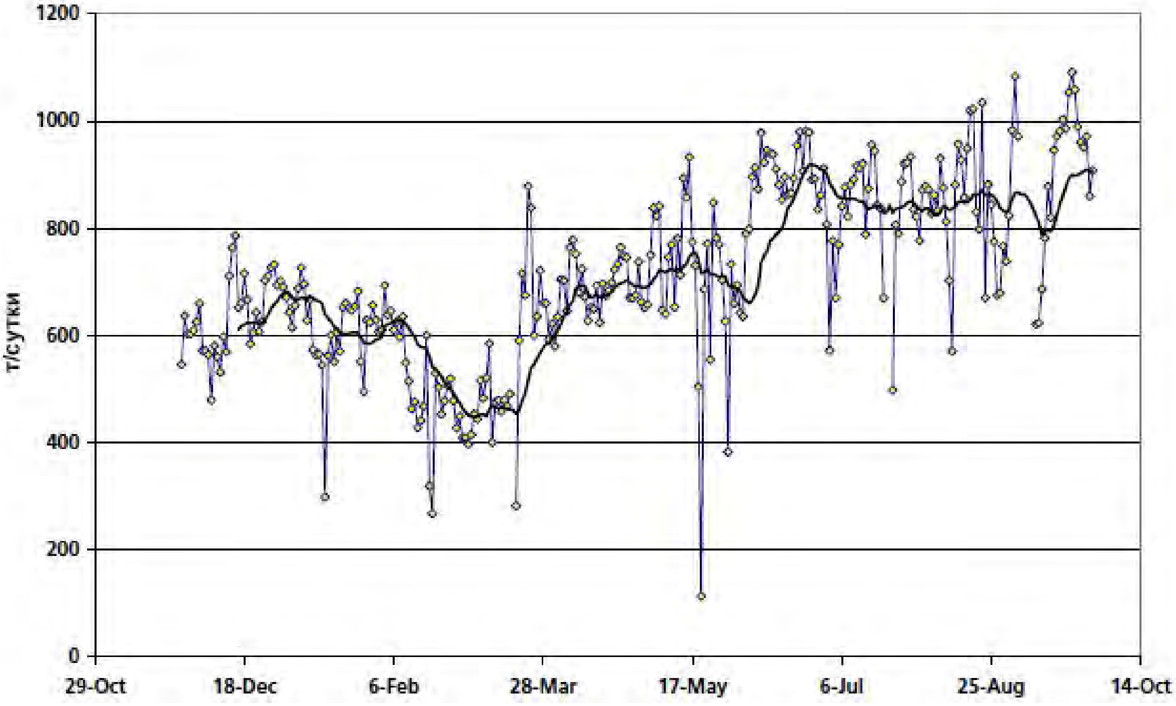


Рисунок 10. Производительность мельницы полусамоизмельчения

Для доизмельчения была выбрана мельница IsaMill™ M500 (показана на рис. 11), поскольку благодаря ее компактности она идеально подходит для проектов расширения действующих производств. С точки зрения минералогических и металлургических характеристик фабрики Cosmos, большую ценность представляла возможность повысить благодаря применению IsaMill™ содержание в концентрате, использовать инертную среду измельчения (чистые поверхности частиц), получить узкий гранулометрический диапазон частиц (в открытом цикле), эффективно измельчать материал при низкой плотности пульпы. Чистые свежие поверхности минеральных частиц, которые могут быть получены при использовании инертных сред, позволяют оптимизировать разделение минералов, снизить расход реагентов и ускорить кинетику флотации (Côté and Adante, 2009; Finch, Rao, and Nesset, 2007; Huang, Grano, and Skinner, 2006). Пример плохо высвобожденного пентландитового композита, требующего доизмельчения, показан на рис. 12.

Рисунок 11. Мельница IsaMill™ с двигателем 200 кВт на обогатительной фабрике Cosmos

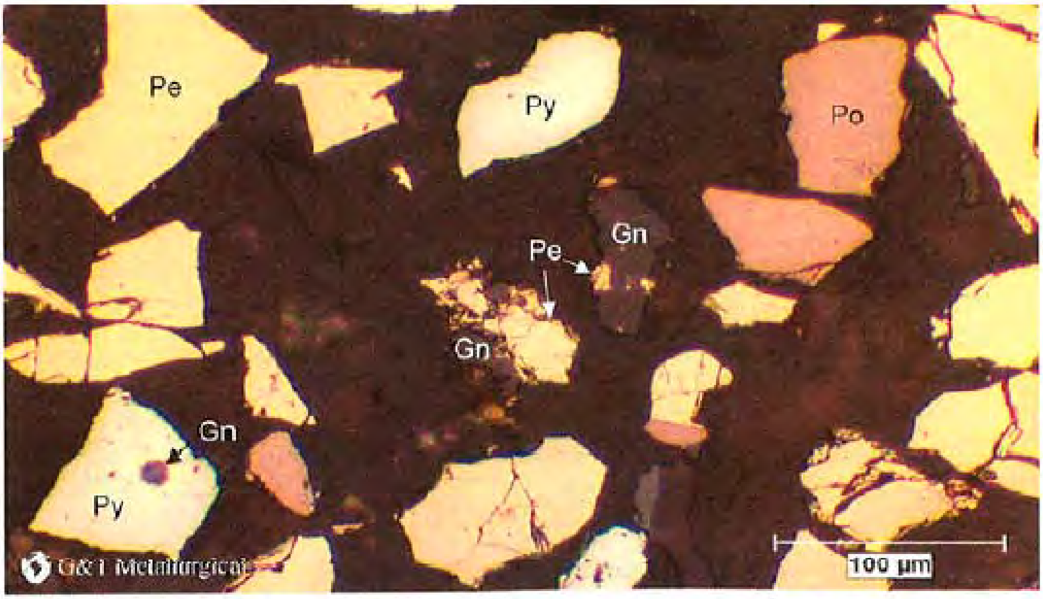


Рисунок 12. Микрофотография плохо высвобожденных частиц пентландита

Лабораторные пробы выбранных концентратов были доизмельчены и подвергнуты флотации, чтобы проверить предположение. Было установлено, что при более мелком питании флотации кривая зависимости содержания никеля от извлечения смещалась вверх (рис. 13). IsaMill™ обеспечивает намного более тщательное удаление MgO и, при создании благоприятных химических условий в цикле перечистной флотации, -оптимальное удаление As.

Характерная гранулометрическая характеристика продукта IsaMill™ M500 показана на рис. 14. F_{80} и P_{80} составляют 40 и 20 мкм, соответственно. Удельный расход энергии на такое измельчение составляет примерно 20 кВт-ч/т. Примечательно, что гранулометрическая характеристика продукта круче гранулометрической характеристики питания. Мельница IsaMill™ направляет энергию измельчения на получение крупных частиц, а не на увеличение доли мелких. На это указывает минимальное изменение крупностей классов P_{10}, P_{20} и P_{30}. Для доизмельчения применяется керамическая среда крупностью 2 мм Keramax® MT1™ производителя Magotteaux.

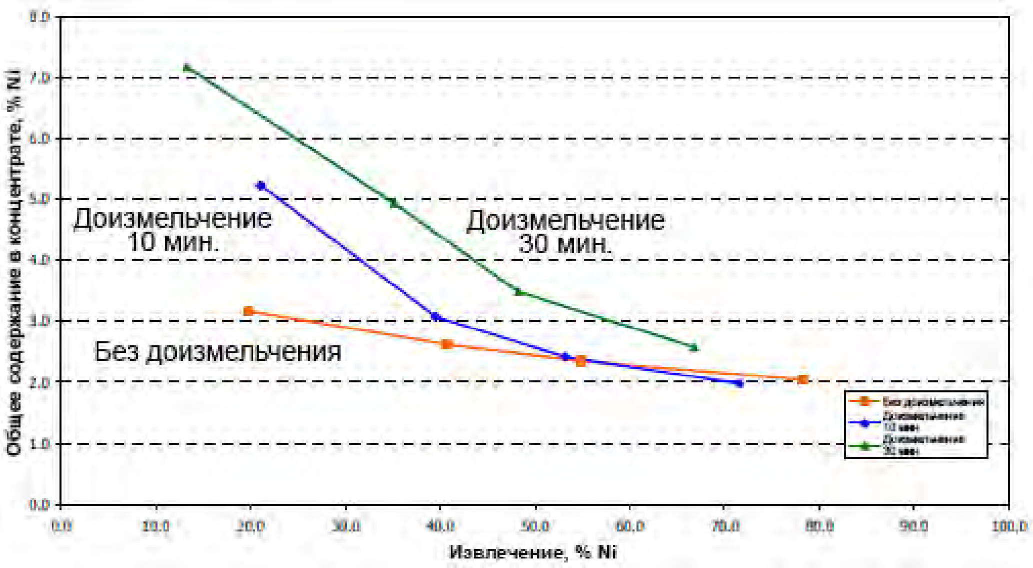


Рисунок 13. Пример положительного влияния доизмельчения на характеристику зависимости содержания от извлечения

Рисунок 14. Характерный гранулометрический состав питания и продукта мельницы IsaMill™ M500

В связи с более высоким исходным содержанием и мощностью производства потребовалась установка дополнительных мощностей флотации. В первую очередь была установлена флотомашина Джеймсон Z1600 для переработки верхнего продукта циклона мельницы полусамоизмельчения в качестве флотомашины скальпирования основной флотации, а цикл флотации был сделан полностью открытым (концентраты перечистной флотации поступают в конечные концентраты, а хвосты перечистной флотации – в отвальные хвосты). После огрубления верхнего продукта циклона мельницы полусамоизмельчения уменьшились содержание высвобожденного пентландита в нижнем продукте циклона и извлечение в SkimAir® (рис. 15). После этого флотомашина SkimAir® стала использоваться в качестве флотомашины основной флотации, получающей питание из верхнего продукта циклона.

Рисунок 15. Извлечение в SkimAir® в зависимости от крупности верхнего продукта циклона мельницы полусамоизмельчения

Во вторую очередь была установлена флотомашина Wemco® SmartCell™ емкостью 40 м^3 для увеличения мощности основной флотации. Сперва она была установлена после флотомашины Джеймсон с целью получения достаточной мощности основной флотации, чтобы остальные первоначально установленные флотомашины могли быть включены в соответствующие циклы перечистной флотации. После пуска флотомашины Wemco® SmartCell™ было установлено, что вспениватель, попадающий из предшествующих технологических стадий, вызывает чрезмерное пенообразование, а также проблемы при перекачивании концентрата. Существующая трубная обвязка позволила подключить флотомашину Wemco® SmartCell™ перед флотомашиной Джеймсон. Такая схема подключения используется по настоящее время.

Флотомашина Джеймсон была выбрана для фабрики Cosmos благодаря ее компактности и высокой производительности. Флотомашина Джеймсон, установленная на фабрике Cosmos, показана на рис. 16. Подобно мельнице IsaMill™, компактная конструкция облегчает установку при переоборудовании действующих предприятий.

Рисунок 16. Флотомашина Джеймсон Z1600 на обогатительной фабрике Cosmos

Флотомашина Джеймсон была запущена в конце марта 2009 года, что совпало с увеличением содержания никеля в концентрате на 2 единицы при прежнем извлечении – 90% (таблица 4). Как следует из рис. 17, содержание в концентрате флотомашины Джеймсон стабильно выше содержания в конечном концентрате. Кроме того, для уменьшения вовлечения несульфидных жильных минералов применяется промывка пены, как показано на рис. 18. Следует отметить, что флотомашина Джеймсон была пущена на верхнем продукте циклона, за несколько недель до реконфигурации флотомашины SkimAir®.

Таблица 4. Металлургические балансы за различные месяцы

Месяц	Конечный продукт	Масса, %	Содержание, %					Распределение, %				
			Ni	Cu	Fe	S	As	Ni	Cu	Fe	S	As
Июль 08	Концентрат	22,2	19,0	0.71	31,9	28,9	0,2	92	91	58	74	58
	Хвосты	77,8	0,5	0,02	6,5	2,9	0,04	8	9	42	26	42
Август 08	Концентрат	18,7	18,7	0,74	29,9	29,6	0,28	91	92	56	76	77
	Хвосты	81,3	0,4	0,01	5,4	2,2	0,02	9	8	44	24	23
Декабрь 08	Концентрат	19,9	19,6	0,65	25,7	24,5	0,25	90	87	50	67	46
	Хвосты	80,1	0,5	0,02	6,4	3,0	0,07	10	13	50	33	54
Февраль 09	Концентрат	22,8	19,9	0,68	24,8	23,5	0,19	90	81	44	60	45
	Хвосты	77,2	0,7	0,05	9,3	4,6	0,07	10	19	56	40	55
Апрель 09	Концентрат	22,5	21,8	0,75	30,6	27,6	0,11	90	88	36	50	44
	Хвосты	77,5	0,7	0,03	15,5	8,0	0,04	10	12	34	50	56

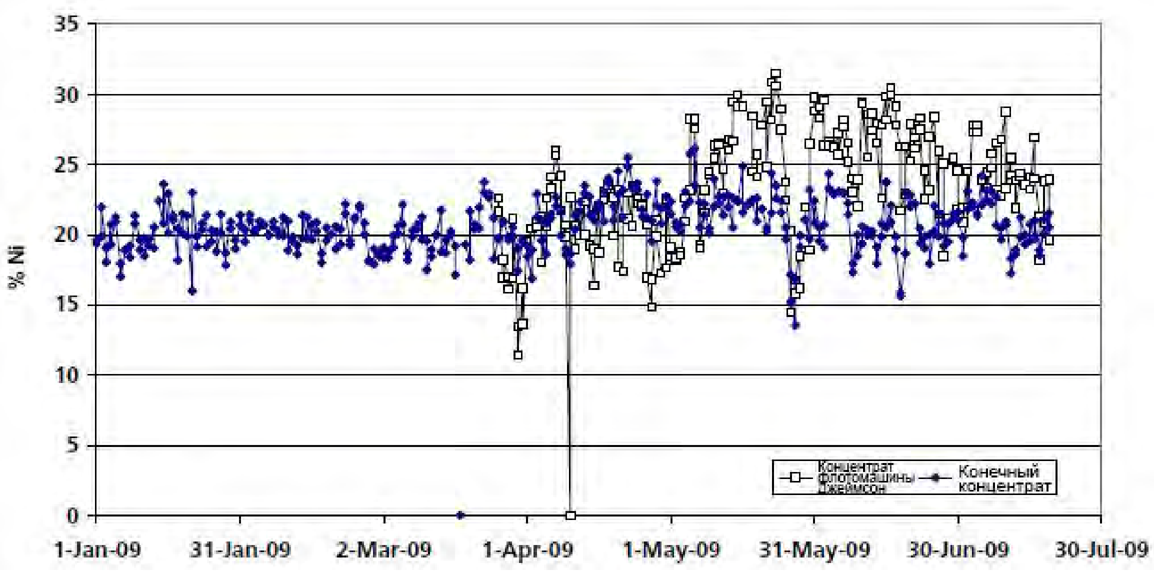


Рисунок 17. Содержание в концентрате флотомашины Джеймсон и в конечном концентрате

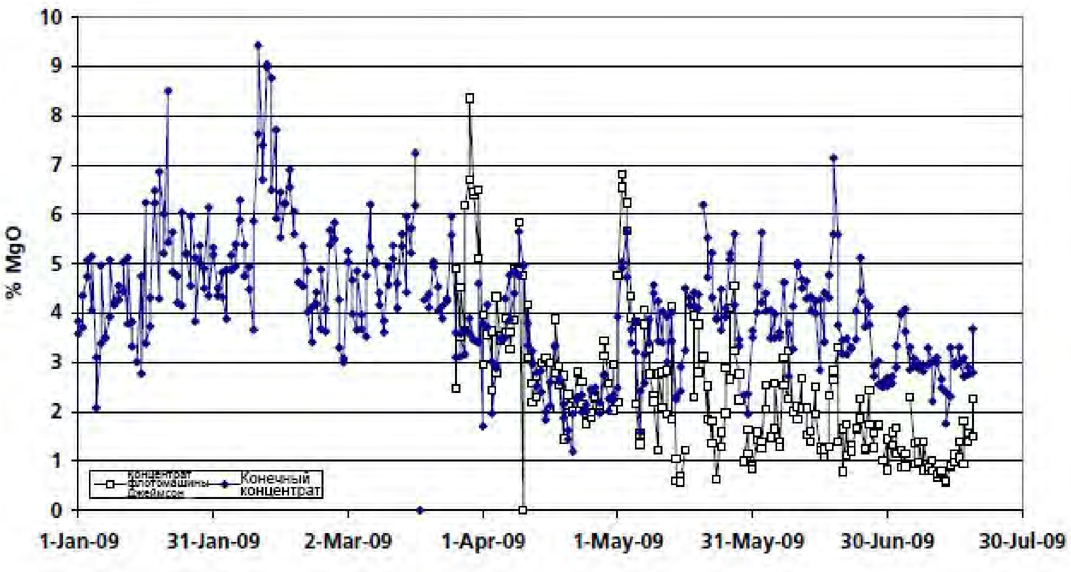


Рисунок 18. Содержание MgO в концентрате флотомашины Джеймсон и в конечном концентрате

В последнюю очередь флотомашины, первоначально работавшие в цикле основной/контрольной флотации, были подключены к циклу перечистной флотации. Организация открытого цикла флотации позволила более четко определить последовательные стадии флотации. Схема трех основных компонентов цикла показана на рис. 19. В данной схеме флотомашины основной и контрольной флотации обеспечивают максимальное извлечение никеля. На стадии перечистной флотации происходит депрессия и удаление несульфидных жильных минералов при помощи доизмельчения, перечистной флотации с промывкой/разубоживанием и применения депрессантов. При необходимости может применяться вторая перечистная флотация для избирательного удаления мышьяка посредством контроля pH и добавления цианида.

Данная схема имеет три преимущества. Во-первых, благодаря последовательному введению, реагенты добавляются точно на нужной стадии, а не до нее через рециркулирующий поток. Так, прошлые исследования показали, что добавление цианида и сульфата меди замедляет кинетику пентландита. Во-вторых, благодаря разделению функций циклом проще управлять. В-третьих, большее время пребывания позволяет увеличить производительность по руде и никелю.

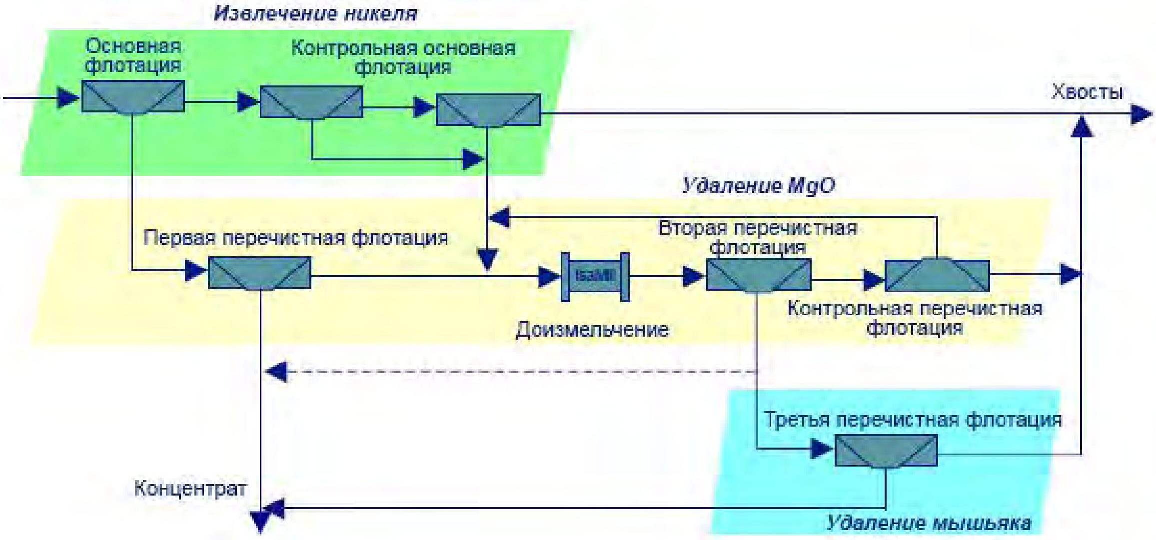


Рисунок 19. Последовательные стадии флотации

Анализ извлечения в конечный концентрат в зависимости от крупности после устранения узких мест указывает на хорошие показатели извлечения высвобожденного пентландита в цикле флотации (рис. 20), хотя по-прежнему наблюдается существенный унос материала мельче 10 мкм. На момент исследования, проведенного в апреле 2009 года, доработка цикла перечистной флотации не была завершена, поэтому время увеличенное пребывания сопровождалось высокой плотностью пульпы. Перечистка с разубоживанием в доработанном цикле позволит извлекать больше ценной мелкой фракции и уменьшить вовлечение высвобожденных несульфидных жильных минералов. Показатели извлечения невысвобожденного пентландита также улучшились (рис. 21). Новый цикл флотации также отличается более эффективным удалением высвобожденного пирротита. Как показано на рис. 22, высвобожденный пирротит составил менее 9% конечного концентрата – по сравнению с 15% в июле 2008 года в цикле до доработки.

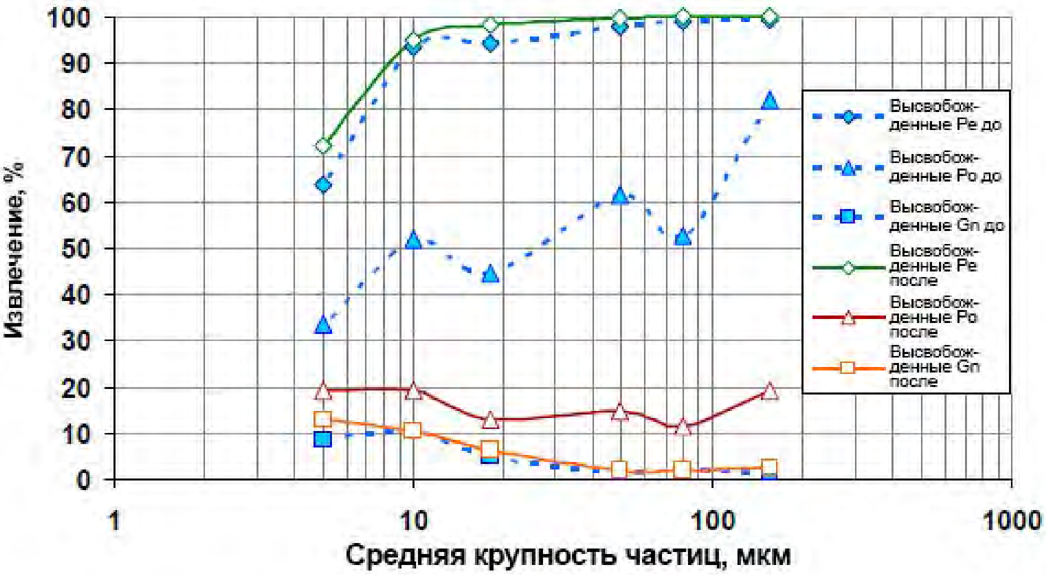


Рисунок 20. Извлечение в конечный концентрат в зависимости от крупности и класса

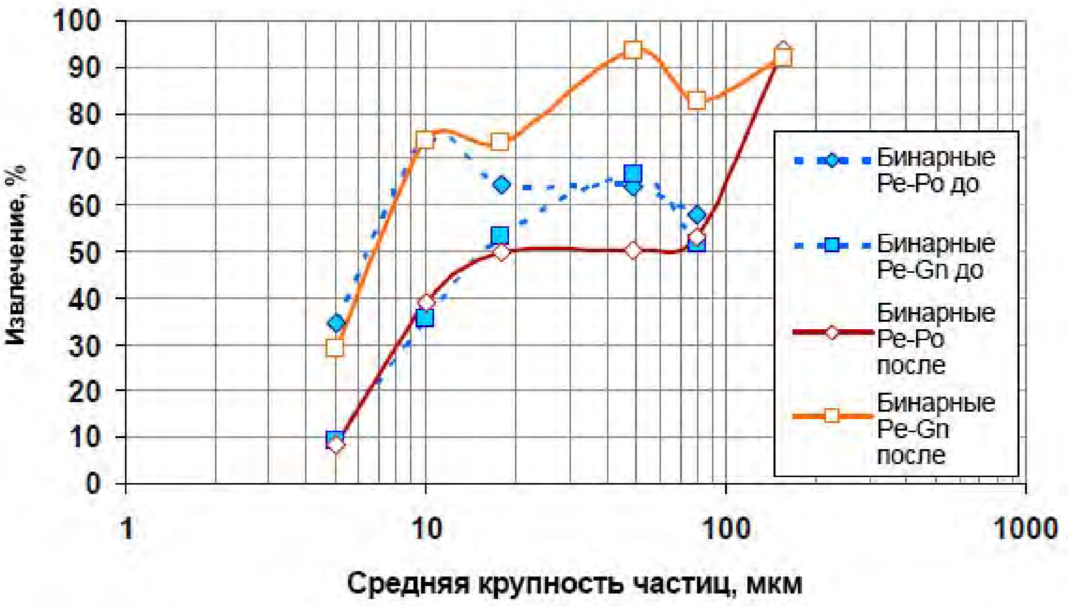


Рисунок 21. Извлечение бинарных частиц пентландита в конечный концентрат

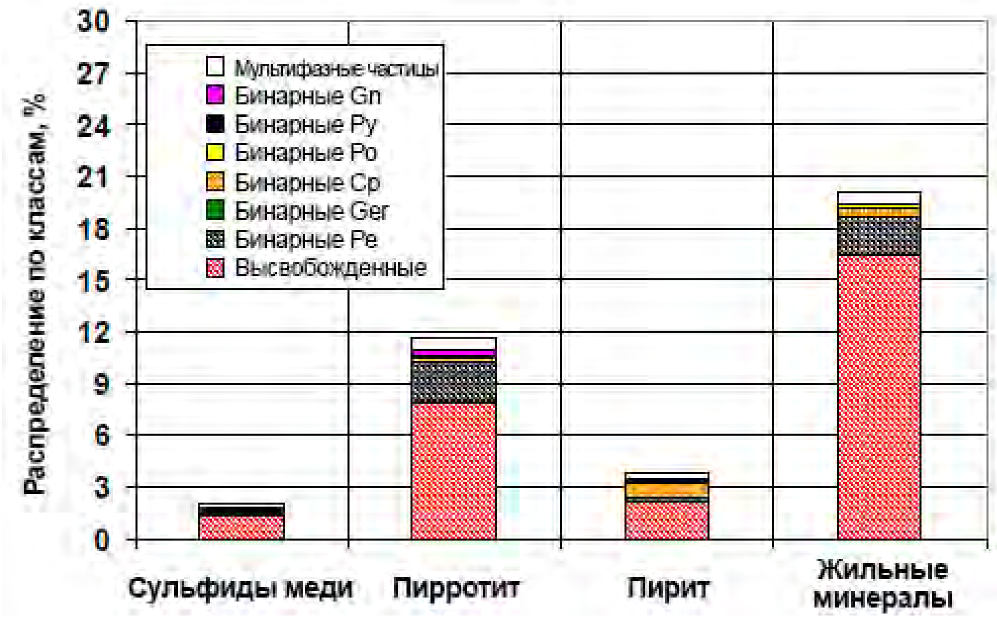


Рисунок 22. Состав разубоживающих минералов в конечном концентрате, апрель 2009 года

Исходя из теоретической характеристики зависимости между содержанием и извлечением, представленной на рис. 23, можно заключить, что, несмотря на некоторое улучшение относительно исходных показателей остаются возможности повышения содержания в концентрате без ущерба для извлечения никеля.

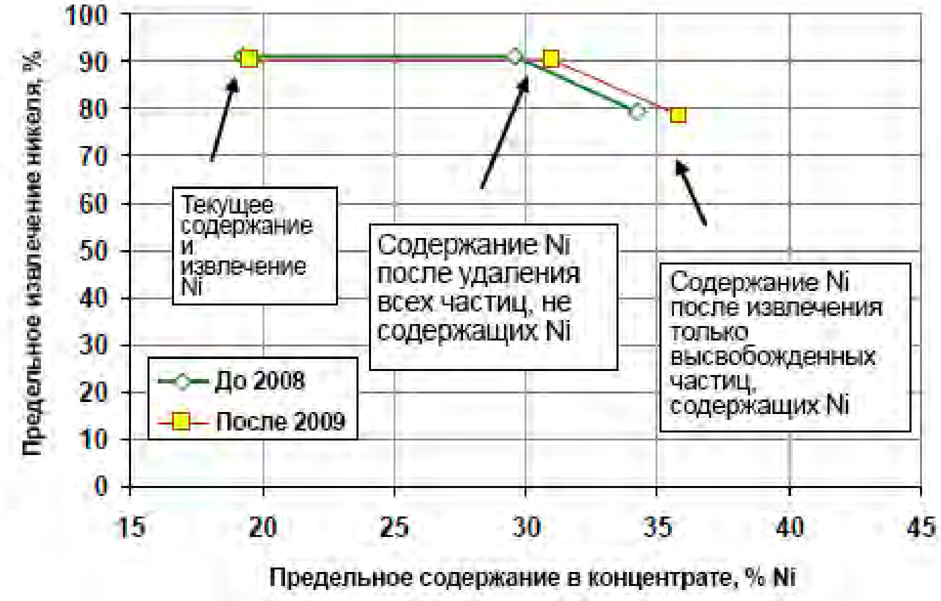


Рисунок 23. Предельные минералогические характеристики содержания и извлечения, до и после

Как показано на рис. 24 и в таблице 5, унос никеля в отвальные хвосты в апреле остается сопоставимым с результатами предыдущего исследования (рис. 9).

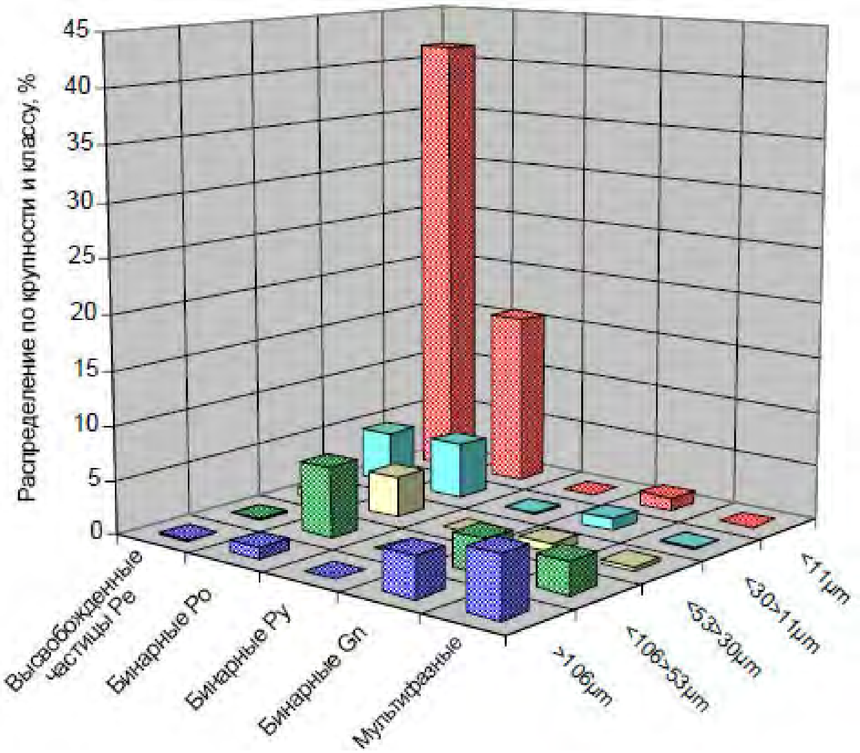


Рисунок 24. Распределение уноса никеля с хвостами по крупности и классу, апрель 2009 года

Таблица 5. Общее распределение уноса никеля в отвальные хвосты

Класс минералов	**Распределение по крупности и классу**					**Унос по классам**
	>106 мкм	**<106>53 мкм**	**<53>30 мкм**	**<30>11 мкм**	**<11 мкм**	
	Среднее за 2008 год					
Высвобожденные	0,1	0,2	0,4	1,0	3,1	4,8
Бинарные Po	0,0	0,2	0,2	0,1	0,2	0,7
Бинарные Py	0,0	0,0	0,0	0,0	0,0	0,0
Бинарные Gn	0,7	0,8	0,2	0,3	0,3	2,3
Мультифазные	0,4	0,4	0,1	0,1	0,0	1,0
Унос по крупности	1,2	1,6	0,9	1,5	3,6	8,8
	Апрель 2009 года					
Высвобожденные	0,0	0,0	0,1	0,5	4,0	4,5
Бинарные Po	0,1	0,6	0,4	0,5	1,5	3,1
Бинарные Py	0,0	0,0	0,0	0,0	0,0	0,0
Бинарные Gn	0,4	0,3	0,1	0,1	0,1	1,0
Мультифазные	0,5	0,3	0,0	0,0	0,0	0,9
Унос по крупности	1,0	1,2	0,5	1,1	5,6	9,4

ОКОНЧАТЕЛЬНАЯ ТЕХНОЛОГИЧЕСКАЯ СХЕМА

Предполагаемая окончательная технологическая схема показана на рис. 25. Доработка позволит завершить программу увеличения производительности по никелю и снижения производственных затрат.

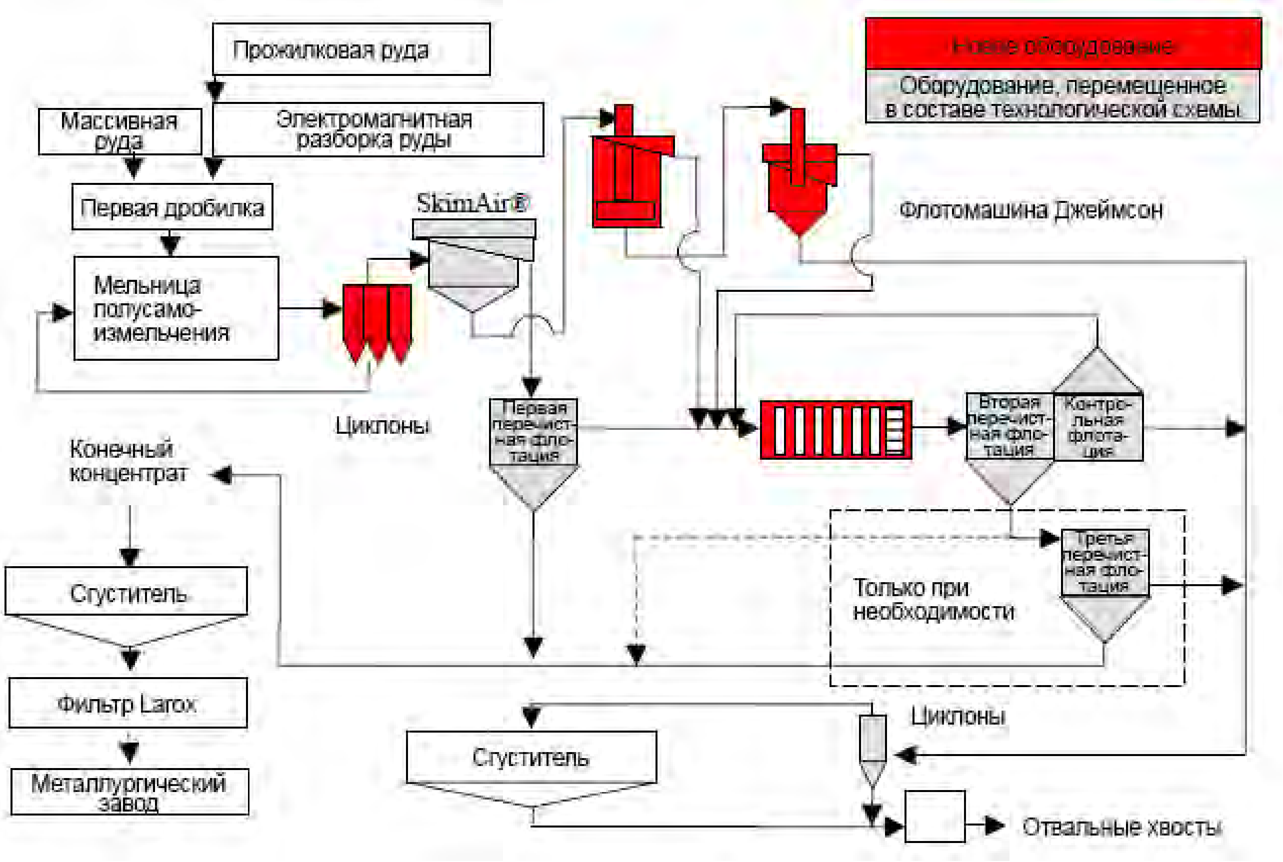


Рисунок 25. Новая технологическая схема

ЗАКЛЮЧЕНИЕ

Производственные показатели обогатительной фабрики Cosmos, в целом, соответствуют плановым показателям, принятым в программе устранения узких технологических мест. Общие показатели содержания в концентрате и извлечения по фабрике сохранены несмотря на то, что на переработку поступает более труднообогатимая руда (исходное содержание никеля в питании уменьшилось, отношение никеля к мышьяку в питании повысилось, отношение массивной руды к вкрапленной понизилось – все эти факторы негативно влияют на извлечение никеля), а производительность фабрики увеличена.

С минералогической точки зрения, рисунки 27 – 29 указывают на увеличившееся извлечение пентландита во всех фракциях. Извлечение несульфидных жильных минералов осталось. в целом, прежним, однако этот показатель станет основным объектом внимания при планируемой доработке стадии перечистной флотации в новой технологической схеме.

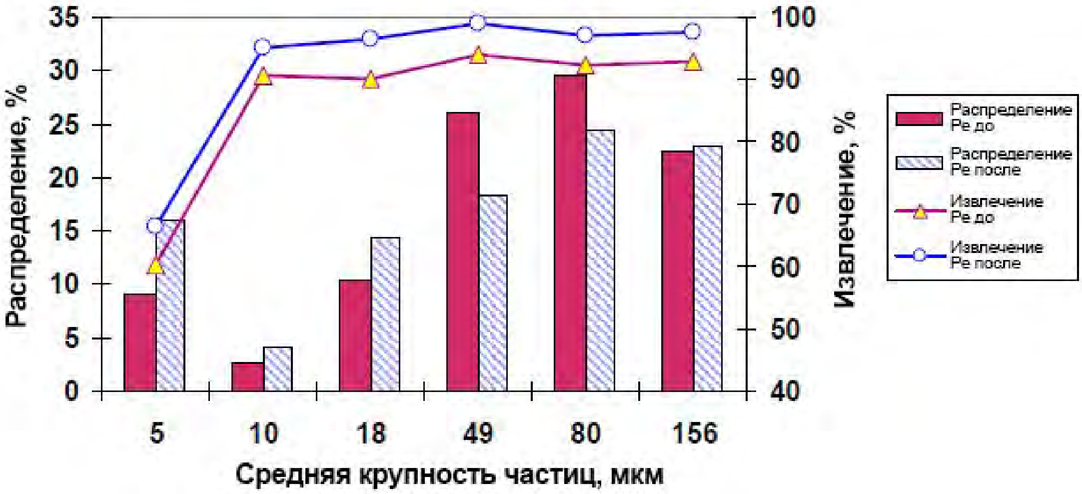


Рисунок 26. Сравнение показателей извлечения пентландита

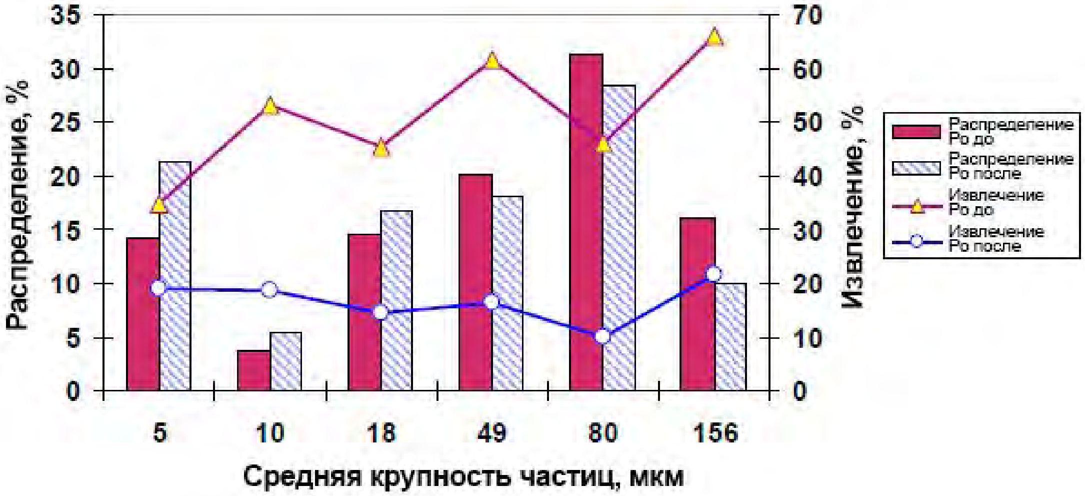


Рисунок 27. Сравнение показателей извлечения пирротита

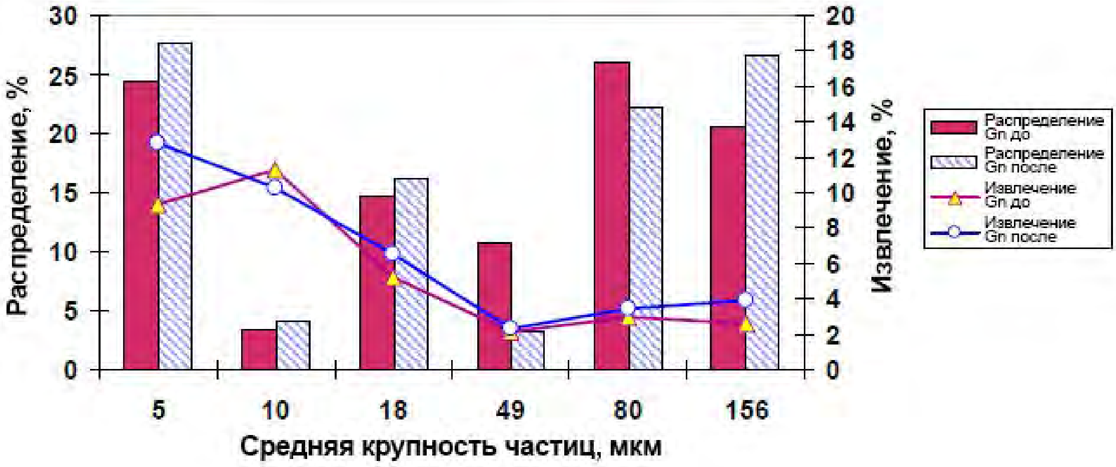


Рисунок 28. Сравнение показателей извлечения несульфидных жильных минералов

Большая часть принятых к настоящему времени мер по устранению технологических мест была направлена на доработку существующего оборудования и установку основного дополнительного оборудования, такого как мельница IsaMill™ M500, флотомашина Джеймсон Z1600 и флотомашина Wemco® SmartCell™, преимущественно с целью увеличения производительности обогатительной фабрики. Как показано на рисунках 27 – 29, эта задача была выполнена без ущерба для производственных показателей фабрики.

Предстоящая доработка цикла перечистной флотации позволит в полной мере реализовать принцип “нужного оборудования в нужном месте” для повышения металлургической эффективности. Аналитические средства, такие как модальный анализ и расчеты извлечения по фракциям крупности, были и остаются незаменимым инструментом определения наилучшего применения имеющихся технологических средств для получения оптимальных показателей разделения минералов. Новый цикл позволит обогатительной фабрике Cosmos сохранить в будущем способность производить никелевый концентрат с минимальной себестоимостью.

БЛАГОДАРНОСТИ

Авторы выражают признательность компании Xstrata Nickel Australasia за разрешение на публикацию настоящей работы.

СПИСОК ИСПОЛЬЗОВАННОЙ ЛИТЕРАТУРЫ

Shouldice, T. and Lafreniere, T. (2008) G&T Metallurgical Report of the July 2008 Survey of the Cosmos Concentrator (Internal)

Côté, I. and Adante, J. (2009) Design, Start-Up and Optimization of Pb & Zn Flotation Circuits Utilizing Ultra-Fine Grinding Technology, 41st Annual Meeting of the Canadian Mineral Processors, Ottawa, Canada, January 20 to 22.

Huang, G., Grano, S. and Skinner, W. (2006) Galvanic Interaction between Grinding Media and Arsenopyrite and its Effect on Flotation: Part II. Effect of Grinding on Flotation, Int. J. Miner. Process. 78, pp. 198–213.

Finch, J. Rao, S., and Nesset, J. (2007) Iron Control in Mineral Processing, 39th Annual Meeting of the Canadian Mineral Processors, Ottawa, Canada, January 23 to 25

Shouldice, T. and Ma, W. (2009) G&T Metallurgical Report of the April 2009 Survey of the Cosmos Concentrator (Internal)

NECESSITY DRIVING CHANGE AND IMPROVEMENT TO THE CLEANER CIRCUIT AT LUMWANA COPPER CONCENTRATOR

*R. Araya[1], G. Cordingley[2], A. Mwanza[2] and L. Huynh[3]

[1]Xstrata Technology Canada
Level 10- 700 West Pender Street
Vancouver, Canada V6C 1G8
*(*Corresponding author:RAraya@xstratatech.com)*

[2]Barrick Copper Business Unit
P.O. Box 110199, Mwinilunga Road
Lumwana East, Solwezi, Zambia

[3]Xstrata Technology Australia
Level 10 – 160 Ann Street
Brisbane, Australia QLD 4000

ABSTRACT

Lumwana is an open pit Copper mine located in the Northwestern Province of Zambia approximately 80 km west from the provincial capital of Solwezi, approximately 300 km northwest from the Copperbelt and 700 km northwest from the national capital of Lusaka. The mine has two major Copper deposits: Malundwe and Chimiwungo. Copper mineralization includes Chalcopyrite, Bornite and Chalcocite. In addition to Copper, these deposits also contain the undesirable element Uranium contained in vein hosted and disseminated Uranite. The processing plant was designed using conventional technologies: semi-autogenous grinding (SAG) and ball mill for grinding, and mechanical cells for rougher and cleaner flotation. Soon after commissioning in early 2009, it was recognised that two stages of conventional cleaning were unable to produce plant final Copper concentrates with the desired Uranium grades. Studies have shown that a significant quantity of liberated and fine Uranium particles (up to 80-90%) is recovered with the concentrate by entrainment. Lumwana Operations sought the use of the Jameson Cell technology after observing successful implementation at other sites in rectifying similar issues. In late 2009, Jameson Cell pilot-plant test work was conducted at the site and this quickly led to the installation of a full-scale B5400/18 model into the existing cleaner circuit. Inclusion of the Jameson Cell resulted in the plant being able to effectively produce clean and saleable concentrate with Uranium contents consistently below acceptable levels. Initially designed to be used as a cleaner scalper cell at the head of the existing conventional cell circuit, subsequent operational issues with the existing recleaner cells forced reconfiguration of the cleaner circuit and reintroduced the Jameson Cell for the recleaning duty. This paper reports the Lumwana Operation from the first two years of operation where it struggled to produce a saleable concentrate with Uranium grades below 150 ppm, which is the maximum allowable limit set out in the smelting contracts, to steps taken in the subsequent years to implement a satisfactory solution. The plant can now consistently produce a clean Copper concentrate with Uranium grades well below 130 ppm and the overall improvements implemented at site have seen the overall plant recovery improve by 1.3%.

KEYWORDS

Flotation, Jameson Cell, Uranium, Copper, Cleaner Circuits, Zambia.

INTRODUCTION

Lumwana is an open pit Copper mine located in the Northwestern Province of Zambia approximately 80 km west from the provincial capital of Solwezi, approximately 300 km northwest from the Copperbelt and 700 km northwest from the national capital of Lusaka (Figure 1). The Lumwana Project was developed by the former Canadian mining company Equinox Minerals Limited (“Equinox”), through its Zambian subsidiary Lumwana Mining Company Limited. The mine was commissioned in 2008 and commercial Copper production started in the second quarter of 2009. In June 2011, Barrick Gold Corporation (“Barrick”) acquired Equinox whose primary asset was the Lumwana Mine.

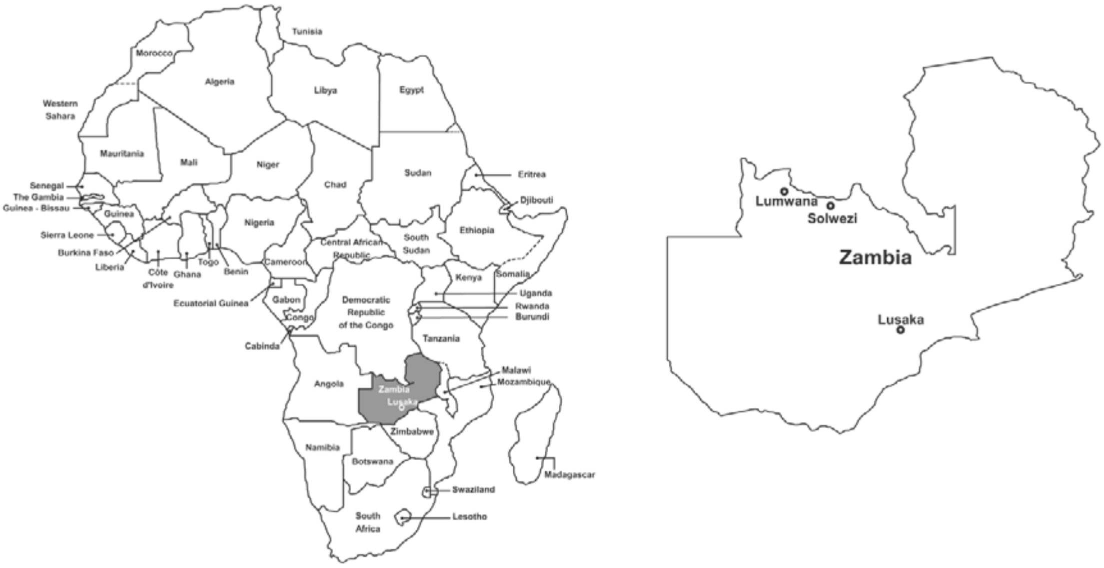


Figure 1 – Map of Africa and Zambia showing location of the Lumwana operation

The Lumwana mine contains two major Copper deposits: Malundwe and Chimiwungo, that are hosted almost entirely within Muscovite-phlogopite-quartz-kyanite schists with disseminated sulphides dominated by Chalcopyrite and Bornite (Davis et al., 2006). Malundwe, the smaller deposit, was mined first and has high Copper grades along with Gold, Uranium and sporadic high Cobalt grades. The Chimiwungo deposit is much larger in size but with lower Copper grades, high Cobalt grade zones and Uranium mineralization (Davis et al., 2006). Mining of the Chimiwungo deposit commenced in August 2012. The expected life of mine is 20 years for mining and 33 years for processing.

The processing plant (Figure 2) was designed to treat 2,500 tonnes per hour (tph) of ore with head grades between 0.4 to 0.6% Cu using a SAG mill (38’x18’, 18 MW) and ball mill (26’x40’, 16 MW) to produce a primary grind with D_{80} of 300 µm for subsequent flotation. The flotation circuit consists of two parallel lines of seven 160 m^3 Wemco cells for rougher flotation followed by two stages of conventional cleaning using four 50 m^3 Dorr-Oliver cells for the first stage of cleaning, three 50 m^3 Dorr-Oliver as cleaner scavengers and five 17 m^3 Dorr-Oliver for recleaning. A second ball mill (13’x23’, 1.25 MW) is used to regrind the rougher (and cleaner scavenger) concentrate to a D_{80} of 106 µm for cleaning. The plant, commissioned in early 2009, is now treating a higher tonnage (3,200 t/h) than the nameplate design. The Copper concentrate produced from the plant (typically between 35 to 45% Cu depending on Copper mineralogy) is sold to different smelters where the maximum allowable Uranium limit in the contracts is 150 ppm.

This paper chronicles the Lumwana Operation from the first two years of operation where it struggled to produce a saleable concentrate within the acceptable Uranium limits to steps taken in the subsequent years to implement a solution.

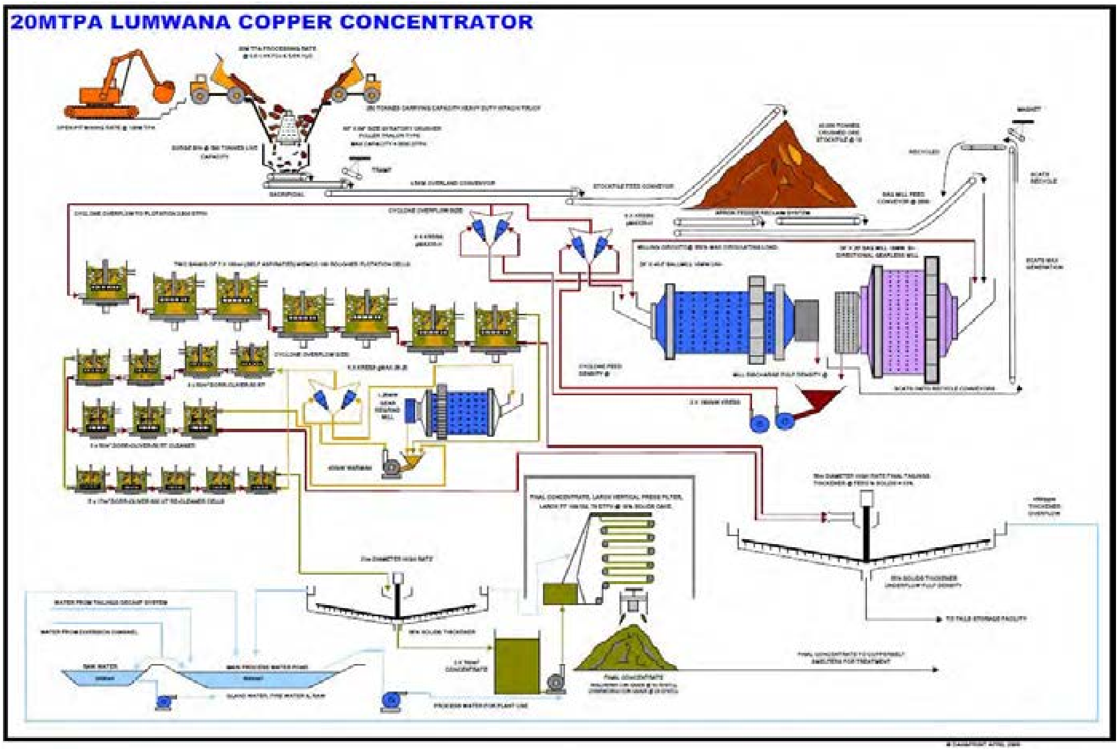


Figure 2 – Flowsheet of the original design of the ore processing plant at Lumwana

PROBLEM DEFINITION

Several months into normal operation and following the commissioning of the concentrator it was recognised that two-stages of conventional cleaning were unable to consistently produce final plant Copper concentrates within the acceptable Uranium limits. Figure 3 shows the Uranium grade of the plant final Copper concentrate and of the feed over the first two full years of operation from 2009 to 2011. Firstly, it shows that the level of Uranium in ore treated is extremely variable and can be below 10 ppm, but can also be as high as 400 ppm. Similarly, the Uranium grades in the final concentrate are just as variable and in many instances exceed the highest acceptable limit of 150 ppm. ‘Off spec’ concentrate need to be stockpiled and blended with the cleaner concentrates with lower levels of Uranium. Figure 3(B) shows the Uranium grades in the final Copper concentrate is fairly evenly spread across a wide range of values which highlights the poor ability of the plant to consistently produce a clean concentrate and to effectively deal with the feed variations. Post July 2010, the Uranium grades in the feed appeared lower than the first year of operation but strangely, the Uranium grade of the final Copper concentrate does not follow this trend and remained on the same level which emphasises the inefficiency of the cleaning circuit to reject Uranium.

Uranium contained in the Lumwana deposit is mostly present as Uranite having particle sizes as fine as 10 µm and is associated with the same lithology that hosts the Copper mineralization (Smith, 2013). Mineralogical analysis and diagnostic tests on plant cleaner samples (from bench scale flotation tests) have showed that the majority of the Uranium particles are very fine, liberated and are admitted into the froth by entrainment. So in theory, much of the Uranium can potentially be rejected, but the problem was that the conventional cleaner cells in the plant were ineffective in controlling entrainment, exacerbated by the changing amounts of Uranium in the ore. In fact, plant data (Figure 4) showed that the recleaner stage is totally inadequate as it did not appear to upgrade Copper, nor reject any Uranium as evidenced by the fact that the Copper and Uranium grades in the first cleaner concentrate and the recleaner (final plant) concentrate were similar in value.

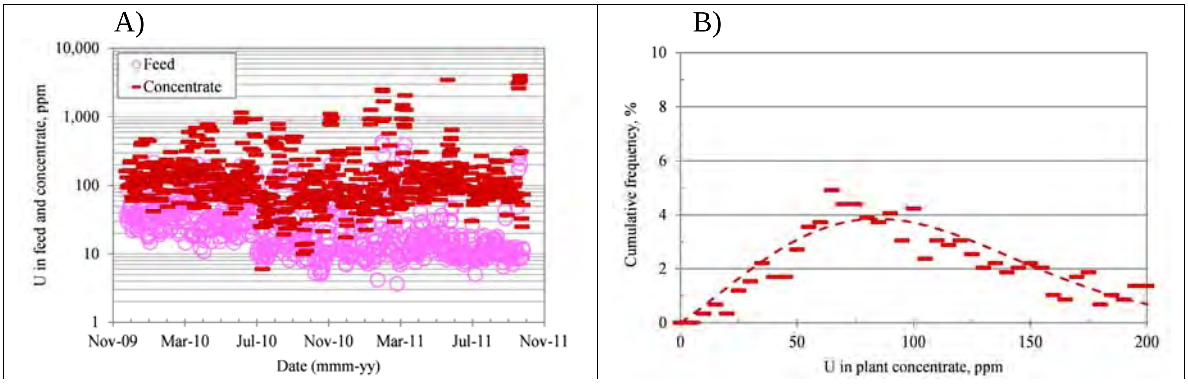


Figure 3 – Uranium grade in the flotation feed and final Copper concentrate produced by the plant over the first two years of operation

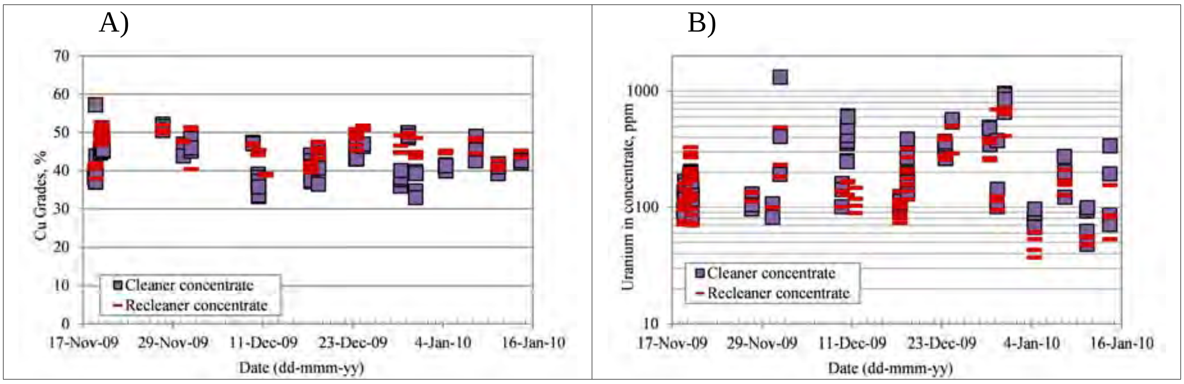


Figure 4 – Grade of Uranium in the plant cleaner and recleaner concentrate shown consecutive months of operation. (A) Copper grade, (B) Uranium grade

Lumwana sought the use of the Jameson Cell technology to address its Uranium issue after seeing the successful implementation at OZ Mineral's Prominent Hill operation which produces high grade and clean concentrate by rejecting the penalty element, Fluorine (Barns et al., 2009). Since then, the Jameson Cell technology has been retrofitted to a number of Copper concentrators to solve process challenges in existing conventional cleaner circuits (Araya et al., 2013). These include PanAust's Phu Kham operation in Laos as described by Bennett et al. (2012) and Newcrest's Telfer operation in Western Australia as described by Seaman et al. (2012). Other reasons for the success of the Jameson Cell technology in these applications reflect its high productivity in a small footprint and simple integration into an existing plant resulting in highly desirable minimal disruption to production

THE JAMESON CELL

The Jameson Cell is a robust, efficient and innovative flotation technology driven by fluid mechanics. Commercialised in the late 1980s, it has been continuously developed over two decades and now has more than 320 installations worldwide across many industries including coal, base metals, industrial minerals, potash and oil sands.

In a Jameson Cell feed slurry is first pumped through a restriction (the slurry lens orifice) to create a high pressure jet which then enters a cylindrical device called a downcomer (Figure 5). The jet of liquid first shears and then entrains air from the atmosphere. Removal of air into the jet causes a vacuum to be generated inside the downcomer. When a hydraulic seal is formed at the bottom of the downcomer, the vacuum causes a column of slurry to be drawn up inside the downcomer. The jet of slurry plunges onto the liquid surface and the high kinetic energy of the jet then disseminate the entrained air into very fine bubbles (Evans et al., 1995). In this zone, the high intensity of the system creates a very favourable environment for the bubbles and particles to collide and attach. The air bubbles and mineral particles move continuously down the downcomer before exiting into the tank. The particle laden bubbles then float to the top to form the froth whilst the hydrophilic particles remain in the pulp phase to be removed as tailings. To ensure consistent operation, tailings recycle is employed. This dampens feed fluctuations to the cell allowing the downcomer to operate at a constant feed pressure and flowrate (Cowburn et al., 2005). The high rate of mixing from the high pressure slurry jet and the fact that the air is self-aspirated, allows the Jameson Cell to have no moving parts other than the feed pump. No agitators or compressors are required.

Due to rapid kinetics and a separate contact zone in the downcomer, the tank is not sized for residence time, so tank volumes tend to be very much smaller than equivalent capacity mechanical and column cells (Harbort et al., 2003).

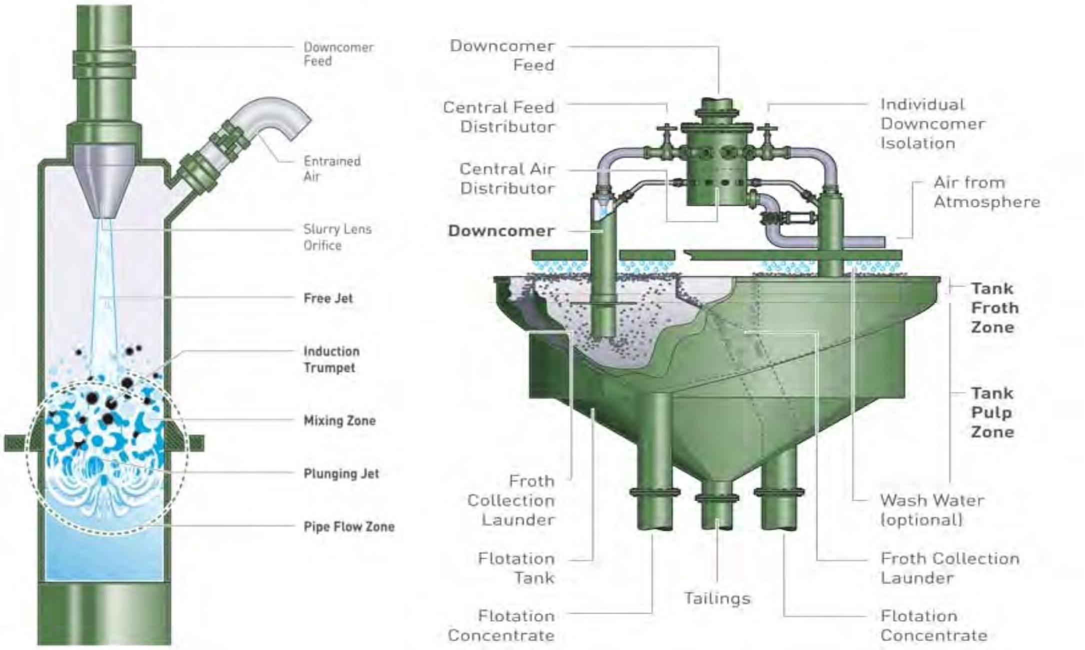


Figure 5 – The Jameson Cell showing downcomer (on left) and overall assembly in tank (on right)

JAMESON CELL PILOT PLANT TESTWORK

In late 2009, a Jameson Cell L500 pilot plant test rig was brought onto the site to allow continuous test work to be conducted. This pilot plant (Figure 6) is a self-contained unit which has a

100 mm diameter downcomer in a 500 mm diameter flotation tank, a washwater system, a pump box incorporating the tailings recycle mechanism and a feed pump. The feed to the cell is a bleed stream from the plant that can be varied from 3 to 8 m^3/h of slurry to provide the desired quantity of tailings recycle (30 to 80%) for testing. The control panel of the pilot unit has pressure gauges to measure feed and vacuum, rotameters for controlling air and wash water flow rates and automatic control of cell level and hence, control of the froth depth.

Figure 6 – Jameson Cell L500 pilot plant test rig at Lumwana site

The pilot-scale rig was operated in continuous mode to simulate Jameson Cell performance in a full scale and the test campaign was carried out over a two-month period from November 2009 to January 2010. Different streams within the cleaner circuit were investigated but the focus was on treating the cleaner feed stream as it was envisaged that the Jameson Cell would be installed in a cleaner scalper duty in the full-scale plant. The results in Figure 7 show that, firstly, the Jameson Cell can produce the required final grade Copper concentrate (typically between 35 to 45% Cu depending on mineralogy) in a single stage of flotation. Copper recovery ranged from 40 to 90% which was controlled by the process variables and settings used in each test. The Copper grade/recovery curve is flat across the entire recovery range indicating that the Copper minerals recovered must be well liberated. The selectivity achieved between Cu and U recovery is rather impressive. It shows that 80-90% of the Uranium in the cleaner feed stream can be rejected. The selectivity remains good even up to a Copper recovery of around 80-90%, remembering again that this is achieved in a single stage of flotation.

A)	B)

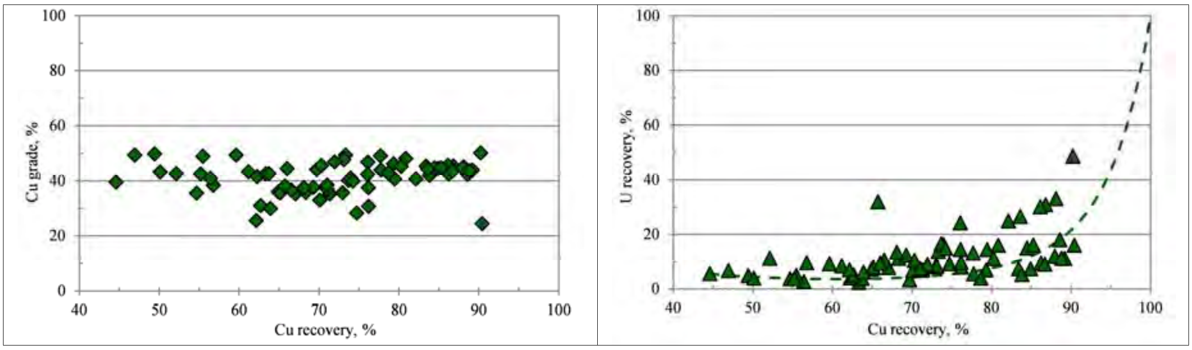


Figure 7 – (A) Copper grade/recovery curve and (B) Copper/Uranium selectivity achieved by the Jameson Cell pilot plant treating the cleaner feed stream

Figure 8 compares the Uranium grades of the Copper concentrate produced by the Jameson Cell in a single stage of flotation versus two stages of conventional cleaning in the plant when treating the same feed. The Copper grades in the concentrate are similar in both cases (not shown) but clearly, the Jameson Cell is far superior in rejecting entrained particles of Uranite. Figure 8 also highlights the fluctuating nature of Uranium grade in the cleaner feed stream which perhaps, explains the challenge for the plant to be able to consistently produce ‘in spec’ concentrates. However, the Jameson Cell can consistently produce a Copper concentrate with Uranium levels well below 130 ppm by effectively removing the entrained fine and liberated Uranite particles. Tests showing Copper concentrates with the higher Uranium grades, say above 100 ppm, in Figure 8 can be attributed to the remaining Uranium that is locked within Copper mineral particles. Otherwise, Uranium grades in the Copper concentrates are typically much lower (20-80 ppm).

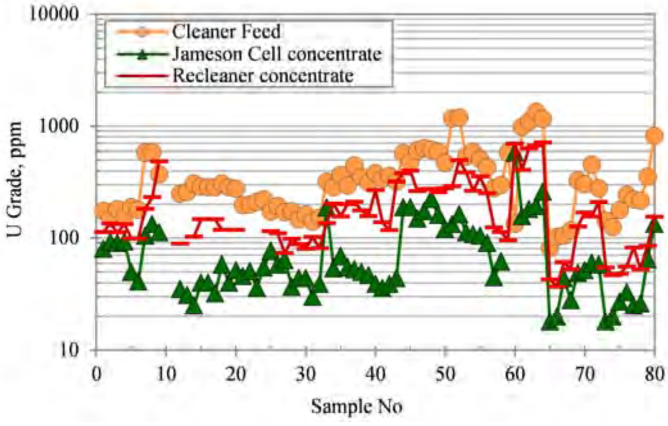


Figure 8 – Uranium grade comparison for single stage Jameson Cell pilot plant versus two stage of conventional cell cleaning in the plant

FULL SCALE JAMESON CELL INSTALLATION

The positive outcome of the pilot plant test work convinced Lumwana to purchase a full scale Jameson Cell in May 2010. The location of the installed Jameson Cell is shown in Figure 9. A model B5400/18 Jameson Cell (which is a 5.4 diameter circular cell with 18 downcomers) was appropriate to treat the tonnage and volumetric flowrate of the cleaner feed stream and this cell was to be installed at the head of the existing cleaner circuit (Figure 10). Lumwana chose an external engineering company for drafting and designing the structure and layout but the project was managed in-house by Lumwana site. The Jameson Cell was installed in a stand-alone structure next to the existing cleaner circuit with the cell elevated to allow the concentrate produced from the Jameson Cell to gravity flow to the existing final concentrate hopper and the tailings from the Jameson Cell to gravity flow into the distribution box at the head of the plant cleaner circuit. The higher cost of elevating the cell was offset by the lower operating and maintenance costs compared to pumping concentrate and tailings.

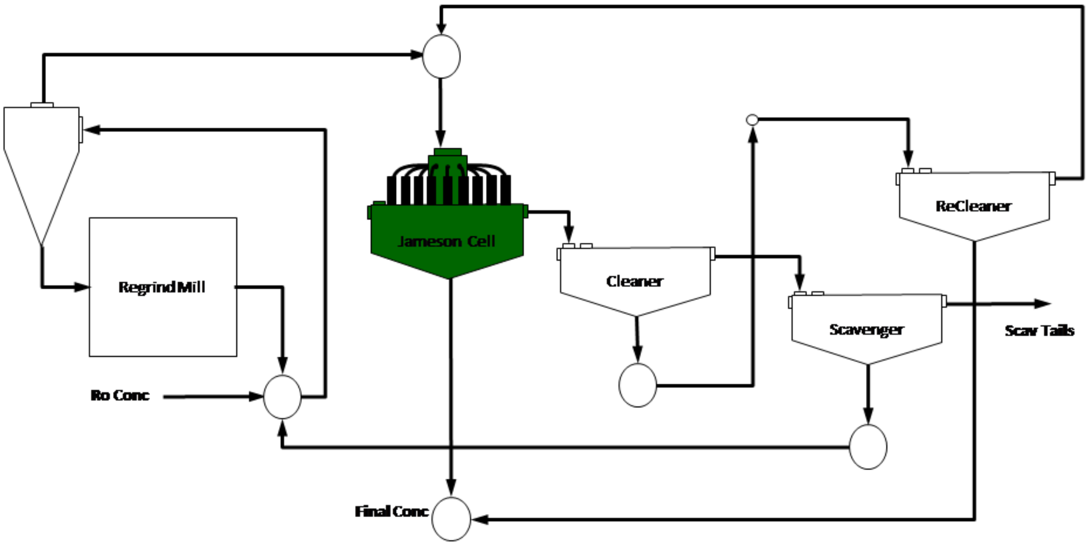


Figure 9 – Flowsheet showing the inclusion of a Jameson Cell in a cleaner scalper duty at the head of the existing circuit


A)

B)

Figure 10 - The full scale model B5400/18 Jameson Cell installed at Lumwana

Due to some minor problems on the project caused by third party contractors (relating to logistics and local fabrication of parts), the installation of the cell was several months late and commissioning of the Jameson Cell commenced in July 2011. This process is straightforward and usually takes less than a week to complete but in this case, other issues arose which slowed progress. On start-up, the feed pump delivered too high a feed pressure (200 to 220 kPa) and excessive volumetric flow rate to the Jameson Cell. The pulley on the motor needed to be changed to reduce the feed pressure back to the correct design value (150-160 kPa). The polyethylene pipes connecting the Jameson Cell tails outlet to the External Recycle Mechanism (ERM) box, which controls tailings recycle, was of the wrong schedule, i.e., too small an internal diameter, which restricted the flow of tailings from the Jameson Cell. The whole assembly had to be dismantled and pipe sections needed re-working (ground out) to increase the diameter. Due to a design error by the engineering company contracted to design the circuit, the concentrate pipes from the Jameson Cell tank outlets to the final concentrate hopper were erroneously reduced to 150 mm when they should be 400 mm. This caused the concentrate (which is frothy) produced by the Jameson Cell to back up into the launders. These pipes were removed and replaced with the larger ones (400 mm). All these issues were quickly resolved but the modifications required several months to rectify so the commissioning process took much longer than expected and the Jameson Cell was not fully operational until August 2011.

On handover, the metallurgical performance of the full-scale Jameson Cell was fully assessed by the operations personnel to see if it matched the performance demonstrated from the pilot plant test work as assured by Xstrata Technology. Figure 11 compares the performance of the full scale Jameson Cell to that of the pilot plant. Results are almost identical in terms of the Copper recovery that can be achieved and more importantly, the excellent selectivity between Copper and Uranium.

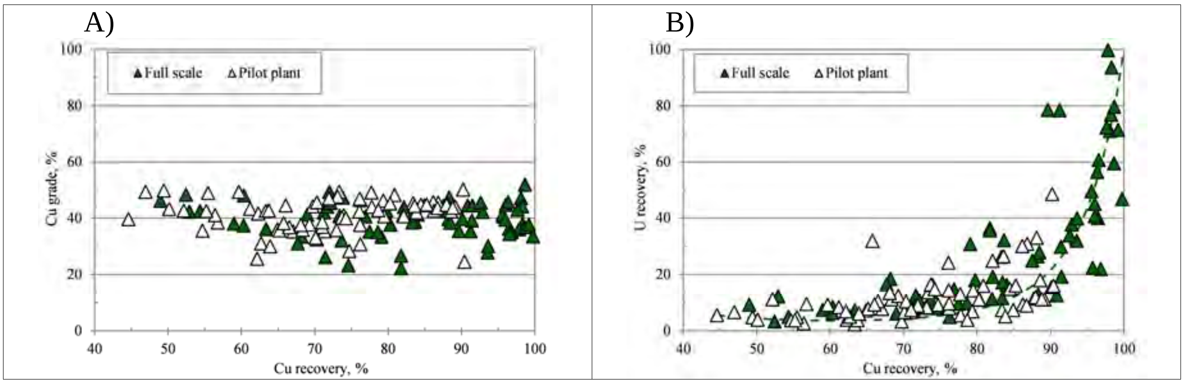


Figure 11 – Comparison of metallurgical performance of the full scale Jameson Cell to the pilot plant. (A) Copper grade/recovery, (B) Copper/Uranium selectivity

The high Copper recovery that the Jameson Cell is able to achieve while still being able to reject Uranium (up to 80-90%) at the head of the cleaner circuit should significantly lessen the load in the cleaner circuit downstream. Although the cleaner circuit is now treating a stream with higher levels of U (tails from the Jameson Cell), it needs to recover much less Copper, and consequently it can be pulled much more slowly, i.e. by increasing the froth depth and using less air, thereby allowing the froth to drain more thoroughly which will decrease entrainment. The lower solids content of the slurry feeding the conventional cleaner circuit will also be beneficial in reducing entrainment by dilution cleaning. Figure 12 illustrates the Uranium grades in the Copper concentrate produced by the Jameson Cell and the recleaner conventional cells over a two month period in 2011. These two concentrates are combined to give the overall plant final concentrate which is also shown in the graph. The figure on the right shows the whole data set: sorted based on frequency of occurrence, organized in bins with sequential increments of 5 ppm and presented as cumulative distribution which is a powerful tool to compare concentrates of different quality. A perfect flotation system would generate a vertical line, meaning that 100% of the time the same U grade is achieved. Deviation from the vertical line indicates less than perfect performance, which is more realistic, and the spread of the probability distribution. In this case, it is highlighted the superior performance of the Jameson Cell compared to the recleaner cell. The Jameson Cell curve has a much steeper slope meaning it can produce Copper concentrates within a narrow range of Uranium grades. It can produce Copper concentrates with Uranium grades below 100 ppm for about 90% of the time. In comparison, the curve for the recleaner cell is very broad. The concentrate produced by the recleaner mechanical cells is very high in Uranium and when combined with the clean low Uranium grade concentrate from the Jameson Cell to give the final plant concentrate, the overall benefit provided by having a Jameson Cell in the circuit is diminished.

A)	B)

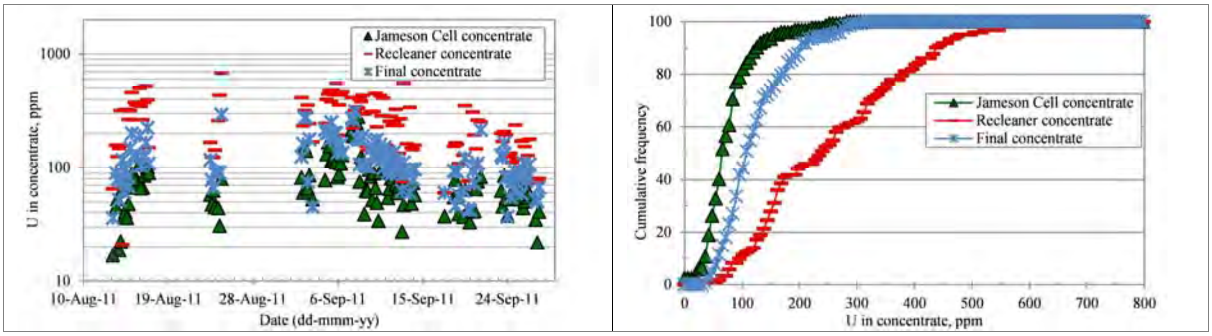


Figure 12 – U grade of the Copper concentrate produced by the full scale Jameson Cell, the recleaner cells and the combined plant final concentrate
(Note: The same data is presented as date [A] and cumulative frequency of occurrence [B])

The poor performance of the recleaner cells was identified as an operational issue long before the Jameson Cell was installed. Essentially, the dart valves on this bank of cells do not function properly so the froth depth and hence, pulling rate of these cells cannot be controlled. This in part explains the lack of ability of these cells to reject Uranite, as shown in Figure 4 (for the original plant) and Figure 12 (after the Jameson Cell was installed). Lumwana Operations management was initially keen to remove this bank of recleaner cells from the circuit and as the installation of the Jameson Cell was primarily intended to produce clean concentrate rather than to add more capacity, the logical move was to re-allocate the duty of the Jameson Cell to recleaning (see Figure 13).

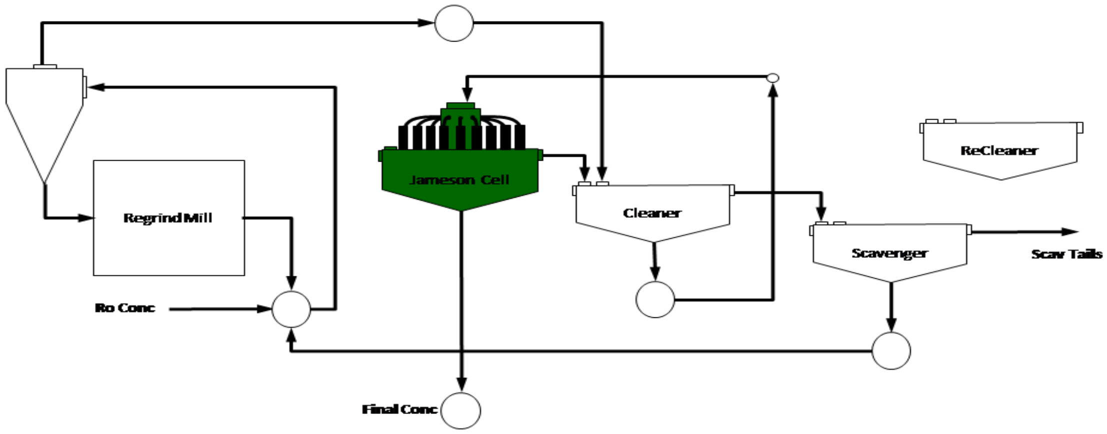


Figure 13 – Revised cleaner circuit flowsheet showing the Jameson Cell operating in the recleaner duty. (Note: the original plant recleaner cells are taken offline)

An initial concern was whether this cleaner circuit configuration would affect the overall plant recovery as the Jameson Cell was not originally intended to make the entire plant final concentrate. With pipework in place to be able to change back the Jameson Cell to the cleaner scalper duty if required, trials on this new circuit were commenced in early 2012. With the Jameson Cell now treating the cleaner concentrate, it was shown to be able to consistently produce a very clean concentrate, low in Uranium. Several weeks of continuous operation resulted in no detrimental effect on the overall plant recovery indicating there was sufficient capacity in the installed Jameson Cell to produce the entire plant concentrate. The plant team then decided to continue using the Jameson Cell permanently in the recleaner duty.

Figure 14(A) shows the Uranium grades in the plant Copper concentrate from plant daily samples collected in the first two years of normal operation (November 2009 to 2011), followed by the period from then up to August 2013 when the Jameson Cell replaced the conventional recleaner bank. There is clear reduction in the Uranium grade in the Copper concentrate after the Jameson Cell was installed. Figure 14(B) shows the frequency distribution, i.e. a function that shows the number of observations within a defined interval, of the Uranium grades in the Copper concentrate for the two separate periods (the frequency distribution after the Jameson Cell was installed is overlaid to the graph shown in Figure 3(B)). The difference is impressive and shows a significant shift in the curve to the left where the average Uranium grade in the Copper concentrate produced is now much lower (65 ppm compared to 85 ppm before the Jameson Cell was installed) and the narrowing of the distribution and higher relative frequency around the average which means the plant is able to produce a lower Uranium grade Copper concentrate more of the time. Also, the 'tail' on the right is reduced to zero around 130 ppm which means the plant is always producing final grade Copper concentrate with Uranium grades below this level. This is a vast improvement compared to the original two-stage conventional cleaner circuit where the Uranium grades in the final Copper concentrate appeared to be highly indiscriminate (as evidenced by the wide deviation around the average as shown in Figure 14(B)) and as random as the Uranium grades in the feed treated.

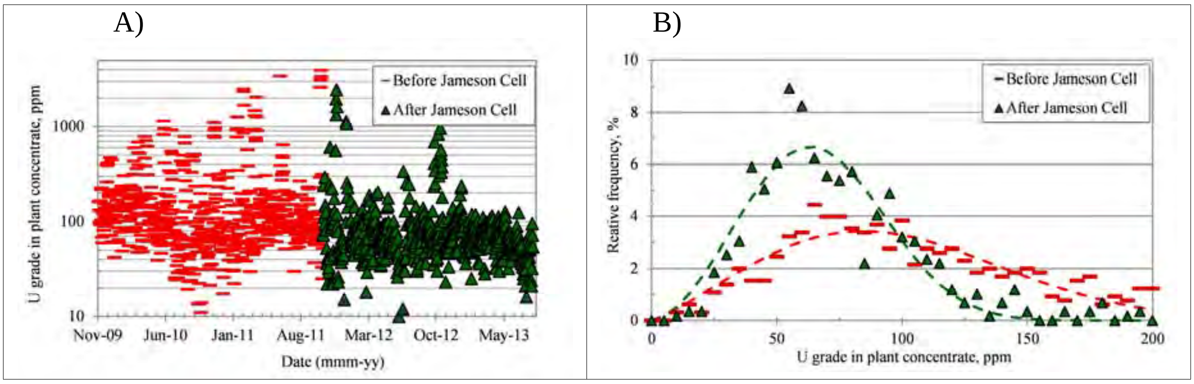


Figure 14 – Comparison of the Uranium grades in the plant final Copper concentrate for the periods before and after installation of the Jameson Cell

Figure 15 shows the frequency distribution for Copper recovery in the plant for the periods prior to and after the Jameson Cell was installed. It can be seen that statistically, the two periods form two independent populations. The period after the Jameson Cell shows the average Copper recovery has increased by 1.3%, which is significant. However, this improvement cannot be solely attributed to replacing the conventional recleaner cell bank with a single Jameson Cell. The two years following the introduction of the Jameson Cell has seen Lumwana Operations implement a number of process improvement strategies including ways to achieve more consistent feed through the plant and reagent-focussed initiatives including trials and optimisation.

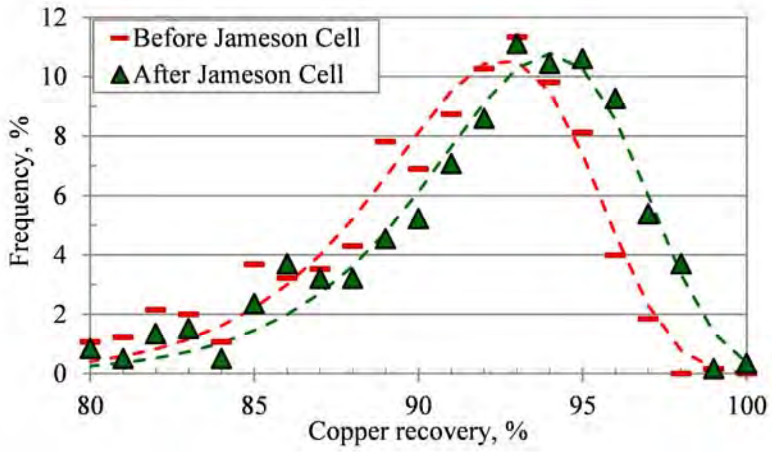


Figure 15 – Comparison of the overall plant Copper recovery for the periods before and after installation of the Jameson Cell

CONCLUDING REMARKS

In the first two years of operation, the Lumwana concentrator struggled to consistently produce a saleable Copper concentrate with acceptable levels of Uranium below 150 ppm. The initially installed two-stage conventional cleaner circuit was unable to handle the varying nature of Uranium content in the ore treated and proved ineffective in rejecting fine Uranite particles. An urgent solution was sought which commenced with Jameson Cell pilot plant testing at the plant site. The results obtained confirmed that the majority of the Uranium (around 80-90%) enters into the froth by entrainment and could be rejected to produce a high grade Copper concentrate with Uranium grades well below the desired limits.

A full-scale Jameson Cell was purchased and installed in a cleaner scalper duty at the head of the existing cleaner circuit. The Jameson Cell produced a very clean Copper concentrate consistently low in Uranium and the performance of the full scale cell was identical to that shown during piloting. However the poor performance of the recleaner conventional cells which produces a 'dirty' concentrate very high in Uranium diminished the overall benefit provided by the Jameson Cell and as a consequence the circuit was altered. The recleaner bank was taken offline and the duty of the Jameson Cell was changed to recleaning. This rearrangement has allowed the plant to consistently produce a clean saleable final grade cleaner concentrate with Uranium grades averaging 65 ppm and always below 130 ppm. While the Jameson Cell was specifically employed to solve a concentrate grade issue at Lumwana rather than adding cleaning capacity, its installation and other initiatives undertaken by site personnel over the past two years has seen recovery at the plant increase by 1.3%.

ACKNOWLEDGEMENTS

The authors would like to thank Barrick for allowing us to publish this paper and to all the personnel from Lumwana Operations who has assisted in this project over the years. We would like to acknowledge Matthew McInnes, formerly from Equinox and Lyndon Ryan, formerly from Xstrata Technology Australia, for their contribution at the early stages of the project. Finally, we would like to thank Dave Osborne, from Xstrata Technology Australia, for his valuable suggestions and feedback in the preparation of this paper.

REFERENCES

Araya, R., Huynh, L., Young, M. & Arburo, K. (2013). Solving challenges in copper cleaning circuits with the Jameson Cell. *10th International Mineral Processing Conference.* (pp. 261-271). Santiago: GECAMIN.

Barns, K., Colbert, P. & Munro, P. (2009). Designing the Optimal Flotation Circuit - The Prominent Hill Case. *Tenth Mill Operators' Conference*, (pp. 173-182). Adelaide.

Bennett, D., Crnkovic, I. & Walker, P. (2012). Recent process developments at the Phu Kham copper-gold concentrator, Laos. *11th Mill Operators' Conference* (pp. 257-272). Hobart, Tasmania: AusIMM.

Cowburn, J., Stone, R., Bourke, S., & Hill, B. (2005). Design developments of the Jameson Cell. *Centenary of Flotation Symposium* (pp. 193-199). Brisbane: AusIMM.

Davis, M., Bertinshaw, R., Miller, T. & Hanbury, R. (2006). *Lumwana Copper Project, North West Province, Republic of Zambia, Technical Report.* Retrieved from SEDAR: http://www.sedar.com

Evans, G., Atkinson, B., & Jameson, G. (1995). The Jameson Cell. In e. K. Matis, *Flotation Sci Eng* (pp. 331-363). New York: Marcel Dekker.

Harbort, G., De Bono, S., Carr, D. & Lawson, V., (2003), *Jameson Cell fundamentals - a revised perspective, Minerals Engineering, (pp. 1091-1101)*

Seaman, D., Burns, F., Adamson, B. & Manton, P. (2012). Telfer processing plant upgrade - The implementation of additional cleaning capacity and the regrinding of copper and pyrite concentrates. *11th Mill Operators' Conference* (pp. 373-381). Hobart, Tasmania: AusIMM.

Smith, S. (2012). Mineralogical analysis of Lumwana Cu/U Samples. Johannesburg, South Africa: ALS Laboratory Group.

DESIGN AND PERFORMANCE ASPECTS OF COAL FLOTATION – EXPERIENCES WITH THE JAMESON CELL

L. Huynh, I. Kohli and D. Osborne
Xstrata Technology, Australia

H. De Waal, C. Walstra
Xstrata Technology, South Africa

ABSTRACT

Flotation is commonly used to treat fine coal (typically below 500 microns in size) and is a complex, three-phase process that is controlled by factors which can be divided into three facets: coal, chemistry and machine. It is most often used for treating metallurgical coal fines where the value of the product can justify the added treatment cost of cleaning and dewatering the product component. However, because of improvements in flotation technology and the dewatering of both the product and the tailings this avenue is becoming increasingly attractive for treating thermal coals.

The Jameson Cell technology is an established robust and efficient high intensity flotation technology which has been continuously developed and improved over two decades. Its high capacity, small footprint and low maintenance requirements have made it more or less a standard flotation technology in the Australian coal industry.

This paper will review how the fundamental characteristics of a flotation machine translate to economic benefits. The performance of the Jameson Cell and continued operational challenges of fine coal circuits will also be discussed.

INTRODUCTION

Flotation is a complex multifaceted process that can be separated into three main areas: the coal, the chemistry and the machine, as shown in Figure 1. To solve plant issues it is important to understand how different factors within these areas affect and control flotation performance for a particular system. Factors within the coal and chemistry areas are dynamic and hence need to be dealt with by personnel on an ongoing basis in normal plant operations. However, factors associated with the machine are generally a characteristic of the machine type itself as this relates to the fundamental design of the technology. One of the most important characteristics of any flotation technology is air bubble generation and the size of air bubbles produced as this controls flotation kinetics and also, it dictates the carrying capacity of the machine. Another crucial component is how the machine effects collision and contact between air bubbles and particles.

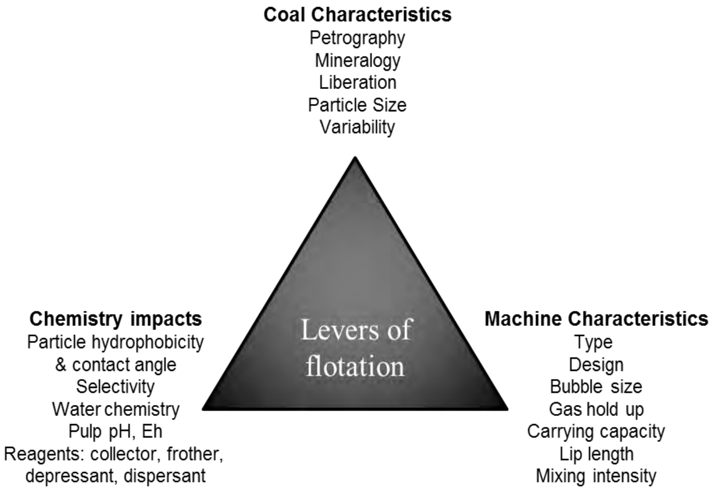


Figure 1: Flotation 'triangle' showing the factors affecting performance.

The Jameson Cell technology was first introduced in the late 1980s to overcome the design and operating inadequacies that were being experienced with column and conventional flotation cells. From the first commercial installation at Mt Isa in 1989, which was for sulphide treatment followed by the first coal installation at Newlands in 1990, it has been continuously developed and improved to improve performance and make it more robust and easier to use. Over 150 Jameson Cells are now operational in coal applications worldwide, with the current largest installation being at Wesfarmer‟s Curragh Mine in Central Queensland which treats over 5 million tonnes of coal fines per year using twelve cells. Long-established coal-producing countries like Kazakhstan and South Africa are realising the benefits of the Jameson Cell over conventional cells and emerging coal regions such as Mozambique and Mongolia are now beginning to use the Jameson Cell for metallurgical coal applications.

The paper explains how fundamental characteristics of a flotation technology translate into an economic advantage. The performance of the Jameson Cells and ongoing operational challenges in flotation and fine coal circuits will also be discussed.

JAMESON CELL PRINCIPLE OF OPERATION

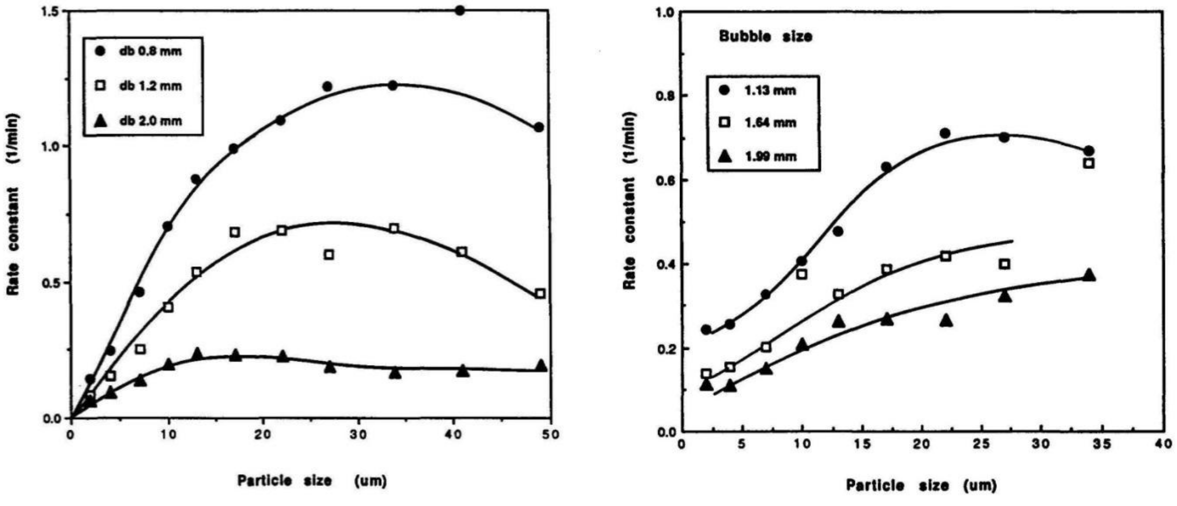
The fundamentals of Jameson Cell operation have been previously described by a number of other authors including Evans, Atkinson & Jameson (1993). To summarise, in a Jameson Cell feed slurry is first pumped through a restriction (the slurry lens orifice) to create a high pressure jet which then enters a cylindrical device called a downcomer. The jet of liquid first shears and then entrains air from the atmosphere. Removal of air into the jet causes a vacuum to be generated inside the downcomer. When a hydraulic seal is formed at the bottom of the downcomer, the vacuum causes a column of slurry to be drawn up inside the downcomer. The jet of slurry plunges onto the liquid surface and the high kinetic energy of the jet then disseminates the entrained air into very fine bubbles. In this zone, the high intensity of the system creates a very favourable environment for the bubbles and particles to collide and attach. The air bubbles and mineral particles move continuously down the downcomer before exiting into the tank. The particle laden bubbles then float to the top to form the froth whilst the hydrophilic rock and mineral particles remain in the pulp phase to be removed as tailings. To ensure consistent operation, tailings recycle is employed. This dampens feed fluctuations to the cell allowing the downcomer to operate at a constant feed pressure and flowrate. The high rate of mixing from the high pressure coal laden slurry jet and the fact that the air is self aspirated, allows the Jameson Cell to have no moving parts other than the feed pump. No agitators or compressors are required.

Due to rapid kinetics and a separate contact zone in the downcomer, the tank is not sized for residence time, so tank volumes tend to be very much smaller than equivalent capacity mechanical and column cells.

BUBBLE SIZE

Flotation Kinetics

Bubble size is one of the most important factors in any flotation system as it has a strong influence over flotation kinetics. Fine bubbles increase the flotation kinetics across *all* particle sizes (Diaz-Penafiel & Dobby, 1994; Ahmed & Jameson, 1985), and not just recovery of fine particles as has often been hypothesised. This is clearly demonstrated in Figure 2 which shows the flotation rate of two different systems: silica and pyrite particles, which are floated at three different bubble sizes. The bubble sizes in this investigation (1 to 2 mm) are more or less the same as those encountered in many current industrial machines. In coal flotation, fine bubbles also improve separation as they intensify the difference in the kinetics of the coal from non-coal particles, thus allowing concentrates with lower ash content to be produced without loss in yield.



(a) (b)

Figure 2: Rate constant as a function of particles size, for three bubble sizes at constant J_g of 1.53 cm/s: (a) Silica and (b) Pyrite (from Diaz-Penafiel & Dobby, 1994).

Carrying capacity

Bubble size dictates the carrying capacity of a flotation machine. Finer bubbles increase the carrying capacity (often measured as the mass flow rate (tonne/h) of concentrate per m^2 of surface area of the flotation machine) as there is more bubble surface area per volume of air added for particles to attach. Essentially, this means that if two different types of flotation machine are used to float the same coal using the same amount of air, the one generating air bubbles (measured as a distribution) that is half the size of the other machine will have double the bubble surface area available for flotation. This is therefore a very important consideration, especially for coal feeds offering very high potential yields, i.e., up to 80-90% mass; so the more bubble surface area that is available from the machine, the lower will be the cross-sectional area required and fewer cells will be required to recover all the coal.

The Jameson Cell is able to produce fine bubbles via the shearing action of a plunging jet (Evans, Jameson & Atkinson, 1992). It is this fundamental characteristic that allows the Jameson Cell to float particles quickly, attain superior selectivity and have high productivity (carrying capacity).

The air bubbles generated by the Jameson Cell are in the range of 300 to 700 µm (Sauter mean diameter, D_{32}) (Evans, Atkinson & Jameson, 1993). Figure 2 compares the bubble size of a range of industrial mechanical and columns cells (Nesset, Finch & Gomez, 2007) to that of the Jameson Cell (Osborne et al., 2013). All results collated in Figure 2 were determined by the same bubble size measurement technique as developed by McGill University and described by Chen, Gomez & Finch (2001) and Gomez & Finch (2007).

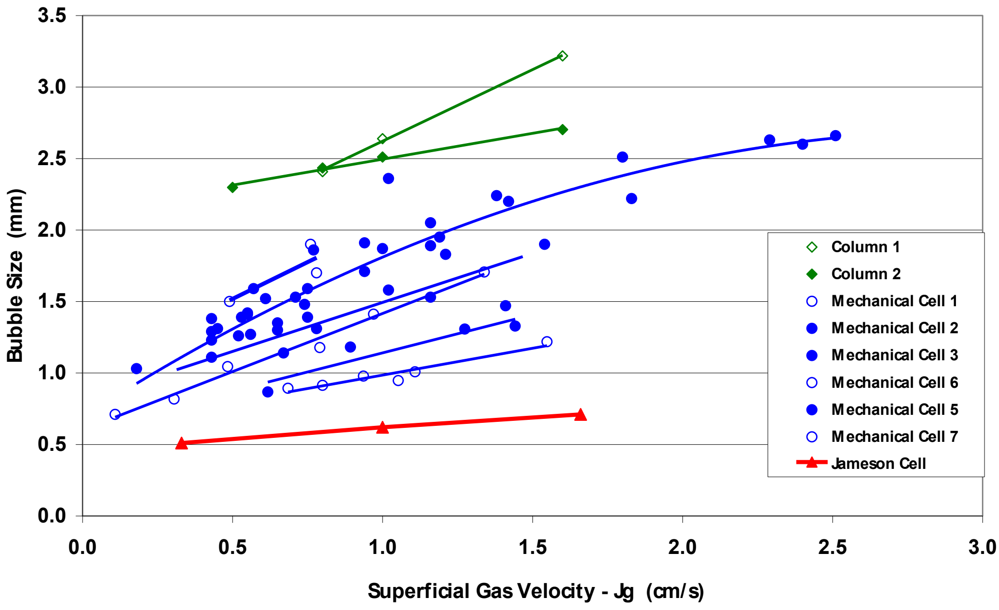


Figure 3: Bubble size as a function of J_g (superficial gas velocity, a measure of air flow rate) for different flotation technologies.

FROTH WASHING

A flotation technology that has froth washing capabilities has advantages over those that do not, as froth washing is one of the most effective methods employed to reduce entrainment thereby enabling concentrates with the lowest ash content to be produced. The amount of wash-water used is therefore an important process variable so the system must be able to be operated over the desired flowrate range for that sized flotation cell based on the designed concentrate tonnages and solids content of the concentrate. Wash-water addition is commonly measured as a ratio with the water in the concentrate. A wash-water ratio of 1.0 means that, theoretically, all the „dirty" water in the concentrate containing the entrained ash particles is replaced by clean wash-water. In practice, the wash-water addition is dependent on the process operation and in particular, factors such as the structure and stability of the froth which in turn is influenced by factors such as particle size and hydrophobicity of the particles recovered in the froth, and on frother concentration. A stable froth allows froth washing to be effective in producing a clean concentrate without having a detrimental effect on combustibles recovery.

Jameson Cells are installed with either wash-water ring or tray systems. Design considerations include the hole size, distribution of the holes to ensure water is spread across the entire surface of the flotation cell, placement of the system at an appropriate distance above the cell lip, and most importantly, the robustness and ease with which it can be maintained as the quality and cleanliness of the water used for froth washing in most plants is often poor which can lead to frequent blockages.

PROJECT ECONOMICS

To be of value for any project the fundamental characteristics of a flotation machine must translate into actual economical benefits. Common measures are capital, operating and maintenance costs. Capital costs must not only include the flotation machine, but also auxiliary equipment (such as pumps, blowers and compressors), structure, piping, civil works and electrical items. Table 1 lists summarises the fundamental characteristics and cost components associated with different flotation technologies.

Table 1: Fundamental characteristics and cost components of different flotation technologies.

Component	Jameson Cell	Column Cell	Conventional Cell
Fundamental			
Bubble Size	0.3 – 0.7 mm	2 – 3 mm	1 – 2 mm
Carrying capacity	High	Low	Low
Mixing intensity	High	Low	Medium

Capital Costs

Equipment	Single cells. Numbers of cells as required to suit tonnage and flows	Single cells. Numbers of cells as required to suit tonnage and flows.	Installed in banks. Number of banks as required to suit tonnage and flows.
Structure & Civil Works	Cell sits on a steel structure at a high level to allow concentrate and tails to gravity flow. Small volume tanks. Integrated into plant.	Due to big volume and weights, need large amount of concrete and foundations for structural support. Need to sit outside plant due to large heights (10 -17 m).	Requires large footprint. Integrated into plant.
Auxiliary Equipment	Feed pump.	Air Compressors and/or recirculation pumps. May need feed pump.	Agitators and Blowers. May need feed pump.

Operating Costs

Power	Feed pump.	Air compressors and recirculating pumps. Feed pump.	Agitators, rotors and blowers. Feed pump.
Reagents	Same across different technology		

Maintenance Costs

Labour	Low. No moving parts.	Medium/high. No moving parts but spargers and compressors to maintain.	High. Rotors, stators and motors to maintain.
Wear part/spares replacements	Highest wear part (slurry lens orifice) lasts 10 years in operation.	Variable. Spargers may need replacing 6 months to 5 years depending on the column technology used.	Variable. Rotors and stators may need replacing 2 - 4 years depending on the make and model of cell.

Jameson Cells are generally very cost-competitive in most cost components as the only auxiliary equipment required is a feed pump so the tie-in to a plant is simple and power use is low. The cell has no moving parts making it easy to operate and with minimal spares requirement it is also economical to maintain.

JAMESON CELL PERFORMANCE

The Jameson Cell has become a standard in the Australian coal industry as it has been proven to be able to consistently produce low ash concentrates and achieve high combustibles recovery whilst being tolerant of commonly encountered variations in feed type and quality variations (Harbort, Manlapig & Jackson, 1992; Atkinson, Conway & Jameson, 1993; Manlapig et al., 1993). There are over a hundred Jameson Cells installed in the coking coal region of the Bowen Basin in Central Queensland (Figure 4) and also in the Hunter Valley region of New South, where thermal coal preparation plants are more dominant.

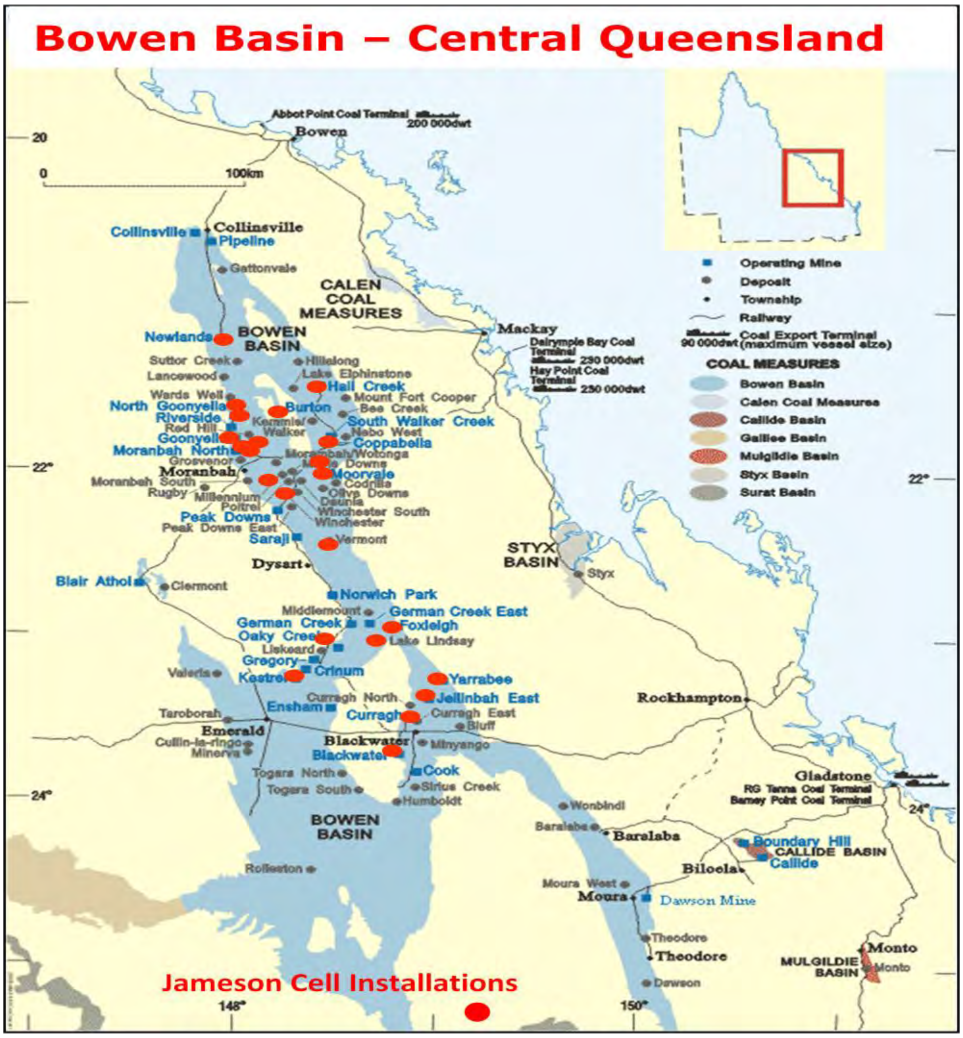


Figure 4: Jameson Cell installations in the Bowen Basin region of Central Queensland, Australia.

Jameson Cells are designed to be able to operate across the entire ash/yield curve for any coal type treated as demonstrated in Figure 5. The solid line is a coal characterisation test, such as the „tree flotation test‟ (ISO 8858-2) widely used in Australia, or the „release analysis‟ (ISO 8858-1) which is perhaps more commonly used in other regions of the world. The data points shown in Figure 4 come from operating shift samples collected from a plant over a six month period. The points overlap the characterisation curve indicating the Jameson Cell is producing a concentrate with the desired or expected, ash content. The large variations in combustible recovery are due to operation of the cell. The low combustibles range is probably due to insufficient reagent additions but clearly, high combustibles recovery, at the knee of the curve, is possible if the process is properly optimised.

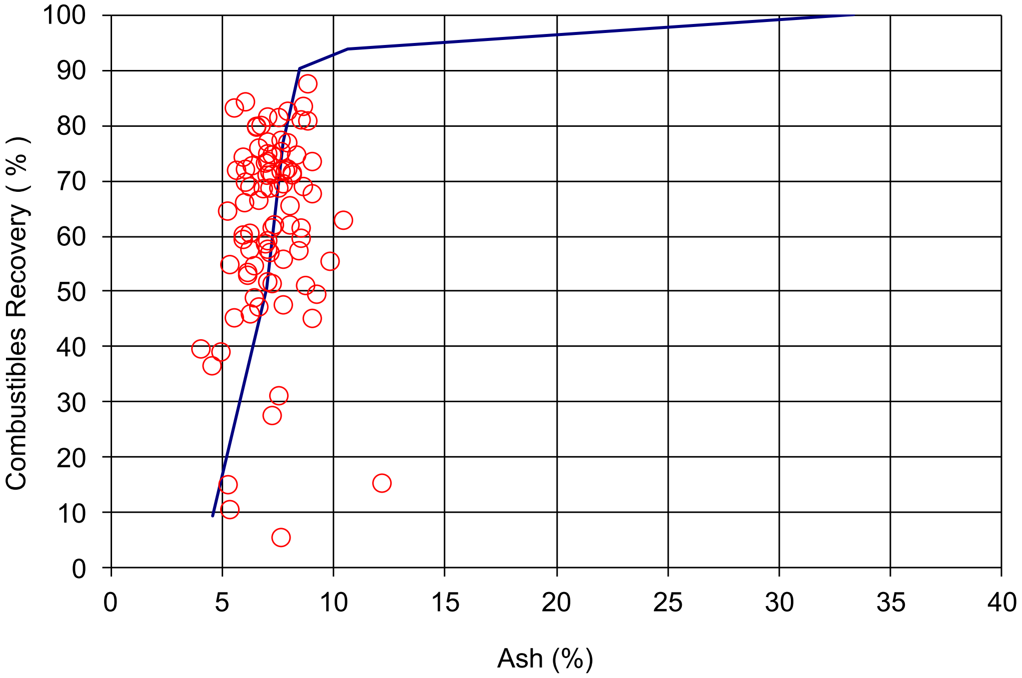


Figure 5: Jameson Cell performance from an operating plant showing its ability to operate across the entire ash-yield curve.

It is important for coal preparation plant operators to be trained to understand that flotation needs to be treated differently to gravity concentration processes because it is based on surface properties of coal and non-coal particles, so reagents are essential to the process. Collectors are necessary to render the coal particles hydrophobic. Too little and some coal particles will not float and too much will result in reduced selectivity leading to a concentrate with a higher than desired ash content. Frothers have a completely different function. They are required to prevent the bubbles generated from a flotation machines from coalescing. Regardless of the technology used, frothers should be used at dosages above the critical concentration of coalescence (CCC), as this is the minimum concentration required to prevent coalescence (Finch, Nesset & Acuna, 2008).

The graph in Figure 6 shows bubble size as a function of frother concentration for the commonly used MIBC (methyl isobutyl carbinol) where the CCC is about 15 ppm. Whilst the CCC is a property of the frother, the actual minimum bubble size generated is dependent on the machine as shown in Figure 3. In operation, the use of reagents should always be investigated ahead of process parameters such as air flowrate and froth depth as these variables are secondary to controlling performance compared to reagents. In addition, froth washing should only be used if target concentrate quality cannot be achieved. It is not mandatory to the flotation process itself but simply an additional tool, albeit an extremely effective one, that can be utilised purely for controlling concentrate grade.

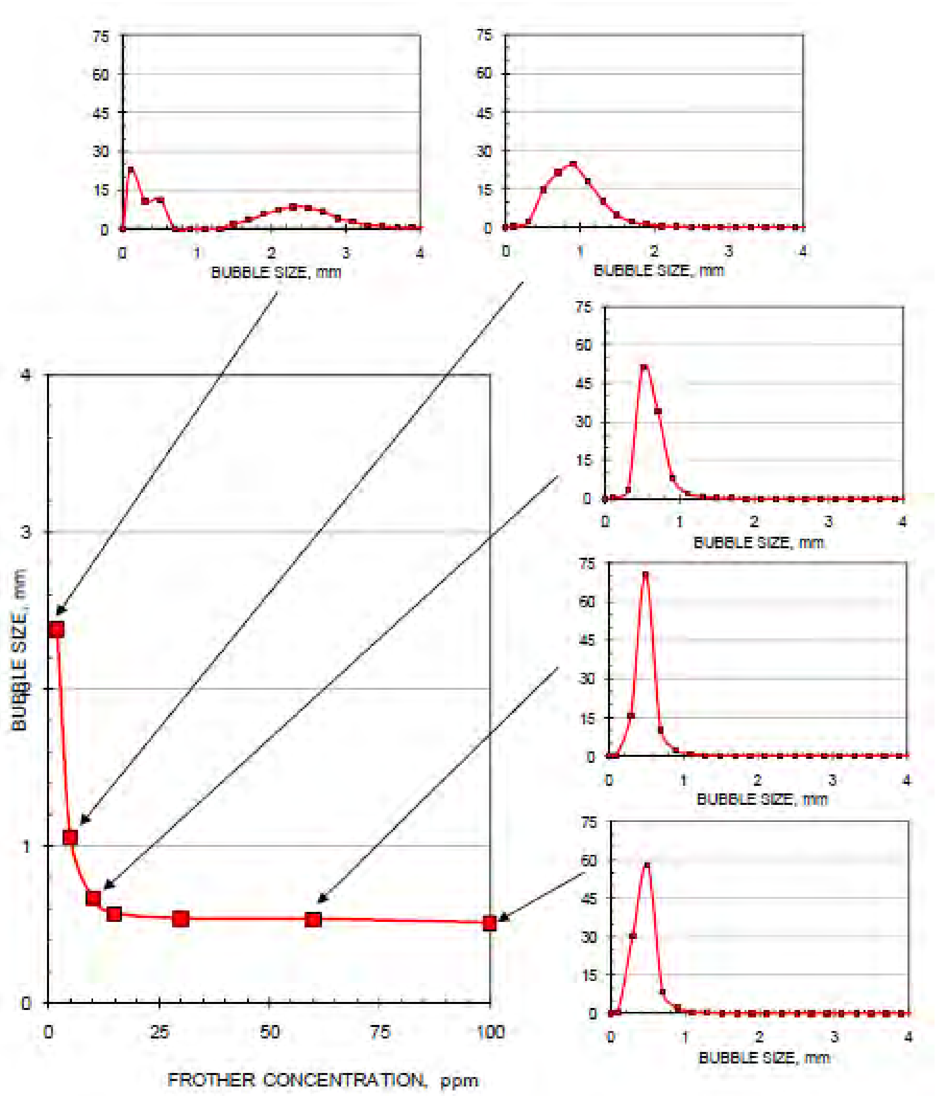


Figure 6: Bubble size as a function of frother concentration (Zou, 2010).

When a Jameson Cell is optimised with respect to the three facets shown in Figure 1, very consistent performance can be produced as shown in Figure 7. The data used in this example are from a plant performance survey carried out at the Goonyella (1,800 tonne/h plant)) coking coal operation in Central Queensland and owned by BHP Billiton Mitsubishi Alliance (BMA) (Caretta, Graham & Dawson, 1997). The graph shows a comparison of the performance of the Jameson Cell with the original mechanical cell circuit it replaced and illustrates that the Jameson Cell can consistently produce a lower ash concentrate at a high combustible recovery. In comparison, the mechanical cells produced a higher ash product and the scatter in the data indicates more inconsistent performance. In this plant, replacement of the 32 mechanical cells with 8 Jameson Cells contributed to an overall plant yield increase of ~3.5% and was reported to have led to production records (Caretta, Graham & Dawson, 1997).

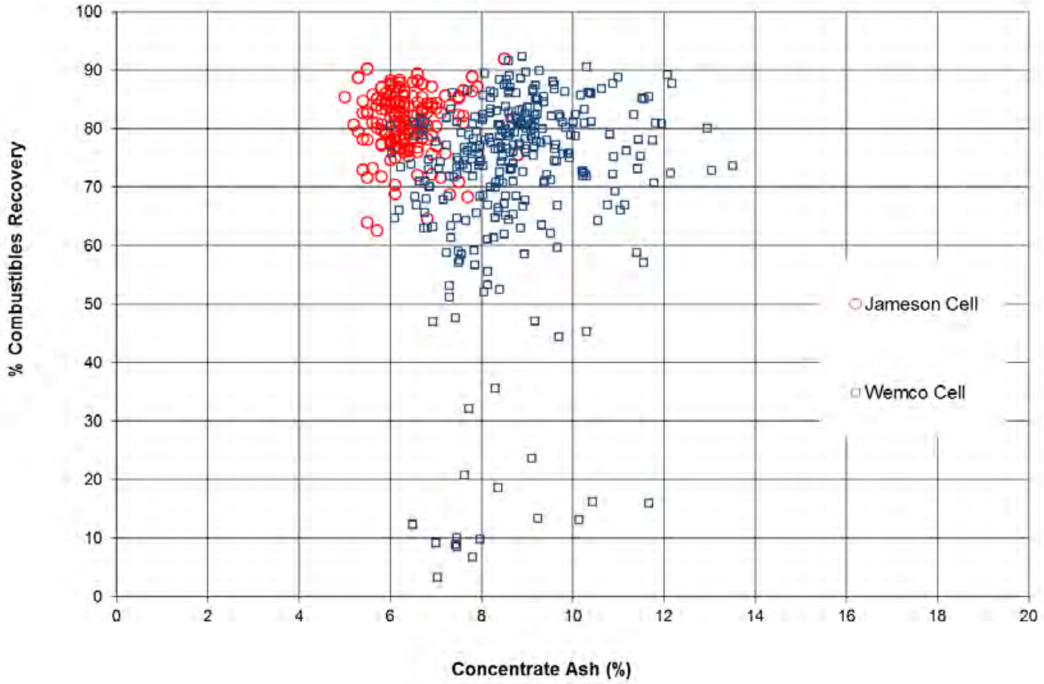


Figure 7: Full scale Jameson Cell performance at BMA's Goonyella mine compared to old mechanical (Wemco) cell circuit.

CHALLENGES IN COAL FLOTATION AND FINE COAL PROCESSING

As mentioned earlier, the flotation machine is only one of three facets important to the overall process, but it tends to be the one that often receives the most attention, so it is often blamed when performance of the fines circuit is poor. However, the variability of the coal feed and flotation reagent control are two factors which are perhaps surprisingly often overlooked.

The greater the number of different coal seams and/or sources that are treated, the more challenging is the task of achieving effective flotation and the targeted qualities and recoveries. Operators must therefore be properly trained to respond to changes in tonnage, particle size distribution and flotation behaviour of the different coal types and make the necessary adjustments to reagent dosages and process variables to optimise performance. Also the provision of adequate sampling facilities and effective controls for reacting to changes that are occurring will make the operation much more capable of achieving expected outcomes.

In many plants, monitoring of flotation performance is irregular and often a „knee-jerk" change is made when performance has clearly deteriorated. Furthermore, in many plants it may be very difficult and sometimes impossible to conduct surveys because sample points do not exist for the feed, concentrate and tailings. Even in the more modern (recently built) plants, the flotation feed often cannot be easily collected as it usually consists of more than one stream which flows by gravity into a large inaccessible collecting sump. In several Jameson Cell installations, operators have often been observed to unknowingly collect the downcomer feed, and use the result from a "rapid ash" analysis together with a two-product formula to calculate yield and combustibles recovery. This is erroneous as the downcomer feed is an internal stream. Sampling points are an essential part of good design practice of the fines circuit.

Even though flotation is the „separation process‟, the overall fines circuit performance is often dictated by concentrate and/or tailings dewatering or treatment capacity as this often proves to be the bottleneck in the process. Many flotation circuits have to be „scaled back‟ to suit the capacity of the dewatering device leading to large losses in coal fines which in turn may then lead to issues in the tailings thickener. Another area to address is the type of dewatering device used for concentrate dewatering. The technology chosen needs to carefully consider the particle size and type of coal treated and not use capital cost as the driver for decisions. Many flotation feeds are becoming much finer in size distribution and therefore the demands in the flotation technology increase. Jameson Cells have been applied effectively to raw coal tailings streams with D_{80} (80% passing) values of as low as 50 microns. Clearly, such challenges can only be met by flotation processes that offer very small bubble sizes.

CONCLUDING REMARKS

The Jameson Cell is a proven, robust and efficient high intensity flotation technology that has been continuously developed and improved over two decades. It is already regarded as a standard in the Australian coal industry and interest is growing in many other coal producing countries. In the coal industry, flotation continues to be a challenge in many coal preparation plants as it does not „fit‟ into what is essentially a gravity separation flowsheet. Operators need to have a sound understanding of the flotation process in order to trouble-shoot and optimise the performance of the whole plant. The importance of reagents to the flotation process has to be realised. Fines circuit operation continues to be an ongoing challenge for coal producers and in too many plants, concentrate dewatering is a bottleneck preventing flotation circuits to be operated to their full potential.

Until such time that all the levers available for the flotation circuit can be fully controlled and measured, the expectancy of adding a desired flotation concentrate component to the product will not be realised. It is therefore essential to focus on flotation as the first priority and ensure that the feed can be controlled and concentrate and tailing treatment are both adequately catered for.

REFERENCES

Ahmed, N. & Jameson, G.J., 1985, The effect of bubble size on the rate of flotation of fine particle‟, *International Journal of Mineral Processing*, vol. **14**, pp. 195-215.

Atkinson, B.W., Conway, C.J. & Jameson, G.J., 1993, Fundamentals of Jameson Cell operation including size-yield response, *6th Australian Coal Preparation Society Conference*, Mackay, Queensland.

Caretta, M.F., Graham, J.N., Dawson, W.J., 1997, Jameson Cell scale-up experiences at BHP Coal‟s Goonyella coal preparation plant, *Coal Prep '97*, Lexington, Kentucky.

Chen, F., Gomez, C.O. & Finch, J.A., 2001, Bubble size measurement in flotation machines, *Minerals Engineering*, vol. **14** (no.4), pp. 427-432.

Diaz-Penafiel, P. & Dobby, G.S., 1994, Kinetic studies in flotation columns: Bubble size effect, *Minerals Engineering*, vol. **7** (no. 4), pp. 465-478.

Evans, G.M., Atkinson, B.W. & Jameson, G.J., 1993, The Jameson Cell, *Flotation Science and Technology*, Marcel Dekker, New York, pp. 331-363.

Finch, J.A., Nesset, J.E. & Acuna, C., 2008, Role of frother on bubble production and behaviour in flotation, *Minerals Engineering*, vol. **21**, pp. 949-957.

Gomez, C.O. & Finch, J.A., 2007, Gas dispersion measurements in flotation machines, *International Journal of Mineral Processing*, vol. **84**, pp. 51-58.

Harbort, G.J., Manlapig, E.V., Jackson, G.J., 1992, Jameson Cell use in fine coal flotation, *ASEAN-Pacific Coal Conference '92*, Cebu, Philippines.

Manlapig, E.V., Jackson, B.R., Harbort, Cheng, C.Y., 1993, Jameson Cell coal flotation, *10th International Coal Preparation Exhibition Conference*, Lexington Kentucky.

Nesset, J.E., Finch, J.A. & Gomez, C.O., 2007, Operating variables affecting bubble size in forced-air mechanical flotation machines, *Proceedings of the Ninth Mill Operators' Conference*, AusIMM, Fremantle, Western Australia, pp. 55-65.

Osborne, D., Huynh, L., Kohli, I., Young, M. & Mercuri, F., 2013, Two decades of Jameson Cell installations in coal, *Proceedings of International Coal Processing Congress*, Istanbul, Turkey, *In submission.*

Zou, J.S., 2010, Effect of frother on the gas dispersion characteristics of Jameson Cells, *McGill University Internal Report*.

Procemin 2013 | 15 - 18 October, Santiago, Chile
procemin.com | procemin@gecamin.com | +56 2 2652 1523

Solving challenges in copper cleaning circuits with the Jameson Cell

Rodrigo Araya*
Xstrata Technology, Canada

Le Huynh and Michael Young
Xstrata Technology, Australia

Karina Arburo
Xstrata Technology, Chile

ABSTRACT

Flotation is a separation process that exploits the hydrophobicity of particles and the buoyancy of air bubbles to recover valuable minerals from liquid-solid suspensions. Flotation is a complex multifaceted process that can be separated into three main areas: ore, chemistry and machine. This paper will focus on factors associated with the machine.

Bubble generation and mixing for bubble-particle contact are fundamental aspects of any flotation machine and different technologies on the market have varying ways of achieving this. The Jameson Cell technology is a robust and efficient high intensity flotation technology with attributes that make it ideal for recovering fast floating liberated mineral particles. It is proven to achieve very high upgrade ratios to produce final grade concentrates in a single stage of flotation.

This papers describes how it can be used to transform traditional flowsheet designs and shows how it has been successfully used to solve capacity and grade issues, associated with non-sulfide gangue and penalty elements, in a number of major copper concentrators around the world. Case studies include operations in Australia (Telfer), Laos (Phu Kham) and Zambia (Lumwana).

*Corresponding author: Process engineer, 10th Floor, 700 West Pender St., Vancouver, British Columbia V6C 1GB, Canada. Phone: +1 604 699 6429. Email: raraya@xstratatech.com

INTRODUCTION

Flotation is a complex multifaceted process that can be separated into three main areas: ore, chemistry and machine, as shown in Figure 1. To solve plant issues it is important to understand how each of the areas and then the different factors within these areas affect and control flotation performance for one's system. Factors within the ore and chemistry areas are dynamic so needs to be dealt with by plant personnel in normal plant operations. However, factors associated with the machine used are generally set as it relates to the fundamental design of the technology. The most important characteristics of any technology is how it generates air bubbles and the size of these air bubbles, as this controls flotation kinetics, and it also dictates the carrying capacity of the machine. The other crucial component is how the machine effects collision and contact between air bubbles and particles.

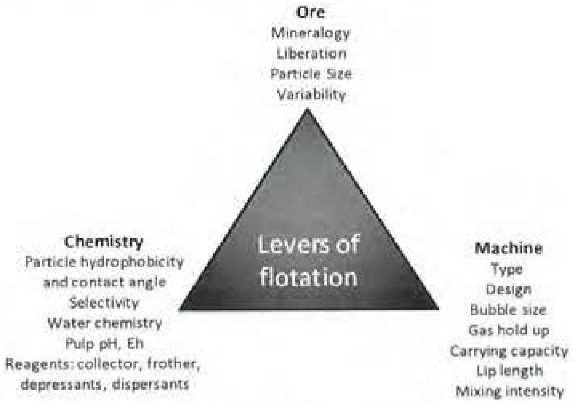


Figure 1 Flotation 'triangle' showing the different factors affecting and controlling performance in all systems

Currently, the majority of the world's copper operations have cleaner circuits use either multi-stages of all mechanical cells or a combination of mechanical cells and flotation columns. The former generally struggles to produce high grade and clean concentrates because of undesirable entrainment of gangue. Upgrade ratios are generally low requiring multiple stages of cleaning to achieve the target grade. The latter circuit suffers from high circulating loads due to the low unit recoveries of columns, exacerbated by spargers (the bubble generation device) that are not properly maintained.

In recent years, the Jameson Cell has been utilised to transform traditional copper cleaner circuits. It is best utilised to recover fast floating liberated mineral particles and is able to produce final grade concentrates in a single stage of flotation (Young et al., 2006). This paper will discuss the fundamental aspects of the Jameson Cell that make it ideal for retrofitting into existing copper cleaner circuits to increase capacity, and for solving issues in plants that are struggling to produce final grade concentrates that have acceptable levels of non-sulfide gangue and/or penalty elements such as uranium, fluorine or mercury.

JAMESON CELL DEVELOPMENT

The Jameson Cell is a high intensity flotation technology jointly developed in the mid-1980s between Mount Isa Mines (MIM, now Xstrata) and Professor Graeme Jameson from the University of Newcastle. What started as a research project to improve the sparger design in the column cleaner cells installed in MIM's zinc circuit culminated into the development of a completely different bubble-generation device called a downcomer, which was originally designed to replace spargers in columns cells. When it discovered that in addition to being a bubble generation device bubble-particle collisions required to effect flotation also occurred inside the downcomer, it became apparent that the large residence time and hence, volumes required for the collection zone in columns was no longer needed. This meant that the downcomers can be placed in much smaller tanks.

JAMESON CELL PRINCIPLE OF OPERATION

The fundamentals of Jameson Cell operation has been described by other authors such as Evans, Atkinson & Jameson (1993). To summarise, in a Jameson Cell feed slurry is first pumped through a restriction (the slurry lens orifice) to create a high pressure jet which then enters into a cylindrical device called a downcomer. The jet of liquid shears then entrains air from the atmosphere. Removal of air into the jet causes a vacuum to be generated inside the downcomer. When a hydraulic seal is formed at the bottom of the downcomer, the vacuum causes a column of slurry to be drawn inside the downcomer. The jet of slurry plunges on the surface of liquid and the high kinetic energy of the jet breaks the entrained air into very fine bubbles. In this zone, the high intensity of the system creates a very favourable environment for the bubbles and particles to collide and attach. The air bubbles and mineral particles move continuously down the downcomer until it exits into the tank. The particle laden bubbles then float to the top to form the froth whilst the hydrophilic mineral particles remain in the pulp phase to be removed as tailings. To ensure consistent operation, tailings recycle is employed. Tailings recycle dampens feed fluctuations to the cell allowing the downcomer to operate at a constant feed pressure and flowrate. The high rate of mixing from the high pressure jet and the fact the air is self aspirated, allows for the Jameson Cell to have no moving parts aside from the feed pump. No agitators or compressors are required.

Due to rapid kinetics and a separate contact zone in the downcomer, the tank is not sized for residence time, so tank volumes are much smaller than equivalent mechanical and column cells.

BUBBLE SIZE

Bubble size is one of the most important factors in any flotation system as it has a strong influence over flotation kinetics. Fine bubbles increase the flotation kinetics across *all* particle sizes (Diaz-Penafiel & Dobby, 1994; Ahmed & Jameson, 1985) and not only improve the recovery of fine particles as often hypothesised. However, fine particles do exhibit slower flotation kinetics (resulting in poorer recovery and selectivity) so the influence of fine bubbles to improve fine particle recovery is, perhaps, more noticeable than for coarse particles which generally floats faster anyway. Fine bubbles improve the separation of minerals as they intensify the difference in the flotation kinetics of the valuable minerals from the gangue minerals, thus allowing higher grade concentrates to be produced. It also increases the carrying capacity (often measured as the tph of concentrate per m^2 of surface area of the flotation machine) as there is more bubble surface area per volume of air added for mineral particles to attach.

The Jameson Cell is able to produce fine bubbles via by the shear action of a plunging jet (Evans, Jameson & Atkinson, 1992). It is this fundamental characteristic that allows the Jameson Cell to float particles quickly, attain superior selectivity and have high productivity (carrying capacity). The latter is particularly beneficial when high mass pulls are required, such as recleaning in metals applications, and in the flotation of metallurgical coal where mass pulls (yield) can exceed 80%.

The air bubbles generated by the Jameson Cell are in the range of 300 to 700 μm (Sauter mean diameter, D_{32}) (Evans, Atkinson & Jameson, 1993). Figure 2 compares the bubble size of a range of industrial mechanical and columns cells (Nesset, Finch & Gomez, 2007; Nesset, 2009) to that of the Jameson Cell (Osborne et al., 2013). All results collated in Figure 2 were determined by the same bubble size measurement technique as developed by McGill University and described by Chen, Gomez & Finch (2001) and Gomez & Finch (2007).

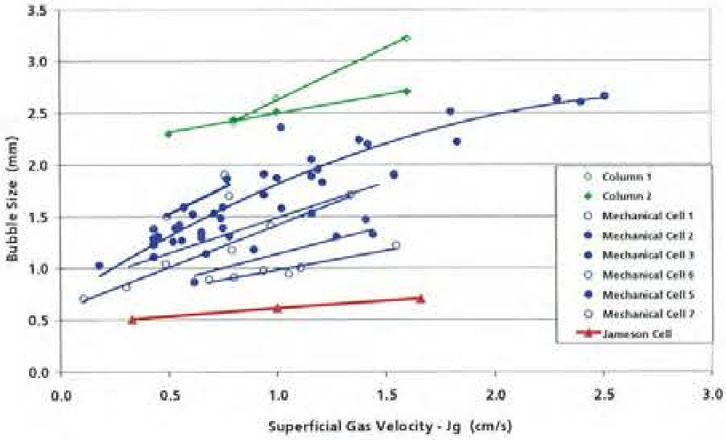


Figure 2 Bubble size as a function of Jg (measure of air flow rate) for different flotation technologies

Gas holdup, ε_g, is defined as the volumetric fraction of gas in the pulp. The gas holdup in a Jameson Cell downcomer is generally in the range of 40 to 60% while gas holdup in the separation zone (tank) is in the 20% to 40% range. In comparison, the gas holdup in conventional machines is found to be much lower and in the range of 10 to 20% (Yianatos et al., 2001; Grau & Heiskanen, 2003). Jameson Cells typically operate between 0.5 to 2 cm/s J_g (superficial gas velocity).

FROTH WASHING

The ore bodies mined today are more complex, lower grade and often need to be ground finer to liberate the value mineral particles. It is well known that effective separation of fine particle systems via flotation is far more challenging compared to coarser ones as fine particles exhibit slower flotation kinetics, oxidise more quickly and require more reagents to float due to larger surface areas, and entrainment of non-floating gangue, which typically occurs below 50 microns (Johnson, 2005), is always a problem.

Froth washing is one of the most effective methods employed to reduce entrainment and allow high grade clean concentrates to be produced. Washwater systems must be designed correctly: have the right hole size, provide good distribution of the water across the entire surface of the flotation cell, be placed at an appropriate distance above the cell lip, and most importantly be robust and easy to maintain as the quality and cleanliness of the water used for froth washing in most plants are poor leading to frequent blockages.

An example of the effect of froth washing on gangue rejection and hence, selectivity, is shown in Figure 3. Data is taken from Jameson Cell pilot plant trials at a site where the cleaner feed stream has a D_{80} of 20-25 μm. The decrease in the recovery of non-sulfide gangue (NSG) with addition of wash water is dramatic and results in a completely different selectivity curve. For any flotation system, the effect of washwater is highly dependent on the properties and structure of the froth. The froth must be stable if the water added is to penetrate through the froth to remove entrained gangue particles to attain a high grade product, whilst minimising its effect on value particles attached to the air bubbles affecting recovery.

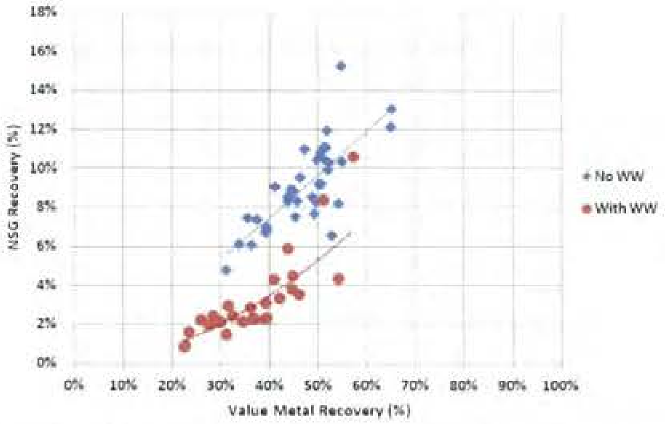


Figure 3 Demonstration of a flotation system in the absence and presence of washwater (WW) to minimise the entrainment of fine non-sulphide gangue (NSG)

RECENT INSTALLATIONS AND CASE STUDIES

The ability of the Jameson Cell to produce high grade clean concentrates have seen it successfully utilised as a cleaner scalper cell (at the head of existing cleaner circuits) in more and more operations. The recovery achieved in a single cell is dependent on the quantity of liberated material in the feed but typically ranges from 40% to 90%. Final grade concentrates are dependent on copper mineralisation in the feed and ranges from 28-32% Cu for chalcopyrite systems to 45-60% Cu for ores containing secondary copper minerals (chalcocite, bornite). Three specific case studies are discussed in this paper.

Case Study 1: Phu Kham

Phu Kham is a 12 Mtpa copper-gold concentrator owned by PanAust and is located approximately 100 km east of the capitol of Laos. Plant feed grades are 0.75% Cu, 0.33 g/t Au and 3.8 g/t Ag. The process plant comprises of a SAG and ball mill grinding circuit followed by rougher flotation. Rougher concentrate is reground using a M10,000 IsaMill and three stages of conventional cleaning is used to achieve a final concentrate grade of 25 % Cu, 7 g/t Au and 60 g/t Ag as described by Bennett, Crnkovic & Walker (2012).

Soon after the plant was commissioned in 2008, a debottlenecking project was initiated to look at ways to increase cleaner capacity as the recovery across this circuit was seen to drop off significantly when the cleaner feed exceeded 150 tph. The study concluded that the best option was to install additional cleaner capacity as the head of the cleaner circuit. A B6500/24 model Jameson Cell was chosen for this duty and installed in March 2011. The simple tie-in of the Jameson Cell to the rest of the plant meant minimal disruption to normal production and commissioning only required one week. The Jameson Cell produces a high grade concentrate typically around 26-28% Cu at 50-60% recovery and has allowed the overall cleaner circuit at

Phu Kham to maintain a consistently high recovery when the cleaner feed tonnage increases from 150 to 300 tph. Tests conducted with the Jameson Cell online and offline in February 2012 showed that the benefit of having the Jameson Cell was an overall plant increase of 0.8% copper recovery (Bennett, Crnkovic & Walker, 2012).

Case Study 2: Telfer

Telfer is a 21 Mtpa copper-gold concentrator owned by Newcrest and located in the South Western area of the Great Sandy Desert in Western Australia. As described by Seaman, Manton & Griffin, 2011, ore is mined from open pit and underground sources which have significant variations from mainly chalcocite in the open pit to chalcopyrite in the underground ore. The concentrator is set up so that these two different ores are treated in two parallel trains. Train 1 treating a blend of open pit and underground ore and Train 2 which only treats open pit ore. The ore is processed in different configurations. One of the most common configurations is sequential flotation where copper concentrate is produced first, then re-activation and flotation of pyrite followed by cyanide leaching to recover gold. Typical head grades are 0.1 to 0.2 % Cu, 1 to 2 g/t Au, 1 to 5 % Sulphur. Gold is present as free gold and also contained within copper minerals and pyrite.

Review of historical data showed that a large proportion of the liberated, non-sulfide non-floating gangue was being recovered to the copper concentrate by entrainment (Seaman et al., 2012). Dilution batch tests were performed to assess the NSG rejection potential of plant cleaner feed streams. Modelling of different cleaner circuit configurations predicted that with Jameson Cells (the chosen technology and whose selectivity performance was simulated by the dilution batch tests) installed at the head of the existing cleaner circuit copper concentrate grades could be increased to 20-25% Cu (from 16-18% Cu) by rejecting half the NSG that is currently reporting the to the concentrate. Removing the load off the circuit allows operation of the existing cleaner cells at lower wt% solids and addition of water to control pulp density, to allow better cleaning in the mechanical cells. With mineralogical circuit analysis, dilution cleaning tests and modelling outputs, pilot plant testing were considered unnecessary and capital expenditure was approved to install one Jameson Cell per train (Seaman et al. 2012).

Two E3432/8 model Jameson Cells were ordered in May 2011 and the project was fast-tracked to allow commissioning in November 2011. Performance of the cells matched the results from dilute batch flotation tests carried out during the project development phase. The Jameson Cells produces copper concentrate with 20 to 25 % Cu and is able to maintain these high grades up to 80% recovery, which exceeded the design of 50% copper recovery. Recovery by size data shows the Jameson Cell can achieve good recovery of copper (and gold) across all size fractions and has good gangue rejection down to very fine fractions below 10 microns. The payback for the project was found to be between two and seven months (Seaman et al. 2012).

Case Study 3: Lumwana

Lumwana is a 20 Mtpa copper concentrator owned by Barrick and located in Zambia's Mwinilunga District, 65 km from the provincial capital of Solwezi in the North Western Province of Zambia and approximately 660 km from the country's capital, Lusaka. Lumwana has two major copper deposits: Malundwe and Chimiwungo. Copper mineralization includes chalcopyrite, bornite and chalcocite. In addition to copper, these deposits also contains the penalty element, uranium (U) contained in vein hosted and disseminated uranite. The processing plant consists of SAG and ball milling for primary grinding. The flotation plant consists of two parallel train of rougher/scavenger followed by a single train of two stages of conventional cleaning. A ball mill is used for rougher concentrate regrind.

Soon after commissioning in early 2009, it was discovered the installed cleaner circuit could not produce final grade copper concentrates with uranium levels below acceptable limits. Studies showed the fine liberated U particles were recovered into the froth by entrainment. Based on the successful application of the Jameson Cell to reject fluorine, another penalty element, at Prominent Hill (Barns, Colbert & Munroe, 2009), Lumwana operations trialled the Jameson Cell on various cleaner streams at their site. Several weeks of continuous pilot plant testing (in November 2009) using a L500 Jameson Cell rig clearly demonstrated that the majority of U particles in the cleaner feed stream can be rejected allowing final grade copper concentrates with acceptable U levels to be consistently produced. Figure 4 shows the resulting copper/uranium selectivity curve produced from these tests. After piloting, a full scale B5400/18 model Jameson Cell was installed in a cleaner scalper duty in the plant with commissioning occurring in July 2011. Addition of the Jameson Cell has allowed Lumwana operations to produce a clean and saleable copper concentrate that has U levels below acceptable limits. This is particularly impressive given the varying nature of the copper mineralogy and uranium content of the plant feed. Although designed for a cleaner scalper duty, operational issues with the recleaner cells at Lumwana has forced reconfiguration of the cleaner circuit and the duty of the Jameson Cell altered to recleaning. The plant is still producing final clean concentrates below target U levels. A project is underway to optimise the cleaner capacity at Lumwana (Araya et al., 2013).

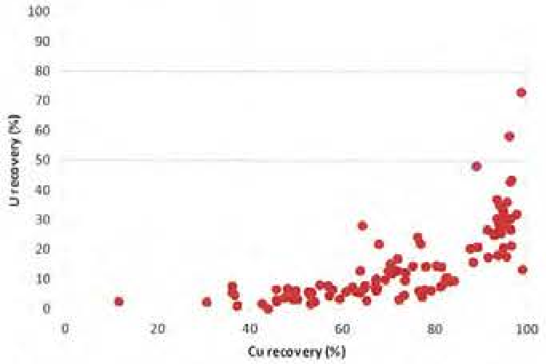


Figure 4 Selectivity of copper uranium recovery determined from Jameson Cell pilot testwork at Lumwana

CLEANER CIRCUIT DESIGN

The attributes that makes the Jameson Cell ideal for a cleaner scalper duty also make it ideal for final cleaning. An optimum cleaner circuit is shown in Figure 5. This design would allow higher plant final grade concentrates to be produced, whilst using lesser number of cells and in a smaller footprint, compared to an all conventional cell circuit. There are benefits in capital (less equipment), operational (less power) and maintenance (labour and spare parts). In fact, this circuit design with Jameson Cells has been operating at Northparkes (New South Wales, Australia) since 1994.

In addition to superior grade, a major advantage of using a Jameson Cell as a recleaner instead of conventional cells is that it may allow operations to diagnose 'online' plant liberation issues. That is, if the Jameson Cell cannot, in this duty, produce decent grade concentrates with froth washing to eliminate or minimise entrainment, then the value mineral particles recovered in this stream must not be sufficiently liberated.

Perhaps one could argue that other flotation technologies can be used instead of the Jameson Cell in the circuit proposed in Figure 5. Indeed this is possible, however it's important to match the attributes of different technologies for the different duties. Fast flotation kinetic and high carrying capacity afforded by the Jameson Cell are the main advantages, but practical considerations such as ability to control flotation performance, very fast response to process changes, whether they are deliberate or part of normal plant fluctuations, and rapid start-up and shut-down, may be just as important. Add to this its simplicity to maintain, it makes the Jameson Cell the best option.

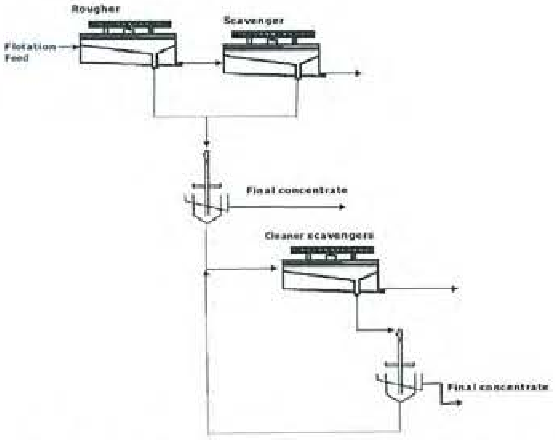


Figure 5 Ideal cleaner flowsheet using Jameson Cells in cleaner scalper and recleaner duties to produce final concentrate and mechanical cells only as cleaner scavenger

CONCLUDING REMARKS

The Jameson Cell has been used to successfully solve cleaner circuit issues in several major copper concentrators around the world. It is ideal for inclusion into existing circuits to increase cleaner capacity. Its small footprint, simple tie-in to existing structures and straight forward commissioning requirements means it can be easily incorporated into plants with minimal disruption to operations. It has found applications in existing operations where entrainment of gangue and/or penalty elements is an issue. The Jameson Cell produces clean concentrate in a single stage of flotation, through superior flotation selectivity and by applying froth washing.

The Jameson Cells are best utilised for cleaner scalper and recleaner duties. The ideal cleaner circuit design would see Jameson Cells used in these duties to produce all the final concentrate from a plant and mechanical cells used for cleaner scavengers only. These cleaner circuits would allow plants to produce the highest grades possible and achieve this using lesser number of cells and in a smaller footprint, compared to traditional designs.

ACKNOWLEDGEMENT

The author's would like to acknowledge PanAust, Newcrest and Barrick for sharing information about their Jameson Cell projects with Xstrata Technology and permission to publish this paper.

REFERENCES

Ahmed, N. & Jameson, G.J. (1985) 'The effect of bubble size on the rate of flotation of fine particles', *International Journal of Mineral Processing*, vol. 14, pp. 195-215.

Araya, R., Mwanza, A., Cordingley, G. & Huynh, L. (2013) 'Necessity driving change and improvement to the cleaner circuit at Lumwana Concentrator', *In preparation.*

Barns, K.E., Colbert, P.J. & Munro, P.D. (2009) 'Designing the optimal flotation circuit – The Prominent Hill case', *Proceedings of the Tenth Mill Operators' Conference*, AusIMM, Adelaide, South Australia, pp. 173-182.

Bennett, D., Crnkovic, I. & Walker, P. (2012) 'Recent process developments at the Phu Kham copper-gold concentrator, Laos', *Proceedings of the 11th Mill Operators' Conference*, AusIMM, Hobart, Tasmania, pp. 257-272.

Chen, F., Gomez, C.O. & Finch, J.A. (2001) 'Bubble size measurement in flotation machines', *Minerals Engineering*, vol. 14, no. 4, pp. 427-432.

Diaz-Penafiel, P. & Dobby, G.S. (1994) 'Kinetic studies in flotation columns: Bubble size effect', *Minerals Engineering*, vol. 7, no. 4, pp. 465-478.

Evans, G.M., Atkinson, B.W. & Jameson, G.J. (1993) 'The Jameson Cell', *Flotation Science and Technology*, Marcel Dekker, New York, pp. 331-363.

Evans, G.M., Jameson, G.J. & Atkinson, B.W. (1992) 'Prediction of the bubble size generated by a plunging liquid jet bubble column', Chemical Engineering Science, vol. 47, no. 13/14, pp. 3265-3272.

Gomez, C.O. & Finch, J.A. (2007) 'Gas dispersion measurements in flotation machines', *International Journal of Mineral Processing*, vol. 84, pp. 51-58.

Grau, R.A. & Heiskanen, K. (2003) 'Gas dispersion measurement in a flotation cell', *Minerals Engineering*, vol. 16, pp. 1081-1089.

Johnson , N.W. (2005) 'A review of the entrainment mechanism and its modelling in industrial flotation processes', *Proceedings of the Centenary of Flotation Symposium*, AusIMM, Brisbane, Queensland, pp. 487-496.

Nesset, J.E., Finch, J.A. & Gomez, C.O. (2007) 'Operating variables affecting bubble size in forced-air mechanical flotation machines', *Proceedings of the Ninth Mill Operators' Conference*, AusIMM, Fremantle, Western Australia, pp. 55-65.

Nesset, J.E. (2009) 'Flotation machine diagnostics', *Mineral Processing Systems Course*, Montreal, Canada.

Osborne, D., Huynh, L., Kohli, I., Young, M. & Mercuri, F. (2013) 'Two decades of Jameson Cell installations in coal', *Proceedings of International Coal Processing Congress*, Istanbul, Turkey, *In submission.*

Seaman, D.R., Burns, F., Adamson, B., Seaman, B.A. & Manton, P. (2012) 'Telfer processing plant upgrade – The implementation of additional cleaning capacity and the regrinding of copper and pyrite concentrates', *Proceedings of the 11th Mill Operators' Conference*, AusIMM, Hobart, Tasmania, pp. 373-381.

Seaman, D., Manton, P. & Griffin, P. (2011) 'Separation efficiency improvement of a low grade copper-gold flotation circuit', *Proceedings of Procemin 2011*, Gecamin, Santiago, pp. 263-272.

Yianatos, J., Bergh, L., Condori, P. & Aguilera, J. (2001) 'Hydrodynamic and Metallurgical Characterization of Industrial Flotation Banks for Control Purposes', *Minerals Engineering*, vol. 14 , pp. 1033-1046.

Young, M.F., Barnes, K.E., Anderson, G.S. & Pease, J.D. (2006) 'Jameson Cell: The Comeback in Base Metals Applications Using Improved Designed Flow Sheets', *Proceedings of the 38th Annual Meeting of the Canadian Mineral Processors*, Ottawa, Ontario, pp. 311-332 .

Опыт двух десятилетий эксплуатации флотомашин Джеймсон в угольной промышленности

Дейв Осборн[1], Ле Хьюн[1], Исхант Коли[1], Майкл Янг[1] и Фрэнк Меркьюри[2]

1 Xstrata Technology, Брисбен, Квинсленд, Австралия

2 Xstrata Coal, Брисбен, Квинсленд, Австралия

Автореферат

Флотомашина Джеймсон была впервые представлена в конце 1980-х годов как решение конструктивных и эксплуатационных недостатков, присущих колонным и традиционным флотомашинам. Начиная с первого опыта промышленной флотации полиметаллических руд на руднике Маунт Айза в 1989 году, за которым последовал первый опыт применения на углеобогатительной фабрике Ньюлэндс в 1990 году, флотомашина непрерывно совершенствовалась, повышая производительность и становясь более простой в эксплуатации. В настоящее время в разных отраслях и для различных применений в эксплуатации находится более 320 флотомашин этого типа, из которых примерно 45% применяется в угольной промышленности. Новейшие модели сочетают в себе изначально присущие этой машине высокую производительность и компактность со значительно минимизированными эксплуатационными затратами и удобством эксплуатации. В настоящей работе рассматривается развитие флотомашины Джеймсон на протяжении последних двух десятилетий и причины перехода с двухстадийной схемы на одностадийную, а затем обратно на двухстадийную. Хотя большинство флотомашин Джеймсон используется в настоящее время для коксующегося угля, они все чаще применяются для переработки ультратонкого энергетического угля, в том числе с высоким содержанием в рядовом угле золы, преимущественно глины. В настоящей работе также описано последнее пилотное тестирование и полученные результаты, которые можно использовать для проекта промышленного масштаба.

Ключевые слова: флотомашина Джеймсон, уголь, флотация, вспениватель

1. Введение

Флотомашина Джеймсон основана на абсолютно другой технологии флотации, нежели механические и колонные флотомашины, и была впервые предложена профессором Грэмом Джеймсоном, Университет Ньюкасл, шт. Новый Южный Уэльс. Флотомашина Джеймсон была впервые испытана и применена в промышленном масштабе на углеобогатительной фабрике разреза Ньюлэндс компании Xstrata Coal (Jameson et al., 1991). Потоком питания тонких классов был слив гидроциклона, который ранее переходил в отходы (крупность минус 20-25 микрон при зольности от 15 до 50%). Пилотное тестирование подтвердило возможность добиться извлечения горючей массы свыше 90% при нормативной зольности продукта 10%. В результате в 1990 году были установлены промышленные флотомашины Джеймсон первого поколения (Mark I Jameson Cells). Эти машины находились в непрерывной эксплуатации в течение более чем 15 лет, пока в 2006 году не была построена новая углеобогатительная фабрика взамен старой. На новой фабрике также применяются флотомашины Джеймсон, а именно четыре новых машины B6000/20 модели Mark III. После первоначальной установки на фабрике Ньюлэндс технология была испытана на многих предприятиях, и было доказано, что эта технология способна обеспечить стабильное получение продукта низкой зольности при высоком извлечении, а также нетребовательна к зольности питания (Harbort et al., 1992; Atkinson et al., 1993; Manlapig et al., 1993). Обогатительная фабрика коксующегося угля Гуньелла компании BHP Coal (в настоящее время - BHP Billiton Mitsubishi Alliance (BMA)) производительностью 1 800 т/ч, расположенная в центральной части шт. Квинсленд, провела тестирование современных технологий флотации и по их результатам полностью заменила 32 механические флотомашины (Wemco) восемью флотомашинами Джеймсон, установленными в две стадии (Caretta et al., 1997). На рис. 1 сопоставлены показатели эффективности флотомашин Джеймсон на этом предприятии после пусконаладки с показателями эффективности старых механических флотомашин. Стабильное получение продукта низкой зольности при высоком извлечении во флотомашине Джеймсон позволило повысить выход по фабрике примерно на 3,5%, что, в свою очередь, позволило поставить новые производственные рекорды.

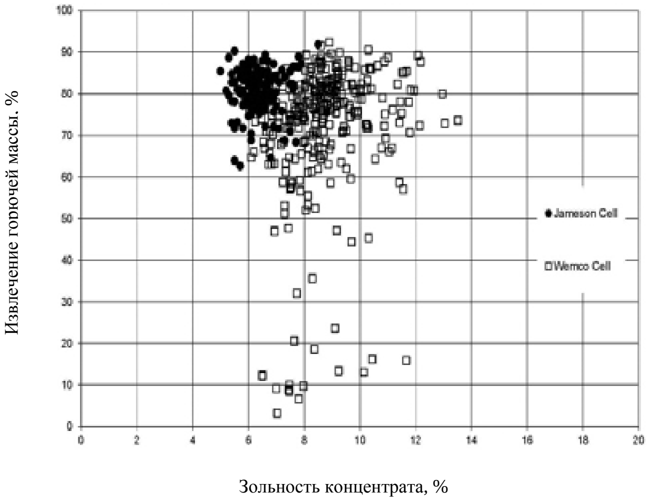


Рисунок 1. Показатели эффективности флотомашин Джеймсон на фабрике Гуньелла относительно показателей старых механических флотомашин (Wemco)

Среди основных преимуществ новой технологии - возможность промывки пены и простота флотомашины Джеймсон, удобство эксплуатации и обслуживания, отсутствие движущихся частей, отсутствие потребности во вспомогательном оборудовании (кроме насоса питания). В настоящее время по всему миру в эксплуатации находится более 150 флотомашин Джеймсон. Крупнейшей является установка на фабрике Керрэг компании Wesfarmer в бассейне Боуэн в центральной части шт. Квинсленд, перерабатывающая более 5 млн. тонн угольных шламов в год всего на двенадцати флотомашинах. Страны с давними традициями добычи угля, в частности, Казахстан, также используют преимущества, предлагаемые флотомашиной Джеймсон по сравнению с традиционными флотомашинами, а новые угледобывающие страны, такие как Мозамбик и Монголия, начинают применять флотомашины Джеймсон для обогащения коксующегося угля.

Далее в этой работе будет рассмотрен процесс развития технологии, лежащей в основе флотомашины Джеймсон, на протяжении последних двух десятилетий, процесс эволюции цикла флотации, проблемы, существующие в переработке угольных шламов, а также в качестве примера рекомендуемого подхода к проектированию будет представлен последний опыт масштабирования производства энергетического угля от пилотной до промышленной установки.

2. Развитие флотомашины Джеймсон

С момента первого промышленного применения в 1989 году флотомашина Джеймсон прошла четыре основных этапа последовательного совершенствования (рис. 2). Результатом двух десятилетий развития этой технологии стала последняя модель флотомашины Джеймсон, Mark IV Jameson Cell, которая имеет следующие характерные особенности:

- Система рециркуляции питания; для обеспечения стабильного режима работы флотомашины и поддержания оптимальной эффективности независимо от колебаний в питании.
- Низкий износ, высокий коэффициент разгрузки, сопло пульпы.
- Гибкая насадка питания упрощает процедуру проверки и позволяет точно отрегулировать падающую струю для максиамльного повышения технологических показателей.
- Улучшенная система промывки над пеной и в пене.

С выходом в 2000 году аэратора Mark III для образования струи вместо диафрагмы используется сопло пульпы (рис. 3). Преимущества этой конструкции заключаются поступлении пульпы под небольшим углом, оптимизирующим поток пульпы и увеличивающим срок службы деталей, высоком коэффициенте разгрузки, улучшенных условиях образования струи с более равномерным вовлечением воздуха и вакуумом.

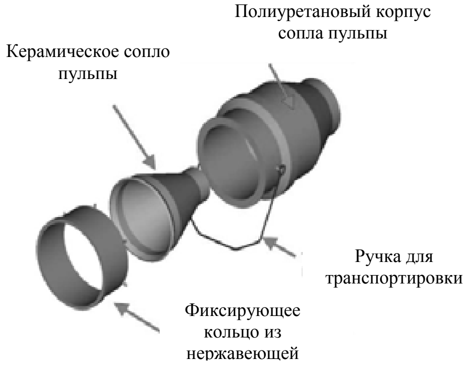


Рисунок 3. Сопло пульпы Mark IV

Доступ к соплу при обслуживании значительно проще по сравнению с прежней конструкцией. Если сопло пульпы закупорено или требует замены, обслуживание занимает всего 10 минут и не требует остановки флотомашины: достаточно изолировать соответствующий аэратор. Износоустойчивость сопла превзошла все ожидания. С момента внедрения в 2000 году на углеобогатительных фабриках установлено 1500 единиц, но ни одно из них до сих пор не потребовало замены по причине износа.

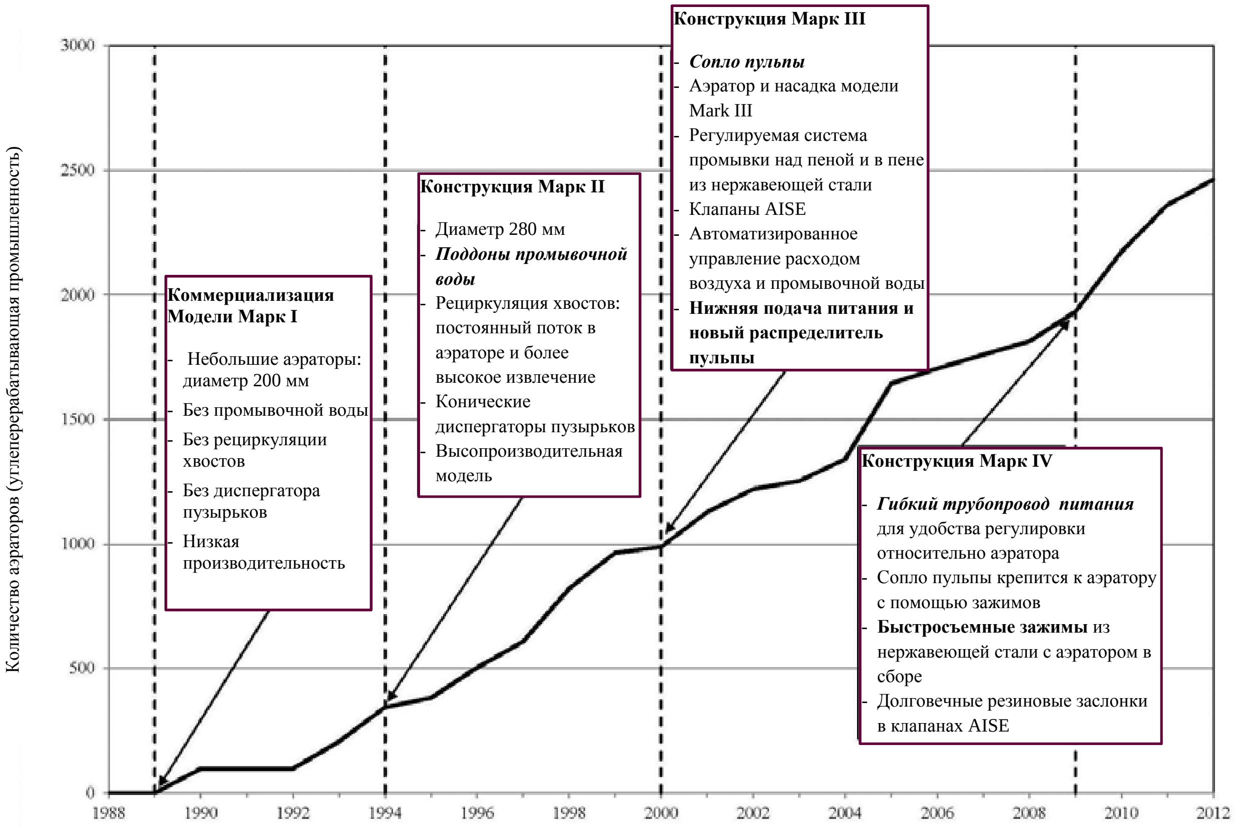


Рисунок 2. Развитие флотомашины Джеймсон

3 Развитие технологических схем с флотомашинами Джеймсон

Первое поколение флотомашин Джеймсон на австралийских углеобогатительных фабриках в начале 1990-х годов устанавливалось исключительно в две стадии: первичная и вторичная флотомашины производили раздельные концентраты, которые затем объединялись для получения готового продукта. Эта схема работала достаточно хорошо, но на эффективность флотации часто влияли колебания в питании цикла флотации. Следствием непостоянного давления питания и колебаний в объемном расходе питания стало , нестабильное вовлечение воздуха и меняющиеся гидродинамические условия внутри аэратора. Проблему удалось решить в середине 1990-х годов, когда в цикле была применена рециркуляция хвостов. Хотя изначально она предназначалась для компенсации колебаний питания, было установлено, что за счет увеличения рециркуляции хвостов до 40-50% и уменьшения подачи свежего питания, повысилось извлечение во флотомашине, благодаря чему появилась возможность получить в установке, состоящей из одной флотомашины Джеймсон, уровень извлечения горючей массы, сравнимый с извлечением в некоторых двухстадийных схемах (рис. 4). Поэтому такие одностадийные машины, как правило, спроектированы из расчета меньшей подачи свежего питания и рециркуляции хвостов на уровне около 40-50%. Для коксующегося угля количество одностадийных флотомашин, необходимое для получения высокого выхода (до 85-90%) при требуемом уровне качества, должно обеспечивать достаточную несущую способность (площадь поверхности пузырьков). Для аналогичной двухстадийной схемы потребовалось бы такое же количество флотомашин, что и для одностадийной, но машины должны были бы быть установлены последовательно, а не параллельно, и эксплуатировались бы при значительно более низкой рециркуляции хвостов (10-20%). В такой схеме рециркуляция хвостов используется исключительно для сглаживания колебаний в подаче свежего питания. Очевидное преимущество одностадийной схемы флотации заключается в меньших капитальных и эксплуатационных затратах, поскольку она может быть реализована с минимальным числом емкостей и насосов питания. Так, для двух флотомашин, установленных параллельно, требуется одна емкость и один насос питания, в то время как для аналогичной двухстадийной схемы потребуется две емкости и два насоса (меньшего размера). Однако объем экономии уменьшается по мере расширения схемы, поскольку в двухстадийных схемах с более чем двумя флотомашинами могут применяться общие емкости и более мощные насосы, поскольку и на первой, и на второй стадии питание может быть общим для двух и более флотомашин.

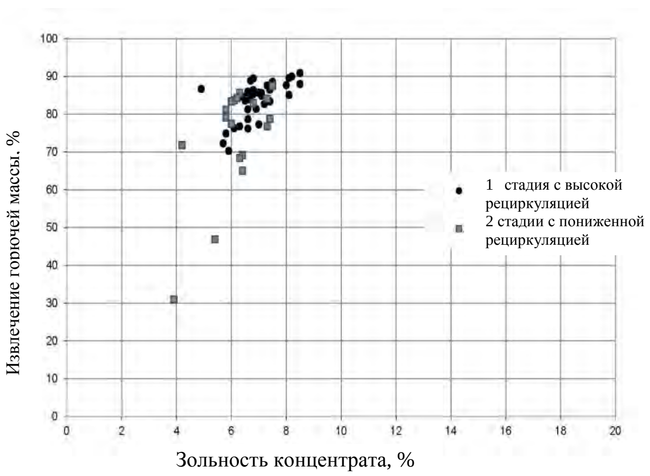


Рисунок 4. Эффективность флотации во флотомашине Джеймсон, установленной в одну или две стадии

Начиная с 1995 года, все флотомашины Джеймсон устанавливаются в одну стадию с рециркуляцией хвостов на уровне 40-50%. На протяжении длительного времени многие предприятия сталкиваются с проблемами пенообразования, вызванными остаточным вспенивателем в рециркулируемом сливе сгустителя хвостов. Эта проблема сложна, по-разному проявляется на разных предприятиях и нередко требует индивидуальных решений. Чувствительность всей фабрики к остаточному вспенивателю также зависит от быстроты повторного использования воды, проектирования водного цикла и сочетания реагентов, применяемых на фабрике. Среди наиболее распространенных проблем - неустойчивость тяжелосредного цикла из-за вовлечения воздуха, недостаточное обезвоживание вследствие сложностей управления пенообразованием и проблемы с осветлением воды.

Последствия уменьшения дозировки вспенивателя, подаваемого на флотацию в одностадийной схеме, достаточно серьезны, как показано на рис. 5. На данной фабрике по производству коксующегося угля извлечение горючей массы составляло всего 35-45%, т.е. значительно ниже перегиба кривой, построенной по результатам стандартного характеризационного теста для угля (черная линия) при концентрации вспенивателя 9 ппм. Расход воздуха и промывочной воды определяли глубину пены, но не оказывали существенного влияния на извлечение. Однако когда дозировку вспенивателя увеличили до 14 ппм, извлечение горючей массы выросло до 75-82%, достигнув максимума на перегибе кривой.

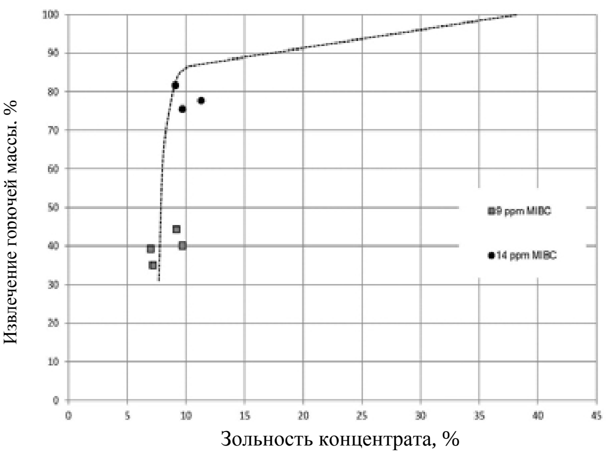


Рисунок 5. Влияние дозировки вспенивателя на эффективность флотации

Резкое изменение показателя извлечения горючей массы, показанное на рис. 5, можно объяснить следующим образом. В одностадийных схемах с использованием как флотомашины Джеймсон, так и колонной флотомашины, несущая способность флотомашины полностью зависит от способности машины обеспечить площадь поверхности пузырьков, достаточную для флотации всего угля. Во флотомашине любого типа максимальная площадь поверхности пузырьков возможна только при условии введения достаточного количества вспенивателя, предотвращающего коагуляцию. Минимальная концентрация, необходимая для предотвращения коагуляции, известна как критическая концентрация коагуляции (ККК) и зависит от типа вспенивателя. На рис. 6 показана зависимость между размером пузырька и дозировкой MIBC, вспенивателя, наиболее часто используемого во флотации угля. ККК вспенивателя MIBC составляет примерно 15 ппм.

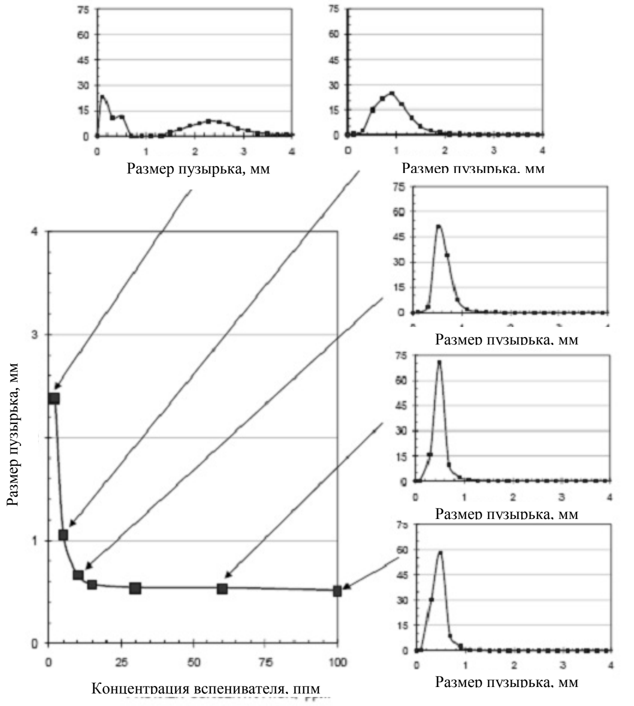


Рисунок 6. Размер пузырька в зависимости от концентрации вспенивателя (Zou, 2010)

Так как вспениватель определяет концентрацию, его нужно использовать в процессе флотации; минимальный размер пузырька же зависит от флотомашины. На рис. 7 показан размер пузырька во флотомашинах различного типа (Джеймсон, механическая или колонная) в зависимости от поверхностной скорости газового потока J_g (показатель расхода воздуха). Небольшой размер пузырьков во флотомашине Джеймсон (300-600 микрон) объясняет, почему эта машина имеет более быструю кинетику флотации и более высокую несущую способность, а следовательно и производительность, по сравнению с флотомашинами других типов.

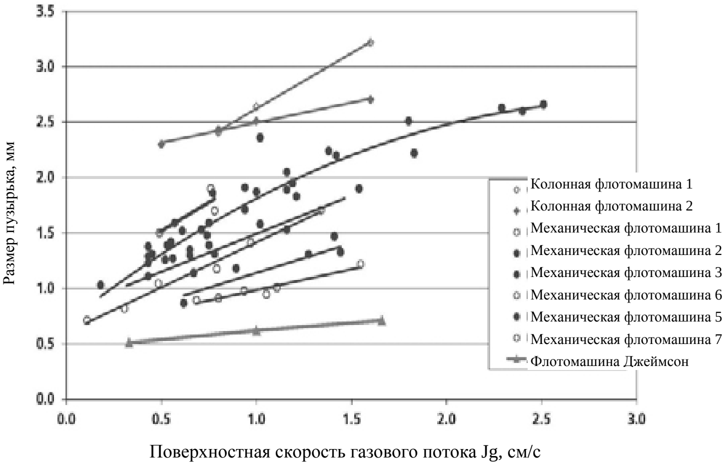


Рисунок 7. Размер пузырька в зависимости от J_g (показатель расхода воздуха) в различных технологиях флотации (Nesset et al., 2007; Gomez, 2012, Zou, 2010)

Проектирование углеобогатительной фабрики с интеграцией отдельных водных циклов для крупных классов и шламов не представляется целесообразным решением проблем пенообразования. Вероятно, единственное практически осуществимое решение заключается в реорганизации цикла флотации и возврате к двухстадийной схеме. Двухстадийная схема может повысить производительность системы, поскольку первичная и вторичная стадии могут эксплуатироваться отдельно с целью оптимизации эффективности. Увеличение дозировки вспенивателя может повысить эффективность первичной флотомашины, а остаточный вспениватель в хвостах первичной машины может быть использован для флотации дополнительного объема угля во вторичной флотомашине. Предприятие Уоркуорт компании RTCA[1] (Новый Южный Уэльс, Австралия) перевело свои флотомашины Джеймсон на двухстадийную схему и сообщает об увеличении выхода и о возможности использования более высокой концентрации вспенивателя без возникновения избыточного пенообразования (Lambert and Revell, 2008). На новых проектах, таких как углеобогатительная фабрика Мударбен компании Yancoal (Новый Южный Уэльс) и Кестрел компании RTCA (Квинсленд), также используется двухстадийная схема флотации.

Обогатительные фабрики, работающие по одностадийной схеме, могут быть переоборудованы в двухстадийные. На фабрике Джерман Крик компании AAMC[2] (Квинсленд) флотомашина Джеймсон была недавно установлена в качестве машины вторичной стадии для переработки хвостов двух флотомашин Microcel, установленных параллельно (рис. 8). Присутствовало ограничение в установочной площади, но проблем с монтажом не возникло, т.к. флотомашина Джеймсон компактна и проста, благодаря легкости присоединений к трубопроводам и отсутствию необходимости во вспомогательном оборудовании (кроме насоса питания).

---

[1] RTCS - Rio Tinto Coal Australia

[2] AAMC - Anglo American Metallurgical Coal

Рисунок 8. Недавно установленная флотомашина Джеймсон (слева) и две колонные флотомашины на предприятии Джерман Крик компании AAMC

4. Проблемы флотации угля и переработки угольных шламов

Флотация является физико-химическим процессом разделения и отличается от других используемых на углеобогатительной фабрике процессов, в которых разделение происходит за счет гравитации. Флотация представляет собой трехфазную систему, зависящую от многих факторов, которые можно условно разделить на три группы: характеристики угля, химические факторы и характеристики оборудования (рис. 9).

Рисунок 9. Треугольник флотации, изображающий факторы, влияющие на эффективность

Флотомашина - это только один из факторов, определяющих весь процесс, однако она привлекает наибольшее внимание и первой становится предметом подозрений, когда эффективность цикла флотации мелких классов падает. Однако удивительно, что при этом часто забывают о таких факторах, как изменчивость характеристик угля и характеристика подачи флотореагентов. Чем больше число пластов и источников, из которых уголь поступает на переработку, тем сложнее добиться эффективной флотации, желаемого качества и извлечения. Поэтому операторы должны уметь реагировать на изменения в тоннаже, гранулометрии и поведении углей различных типов при флотации и вносить необходимые корректировки в дозировку реагентов и переменные технологические параметры для оптимизации эффективности.

На многих обогатительных фабриках контроль эффективности флотации нерегулярен, а меры принимаются с опозданием, когда эффективность уже заметно снизилась. Более того, на многих фабриках контроль вообще невозможен, потому что точки отбора проб питания, концентрата и хвостов не предусмотрены. Даже на более современных (недавно построенных) фабриках получить пробу питания непросто, потому что, как правило, питание состоит из нескольких технологических потоков, поступающих самотеком в общий приемник большого объема. На многих фабриках, где установлены флотомашины Джеймсон, персонал, не подозревая об этом, отбирает пробы питания аэратора и использует результаты экспресс-анализа на зольность и двухпродуктную формулу для расчета выхода и извлечения горючей массы. Но это неверно, т.к. питание аэратора является внутренним потоком. Поэтому точки отбора проб являются необходимым элементом правильного проектирования секции обогащения мелких классов.

Хотя флотация является процессом разделения, общая эффективность секции обогащения мелких классов нередко определяется возможностями обезвоживания концентрата, что часто становится узким местом процесса. Зачастую производительность цикла флотации приходится уменьшать, чтобы согласовать ее с производительностью обезвоживающей установки, в результате чего возникают крупные потери мелкой угольной фракции, влекущие, в свою очередь, проблемы в работе сгустителя хвостов. Другим вопросом, требующим внимания, является выбор типа обезвоживающей установки для концентрата. При выборе технологии в качестве определяющего фактора необходимо учитывать гранулометрию и тип перерабатываемого угля, а не капитальные затраты.

5. Пилотная установка на фабрике Балга

Флотация традиционно используется в качестве технологии обогащения при переработке мелких классов угля, хотя в прошлом она преимущественно использовалась для коксующегося угля. Тем не менее, в ЮАР эксплуатируется ряд флотационных установок, предназначенных для извлечения коммерческого энергетического угля из рядовых хвостов, которые в противном случае были бы направлены в отходы. Результаты эксплуатации этих установок неоднозначны, поскольку уголь этого региона плохо поддается флотации, однако они одновременно указывают на необходимость выбора эффективных циклов обезвоживания для переработки угольных концентратов (Power, 2010).

Xstrata Technology совместно с Xstrata Coal занимаются разработкой технологической схемы с использованием флотомашины Джеймсон для переработки энергетического угля на фабрике Балга у поселка Броук, Новый Южный Уэльс. Фабрика будет перерабатывать необработанные рядовые хвосты и хвосты из соседнего хвостохранилища, содержащие, по оценкам, 3 млн. т извлекаемого угля.

Фабрика перерабатывает уголь из 12 пластов, а также уголь, поступающий из открытых разрезов и шахт, с зольностью питания флотации от 30 до 65%. При использовании любой технологии непросто получить продукт стабильного качества при столь большом разбросе характеристик питания.

Пилотная флотационная установка с использованием флотомашин Джеймсон была запущена в эксплуатацию на площадке для сбора проектной информации, необходимой для расширения фабрики. Испытания пилотной флотационной установки показали стабильно высокое извлечение из питания линии обогащения энергетического угля: до 70-90% горючей массы при зольности продукта 8-12%. При переработке более окисленного угля потребовалось увеличить дозировку коллектора для получения достаточно высоких показателей извлечения.

Пилотная установка состояла из двух флотомашин Джеймсон L500, установленных последовательно с целью смоделировать промышленный цикл флотации, который будет установлен в будущем. Предлагаемая схема промышленной установки с использованием флотомашин Джеймсон показана ниже на рис. 10. Эта компактная схема позволяет разместить флотационную установку в условиях расширения существующего предприятия. Были рассмотрены различные варианты обезвоживания и других способов переработки, включая применение центрифуги и установки Centribaric с последующим брикетированием продукта более высокой стоимости. Это оборудование будет также испытано в составе пилотной установки, а после завершения работ данный цикл флотации будет спроектирован как стандартный для будущих углеобогатительных фабрик Xstrata Coal, производящих как коксующийся, так и энергетический уголь.

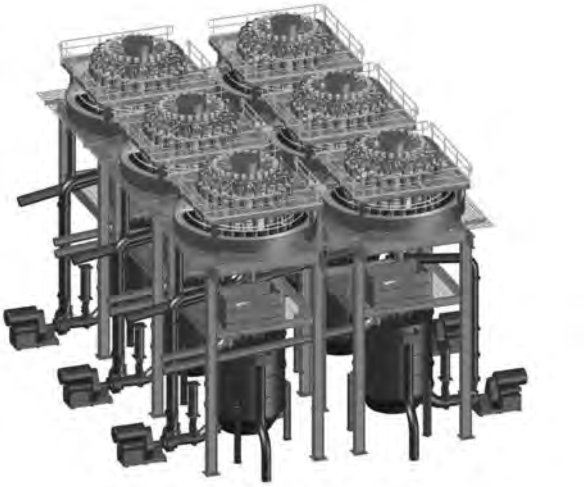


Рисунок 10. Схема установки шести флотомашин Джеймсон модели B6500/24 (занимаемая площадь 22 x 26 м; высота 17 м)

6. Заключение

За немногим более десятилетия флотомашина Джеймсон стала стандартной технологией флотации в австралийской угольной промышленности. Более 110 флотомашин, находящихся в эксплуатации, представляют собой важный источник информации об эксплуатации, обслуживании и совершенствовании установок в будущем. Флотомашина Джеймсон все чаще применяется в углеперерабатывающей промышленности других крупных угледобывающих стран, включая такие развивающиеся страны, как Монголия и Мозамбик. Основные преимущества этой машины, такие как простота, высокая эксплуатационная готовность, нетребовательность к обслуживанию, а также доказанная производительность и стабильная эффективность, давно признаны эксплуатирующими компаниями.

Своим развитием флотомашина Джеймсон в значительной мере обязана совершенствованию материалов, применяемых в ее компонентах, однако другие факторы также оказывают влияние на общую эффективность флотации. Некоторые из этих факторов, например обезвоживание, отбор проб, управление технологическим процессом и другие, также стали предметом внимания. Новые установки необходимо проектировать с учетом последних достижений в каждой из этих областей, в особенности, если операторам могут быть предоставлены надежные инструменты контроля и управления, позволяющие корректировать условия эксплуатации и на ранних этапах выявлять отклонения от требований к качеству и извлечению.

Xstrata Technology (XT) создала сильный коллектив проектировщиков, инженеров и технологов и благодаря этому последовательно совершенствует технологию флотации Джеймсон на протяжении более чем двух десятилетий, поддерживая тесную связь с ее изобретателем и его коллегами. Подобные отношения являются редкостью в технологическом проектировании, а кроме того, ХТ не просто поставляет оборудование, а строит технологические партнерства с заказчиками, среди которых не только углеперерабатывающие предприятия, но и технологи, и проектировщики, занятые в отрасли цветных металлов, промышленных минералов и в других областях применения. В такой уникальной ситуации все источники информации и все пользователи технологии вовлечены в обмен знаниями и опытом. Вне всякого сомнения, результатом развития технологии в будущем станут новые модели и технологические циклы, обладающие всеми требуемыми характеристиками, такими как анализ в режиме реального времени, отбор проб из пульпы, датчики реагентов и т.д.

7. Список использованной литературы

1. Atkinson, B.W., Conway, C.J., Jameson, G.J. Fundamentals of Jameson Cell operation including size-yield response, 6th Australian Coal Preparation Society Conference, Mackay QLD, September, 1993.
2. Caretta, M.F., Graham, J.N., Dawson, W.J., Jameson Cell scale-up experiences at BHP Coal’s Goonyella coal preparation plant, Coal Prep ’97, Lexington, Kentucky, April, 1997.
3. Gomez, C.O., Personal communication, 2012.
4. Harbort, G.J., Manlapig, E.V., Jackson, G.J., Jameson Cell use in fine coal flotation, ASEAN-Pacific Coal Conference ’92, Cebu, Philippines, 1992.
5. Jameson, G.J., Goffinet, M., Hughes, D., Operating experiences with Jameson Cell at Newland Coal Pty Ltd, 5th Australian Coal Preparation Society Conference, Newcastle NSW, May, 1991.
6. Lambert, L. and Revell, P., 2008. Tripling Flotation Yield at Warkworth CHPP, 12th Australian Coal Preparation Society Conference, Darling Harbour NSW, October, 2008.
7. Manlapig, E.V., Jackson, B.R., Harbort, Cheng, C.Y., Jameson Cell coal flotation, Coal Prep ’93, 10th International Coal Preparation Exhibition Conference, Lexington Kentucky, May, 1993.
8. Nesset, J.E., Finch, J.A., Gomez, C.O., Operating variables affecting bubble size in forced-air mechanical flotation machines, Ninth Mill Operator’s Conference, Fremantle WA, 2007.
9. Power, D., The new highs and lows of coal processing for South African coals, 13th Australian Coal Preparation Society Conference, Mackay QLD, September, 2010.
10. Swanepoel, C., Fine coal processing developments in Anglo American Thermal Coal in South Africa, Challenges in fine coal processing, dewatering and disposal, SME, edited by Klima, M.S., Arnold, B.J., Bethell, P.J., 2012.
11. Zou, J.S., Effect of frother on the gas dispersion characteristics of Jameson Cells, McGill University Internal Report, 2010.

Two decades of Jameson Cell installations in coal

Dave Osborne[1], Le Huynh[1], Ishant Kohli[1], Michael Young[1] and Frank Mercuri[2]

1 Xstrata Technology, Brisbane, QLD, Australia

2 Xstrata Coal, Brisbane, QLD, Australia

ABSTRACT

Jameson Cell technology was first introduced in the late 1980s to overcome the design and operating inadequacies of column and conventional flotation cells. From the first commercial base metals installation at Mt Isa in 1989 followed by the first coal installation at Newlands in 1990, it has been continuously developed and improved to make it more robust and easier to use. There are now over 320 units installed in a wide range of applications and industries of which about 45% are employed for coal. The latest designs combine the original advantages of high productivity and small footprint with significantly reduced maintenance costs and much more operator-friendly features. This paper will describe the development of the Jameson Cell technology over two decades and drivers for changing circuit configurations from 2-stage to single stage and then back to 2-stage over the years. Although most Jameson Cell installations are currently utilised for recovery of metallurgical coal, there are a growing number of applications emerging for treating ultrafine thermal coal, many with high raw coal ash contents, predominantly clay. The paper will also describe recent pilot-scale test work and the results obtained from this to support the commercial scale project.

Keywords: Jameson Cell, Coal, Flotation, Frother

1. Introduction

The Jameson Cell is a fundamentally different flotation technology to mechanical and column cells, having been invented by Prof Graeme Jameson at the University of Newcastle in NSW. They were first tested and commercially installed in a coal washing plant at Xstrata Coal's Newlands mine (Jameson et al., 1991). The fines stream was cyclone overflow material which was previously discarded (minus 20-25 microns in particle size with 15 to 50% ash content). Pilot plant testing showed it was possible to achieve greater than 90% combustibles recovery with a product target of 10% ash. This led to the installation of the first generation full scale, Mark I Jameson Cells in 1990. These cells were in continuous operation for over 15 years until a new washing plant was built to replace the old plant in 2006. The new plant also uses Jameson Cells and has four new B6000/20 model Mark III cells installed. Following the initial Newlands installation, many sites have tested the technology which was shown to consistently produce low ash concentrates and achieve high combustibles recovery whilst being forgiving to variations in feed ash (Harbort et al., 1992; Atkinson et al., 1993; Manlapig et al., 1993). BHP Coal's (now BHP Billiton Mitsubishi Alliance - BMA) Goonyella 1,800 tph coking coal operation in Central Queensland tested the advanced flotation technologies and subsequently replaced the entire 32 mechanical (Wemco) cell circuit with 8 Jameson Cells operating in a 2-stage configuration (Caretta et al., 1997). Figure 1 compares the performance of the Jameson Cells at this plant after commissioning to the old mechanical cell circuit. The ability of the Jameson Cell to consistently deliver a low ash product at high combustibles recoveries contributed to an overall plant yield increase of ~3.5% and led to production records.

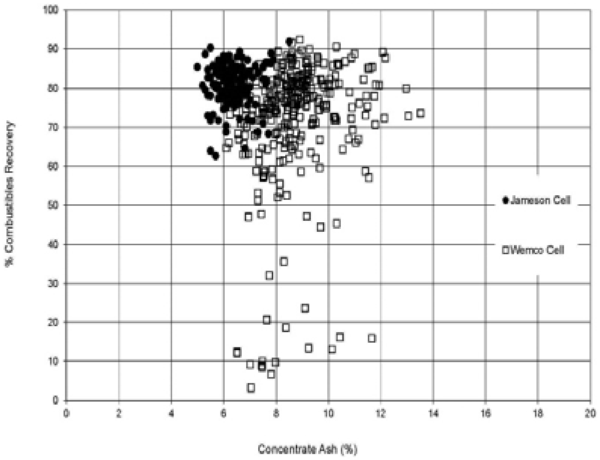


Figure 1: Full scale Jameson Cell performance at Goonyella mine compared to old mechanical (Wemco) cell circuit

Amongst the key benefits is froth washing and the simplicity afforded by the Jameson Cell, it being easy to operate and maintain; with no moving parts, and needing no auxiliary equipment except for the feed pump. Over 150 Jameson Cells are now operational on coal worldwide, the current largest installation being at Wesfarmer's Curragh Mine in the Bowen Basin of Central Queensland which treats over 5 million tonnes of coal fines per year using only twelve cells. Long-established coal producing countries like Kazakhstan are also realising the benefits of the Jameson

Cell over conventional cells and emerging coal regions such as Mozambique and Mongolia are now beginning to use the Jameson Cell for metallurgical coal applications.

The remainder of this paper will describe the development of the Jameson Cell technology over two decades, the evolution of flotation circuit designs, continued challenges of fine coal processing in the coal industry; and a recent case study for the development of a thermal coal project from pilot plant to full scale is included to illustrate the recommended design approach.

2. Jameson Cell Development

The Jameson Cell development path has seen four main phases of progressive improvement (Figure 2) since its first commercial installation in 1989. Two decades of significant advances have culminated in the latest model, the Mark IV Jameson Cell, which incorporates the following features:

- Feed recycle system; to ensure stable cell operation maintaining optimum performance independent of feed fluctuations.

- Low wear, high discharge coefficient, slurry lens orifice.
- Flexible feed nozzle allowing quick and easy inspection and precise alignment of the plunging jet to maximise metallurgical performance.
- Improved above-froth or in-froth washwater system.

With the release of the Mark III downcomer in 2000, the orifice plates used for jet formation were replaced with the slurry lens orifice (Figure 3). The benefits included a shallow slurry entry angle which optimised slurry flow and maximised component wear life; a high discharge coefficient, and improved slurry jet formation resulting in more consistent air entrainment and vacuum.

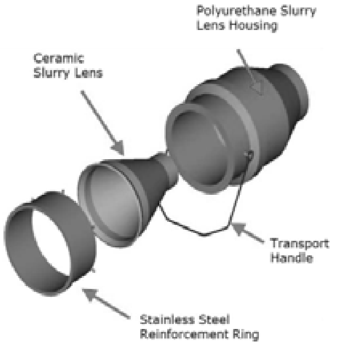


Figure 3: Mark IV Slurry Lens

For maintenance, the slurry lens is significantly easier to access compared to the old design. If a slurry lens becomes blocked or needs replacing, it is now a 10 minute job which can be done online by isolating the specific downcomer while the rest of cell (and other downcomers) remain in operation. The wear performance of the slurry lens orifice has far exceeded expectations. Since its introduction in 2000, 1,500 slurry lenses have been installed in coal applications and no site has so far needed to replace one of these items due to wear.

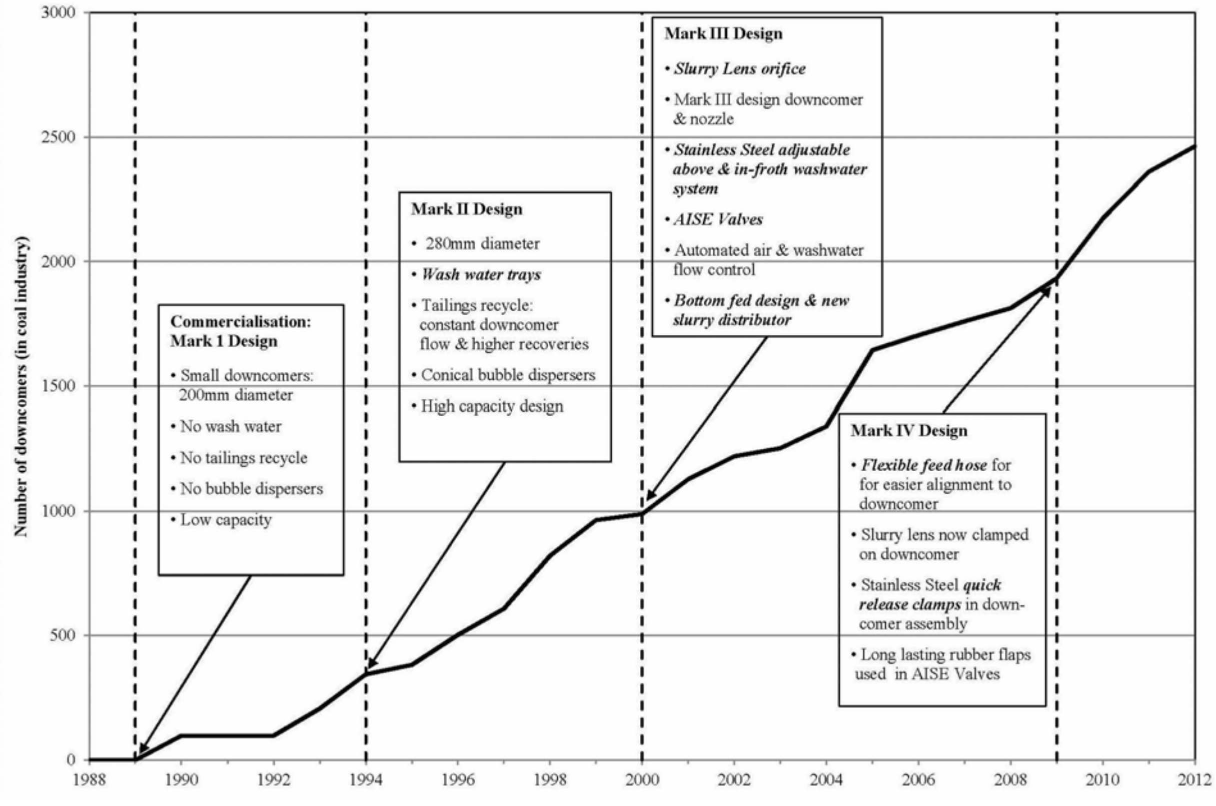


Figure 2: Jameson Cell Development Path

3 Evolution of Jameson Cell Circuit Designs

The first generation of Jameson Cell coal plant installations in Australia in the early 1990s were all 2-stage designs with primary and secondary cells producing separate concentrates that were combined to give the final product. These circuits generally operated well but flotation performance was often affected by wide variations in feed volumetric flow rate to the flotation circuit. This was usually because fluctuation in the flow rate and feed pressure resulted in inconsistent air entrainment and hydrodynamic mixing inside the downcomer. This was overcome in the mid-1990s with the incorporation of tailings recycle into the circuit design. Although originally designed to dampen feed flow fluctuations, by employing higher tailings recycle of between 40-50% and reducing the fresh feed flow rate, it was observed that cell recovery was improved which enabled a single Jameson Cell to achieve a similar level of combustibles recovery as some 2-stage circuits (Figure 4). These single stage cells are therefore typically designed to operate with a lower fresh feed flow rate and higher tailings recycle of about 40-50%. In metallurgical coal applications the number of single-stage cells required to achieve high yields (up to 85-90%) at the required quality must ensure adequate carrying capacity (bubble surface area). An equivalent 2-stage circuit would use the same number of cells as the single stage arrangement but the cells will be installed in series rather than in parallel and will operate at much lower tailings recycle (10-20%). In this type of circuit, the purpose of the tailings recycle is to simply dampen fresh feed fluctuations. The obvious attraction of the single stage flotation circuit is the reduced capital and operating cost as it can be designed using minimal feed sumps and pumps. For example, for two cells in a parallel circuit, only a single feed sump and a single feed pump are required, whereas for an equivalent 2-stage circuit, 2 sumps and 2 (smaller) pumps are required. However, the level of saving is diminished as the size of the circuit increases because 2-stage circuits with more than 2 cells can also make use of common sumps and larger feed pumps as the feed for each stage primary and secondary stage can also be split between two or more cells.

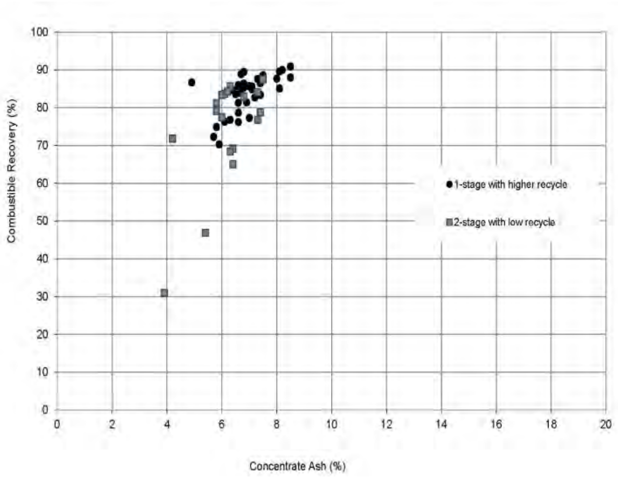


Figure 4: Flotation performance of single stage versus 2-stage Jameson Cell circuit.

From 1995 onwards, all Jameson Cell installations were single stage circuits with 40-50% tailings recycle. Over the years, many operations have experienced frothing issues in the plant caused by residual frother in the recirculated tailings thickener overflow water. This issue is complex and the effects experienced at each site are varied and often require individual solutions. The sensitivity of the overall plant to residual frother is also dependent on how quickly water is reused, the design of the water circuit and the reagent combinations used throughout the plant. Common problems experienced include instability in dense medium circuits through air entrainment, dewatering limitation due to froth handling constraints and water clarification issues.

However, the impact of lowering frother dosage to the flotation recovery in single stage circuits is severe as demonstrated in Figure 5. At this metallurgical coal operation, combustibles recovery was poor at 35-45% and well below the knee of the curve generated from a standard coal characterisation test (black line) when the frother concentration was 9 ppm. Altering froth depth, air flow rate and washwater flow rate had negligible impact on the recovery. However, when the frother dosage was increased to 14 ppm, the combustibles recovery dramatically increased to 75 to 82%, reaching the maximum at the knee of the curve.

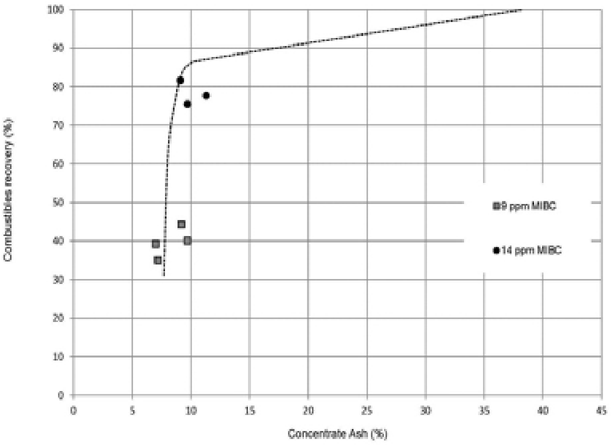


Figure 5: Effect of frother dosage on flotation performance

The reason for the dramatic change in combustibles recovery shown in Figure 5 can be explained as follows. In single stage circuits, whether a Jameson Cell or a column cell is used, the carrying capacity of the cell is entirely dependent on that cell being able to produce sufficient bubble surface area required to float all the coal. For any type of flotation machine, maximum bubble surface area can only be achieved if sufficient frother is added to prevent coalescence. The minimum concentration required for this is called the Critical Concentration of Coalescence (CCC) and is characteristic for each type of frother. Figure 6 shows the change in bubble size with frother dosage for MIBC, the most commonly used frother in coal flotation. The CCC for MIBC is around 15 ppm.

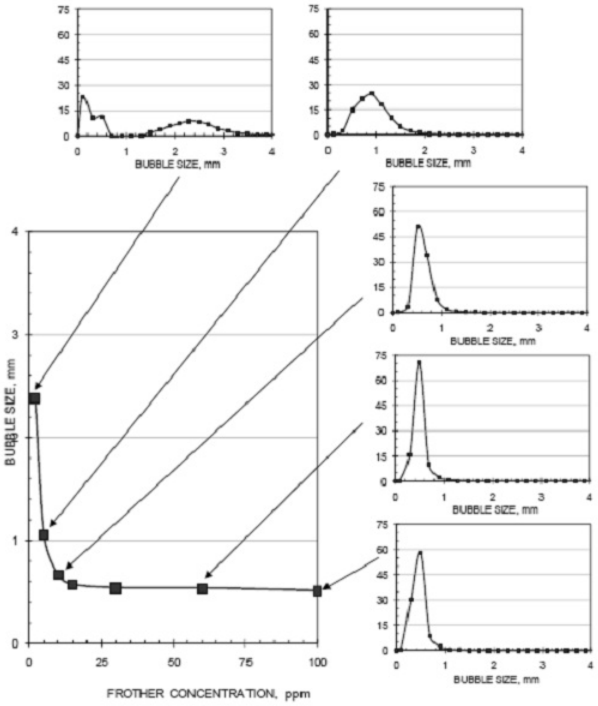


Figure 6: Bubble size as a function of frother concentration (Zou, 2010)

While the frother dictates the concentration it must be used in the flotation process; it is the flotation machine that controls the minimum bubble that can be generated. Figure 7 compares the bubble size of different technologies: Jameson, mechanical and column cells, are included as a function of the superficial gas velocity, J_g (a measure of air flow rate). The fine bubbles generated by the Jameson Cell (300-600 microns) explain why it can achieve faster flotation kinetics and has higher carrying capacity, and therefore productivity, when compared with the other cell types.

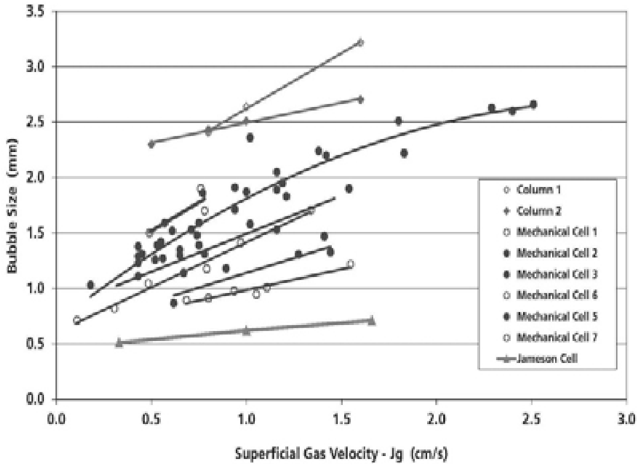


Figure 7: Bubble size as a function of J_g (measure of air flow rate) for different flotation technologies (Nesset et al., 2007; Gomez, 2012, Zou, 2010)

An impractical solution to the frother problem might be to design the entire washing plant to incorporate separate water circuits for the coarse and fines circuits. Perhaps the only practical solution that can be made is to reconfigure the flotation circuit and revert to a 2-stage design. The 2-stage circuit may provide a more robust operation because the primary and secondary stages can be operated separately to optimise performance. Higher frother dosage added upfront would enhance the performance of the primary cells and then the remaining frother in the tailings of this cell can be used to float additional coal in the secondary cells. RTCA's[1] Warkworth operations in NSW, Australia have reconfigured the Jameson Cells to operate in 2-stages and have reported an improvement in yield and higher concentrations of frother can be used without excessive frothing occurring (Lambert and Revell, 2008). New installations such as Yancoal's Moolarben plant (NSW) and RTCA's Kestrel (QLD) have also chosen 2-stage flotation circuits.

There are opportunities for washing plants operating with single stage flotation circuits to be retrofitted to become 2-stage circuits. AAMC's[2] German Creek operations (QLD) recently installed a single Jameson Cell as a secondary stage to treat the tails from two Microcel units operating in parallel (Figure 8). The footprint was small and installation was straightforward because of the simple tie-ins and no other auxiliary equipment was required other than the feed pump.

Figure 8: Recently installed Jameson Cell (left) next to two column cells at AAMC's German Creek operation

4. Challenges in coal flotation and fine coal processing

Flotation is a physio-chemical separation process and is different to other unit operations in a coal washing plant which all employ gravity based techniques to effect

---

[1] RTCA is Rio Tinto Coal Australia

[2] AAMC is Anglo American Metallurgical Coal

separation. Flotation is a complex 3-phase system that is controlled by many factors which can be categorised into three facets: coal characteristics, chemistry impacts and machine characteristics (Figure 9).

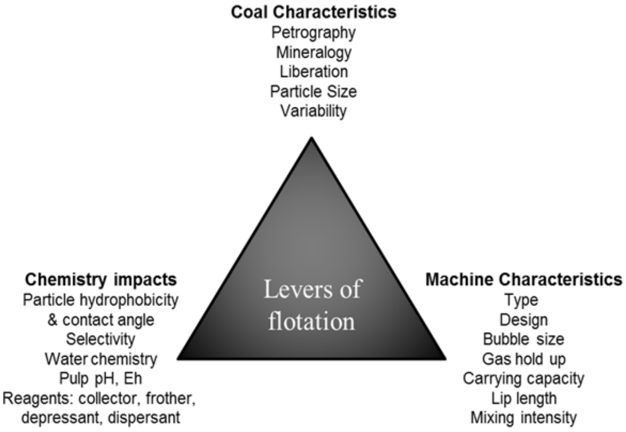


Figure 9: Flotation 'triangle' showing the factors affecting performance

The flotation machine is only one facet important to the overall process, but seems to get the most attention and is often blamed when performance of the fines circuit is poor. However, the variability of the coal and flotation reagent control are two factors which perhaps is surprisingly often overlooked. The greater the number of different coal seams and sources that are treated, the more challenging is the task of achieving effective flotation and the targeted qualities and recoveries. Operators must therefore be trained to respond to changes in tonnage, particle size distribution and flotation behaviour of the different coal types and make the necessary adjustments to reagent dosages and process variables to optimise performance.

In many plants, monitoring of flotation performance is irregular and often a 'knee-jerk' change is made when performance has clearly deteriorated. Furthermore, in many plants it may be impossible to conduct surveys because sample points do not exist for the feed, concentrate and tailings. Even in the more modern (recently built) plants, the flotation feed cannot be easily collected as it usually consists of more than one stream which gravity flows into a large collecting sump. In Jameson Cell installations, many operations unknowingly collect the downcomer feed, and use the result from a "rapid ash" analysis and the two-product formula to calculate yield and combustibles recovery. This is erroneous as the downcomer feed is an internal stream. Sampling points are therefore an essential part of good design of the fines circuit.

Even though flotation is the separation process, the overall fines circuit performance is often dictated by concentrate dewatering capacity as this often proves to be the bottleneck in the process. Many flotation circuits have to be 'scaled back' to suit the capacity of the dewatering device leading to large losses in coal fines which then cause issues in the tailings thickener. Another area to address is the type of dewatering device used for concentrate dewatering. The technology chosen needs to carefully consider the particle size and type of coal treated and not use capital cost as the driver for decisions.

5. Bulga Pilot Plant

Flotation has been traditionally used as a beneficiation process for treating coal fines but in the past it has been mainly used for metallurgical coal. However, in South Africa there has been a number of flotation installations aimed at recovering saleable thermal coal from raw tailings that would otherwise be discharged as waste. These have had mixed success because coals from this region are not readily floated, but they have also demonstrated the parallel need for selecting effective dewatering circuits for treating the clean coal concentrates (Power, 2010).

Xstrata Technology together with Xstrata Coal has been developing the flowsheet for a Jameson Cell installation to treat thermal coal at Bulga mine near Broke in NSW. This plant will eventually treat natural raw coal tailings and also tailings recovered from nearby tailings ponds that contain an estimated 3Mt of recoverable coal.

The plant washes as many as 12 coal seams and coal sourced from both open-pit and underground mining operations with a flotation feed ash varying from 30-65%. It is challenging for any technology to make a consistent quality product with such a large variation in feed.

A Jameson Cell flotation pilot plant was operated on-site to provide the necessary design information for the plant extension. The pilot plant flotation testwork showed consistently high recoveries from the thermal coal feed, achieving 70 to 90% combustible recovery with 8 to 12% ash in the product. The more oxidised coal types required increased collector additions to achieve acceptably high recoveries.

The pilot plant was configured with two L500 Jameson cell units arranged in series to simulate full scale design of the flotation circuit to be installed in future. A layout of a proposed full scale Jameson Cell installation is shown below in Figure 10. This compact design allows for installation of the flotation plant efficiently in a brownfield expansion. Various dewatering and other treatment options have been considered for this plant including the combination of screen-bowl and Centribaric units with subsequent briquetting of the higher value products. These equipment items will also be tested with the pilot-plant to ascertain their suitability and once this work is completed this flotation circuit will be designed as a "standard" for future Xstrata Coal washing plants treating both metallurgical and thermal coal types.

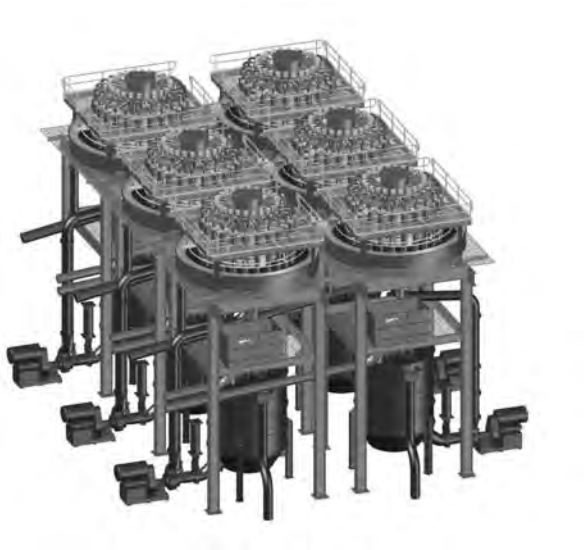


Figure 10: Layout of six B6500/24 model Jameson Cells (Footprint = 22 x 26m; Height = 17m)

6. Concluding Comments

Over little more than a decade Jameson Cells have become a standard flotation technology in the Australian coal industry. With over 110 installations, this industry represents a substantial knowledge base for operating, maintaining and improving installations in the future. There are also a growing number of coal installations worldwide in every major coal region including emerging regions such as Mongolia and Mozambique. Users have long-since recognised the major advantages, i.e. simplicity, excellent availability and low maintenance in addition to proven robust and consistent process performance.

Much of the progress that has occurred in Jameson Cell evolution is owed to improvements in the materials used in components as previously designed, but other factors clearly affect the overall performance of the flotation circuit. Some of these have also been addressed, i.e. dewatering, sampling, process control, etc. Future plants will need to be designed to incorporate further innovations in each of these areas especially where operators can be provided with reliable monitoring and control features that enable changes to be made to operating conditions and early indications of deviations from quality and recovery targets.

Xstrata Technology (XT) has built up a strong team of designers, engineers and process specialists and as a result, has continuously improved the technology over two decades whilst still maintaining close association with its inventor and his team of scientists. Such relationships are rare in process engineering and in addition XT is not just an equipment supplier, their modus operandi being to develop technology ‘partnerships’ with clients which include not only coal users, but process engineers and designers working in base metals, industrial minerals and a variety of other applications. Therefore, this is a unique situation where knowledge transfer from all sources is passed on to all users. There is little doubt that further improvements will result in more new models and circuit designs and the incorporation of desirable features such as on-line analysis, designed-in slurry samplers, reagent sensors, etc.

7. References

1. Atkinson, B.W., Conway, C.J., Jameson, G.J. Fundamentals of Jameson Cell operation including size-yield response, 6th Australian Coal Preparation Society Conference, Mackay QLD, September, 1993.
2. Caretta, M.F., Graham, J.N., Dawson, W.J., Jameson Cell scale-up experiences at BHP Coal’s Goonyella coal preparation plant, Coal Prep ’97, Lexington, Kentucky, April, 1997.
3. Gomez, C.O., Personal communication, 2012.
4. Harbort, G.J., Manlapig, E.V., Jackson, G.J., Jameson Cell use in fine coal flotation, ASEAN-Pacific Coal Conference ’92, Cebu, Philippines, 1992.
5. Jameson, G.J., Goffinet, M., Hughes, D., Operating experiences with Jameson Cell at Newland Coal Pty Ltd, 5th Australian Coal Preparation Society Conference, Newcastle NSW, May, 1991.
6. Lambert, L. and Revell, P., 2008. Tripling Flotation Yield at Warkworth CHPP, 12th Australian Coal Preparation Society Conference, Darling Harbour NSW, October, 2008.
7. Manlapig, E.V., Jackson, B.R., Harbort, Cheng, C.Y., Jameson Cell coal flotation, Coal Prep ’93, 10th International Coal Preparation Exhibition Conference, Lexington Kentucky, May, 1993.
8. Nesset, J.E., Finch, J.A., Gomez, C.O., Operating variables affecting bubble size in forced-air mechanical flotation machines, Ninth Mill Operator’s Conference, Fremantle WA, 2007.
9. Power, D., The new highs and lows of coal processing for South African coals, 13th Australian Coal Preparation Society Conference, Mackay QLD, September, 2010.
10. Swanepoel, C., Fine coal processing developments in Anglo American Thermal Coal in South Africa, Challenges in fine coal processing, dewatering and disposal, SME, edited by Klima, M.S., Arnold, B.J., Bethell, P.J., 2012.
11. Zou, J.S., Effect of frother on the gas dispersion characteristics of Jameson Cells, McGill University Internal Report, 2010.

PROCEEDINGS

XVII. INTERNATIONAL COAL PREPARATION CONGRESS

1-6 OCTOBER 2013
ISTANBUL TURKEY

Edited by
Gülhan ÖZBAYOĞLU & Ali İhsan AROL

Two decades of Jameson Cell installations in coal

Dave Osborne[1], Le Huynh[1], Ishant Kohli[1], Michael Young[1] and Frank Mercuri[2]

1 Xstrata Technology, Brisbane, QLD, Australia

2 Xstrata Coal, Brisbane, QLD, Australia

ABSTRACT

Jameson Cell technology was first introduced in the late 1980s to overcome the design and operating inadequacies of column and conventional flotation cells. From the first commercial base metals installation at Mt Isa in 1989 followed by the first coal installation at Newlands in 1990, it has been continuously developed and improved to make it more robust and easier to use. There are now over 320 units installed in a wide range of applications and industries of which about 45% are employed for coal. The latest designs combine the original advantages of high productivity and small footprint with significantly reduced maintenance costs and much more operator-friendly features. This paper will describe the development of the Jameson Cell technology over two decades and drivers for changing circuit configurations from 2-stage to single stage and then back to 2-stage over the years. Although most Jameson Cell installations are currently utilised for recovery of metallurgical coal, there are a growing number of applications emerging for treating ultrafine thermal coal, many with high raw coal ash contents, predominantly clay. The paper will also describe recent pilot-scale test work and the results obtained from this to support the commercial scale project.

Keywords: Jameson Cell, Coal, Flotation, Frother

1. Introduction

The Jameson Cell is a fundamentally different flotation technology to mechanical and column cells, having been invented by Prof Graeme Jameson at the University of Newcastle in NSW. They were first tested and commercially installed in a coal washing plant at Xstrata Coal's Newlands mine (Jameson et al., 1991). The fines stream was cyclone overflow material which was previously discarded (minus 20-25 microns in particle size with 15 to 50% ash content). Pilot plant testing showed it was possible to achieve greater than 90% combustibles recovery with a product target of 10% ash. This led to the installation of the first generation full scale, Mark I Jameson Cells in 1990. These cells were in continuous operation for over 15 years until a new washing plant was built to replace the old plant in 2006. The new plant also uses Jameson Cells and has four new B6000/20 model Mark III cells installed. Following the initial Newlands installation, many sites have tested the technology which was shown to consistently produce low ash concentrates and achieve high combustibles recovery whilst being forgiving to variations in feed ash (Harbort et al., 1992; Atkinson et al., 1993; Manlapig et al., 1993). BHP Coal's (now BHP Billiton Mitsubishi Alliance - BMA) Goonyella 1,800 tph coking coal operation in Central Queensland tested the advanced flotation technologies and subsequently replaced the entire 32 mechanical (Wemco) cell circuit with 8 Jameson Cells operating in a 2-stage configuration (Caretta et al., 1997). Figure 1 compares the performance of the Jameson Cells at this plant after commissioning to the old mechanical cell circuit. The ability of the Jameson Cell to consistently deliver a low ash product at high combustibles recoveries contributed to an overall plant yield increase of ~3.5% and led to production records.

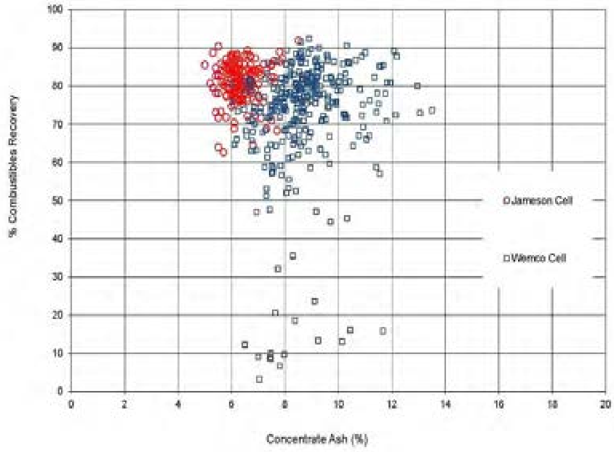


Figure 1: Full scale Jameson Cell performance at Goonyella mine compared to old mechanical (Wemco) cell circuit

Amongst the key benefits is froth washing and the simplicity afforded by the Jameson Cell, it being easy to operate and maintain; with no moving parts, and needing no auxiliary equipment except for the feed pump. Over 150 Jameson Cells are now operational on coal worldwide, the current largest installation being at Wesfarmer's Curragh Mine in the Bowen Basin of Central Queensland which treats over 5 million tonnes of coal fines per year using only twelve cells. Long-established coal producing countries like Kazakhstan are also realising the benefits of the Jameson

Cell over conventional cells and emerging coal regions such as Mozambique and Mongolia are now beginning to use the Jameson Cell for metallurgical coal applications.

The remainder of this paper will describe the development of the Jameson Cell technology over two decades, the evolution of flotation circuit designs, continued challenges of fine coal processing in the coal industry; and a recent case study for the development of a thermal coal project from pilot plant to full scale is included to illustrate the recommended design approach.

2. Jameson Cell Development

The Jameson Cell development path has seen four main phases of progressive improvement (Figure 2) since its first commercial installation in 1989. Two decades of significant advances have culminated in the latest model, the Mark IV Jameson Cell, which incorporates the following features:

- Feed recycle system; to ensure stable cell operation maintaining optimum performance independent of feed fluctuations.

- Low wear, high discharge coefficient, slurry lens orifice.
- Flexible feed nozzle allowing quick and easy inspection and precise alignment of the plunging jet to maximise metallurgical performance.
- Improved above-froth or in-froth washwater system.

With the release of the Mark III downcomer in 2000, the orifice plates used for jet formation were replaced with the slurry lens orifice (Figure 3). The benefits included a shallow slurry entry angle which optimised slurry flow and maximised component wear life; a high discharge coefficient, and improved slurry jet formation resulting in more consistent air entrainment and vacuum.

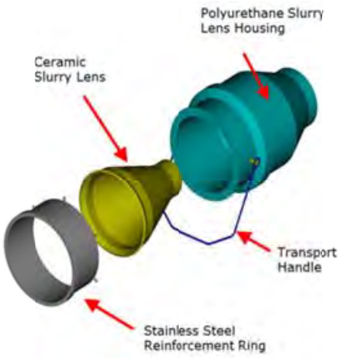


Figure 3: Mark IV Slurry Lens

For maintenance, the slurry lens is significantly easier to access compared to the old design. If a slurry lens becomes blocked or needs replacing, it is now a 10 minute job which can be done online by isolating the specific downcomer while the rest of cell (and other downcomers) remain in operation. The wear performance of the slurry lens orifice has far exceeded expectations. Since its introduction in 2000, 1,500 slurry lenses have been installed in coal applications and no site has so far needed to replace one of these items due to wear.

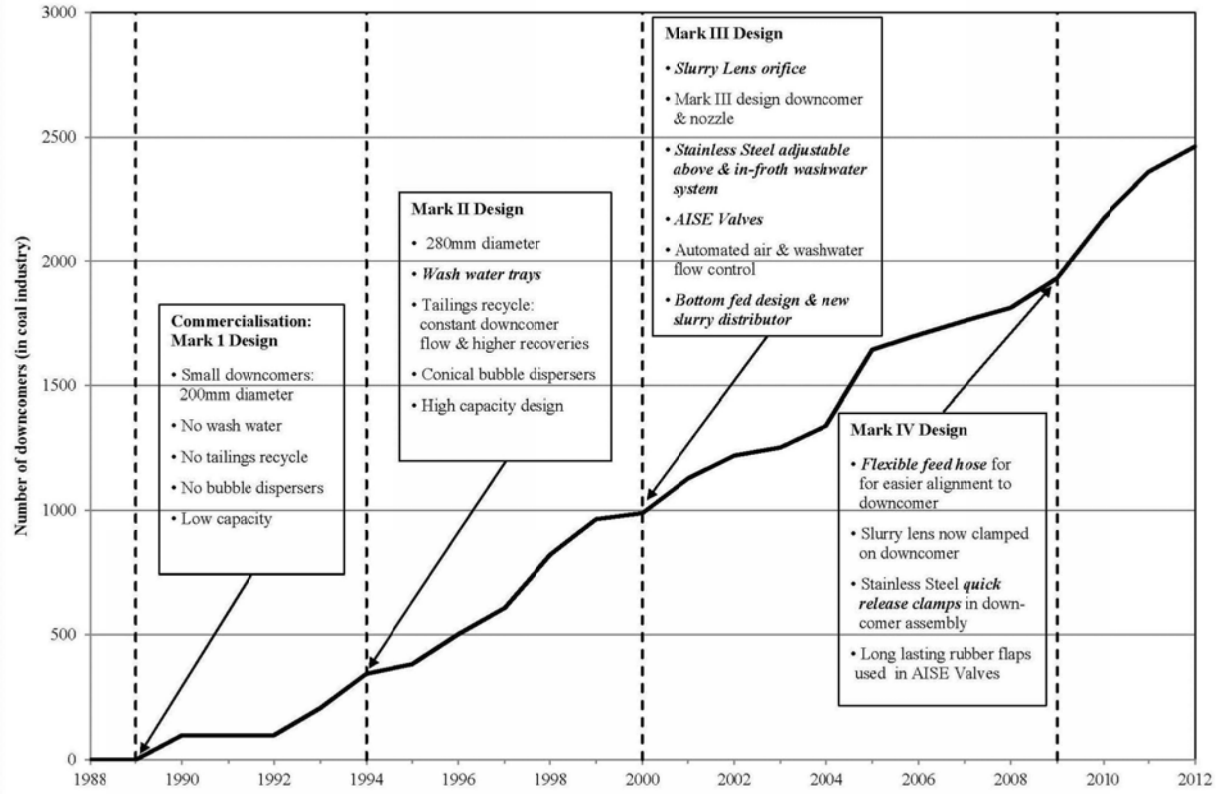


Figure 2: Jameson Cell Development Path

3 Evolution of Jameson Cell Circuit Designs

The first generation of Jameson Cell coal plant installations in Australia in the early 1990s were all 2-stage designs with primary and secondary cells producing separate concentrates that were combined to give the final product. These circuits generally operated well but flotation performance was often affected by wide variations in feed volumetric flow rate to the flotation circuit. This was usually because fluctuation in the flow rate and feed pressure resulted in inconsistent air entrainment and hydrodynamic mixing inside the downcomer. This was overcome in the mid-1990s with the incorporation of tailings recycle into the circuit design. Although originally designed to dampen feed flow fluctuations, by employing higher tailings recycle of between 40-50% and reducing the fresh feed flow rate, it was observed that cell recovery was improved which enabled a single Jameson Cell to achieve a similar level of combustibles recovery as some 2-stage circuits (Figure 4). These single stage cells are therefore typically designed to operate with a lower fresh feed flow rate and higher tailings recycle of about 40-50%. In metallurgical coal applications the number of single-stage cells required to achieve high yields (up to 85-90%) at the required quality must ensure adequate carrying capacity (bubble surface area). An equivalent 2-stage circuit would use the same number of cells as the single stage arrangement but the cells will be installed in series rather than in parallel and will operate at much lower tailings recycle (10-20%). In this type of circuit, the purpose of the tailings recycle is to simply dampen fresh feed fluctuations. The obvious attraction of the single stage flotation circuit is the reduced capital and operating cost as it can be designed using minimal feed sumps and pumps. For example, for two cells in a parallel circuit, only a single feed sump and a single feed pump are required, whereas for an equivalent 2-stage circuit, 2 sumps and 2 (smaller) pumps are required. However, the level of saving is diminished as the size of the circuit increases because 2-stage circuits with more than 2 cells can also make use of common sumps and larger feed pumps as the feed for each stage primary and secondary stage can also be split between two or more cells.

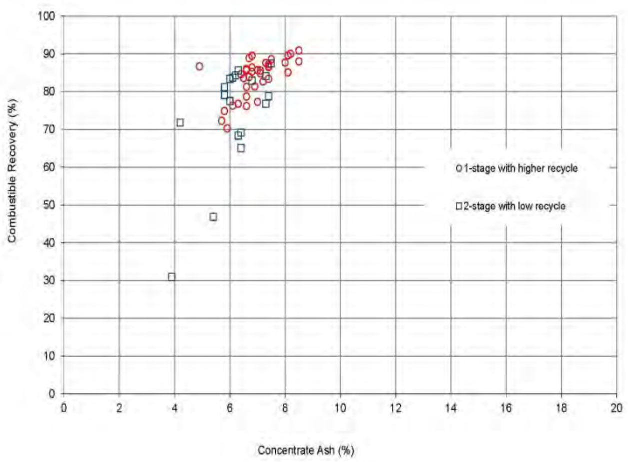


Figure 4: Flotation performance of single stage versus 2-stage Jameson Cell circuit.

From 1995 onwards, all Jameson Cell installations were single stage circuits with 40-50% tailings recycle. Over the years, many operations have experienced frothing issues in the plant caused by residual frother in the recirculated tailings thickener overflow water. This issue is complex and the effects experienced at each site are varied and often require individual solutions. The sensitivity of the overall plant to residual frother is also dependent on how quickly water is reused, the design of the water circuit and the reagent combinations used throughout the plant. Common problems experienced include instability in dense medium circuits through air entrainment, dewatering limitation due to froth handling constraints and water clarification issues.

However, the impact of lowering frother dosage to the flotation recovery in single stage circuits is severe as demonstrated in Figure 5. At this metallurgical coal operation, combustibles recovery was poor at 35-45% and well below the knee of the curve generated from a standard coal characterisation test (black line) when the frother concentration was 9 ppm. Altering froth depth, air flow rate and washwater flow rate had negligible impact on the recovery. However, when the frother dosage was increased to 14 ppm, the combustibles recovery dramatically increased to 75 to 82%, reaching the maximum at the knee of the curve.

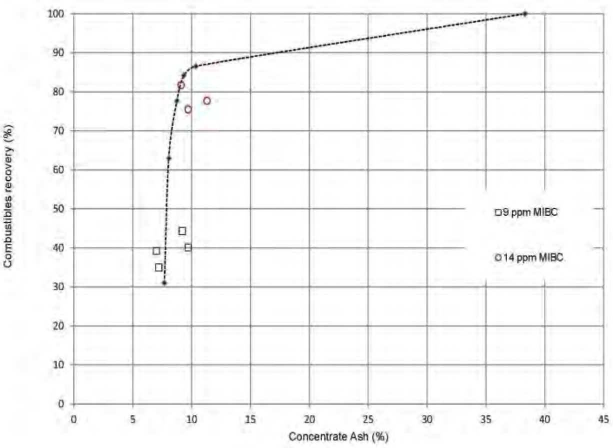


Figure 5: Effect of frother dosage on flotation performance

The reason for the dramatic change in combustibles recovery shown in Figure 5 can be explained as follows. In single stage circuits, whether a Jameson Cell or a column cell is used, the carrying capacity of the cell is entirely dependent on that cell being able to produce sufficient bubble surface area required to float all the coal. For any type of flotation machine, maximum bubble surface area can only be achieved if sufficient frother is added to prevent coalescence. The minimum concentration required for this is called the Critical Concentration of Coalescence (CCC) and is characteristic for each type of frother. Figure 6 shows the change in bubble size with frother dosage for MIBC, the most commonly used frother in coal flotation. The CCC for MIBC is around 15 ppm.

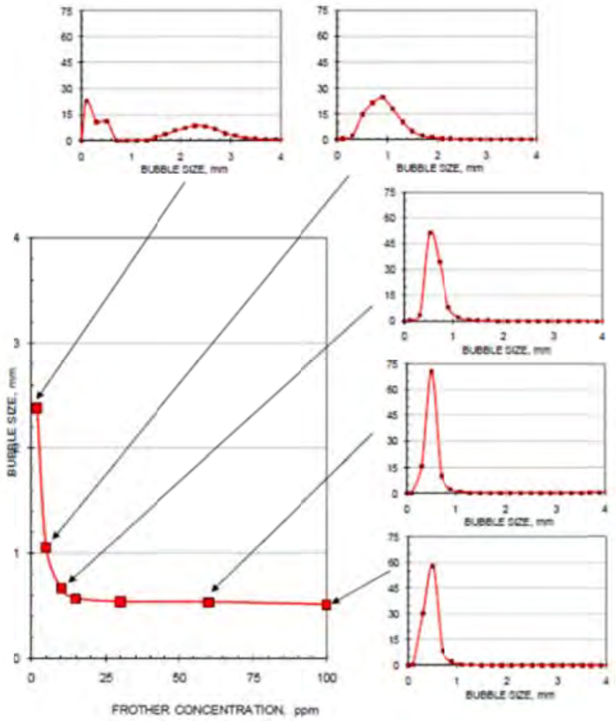


Figure 6: Bubble size as a function of frother concentration (Zou, 2010)

While the frother dictates the concentration it must be used in the flotation process; it is the flotation machine that controls the minimum bubble that can be generated. Figure 7 compares the bubble size of different technologies: Jameson, mechanical and column cells, are included as a function of the superficial gas velocity, J_g (a measure of air flow rate). The fine bubbles generated by the Jameson Cell (300-600 microns) explain why it can achieve faster flotation kinetics and has higher carrying capacity, and therefore productivity, when compared with the other cell types.

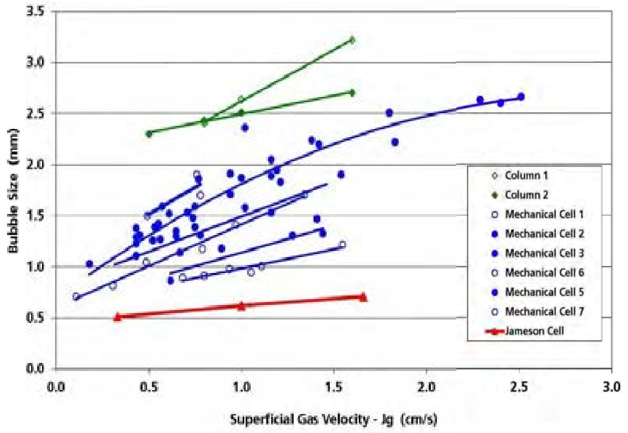


Figure 7: Bubble size as a function of J_g (measure of air flow rate) for different flotation technologies (Nesset et al., 2007; Gomez, 2012, Zou, 2010)

An impractical solution to the frother problem might be to design the entire washing plant to incorporate separate water circuits for the coarse and fines circuits. Perhaps the only practical solution that can be made is to reconfigure the flotation circuit and revert to a 2-stage design. The 2-stage circuit may provide a more robust operation because the primary and secondary stages can be operated separately to optimise performance. Higher frother dosage added upfront would enhance the performance of the primary cells and then the remaining frother in the tailings of this cell can be used to float additional coal in the secondary cells. RTCA's[1] Warkworth operations in NSW, Australia have reconfigured the Jameson Cells to operate in 2-stages and have reported an improvement in yield and higher concentrations of frother can be used without excessive frothing occurring (Lambert and Revell, 2008). New installations such as Yancoal's Moolarben plant (NSW) and RTCA's Kestrel (QLD) have also chosen 2-stage flotation circuits.

There are opportunities for washing plants operating with single stage flotation circuits to be retrofitted to become 2-stage circuits. AAMC's[2] German Creek operations (QLD) recently installed a single Jameson Cell as a secondary stage to treat the tails from two Microcel units operating in parallel (Figure 8). The footprint was small and installation was straightforward because of the simple tie-ins and no other auxiliary equipment was required other than the feed pump.

Figure 8: Recently installed Jameson Cell (left) next to two column cells at AAMC's German Creek operation

4. Challenges in coal flotation and fine coal processing

Flotation is a physio-chemical separation process and is different to other unit operations in a coal washing plant which all employ gravity based techniques to effect separation. Flotation is a complex 3-phase system that is controlled by many factors which can be categorised into

---

[1] RTCA is Rio Tinto Coal Australia

[2] AAMC is Anglo American Metallurgical Coal

three facets: coal characteristics, chemistry impacts and machine characteristics (Figure 9).

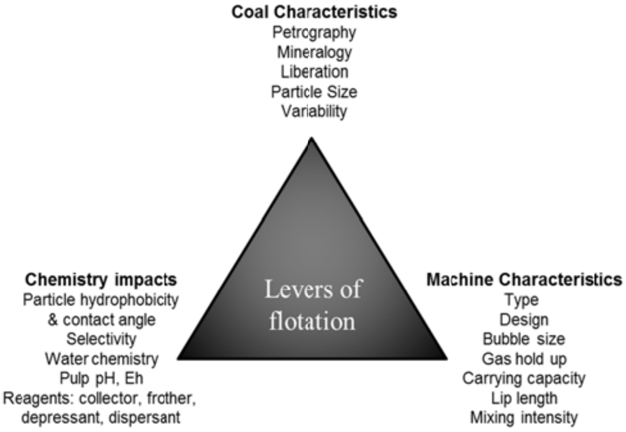


Figure 9: Flotation 'triangle' showing the factors affecting performance

The flotation machine is only one facet important to the overall process, but seems to get the most attention and is often blamed when performance of the fines circuit is poor. However, the variability of the coal and flotation reagent control are two factors which perhaps is surprisingly often overlooked. The greater the number of different coal seams and sources that are treated, the more challenging is the task of achieving effective flotation and the targeted qualities and recoveries. Operators must therefore be trained to respond to changes in tonnage, particle size distribution and flotation behaviour of the different coal types and make the necessary adjustments to reagent dosages and process variables to optimise performance.

In many plants, monitoring of flotation performance is irregular and often a 'knee-jerk' change is made when performance has clearly deteriorated. Furthermore, in many plants it may be impossible to conduct surveys because sample points do not exist for the feed, concentrate and tailings. Even in the more modern (recently built) plants, the flotation feed cannot be easily collected as it usually consists of more than one stream which gravity flows into a large collecting sump. In Jameson Cell installations, many operations unknowingly collect the downcomer feed, and use the result from a "rapid ash" analysis and the two-product formula to calculate yield and combustibles recovery. This is erroneous as the downcomer feed is an internal stream. Sampling points are therefore an essential part of good design of the fines circuit.

Even though flotation is the separation process, the overall fines circuit performance is often dictated by concentrate dewatering capacity as this often proves to be the bottleneck in the process. Many flotation circuits have to be 'scaled back' to suit the capacity of the dewatering device leading to large losses in coal fines which then cause issues in the tailings thickener. Another area to address is the type of dewatering device used for concentrate dewatering. The technology chosen needs to carefully consider the particle size and type of coal treated and not use capital cost as the driver for decisions.

5. Bulga Pilot Plant

Flotation has been traditionally used as a beneficiation process for treating coal fines but in the past it has been mainly used for metallurgical coal. However, in South Africa there has been a number of flotation installations aimed at recovering saleable thermal coal from raw tailings that would otherwise be discharged as waste. These have had mixed success because coals from this region are not readily floated, but they have also demonstrated the parallel need for selecting effective dewatering circuits for treating the clean coal concentrates (Power, 2010).

Xstrata Technology together with Xstrata Coal has been developing the flowsheet for a Jameson Cell installation to treat thermal coal at Bulga mine near Broke in NSW. This plant will eventually treat natural raw coal tailings and also tailings recovered from nearby tailings ponds that contain an estimated 3Mt of recoverable coal.

The plant washes as many as 12 coal seams and coal sourced from both open-pit and underground mining operations with a flotation feed ash varying from 30-65%. It is challenging for any technology to make a consistent quality product with such a large variation in feed.

A Jameson Cell flotation pilot plant was operated on-site to provide the necessary design information for the plant extension. The pilot plant flotation testwork showed consistently high recoveries from the thermal coal feed, achieving 70 to 90% combustible recovery with 8 to 12% ash in the product. The more oxidised coal types required increased collector additions to achieve acceptably high recoveries.

The pilot plant was configured with two L500 Jameson cell units arranged in series to simulate full scale design of the flotation circuit to be installed in future. A layout of a proposed full scale Jameson Cell installation is shown below in Figure 10. This compact design allows for installation of the flotation plant efficiently in a brownfield expansion. Various dewatering and other treatment options have been considered for this plant including the combination of screen-bowl and Centribaric units with subsequent briquetting of the higher value products. These equipment items will also be tested with the pilot-plant to ascertain their suitability and once this work is completed this flotation circuit will be designed as a "standard" for future Xstrata Coal washing plants treating both metallurgical and thermal coal types.

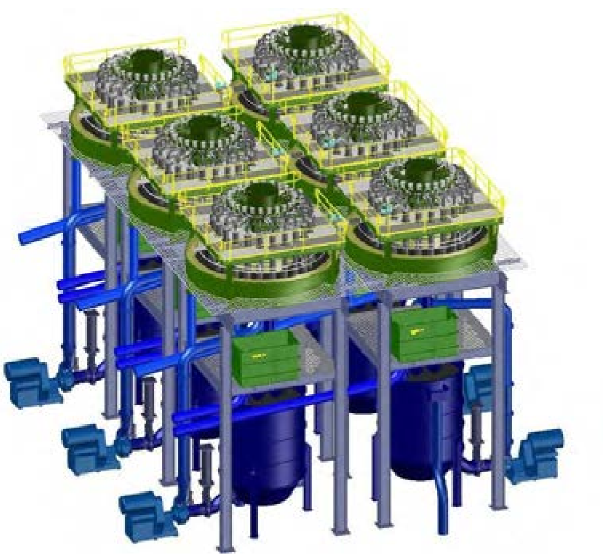


Figure 10: Layout of six B6500/24 model Jameson Cells (Footprint = 22 x 26m; Height = 17m)

6. Concluding Comments

Over little more than a decade Jameson Cells have become a standard flotation technology in the Australian coal industry. With over 110 installations, this industry represents a substantial knowledge base for operating, maintaining and improving installations in the future. There are also a growing number of coal installations worldwide in every major coal region including emerging regions such as Mongolia and Mozambique. Users have long-since recognised the major advantages, i.e. simplicity, excellent availability and low maintenance in addition to proven robust and consistent process performance.

Much of the progress that has occurred in Jameson Cell evolution is owed to improvements in the materials used in components as previously designed, but other factors clearly affect the overall performance of the flotation circuit. Some of these have also been addressed, i.e. dewatering, sampling, process control, etc. Future plants will need to be designed to incorporate further innovations in each of these areas especially where operators can be provided with reliable monitoring and control features that enable changes to be made to operating conditions and early indications of deviations from quality and recovery targets.

Xstrata Technology (XT) has built up a strong team of designers, engineers and process specialists and as a result, has continuously improved the technology over two decades whilst still maintaining close association with its inventor and his team of scientists. Such relationships are rare in process engineering and in addition XT is not just an equipment supplier, their modus operandi being to develop technology ‘partnerships’ with clients which include not only coal users, but process engineers and designers working in base metals, industrial minerals and a variety of other applications. Therefore, this is a unique situation where knowledge transfer from all sources is passed on to all users. There is little doubt that further improvements will result in more new models and circuit designs and the incorporation of desirable features such as on-line analysis, designed-in slurry samplers, reagent sensors, etc.

7. References

1. Atkinson, B.W., Conway, C.J., Jameson, G.J. Fundamentals of Jameson Cell operation including size-yield response, 6th Australian Coal Preparation Society Conference, Mackay QLD, September, 1993.
2. Caretta, M.F., Graham, J.N., Dawson, W.J., Jameson Cell scale-up experiences at BHP Coal’s Goonyella coal preparation plant, Coal Prep ’97, Lexington, Kentucky, April, 1997.
3. Gomez, C.O., Personal communication, 2012.
4. Harbort, G.J., Manlapig, E.V., Jackson, G.J., Jameson Cell use in fine coal flotation, ASEAN-Pacific Coal Conference ’92, Cebu, Philippines, 1992.
5. Jameson, G.J., Goffinet, M., Hughes, D., Operating experiences with Jameson Cell at Newland Coal Pty Ltd, 5th Australian Coal Preparation Society Conference, Newcastle NSW, May, 1991.
6. Lambert, L. and Revell, P., 2008. Tripling Flotation Yield at Warkworth CHPP, 12th Australian Coal Preparation Society Conference, Darling Harbour NSW, October, 2008.
7. Manlapig, E.V., Jackson, B.R., Harbort, Cheng, C.Y., Jameson Cell coal flotation, Coal Prep ’93, 10th International Coal Preparation Exhibition Conference, Lexington Kentucky, May, 1993.
8. Nesset, J.E., Finch, J.A., Gomez, C.O., Operating variables affecting bubble size in forced-air mechanical flotation machines, Ninth Mill Operator’s Conference, Fremantle WA, 2007.
9. Power, D., The new highs and lows of coal processing for South African coals, 13th Australian Coal Preparation Society Conference, Mackay QLD, September, 2010.
10. Swanepoel, C., Fine coal processing developments in Anglo American Thermal Coal in South Africa, Challenges in fine coal processing, dewatering and disposal, SME, edited by Klima, M.S., Arnold, B.J., Bethell, P.J., 2012.
11. Zou, J.S., Effect of frother on the gas dispersion characteristics of Jameson Cells, McGill University Internal Report, 2010.

Опыт применения флотомашины Джеймсон в компании Cleveland Potash Ltd., Северный Йоркшир, Англия

М. Дж. Бернс, Дж. Коутс, Л. Барнард

Cleveland Potash, Ltd. (CPL) является горнодобывающей компанией, занимающейся добычей и переработкой сильвинитовой руды, преимущественно KCl и NaCl, на месторождении эвапоритов в Северном Йоркшире, Англия. На обогатительной фабрике для получения продукта с высоким содержанием KCl - либо ультратонкого концентрата (-100 мкм), либо более грубого концентрата (100-1100 мкм) - применяется пенная флотация. Флотомашина Джеймсон испытывается в обоих циклах.

Хотя аминовый коллектор быстро придает KCl гидрофобные свойства, при извлечении из крупной фракции с высоким содержанием все же возникают затруднения. Более полное извлечение из этой медленно флотируемой фракции +0,85 мм повысило бы общую эффективность фабрики, способствуя отделению пустой породы и сушке концентратов и снижая нагрузку на цикл вторичного измельчения. Основные элементы цикла показаны на рис. 1.

Пилотные испытания

Пилотная флотомашина Джеймсон имела внутренний диаметр 520 мм и высоту примерно 5 м. Питание флотомашины обеспечивал шламовый насос Warman, поддерживающий расход 10 $м^3ч^{-1}$ при 2,5 бар на стороне питания. Время нахождения во флотомашине составляло по расчетам 2 минуты. Конструкция флотомашины схематично представлена на рис. 2.

Флотомашина Джеймсон применялась для флотации четырех разных потоков, представленных в таблице 1.

Флотация концентрата второй перечистной флотации

В тестировании флотации концентрата второй перечистной флотации вспениватель метил изобутил карбинол (MIBC) вводился непосредственно в линию всаса насоса при требуемой концентрации 5 ппм, необходимой для запуска и поддержания флотации. Граница между пульпой и пеной была установлена на уровне 30 см ниже порога флотомашины с помощью датчика пузырьков, реле и пневматического клапана хвостов. На данном этапе пена не промывалась насыщенным раствором.

Расход воздуха во флотомашине был установлен на уровне 50 л $мин^{-1}$, что соответствует коэффициенту скорости поверхностного подъема 0,41 см $с^{-1}$ и коэффициенту воздух-пульпа 1:3,33. Скорость подъема равна частному от деления расхода воздуха на площадь сечения флотомашины, и имеет важное значение для расчета параметров масштабирования.

Флотация концентрата перечистной флотации

Условия флотации концентрата перечистной флотации оставались такими же, как описано выше. Содержание и извлечение были измерены в широком диапазоне содержаний в питании, что было необходимо для выполнения анализа производственного потенциала флотомашины Джеймсон. Плотность пульпы поддерживалась на уровне 15-20% твердой фракции по весу, чтобы облегчить перекачивание и тем самым стабилизировать давление на стороне питания и расход воздуха во флотомашине на уровне 2,5 бар и 50 л $мин^{-1}$, соответственно.

Результаты представлены в таблице 2. Таблица 3 содержит эксплуатационные показатели флотомашины Джеймсон и существующих машин перечистной флотации при питании с крайне низким содержанием. Гранулометрические характеристики флотационных концентратов представлены в таблице 4.

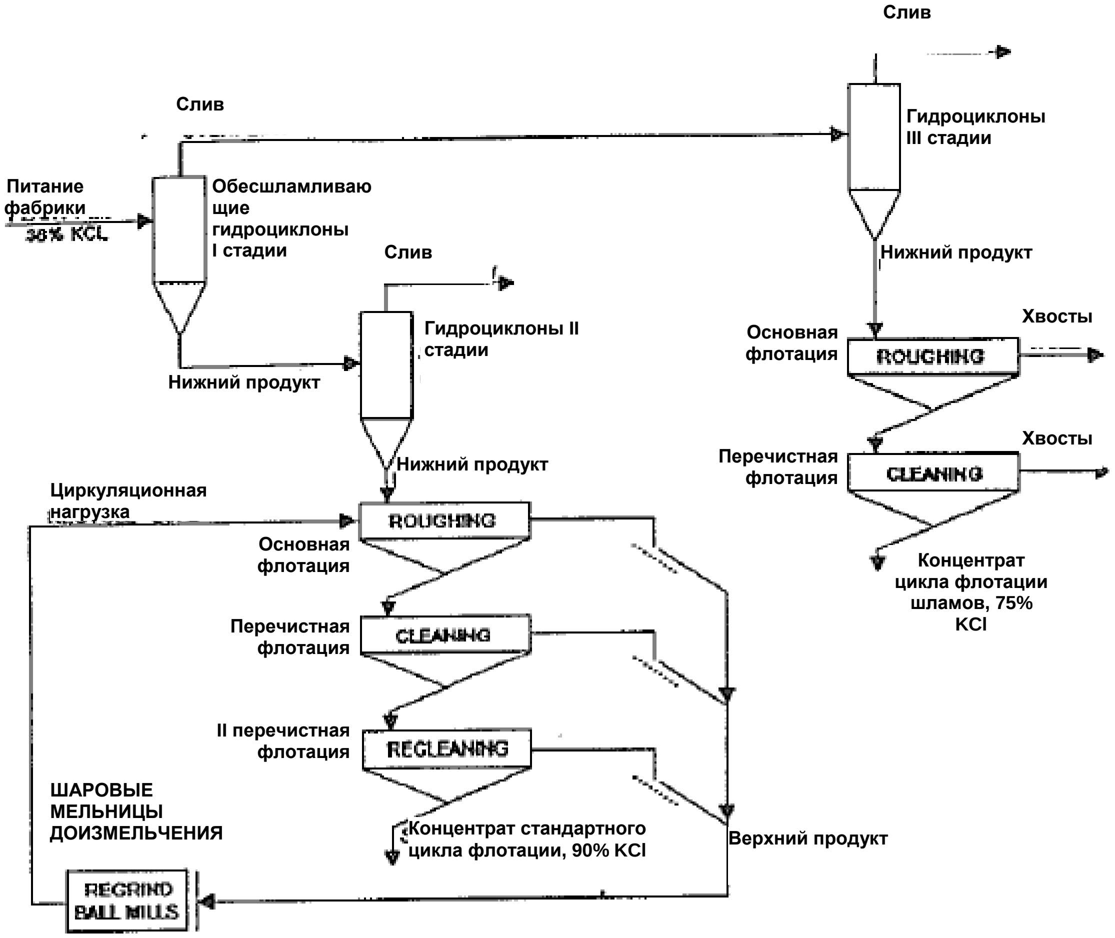


Рис. 1. Схема стандартного цикла флотации и цикла флотации шламов на предприятии Cleveland Potash, Ltd.

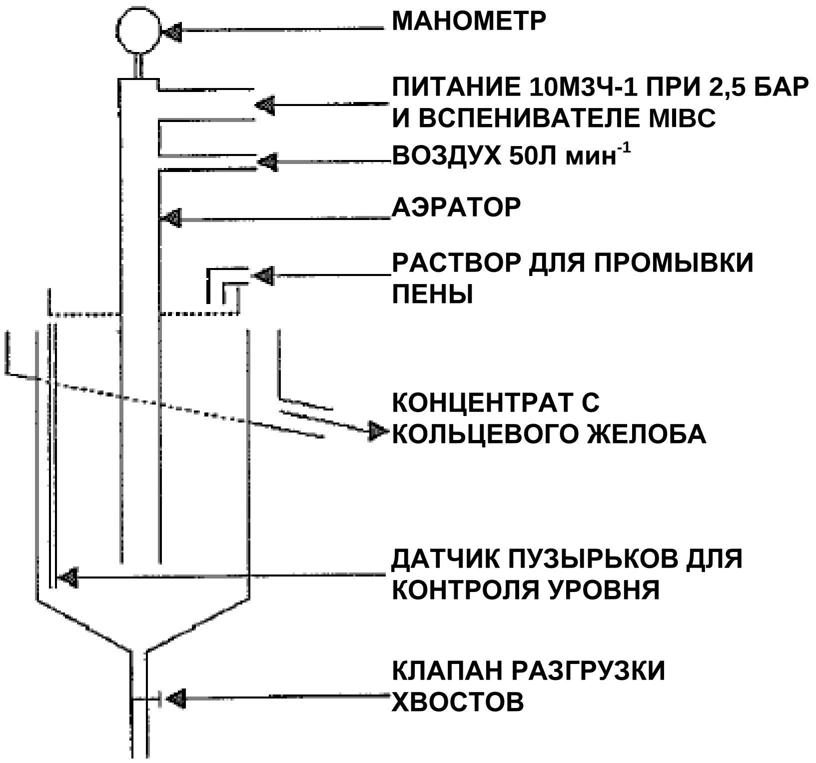


Рис 2.: Схематическая диаграмма Jameson Cell

Таблица 1. Характеристики тестируемых потоков

Крупность, мм	**Концентрат основной флотации**	**Концентрат перечистной флотации**	**Концентрат второй перечистной флотации**	**Питание флотации шламов**
Остаток на сите, % (по весу)				
+0,85	12,4	6,2	2,6	0,0
-0,85 + 0,60	14,3	15,4	14,6	0,3
-0,60 + 0,30	39,3	45,7	51,6	3,0
-0,30 + 0,10	24,6	24,8	26,7	34,6
(-0,10)	9,4	7,9	4,5	62,1
Твердая фракция, % (по весу)	32	33	38	18
Удельная плотность	1,41	1,42	1,45	1,33
KCl, % (по весу)	70-80	80-87	87-90	25-35

Флотация концентрата основной флотации

При исследованиях флотации концентрата основной флотации режим работы флотомашины был изменен, и глубина слоя пены была увеличена до 40 см. Расход воздуха и MIBC остались прежними. Отбор проб был организован так, чтобы сопоставить показатели извлечения во флотомашине Джеймсон с показателями всего цикла перечистной и второй перечистной флотации (таблицы 5 и 6). На данном этапе была использована промывка пены (проходы 23, 24 и 27) для удаления механически выносимой в пенный слой пустой породы.
В проходах 11-21 тесты проводились при глубине слоя пены 30 см без промывки. Прочие эксплуатационные параметры оставались неизменными.
При флотации концентрата основной флотации был прослежен путь извлечения крупной фракции (+0,85 мм) (рис. 3 и 4).

Флотация питания флотации шламов

При флотации питания флотации шламов испытания проводились при глубине слоя пены 40 см без изменения других параметров. Разбавления питания не потребовалось, поскольку было очевидно, что давление на стороне питания и расход воздуха были достаточными. Результаты представлены в таблице 7.

Таблица 2. Эффективность флотомашин на концентрате перечистной флотации KCl

Проход	Дата	Флотомашина Джеймсон				Текущие машины второй перечистной флотации			
		Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %	Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %
1	29/04	75,4	34,7	91,2	87,0	75,4	44,3	89,0	82,1
2	04/05	80,0	22,5	88,1	96,5	80,0	48,1	91,5	84,1
3	05/05	65,5	20,1	89,2	89,5	65,5	34,0	88,1	78,3
4	06/05	75,8	44,3	90,6	81,3	75,8	48,9	89,5	78,2
5	06/05	70,0	32,6	88,6	84,5	70,0	49,2	89,2	66,3
6	06/05	73,7	38,8	88,5	84,3	73,7	48,5	90,6	73,6
7	06/05	68,3	31,9	88,0	83,6	68,3	37,5	90,3	77,1
8	13/05	72,0	18,8	87,4	94,1	72,0	27,0	88,0	90,2
Среднее		72,6	30,5	89,0	87,6	72,6	42,2	89,5	78,7

Таблица 3. Эффективность флотомашин на концентрате перечистной флотации с низким содержанием

Проход	Дата	Флотомашина Джеймсон				Текущие машины второй перечистной флотации			
		Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %	Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %
9	06/05	41,3	16,5	89,7	73,6	41,3	28,1	88,6	46,8
10	06/05	43,2	20,5	88,5	68,4	43,2	26,9	88,5	54,2
Среднее		42,3	18,5	89,1	71,0	42,3	27,5	88,6	50,5

Таблица 4. Гранулометрическая характеристика концентратов, полученных из флотомашины Джеймсон и установленных машин второй перечистной флотации

Номинал сита, мм	Остаток на сите, % (по весу)		KCl, % (по весу)		NaCl, % (по весу)		Нерастворимая фракция, % (по весу)	
	1	2	1	2	1	2	1	2
+0,85	30,4	5,7	90,8	92,7	8,0	6,8	1,2	0,5
-0,85 + 0,60	31,3	21,3	85,1	89,0	10,5	8,0	4,4	3,0
-0,60 + 0,30	27,4	41,9	86,3	91,6	12,3	8,0	1,4	0,4
-0,30 + 0,10	9,7	26,3	91,1	92,1	8,0	7,4	0,9	0,5
(-0,10)	1,2	4,8	71,2	79,7	3,4	6,4	25,4	13,9

1. Концентрат флотомашины Джеймсон
2. Концентрат имеющейся машины второй перечистной флотации

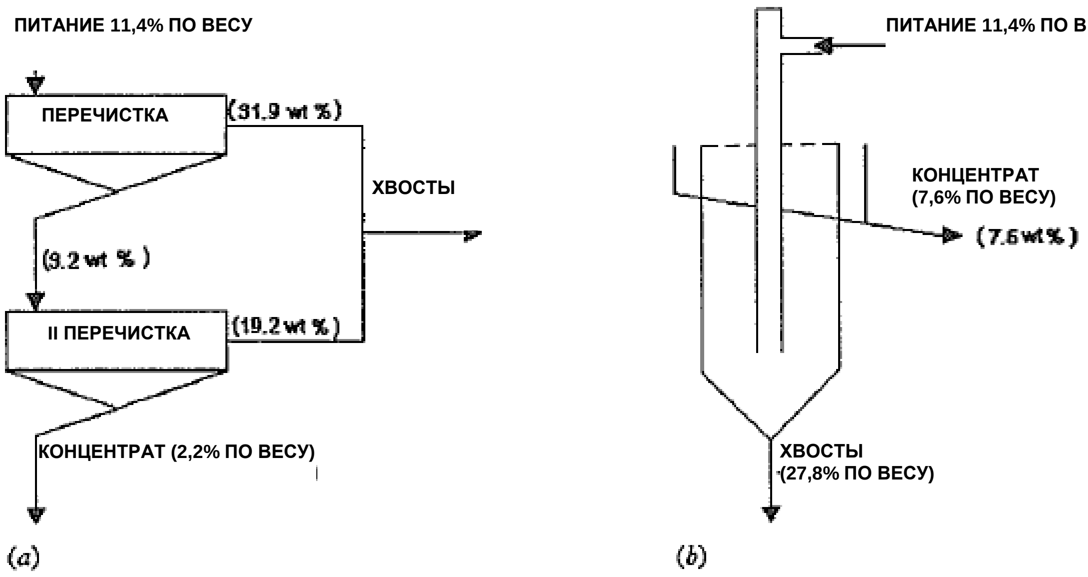


Рис. 3. Вовлечение медленно флотируемых частиц KCl +850 мкм с высоким содержанием в конечный концентрат при тестовом проходе 24: (a) цикл фабрики, (b) флотомашина Джеймсон. На рисунке указан материал, оставшийся на сите. Общее извлечение материала +850 мкм в концентрат в цикле фабрики и во флотомашине Джеймсон составило 14,75 и 53% (по весу), соответственно.

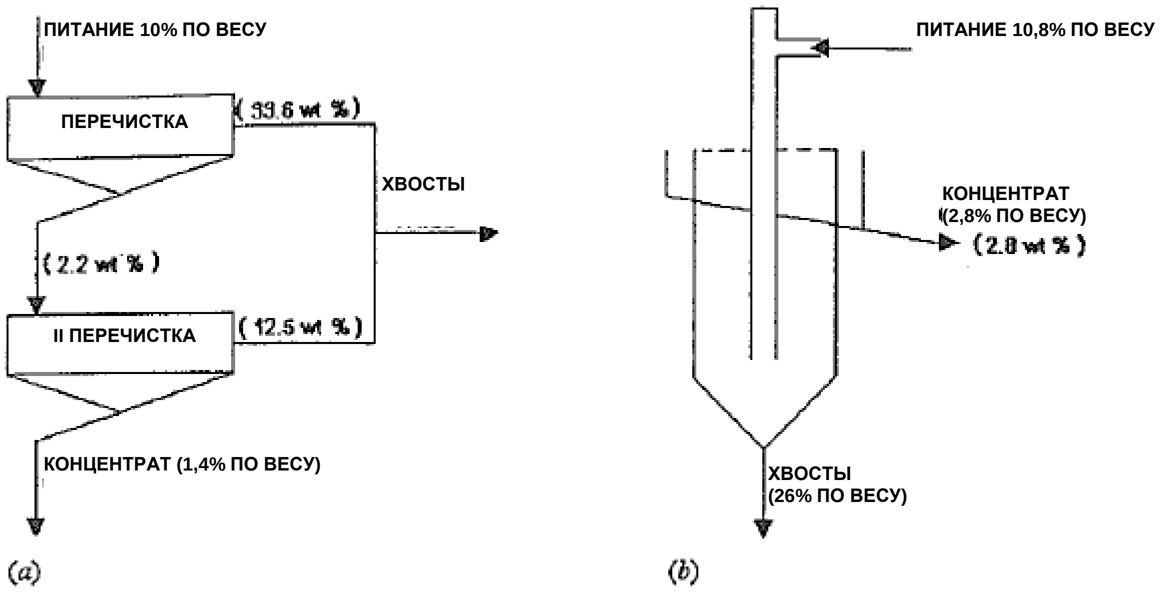


Рис. 4. Вовлечение медленно флотируемых частиц KCl +850 мкм с высоким содержанием в конечный концентрат при тестовом проходе 25: (a) цикл фабрики, (b) флотомашина Джеймсон. На рисунке указан материал, оставшийся на сите. Общее извлечение материала +850 мкм в концентрат в цикле фабрики и во флотомашине Джеймсон составило 8,74 и 17% (по весу), соответственно.

Анализ и выводы

Непосредственно после пуска в эксплуатацию для достижения максимальной эффективности флотомашины не потребовалось значительной корректировки металлургических параметров, таких как расход воздуха, расход вспенивателя, глубина слоя пены, уровень пульпы в аэраторе и промывка пены. Не потребовалось и сложных испытаний, связанных с анализом небольших пошаговых изменений в настройках флотомашины. Большая часть испытаний была направлена на разработку точных методик отбора проб и режима непрерывной круглосуточной эксплуатации при минимальном контроле.

В таблице 1 представлены характеристики исследованных потоков; содержание твердой фракции в пульпе было значительным на протяжении всех пилотных исследований. Все потоки, кроме питания флотации шламов, требовали разбавления питания флотомашины Джеймсон насыщенным раствором в пропорции 1:1. Совокупное объемное извлечение в цикле основной флотации CPL составило 500 $м^3ч^{-1}$. Следовательно, при расходе 1000 $м^3ч^{-1}$ потребовалась бы промышленная установка. Расчеты предварительного масштабирования показывают, что для замещения 28 машин перечистной и второй перечистной флотации Denver 30DR 2,8 $м^3$ потребуется две флотомашины диаметром 2,75 м.

Содержание твердой фракции в концентрате основной флотации составляет, как правило, 32% (по весу) и не делает пульпу слишком вязкой. Однако концентрат содержит значительное количество вовлеченного воздуха, который с трудом выходит из насоса питания флотомашины, что приводит к низкому нестабильному давлению на стороне питания менее 2,5 бар. Столь неблагоприятный режим эксплуатации приводит к низкой аспирации, а следовательно снижает эффективность флотации. Основная функция разбавления насыщенным раствором заключалась, таким образом, в удалении вовлеченного воздуха перед перекачиванием, а вызванное этим снижение вязкости не было значительным. Еще предстоит установить, можно ли подобрать иной метод, помимо дальнейшего разбавления раствором, для уменьшения потерь давления на стороне питания, чтобы сократить размеры и стоимость промышленной установки при увеличении производительности.

Флотация концентрата второй перечистной флотации

В настоящее время на предприятии CPL практически отсутствуют возможности для четвертой стадии флотации. Однако применение флотомашины Джеймсон позволило повысить содержание KCl в концентрате перечистной флотации с 88,5% (по весу) до 91,2% (по весу) при извлечении 94,7% KCl. Это краткое испытание позволило определить обоснованный исходный уровень и точку отсчета для дальнейших тестов.

Флотация концентрата перечистной флотации

В таблице 2 сравниваются показатели эффективности флотомашины Джеймсон и машины второй перечистной флотации CPL при одинаковом питании. Очевидно, что флотомашина Джеймсон обеспечила более высокое содержание и извлечение KCl (89% и 87,6% по весу, соответственно) по сравнению с машинами второй перечистной флотации (89,5% и 78,8% по весу, соответственно). Повышенное извлечение несомненно связано с переходом в концентрат крупных частиц с высоким содержанием в количестве, превышающем аналогичный показатель в машине второй перечистной флотации (таблица 4).

Таблица 5. Эффективность флотомашин на концентрате основной флотации

Проход	Дата	Флотомашина Джеймсон (глубина слоя пены 30 см)				Имеющиеся машины второй перечистной флотации			
		Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %	Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %
11	18/05	67,1	24,0	88,5	88,1	67,1	33,6	79,4	86,6
12	19/05	63,9	32,8	88,5	77,3	63,9	27,7	85,8	83,7
13	19/05	69,9	29,4	90,0	86,0	69,9	32,8	87,9	84,7
14	19/05	76,7	37,4	90,0	87,7	76,7	33,9	87,4	91,2
15	19/05	74,6	36,6	87,4	87,6	74,6	40,4	86,9	85,7
16	20/05	65,5	31,1	87,4	81,5	65,5	34,5	86,3	78,8
17	20/05	61,7	30,4	88,5	77,3	61,7	28,9	83,9	81,1
18	20/05	68,3	29,6	88,5	85,1	68,3	33,8	84,4	84,2
19	20/05	71,2	36,4	89,0	82,7	71,2	40,9	85,8	81,3
20	20/05	66,4	33,8	90,1	78,6	66,4	36,4	86,9	77,7
21	20/05	65,6	40,5	91,2	68,8	65,6	33,8	86,3	79,7
Среднее		68,3	32,9	89,0	81,9	68,3	34,2	85,5	83,2

Таблица 6. Эффективность флотомашин на концентрате основной флотации

Проход	Дата	Флотомашина Джеймсон (глубина слоя пены 40 см)				Имеющиеся машины перечистной и второй перечистной флотации			
		Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %	Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %
22	11/06	78,8	47,0	90,0	84,5	78,8	54,5	90,2	78,0
23*	15/06	80,0	52,9	90,2	82,2	80,0	63,0	92,7	67,0
24*	15/06	79,9	48,1	91,7	83,7	79,9	53,0	91,8	79,6
25	16/06	74,7	47,3	89,6	77,7	74,7	50,7	89,0	74,6
26	16/06	79,4	46,5	92,8	83,0	79,4	47,2	93,9	74,0
27*	24/06	77,5	52,6	91,1	76,0	77,5	53,5	91,1	75,0

Проход	Дата	Флотомашина Джеймсон (глубина слоя пены 40 см)				Имеющиеся машины перечистной и второй перечистной флотации			
		Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %	Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %
Среднее		78,4	49,1	90,9	81,2	78,4	53,6	91,4	74,7

* Промывка пены насыщенным раствором для снижения механического уноса пустой породы

Таблица 7. Эффективность флотомашин на питании флотации шламов

Проход	Дата	Флотомашина Джеймсон				Имеющиеся машины второй перечистной флотации			
		Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %	Питание, % (по весу) KCl	Хвосты, % (по весу) KCl	Концентрат, % (по весу) KCl	Извлечение, %
28	07/07	27,8	4,9	77,4	87,9	27,8	8,5	78,4	77,9
29	08/07	18,4	6,0	75,9	73,2	18,4	6,5	77,9	70,6
30	09/07	23,3	4,6	71,8	85,8	23,3	6,6	82,3	77,9
31	12/07	21,8	4,9	70,0	83,4	21,8	4,3	78,5	84,9
Среднее		22,8	5,1	73,8	82,6	22,8	6,5	79,3	77,8

При отборе проб, данные которого представлены в таблице 4, было установлено, что флотомашина Джеймсон извлекает в концентрат 30,4% (по весу) фракции +0,85 мм, в то время как установленная на фабрике машина второй перечистной флотации извлекает только 5,7% той же фракции. Повышенная эффективность улучшает отделение пустой породы, центрифугирование и сушку концентрата, общее по фабрике извлечение и уменьшает объем материала, поступающего на вторичное измельчение.

Данные таблицы 3 характеризуют эффективность флотомашины Джеймсон при очень низком содержании питания. Очевидно, что флотомашина способна работать с высокой эффективностью в экстремальных эксплуатационных условиях. Флотомашина Джеймсон действительно обеспечивает намного более высокое извлечение, что позволяет рассчитывать на значительное сокращение материала, возвращаемого на основную флотацию при переработке на фабрике материала с низким содержанием.

Флотация концентрата основной флотации

При флотации концентрата основной флотации были выполнены две отдельные программы отбора проб, результаты которых представлены в таблицах 5 и 6. В таблице 5 сопоставляются показатели эффективности флотомашины Джеймсон и установленных на фабрике машин перечистной флотации при одинаковом содержании питания. Средние показатели свидетельствуют о том, что флотомашина Джеймсон обеспечила содержание KCl 89% (по весу) при извлечении 81,9%. Машины перечистной флотации обеспечили содержание KCl 85,5% (по весу) при извлечении 83,2%. Более низкое извлечение во флотомашине Джеймсон связано с высокой границей пена-пульпа. Основное преимущество флотомашины Джеймсон заключается в удобстве корректировки границы пульпа-пена при изменении содержания в питании, которое отличается заметным непостоянством.

В таблице 6 представлены результаты более детального отбора проб, проведенного для прямого сопоставления эффективности флотомашины Джеймсон с эффективностью всего установленного на предприятии цикла перечистной и второй перечистной флотации (показан на рис. 3 и 4). Средние значения по результатам отбора проб (таблица 6) наглядно указывают на разницу в эффективности между флотомашиной Джеймсон и установленными машинами перечистной и второй перечистной флотации. Флотомашина Джеймсон обеспечила приемлемое содержание в конечном концентрате, сопоставимое с показателями машин второй перечистной флотации, при извлечении 81,2% KCl по сравнению с 74,7% в цикле, показанном на рис. 3 и 4. Повышенное извлечение вызвано тем, что крупная фракция поступает в концентрат, а не в хвосты.

В трех случаях слой пены промывался насыщенным раствором (проходы 23, 24 и 27), но это не принесло очевидных преимуществ. Этого следовало ожидать, учитывая природу пены на предприятии CPL, которая не склонна к механическому уносу пустой породы.

Флотация питания флотации шламов

Из таблицы 7, содержащей результаты переработки питания флотации шламов, наглядно следуют два отношения между содержанием и извлечением. Извлечение во флотомашине Джеймсон, составляющее 82,6%, позволяет получать дополнительно 1 т $ч^{-1}$ поташа, что соответствует дополнительному доходу в размере 518 тысяч фунтов в год.

Предположение, что аналогичного извлечения можно добиться на существующих флотомашинах за счет ускоренного прохождения концентрата, в корне ошибочно, учитывая особенности переработки больших объемных расходов пульпы при низкой плотности. При этом неизбежно страдает конечный концентрат шламов.

Компания CPL приняла решение приобрести промышленную установку для переработки питания флотации шламов. Установка будет также использоваться для переработки концентрата основной флотации, чтобы в дальнейшем можно было принять решение о закупке дополнительных флотомашин Джеймсон для полной замены цикла перечистной и второй перечистной флотации. Флотомашина будет иметь диаметр 3,25 м и перерабатывать 325 $м^3ч^{-1}$ пульпы. Фактически она заменит две батареи машин основной флотации и одну батарею машин перечистной флотации, т.е. всего 16 флотомашин объемом по 2,8 $м^3$.

Ожидаемая экономия

Энергопотребление двигателей установленных флотомашин является одной из очевидных возможностей для экономии; расходы на электроэнергию могут быть снижены как минимум на 21 тысячу фунтов в год. Расходы только на запасные части, необходимые для обслуживания 28 флотомашин Denver, составляют почти 20 тысяч фунтов в год (исключая стоимость самих работ). Расход топлива на сушку готового продукта также может быть сокращен с получением экономии порядка 100 тысяч фунтов. Эти оценки сделаны из расчета замены 28 машин перечистной и второй перечистной флотации объемом 2,8 м^3 каждая двумя флотомашинами Джеймсон. Срок окупаемости составит около одного года. Необходимо отметить, что если бы флотомашины были заменены традиционными установками с перемешиванием, экономия, как и срок окупаемости, были бы равны нулю из-за отсутствия технологических преимуществ.
Очевидно, что флотация во флотомашине Джеймсон позволяет не только сократить капитальные и эксплуатационные затраты, но и повысить технологические показатели.

Об авторах

М. Дж. Бернс (член) является инженером-разработчиком в компании Cleveland Potash. Ранее работал в ЮАР в качестве металлурга на предприятии компании Johannesburg Consolidated Industries, где приобрел опыт переработки драгоценных металлов и полиметаллов. Окончил Университет Лидса в Англии и имеет степени бакалавра технических наук (переработка минерального сырья) и магистра наук (инженерная керамика).
Адрес: Cleveland Potash, Ltd., Boulby Mine, Saltburn, Cleveland TS13 4UZ, England.

Дж. Коутс уже шесть лет является техником-металлургом компании Cleveland Potash. Имеет дипломы о высшем образовании в области гражданского строительства, предпринимательства и финансов.

Л. Барнард недавно заняла должность техника-металлурга в компании Cleveland Potash, а до этого работала лаборантом в Национальной физической лаборатории, Теддингтон, гр. Мидлсекс. Окончила Королевский колледж в Лондоне со степенью бакалавра технических наук (материаловедение).

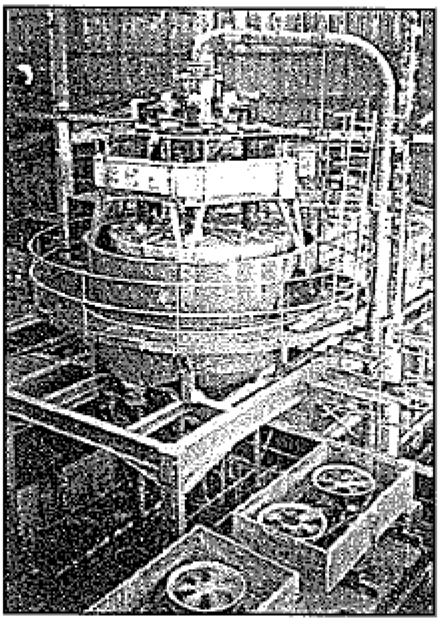
Новая технология флотации промышленных минералов во флотомашине Джеймсон

Стив Холл* и Марк Харрисон**

* Университет Ноттингема, Великобритания
** MIM Technology Marketing Europe, Великобритания

Эффективность пенной флотации зависит от последовательного ряда вероятностей (1). Принципиальное значение имеет обеспечение плотного контакта между пузырьками и частицами материала до связывания частицы ценного минерала с пузырьком, разделения пульпы и пены и, наконец, подачи флотоконцентрата в желоб флотомашины. В разных технологиях флотации применяются разные методики (2, 3, 4) поддержания описанных процессов. Результаты оценки чистой экономии на эксплуатационных расходах при использовании конкретной технологии применимы только при условии аналогичного питания. Затем может быть выполнена экономическая оценка на основании реальной экономии и подтвержденного дополнительного дохода от повышения технологических показателей.

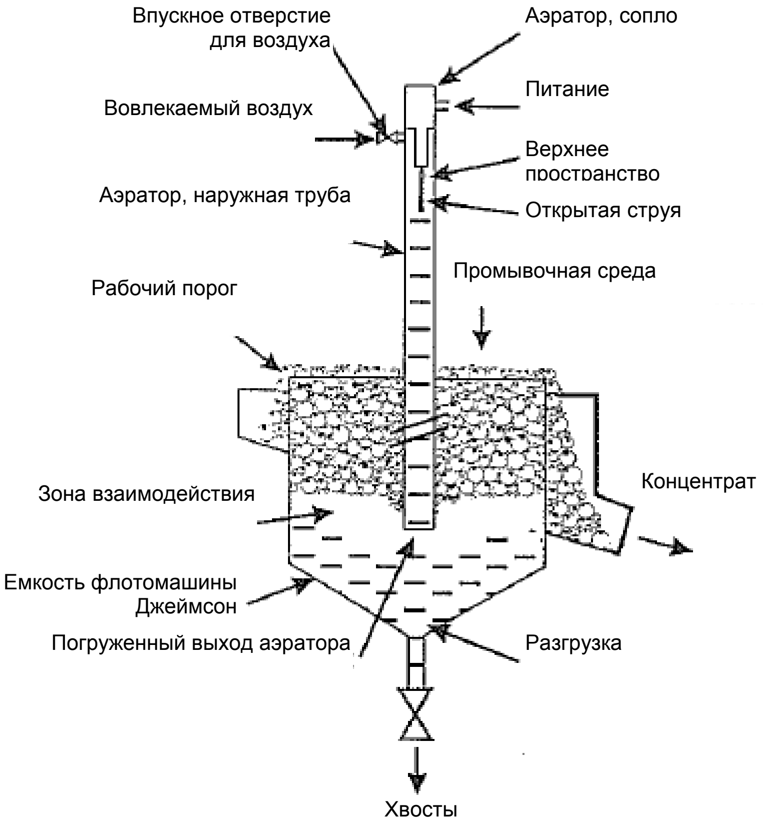
Флотомашина Джеймсон является уникальной машиной пенной флотации, поскольку она имеет гидродинамический принцип работы. Связанные с этим технические и экономические особенности данной немеханической технологии открывают революционные возможности перед промышленной флотацией. К 1995 году в мире эксплуатируется почти 130 промышленных флотомашин Джеймсон на 44 предприятиях флотации. 6 сентября 1994 года компания Cleveland Potash Ltd. (Великобритания) запустила в эксплуатацию первую промышленную флотомашину Джеймсон для флотации промышленных минералов.



Флотомашина Джеймсон: принцип работы

"Сердцем" флотомашины Джеймсон является аэратор. Схема работы одиночного аэратора во флотомашине Джеймсон показан на рис. 1. Аэратор - это простое статичное устройство, состоящее из сопла, установленного на наружной трубе аэратора с единственным впускным отверстием для воздуха. Аэратор установлен вертикально, а его нижний срез находится на определенной глубине в емкости флотомашины Джеймсон.

Рис. 1. Флотомашина Джеймсон в разрезе. Показано расположение и принцип действия одиночного аэратора



При запуске питание поступает в емкость флотомашины Джеймсон в виде нисходящей струи пульпы через сопло. Нижний срез аэратора быстро оказывается погруженным в жидкость, что создает гидравлическое уплотнение, обеспечивающее наполнение аэратора пульпой. Столб пульпы поднимается над рабочим порогом, создавая гидростатическое давление всасывания, а следовательно, область низкого вакуума в верхней части наружной трубы.

Первоначально струя погружена в пульпе аэратора. Но сразу же после открытия впускного отверстия для воздуха, воздух из внешней среды начинает поступать в вакуум. Посредством регулировки расхода воздуха уровень пульпы стабилизируется на определенной высоте внутри аэратора. В результате обнажается часть нисходящей струи пульпы, проходящей через зону низкого вакуума. Падение открытой струи в пульпу в аэраторе создает оптимальные условия для перемешивания и сильное сдвиговое усилие. Воздух непрерывно вовлекается и распространяется в виде мелких пузырьков диаметром от 300 до 600 мкм (среднее арифметическое) (5). В работающем аэраторе поддерживается доля свободного объема от 50 до 60%, что уменьшает толщину пленки между пузырьками и увеличивает площадь межфазной поверхности, доступную для образования контакта с минеральными частицами.

Пульпа находится в аэраторе от 8 до 10 секунд. Создаваемый в жидкости момент обеспечивает движение пузырьков, несущих минеральные частицы, вниз к погруженному выходу аэратора.

У выхода пульпа и пена поступают в зону разделения. Пена поднимается, образуя сплошной слой пены, который промывается по мере необходимости для удаления вовлеченных безрудных минералов перед поступлением флотоконцентрата в желоб верхнего продукта. Хвосты проходят через выпускное отверстие в нижней части

емкости флотомашины Джеймсон. Получаемые в результате характеристики жидкостного потока удовлетворяют всем требованиям к перемешиванию. В общей сложности пульпа находится во флотомашине Джеймсон от 120 до 180 секунд.
В промышленных флотомашинах Джеймсон может использоваться несколько аэраторов, установленных в емкости флотомашины с разделением пены. Отделение поднимающейся пены, несущей минеральные частицы и создаваемой одним или группой аэраторов, поддерживает эффективность флотации при объемном расходе питания, во много раз меньшем проектного максимального расхода. Помимо более высокой металлургической эффективности, эксплуатация флотомашины Джеймсон сама по себе предлагает целый ряд уникальных возможностей для экономии.

Флотомашина Джеймсон: экономические выгоды

18 экономических особенностей флотомашины Джеймсон лучше всего описать, условно разделив их на внутренние и внешние.

Таблица 1. Внутренние особенности флотомашины Джеймсон

Экономическая особенность	Описание
1	Подсос воздуха
2	Образование пузырьков
3	Плотный контакт между частицами и пузырьками за краткое время нахождения
4	Спокойная, поднимающаяся, несущая минеральные частицы пена, доступная для промывки
5	Перемешивание
6	Возможность технического обслуживания без остановки машины благодаря возможности быстрого изолирования изнашиваемых компонентов
7	Возможность эксплуатации при минимальной производительности

Конструктивные особенности

Благодаря аэратору флотомашина имеет семь внутренних особенностей, перечисленных в Таблице 1.

Таблица 2. Компоненты традиционной флотомашины, отсутствующие в конструкции флотомашины Джеймсон

Система	Описание
A	Воздушные компрессоры, воздушные нагнетатели, воздушные подающие магистрали
B	Воздухораспределители и/или распределительные механизмы

Система	Описание
C	Перемешивающие устройства: ротор и статор, трансмиссия и двигатель

В таблице 2 перечислены механические системы, потребность в которых отсутствует благодаря внутренним особенностям аэратора.
Устранение из системы флотации механических узлов делает возможной значительную экономию за счет уменьшения потребляемой энергии и расходов на текущее обслуживание.

Таблица 3. Внешние особенности флотомашины Джеймсон

Экономическая особенность	Описание
8	Компактность оборудования, т.е. меньшая занимаемая площадь
9	Меньше флотомашин на стадию флотации
10	Более простая схема цикла
11	Легкое управление технологическим процессом и контроль параметров
12	Эффективное реагирование на изменения в содержании питания
13	Меньшая потребность в операторах
14	Лучшее понимание процесса операторами
15	Сокращение совокупных затрат на обслуживание
16	Уменьшение занимаемой флотацией площади
17	Значительное уменьшение объема запасных частей
18	Быстрая стабилизация системы после запуска

Внешние особенности

Внешние особенности, перечисленные в таблице 3, обусловлены тем, что эффективность флотации в промышленной флотомашине Джеймсон достигается за очень короткое время пребывания пульпы.

В настоящей статье подробно рассматриваются новые применения флотомашины Джеймсон в переработке промышленных минералов. В настоящее время промышленные флотомашины Джеймсон применяются для поташа и графита. В выводах, основанных на результатах пилотного тестирования, кратко рассматриваются применения для графита, флуорита и циркона.

Промышленные установки

(а) Предприятие 1: производство поташа

Компания Cleveland Potash Ltd (CPL) открыла потенциал флотомашины Джеймсон в 1992 году. Первоначальные лабораторные тесты второй перечистной флотации поташа были выполнены Университетом Ноттингема, выступившим в качестве организации, ответственной за испытания флотомашины Джеймсон. Тесты на пробах, доставленных с предприятий компании CPL, были проведены в пилотной флотомашине Джеймсон J100. Наличие сушильного отделения позволило избежать проблем, связанных с рекристаллизацией поташа в насыщенной раствором пульпе при снижении температуры. Тестирования показали, что аэратор может применяться для поташа и что промывка пены насыщенным раствором из емкости также возможна. Эти выводы послужили толчком к разработке программы экспериментальной эксплуатации на предприятии.

CPL использует флотацию для обогащения сильвина (KCl), отделяемого от галита (NaCl) и нерастворимых веществ, таких как глина и сульфаты. Перед добавлением крахмала (депрессора глины) и амина (коллектор KCl) необходимо тщательное обесшламливание всего питания флотации. Переход в раствор хлоридов сводится к минимуму посредством применения насыщенного раствора в качестве среды обесшламливания и флотации.

Двухстадийное обесшламливание до 100 мкм происходит в батареях гидроциклонов первой и второй стадии. Нижний продукт гидроциклона второй стадии (-1500+100 мкм) поступает к насосу стандартного питания флотации. Верхний продукт гидроциклона первой стадии подвергается дальнейшему обесшламливанию до 30 мкм в гидроциклонах третьей стадии.

Стандартные хвосты основной флотации проходят через обесшламливающие грохоты для извлечения промежуточного продукта и попавшей в них крупной фракции поташа. Верхний продукт крупностью выше номинала (+450 мкм) поступает в три шаровые мельницы доизмельчения. Нижний продукт поступает в центрифугу для хвостов. Стандартные хвосты перечистной флотации поступают самотеком в циклоны LH, а объединенный нижний продукт - в мельницы доизмельчения. Стандартные хвосты второй перечистной флотации поступают на обесшламливающие грохоты. Верхний продукт (+200 мкм) также отправляется на доизмельчение.

Как верхний продукт циклонов LH, так и нижний продукт обесшламливающих грохотов после второй перечистной флотации поступают в гидроциклоны третьей стадии. Получаемый нижний продукт гидроциклонов третьей стадии (-100+80 мкм) поступает к насосу питания флотации шламов. Верхний продукт гидроциклонов третьей стадии поступает в сгуститель хвостов для извлечения раствора из твердой фракции хвостов. В пилотных испытаниях флотомашины Джеймсон на предприятии использовались основные потоки обоих циклов флотации (6, 7). Результаты показали, что наибольшая экономическая выгода может быть получена при использовании флотомашин Джеймсон как можно ближе к началу каждого цикла флотации.

Флотация шламов

6 сентября 1994 года компания CPL запустила в эксплуатацию одну флотомашину Джеймсон 3250/6 (FP) в качестве системы комплексной флотации шламов. Одна флотомашина Джеймсон заменила 16 флотомашин Denver No. 30 DR емкостью 2,8

м^3 каждая, которые использовались для основной и перечистной флотации шламов, т.е. для полного цикла флотации шламов.

Заказчик: CPL

Схема цикла с использованием флотомашины Джеймсон
Количество машин: 1
Количество стадий: 1

Характеристика питания
Источник: нижний продукт циклона третьей стадии
Объемный расход (м^3/ч): 370
Твердая фракция (% по весу): 18
Гранулометрический состав: (-100+30) мкм
Доля общего продукта, приходящегося на флотомашину Джеймсон, %: 10

Экономическая эффективность, средние производственные показатели за 30 дней
Питание (% KCl): 24,5
Сорт концентрата (% KCl): 77,4
Хвосты (% KCl): 5,7
Извлечение, %: 82,8

Консервативная экономическая оценка эксплуатации флотомашины Джеймсон на предприятии CPL кратко представлена в таблице 4.

Таблица 4. Консервативная экономическая оценка (8)

Показатель	**Значение (фунтов стерлингов)**
Срок окупаемости, дней	213
Дополнительный годовой доход от повышенного извлечения	201,939
Годовая чистая экономия на эксплуатационных расходах	209,985

В таблице 5 показано сокращение расходов на энергию на 76,7% в результате использования гидродинамических сил в аэраторе вместо мощности двигателя перемешивающего устройства традиционной механической флотомашины, применяемой в первоначальном цикле флотации шламов.
На предприятии CPL одна флотомашина Джеймсон обеспечивает извлечение тонкого поташа из пустой породы, представленной галитом, глинами и сульфатами. Селективность в отношении тонких частиц сопровождается повышенным извлечением и, по консервативным оценкам, приносит предприятию дополнительный доход в размере 201 939 фунтов стерлингов в год при сроке окупаемости, равном 213 рабочим дням.

Таблица 5. Экономия (%), обеспечиваемая флотомашиной Джеймсон, относительно первоначального цикла флотации шламов

Показатель	**Экономия, %**
Площадь, занимаемая оборудованием	80,9
Общая площадь отделения флотации	81,3
Потребляемая мощность (*), в год	76,7
Техническое обслуживание (**), в год	>79.0

(*) Дополнительный расход энергии на перекачивание при использовании флотомашины Джеймсон относительно расхода энергии двигателями флотомашин первоначального цикла флотации шламов

(**) Максимальный объем обслуживания флотомашины Джеймсон относительно флотомашин первоначального цикла - расходы за предыдущий год эксплуатации, исключая стоимость работ

Таблица 6. Пилотная флотомашина Джеймсон в одностадийной схеме в сравнении с флотомашинами стандартного цикла [{*}]
Питание = стандартный концентрат основной флотации, без промывки пены

Одна флотомашина Джеймсон						**Флотомашины стандартного цикла**			
Питание									
		Концентрат		**Извлечение**		**Концентрат**		**Извлечение**	
% KCl	**Стандарт**	**% KCl**	**Стандарт**	**%**	**Стандарт**	**% KCl**	**Стандарт**	**%**	**Стандарт**
70,6	3,57	89,1	0,94	85,1	3,18				
77,6	2,09	90,8	1,42	81,7	2,17	91,0	2,09	75,5	1,76

[{*} Флотомашины стандартного цикла = (28) флотомашин Denver No. 30 DR 2,8 $м^3$ с субаэрацией в цикле перечистной/второй перечистной флотации]

Таблица 7. Пилотная флотомашина Джеймсон в одностадийной схеме в сравнении с флотомашинами стандартного цикла [{*}]
Питание = стандартный концентрат основной флотации (содержание= 79,9% KCl)

Технология	Количество стадий флотации	Извлечение (по весу) класса +850 мкм	Содержание концентрата, % KCl	Извлечение, %
Одна флотомашина Джеймсон	1	46,1	91,8	83,7
Стандартный цикл	2	13,4	91,7	79,6
[{*} Флотомашины стандартного цикла = (28) флотомашин Denver No. 30 DR 2,8 м^3 с субаэрацией в цикле перечистной/второй перечистной флотации]				

Стандартная флотация

Применение флотации на предприятии CPL свидетельствует о хорошем высвобождении KCl из NaCl в верхних классах крупности. Применение двух циклов флотации требует, чтобы оборудование стандартного цикла флотации извлекало частицы +850 мкм. В таблице 6 сравнивается эффективность пилотной флотомашины Джеймсон и стандартных флотомашин с субаэрацией при переработке стандартного концентрата основной флотации.
Данные, представленные в таблице 6, свидетельствуют о том, что флотомашина Джеймсон способна повысить извлечение в стандартном цикле флотации с 75,5% до 81,7% при содержании KCl в продукте на уровне 90,8% при условии, что результаты пилотных тестов могут быть воспроизведены при промышленной эксплуатации. Количество материала +850 мкм с высоким содержанием, извлеченного из стандартного концентрата основной флотации при содержании KCl 79,9%, показано в таблице 7.
При использовании флотомашины Джеймсон извлечение (по весу) из фракции +850 мкм с высоким содержанием повысилось с 13,4% до 46,1%, в результате чего извлечение KCl при стандартной флотации повысилось на 4,1%. За счет извлечения более крупного материала во флотомашине Джеймсон может быть устранена потребность в доизмельчении примерно трети фракции +850 мкм, поступающей в концентрат основной флотации. В циклах флотации шламов и стандартной флотации на предприятии CPL высокая ионная сила насыщенной рассольной среды обеспечивает следующее (9):

(i) сильное сжатие двойных электрических слоев,
(ii) четырехкратное уменьшение поверхностного натяжения при введении вспенивателя по сравнению с действием аналогичной концентрации такого же вспенивателя, введенного в водную среду.

Медленное движение поднимающейся пены, несущей поташ, в промышленной флотомашине Джеймсон указывает на очень быстрое изменение поверхностного натяжения в аэраторе при образовании контакта между поташом и пузырьками. Поэтому, вероятно, извлечению крупных частиц существенно содействует устойчивое быстрое образование во флотомашине Джеймсон определенной необходимой площади межфазной поверхности.

(b) Предприятие 2: производство графита

Извлечение графита из слива гидроциклона первой стадии (СГПС) считалось необходимым условием экономической рентабельности предприятия в ЮАР. На общую эффективность флотации негативно влияло присутствие в потоке СГПС пустой породы, представленной тонкими шламами, а также низкая эффективность гидроциклона первой стадии. Объединение продукта флотации СГПС с продуктом флотации крупного графита приводило к загрязнению конечного продукта флотации и снижению содержания ниже требуемых 98% С.

Попытки повысить эффективность установленных традиционных флотомашин с субаэрацией не смогли устранить проблему. Предприятие, расположенное в удаленной местности, отправило пробу, содержащую примерно 30 кг СГПС, в Йоханнесбург для выполнения ограниченной программы тестов в лаборатории, использующей флотомашину Джеймсон J100.

На основании результатов тестов во флотомашине Джеймсон была быстро спроектирована, установлена и пущена в эксплуатацию промышленная флотомашина Джеймсон, предназначенная для переработки СГПС. 26 сентября 1994 года была пущена флотомашина Джеймсон 800/1 в качестве машины контрольной флотации тонкой фракции/шламов и перечистной флотации.

Заказчик: конфиденциальная информация

Схема цикла с использованием флотомашины Джеймсон
Количество машин: 1
Количество стадий: 1

Характеристика питания
Источник: СГПС, полученный из массового концентрата основной флотации
Объемный расход (м^3/ч): 11
Твердая фракция (% по весу): 1,5 - 2,5
Гранулометрический состав: d50 = 600 мкм, 10% класса 150 мкм
Доля продукта по фабрике, приходящегося на флотомашину Джеймсон, %: 10

Экономическая эффективность, одиночная проба
Питание (% С): 52
Содержание концентрата (% С): 90,2
Хвосты (% С): 25
Извлечение, %: 72

Включение одной флотомашины Джеймсон в технологическую схему переработки графита позволило сохранить экономическую рентабельность предприятия. Флотомашина Джеймсон производит вязкий флотоконцентрат тонкого/ультратонкого графита, непрерывно подаваемый на перечистную флотацию третьей стадии. Благодаря этому, извлечение по фабрике значительно повысилось при требуем содержании С в продукте 98%.

Таблица 8. Результаты флотации флуорита во флотомашине Джеймсон

Применение	**Количество стадий флотации во флотомашине Джеймсон**	**Питание**	**Концентрат**	
Свинец		% Pb	% Pb	Извлечение, %
PbS, весь цикл	1	6,5	69,8	93,4
Флуорит		% CaF_2	% CaF_2	Извлечение, %
Перечистная флотация 1	1	78,2	94,0	86,3
Перечистная флотация 2	1	93,8	97,8	52,0

Заключение

В гидродинамической флотомашине Джеймсон вместо механических систем традиционных флотомашин для повышения производительности цикла флотации используются статические аэраторы. Промышленная флотомашина Джеймсон, эксплуатируемая Cleveland Potash, обеспечивает хорошую селективность в отношении тонких частиц в нижнем продукте циклона третьей стадии. Одностадийная флотомашина Джеймсон заменяет собой целый цикл флотации шламов производительностью 370 м^3/ч; благодаря этому занимаемая оборудованием площадь уменьшилась на 80%. Чистая экономия на эксплуатационных расходах в размере 209 985 фунтов стерлингов в год обусловлена дополнительным доходом от повышенного извлечения поташа и сокращением на 80% расходов на текущее обслуживание и энергию.

В пилотной флотомашине Джеймсон возможно извлечение крупных частиц поташа +850 мкм из стандартного концентрата основной флотации, что указывает на существующий потенциал увеличения извлечения по фабрике при оптимальной крупности продукта. Также ожидаются технологические улучшения на этапах отделения пустой породы, центрифугирования и сушки концентрата, сопровождаемые снижением нагрузки на цикл доизмельчения. Доказана возможность извлечения в промышленных условиях тонкого графита из неэффективно классифицируемого верхнего продукта циклона первой стадии с увеличением извлечения по фабрике до 98% C и обеспечением экономической рентабельности всего предприятия.

Пилотные тестирования показали, что аэраторы могут успешно применяться в основной флотации циркона, флотации галенита из флуорита и перечистной флотации флуорита. При флотации галенита и флуорита флотомашина Джеймсон производит концентрат с высоким содержанием при значительно меньшем количестве стадий, чем применяемое в настоящее время традиционное оборудование. Необходимы дальнейшие работы на предприятиях с целью уточнения возможностей применения флотомашины Джеймсон для флотации циркона и флуорита. Многочисленные внутренние и внешние особенности флотомашины Джеймсон позволяют уверенно получать поднимающуюся нетурбулентную пену, несущую минеральные частицы и пригодную для непрерывной промывки. Эта важная особенность должна в значительной мере облегчить рентабельное получение продукта необходимого содержания для многих других видов промышленного минерального сырья.

Прочие исследования

(a) Графит: Универститет Ноттингема в качестве организации, ответственной за испытания
Заказчик: конфиденциальная информация

Университету Ноттингема было предложено изучить возможности повышения эффективности флотации графита, в особенности содержания концентрата, с использованием флотомашины Джеймсон.
Перед проведением исследования заказчик заключил соглашение о конфиденциальности в MIM Technology Marketing Ltd и договор о выполнении ТЭО. Графитовая руда, добытая в России, была предоставлена заказчиком в виде крупных кусков и измельчена до -106 мкм. Были изучены флотореагенты различного химического состава и установлено, что флотомашина Джеймсон J100 может обеспечить содержание концентрата, аналогичное производимому традиционной лабораторной флотомашиной Denver (87 - 93% C). Предположительно, в данном случае содержание концентрата был ограничено не эффективностью флотации, а недостаточным высвобождением безрудных минералов из графита. Зачастую в дополнение к флотационным тестам необходимы детальные минералогические исследования.

(b) Графит: пилотные испытания на фабрике в Намибии
Заказчик: конфиденциальная информация

В ходе пилотных тестов на фабрике в начале 1993 года были оценены возможности применения флотомашины Джеймсон в первичной и вторичной перечистной флотации в составе пилотной установки.
Результаты факторного (2^4) эксперимента указали на значительное непостоянство (i) расхода воздуха и (ii) глубины пены. Было установлено, что содержание продукта зависит от выбранной скорости поднятия флотоконцентрата во флотомашине Джеймсон. Флотомашина Джеймсон обеспечивала такое же или более высокое содержание концентрата в сравнении с традиционными флотомашинами или колоннами с субаэрацией. Полученные обнадеживающие результаты должны были привести к тщательным испытаниям флотомашины Джеймсон в качестве машины перечистной флотации конечного продукта. К сожалению, пилотная установка была закрыта из-за отсутствия рынка сбыта для продукции.

(c) Циркон: пилотные тесты на предприятии в Австралии
Заказчик: конфиденциальная информация

Возможности применения флотомашины Джеймсон в основной флотации циркона были предварительно изучены на пилотной установке по переработке минеральных песков. Флотация циркона протекала при контроле pH и без вспенивателя. Сбор амином производился только после депрессии TiO_2. Была поставлена задача добиться при основной флотации циркона содержания ZrO_2 в концентрате 22% при извлечении 90% и максимальной примеси TiO_2 2%.

В одностадийных тестах с содержанием ZrO_2 в питании 10% флотомашина Джеймсон производила концентрат с содержанием >21% ZrO_2 при извлечении 65% (<2% TiO_2) без регулирования pH промывочной среды. Это указывало на возможность основной флотации циркона в две стадии с использованием флотомашины Джеймсон, хотя возможности использования в контрольной флотации не были подтверждены. Кроме того, не было установлено, являются ли примеси SiO_2 и Fe_2O_3 в концентрате вовлеченными или совместно собираемыми компонентами.

(d) Флуорит: испытания на фабрике в Европе
Заказчик: конфиденциальная информация

Применение аэратора во флотации флуорита дало результаты, показанные в таблице 8. Все результаты флотационных тестов были использованы в проектировании технологической схемы флотации и металлургического баланса. Согласно схеме цикла флотации PbS, одна стадия, состоящая из одной флотомашины Джеймсон, должна была заменить целый цикл флотации свинца, т.е. 23 традиционные флотомашины, установленные в одну стадию основной флотации и три стадии перечистной флотации. Для флотации CaF_2 необходимо было установить три стадии, состоящие из одиночных флотомашин Джеймсон в качестве машин перечистной флотации, чтобы обеспечить требуемое содержание конечного продукта и заменить действующую установку из 49 традиционных флотомашин, установленных в шесть стадий перечистной флотации.
Применение флотомашины Джеймсон на данной фабрике в будущем будет зависеть от стабилизации промышленной химии и определения стратегии эксплуатации предприятия с учетом изменений в минералогии питания.

Список использованной литературы

1. Atkinson B.W., Conway C,J. and Jameson G.J. High Efficiency Flotation of Coarse and Fine Coal, paper presented at: High Efficiency Coal Preparation Symposium, SME Annual Meeting, AIME 124th Annual Meeting, March 6-9, Denver, Colorado, USA (1995).
2. Couch G.R., Flotation, p38-45, section 5.1 in: Advanced Coal Cleaning Technology IEACR/44, pp96, IEA Coal Research, London, England (1991).
3. Hall S.T., Developing Flotation Technologies, p(1.1.1-1.1.5) in: Proceedings of Minprep '91 Symposium, Minerals Engineering the Challenges for the '90s, 9-11 April, Doncaster, UK, 09-11 April 1991, pp81, Mining Industry Promotions Ltd, Rickmansworth, England (1991).
4. Skillen A., Froth Flotation: New Technologies Bubbling Under, Industrial Minerals, No.305, February, p47-59 (1993).
5. Evans G.M., Atkinson B.W. and Jameson G.J. The Jameson Cell, p331-363 in Flotation Science and Engineering, pp576, Marcel Dekker, New York, USA (1995).
6. Burns M.J., Coates G. and Barnard L., Use of Jameson Cell Flotation Technology at Cleveland Potash Ltd., North Yorkshire, England, Trans IMM, Sect. C, Vol. 103, C162-167 (1994).
7. Burns M.J., Coates G., Barnard L., The Use of Jameson Cell Flotation Technology at Cleveland Potash, p290-300 in: Proceedings of International Fertilizer Association (IFA) Technical Conference, Amman, Jordan, 2-6 October 1994, pp492, IFA, Paris, France (1994).

8. Burns M.J, Pearson J. and Harrison M.E. Operating Experience and Savings with Jameson Cell Technology at Cleveland Potash, paper in press.
9. Nemedy L., Factors Influencing Froth Stability in Potash Processing, p605-609 in: Proceedings of First International Potash Technology Conference, Saskatoon, Saskatechwan, Canada, 3-5 October 1983, ppS87, Pergammon Press, Oxford, England (1984).

ФЛОТОМАШИНА ДЖЕЙМСОН: ВОЗВРАЩЕНИЕ В ПРИМЕНЕНИЕ ДЛЯ РУД ЦВЕТНЫХ МЕТАЛЛОВ С УСОВЕРШЕНСТВОВАННОЙ КОНСТРУКЦИЕЙ И ТЕХНОЛОГИЧЕСКОЙ СХЕМОЙ

М.Ф. Янг[1], К.Э. Барнс[2] , Дж.С. Андерсон[2] и Дж.Д. Пиз[2]

[1] Xstrata Zinc
Mount Isa, Qld,
Australia

[2] Xstrata Technology
87 Wickham Tce
Brisbane QLD 4000
Australia
+61 7 3833 8500

Ключевые слова: флотация, флотомашина Джеймсон

АВТОРЕФЕРАТ

Флотомашина Джеймсон была разработана на руднике Маунт Айза в конце 1980-х годов в качестве решения, позволяющего преодолеть недостатки, присущие флотационным колоннам. Первые установки для руд цветных металлов имели неоднозначные показатели. Тестирования и экспериментальная эксплуатация показали, что при правильной эксплуатации повышались технологические показатели, однако эффективность первых установок страдала от ряда проблем при эксплуатации и обслуживании. Нужны были более надежные установки, а также требовалось более глубокое понимание того, как успешно интегрировать флотомашины в технологическую схему предприятия.

В 1990-е годы флотомашины Джеймсон получили признание в таких применениях, как флотация угольных шламов и удаление органических примесей в технологии SXEW, и стали стандартом в Австралии для таких применений. Эти установки имели усовершенствованную конструкцию и материалы, улучшенную технологическую схему, низкую степень износа и автоматический контроль, что обеспечивало хорошую эффективность при минимальном вмешательстве операторов.

Эти усовершенствования позволили заново оценить возможность применения флотомашин Джеймсон для руд цветных металлов. Недавние новые установки указывают на значительные преимущества в так называемых "гибридных" циклах, сочетающих преимущества флотомашин Джеймсон и традиционные флотомашины, обеспечивая более высокую общую эффективность цикла при меньшей занимаемой площади по сравнению с любой из технологий, взятой в отдельности.

В настоящей статье рассматривается опыт успешного применения для цветных металлов и объясняется повышенная эффективность цикла с использованием минералогии по классам. Также описываются новые методы проектирования технологических схем.

ВВЕДЕНИЕ

Флотомашина Джеймсон сочетает в себе новый метод взаимодействия между воздухом и пульпой, при котором падающая струя естественным образом вовлекает воздух, обеспечивая высокую концентрацию воздуха относительно объема пульпы, малый размер пузырьков и тесный контакт между пузырьками и частицами материала (рис. 1). Маленькие пузырьки (0,3-0,5 мм) образуются непрерывно, и в аэраторе быстро (6-10 секунд) возникает тесный контакт между пузырьками и частицами. Как следствие, флотомашина Джеймсон обладает большой интенсивностью и высокой скоростью флотации минералов, особенно тонкой фракции. Поскольку контакт между пузырьком и частицей происходит в аэраторе, задача флотомашины заключается в разделении пузырьков и пульпы, благодаря чему ее объем небольшой по сравнению с колоннами. Высокая скорость флотации вследствие интенсивной аэрации означает высокую производительность в пересчете на площадь поверхности и делает промывку пену привлекательным способом повышения содержания в концентрате. Энергопотребление ниже, чем в аналогичных механических или колонных флотомашинах (энергию расходует только насос питания; нет ни нагнетателя, ни компрессора), а основными изнашиваемыми деталями является сопло и насос питания. Основы эксплуатации флотомашины Джеймсон описаны многими авторами, в том числе в Clayton, Jameson and Manlapig (1991)

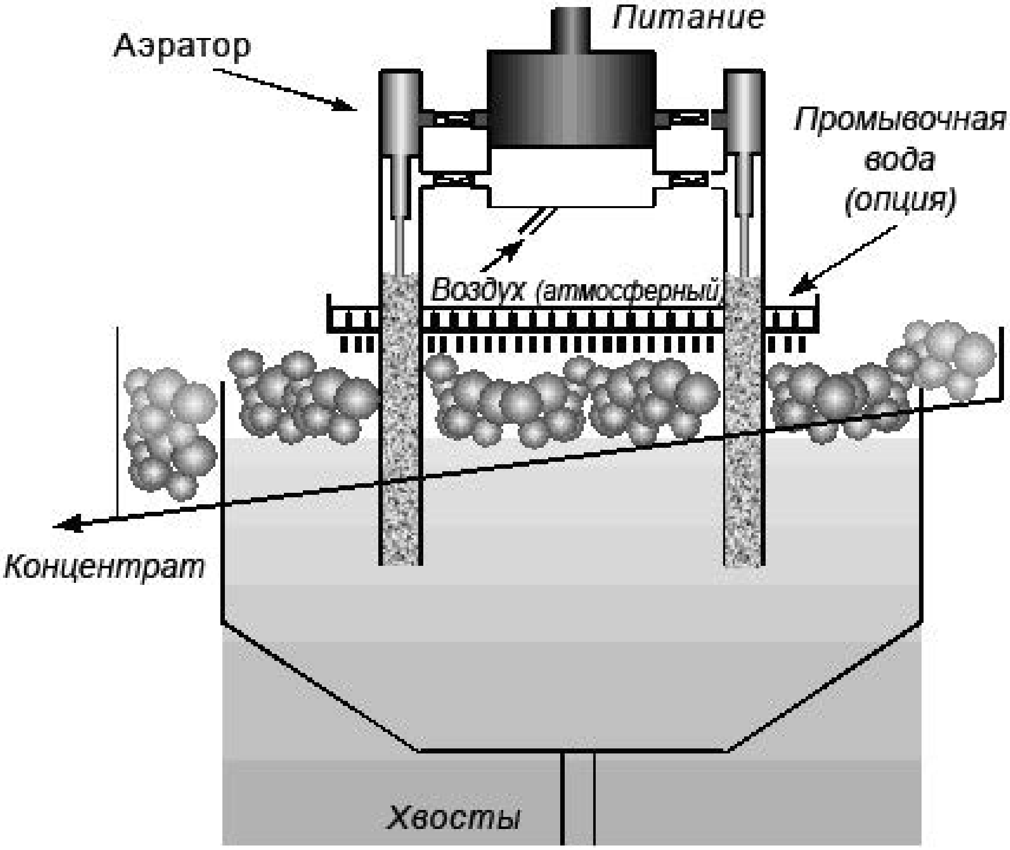


Рисунок 1. Флотомашина Джеймсон

Первыми внедренными промышленными флотомашинами Джеймсон были установки перечистной флотации свинца, описанные в Jameson and Manlapig (1991). Эта установка показала огромную разницу в кинетике флотации между флотомашиной Джеймсон, механической флотомашиной и колонной флотомашиной. Кроме того, размер, занимаемая площадь и стоимость флотомашины Джеймсон были намного меньше по сравнению с традиционной механической флотомашиной и флотационной колонной. Это наглядно показано на рис. 2, на котором представлены установленные рядом колонны и флотомашины Джеймсон близкой производительности на свинцово-цинковой обогатительной фабрике Маунт Айза. Ниже не только стоимость самой машины, но и стоимость ее эксплуатации, поскольку в ней меньше воздухораспределителей, меньше затраты на перекачивание и используется подсасываемый, а не сжатый, как в колоннах, воздух.

Первые флотомашины Джеймсон использовались для перечистной флотации благодаря способности машины производить концентраты с высоким содержанием за одну стадию (Jameson, Harbort, Riches, 1991). Однако в этих установках извлечение было довольно скромным - 50-80%. Для более высокого общего извлечения в цикле флотомашины Джеймсон нужно было эксплуатировать в замкнутом цикле с с другими флотомашинами.

Эти выводы подтверждаются работами Riches (1991) и Harbort (2002), в которых подчеркивается, что за одну стадию флотомашина Джеймсон способна получить из питания основной флотации готовый концентрат с извлечением от 60% до 70%, что значительно выше результатов тестирования механических флотомашин.

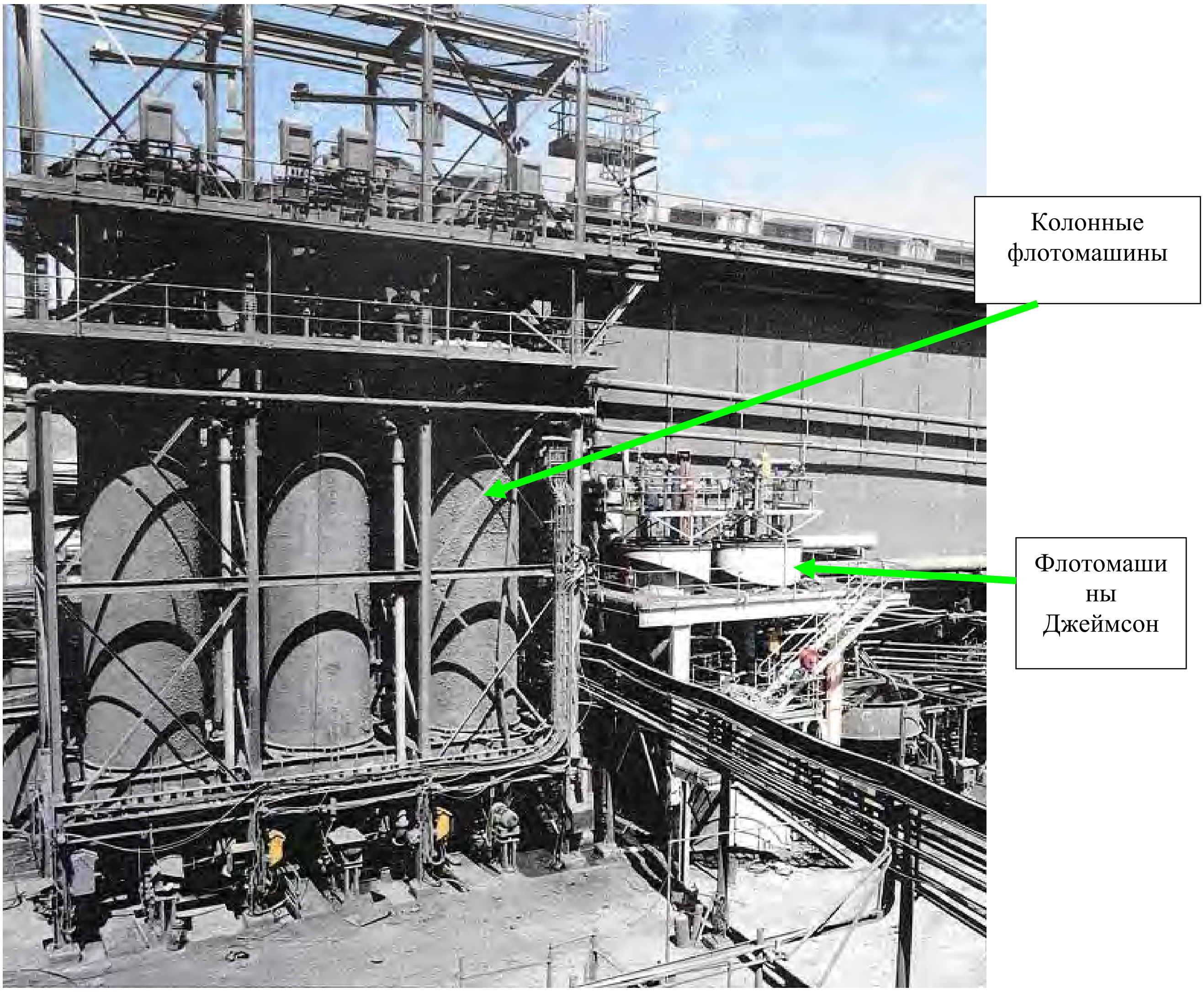


Рисунок 2. Сравнение габаритов колонных флотомашин и флотомашин Джеймсон аналогичной производительности

Несмотря на существенные технологические и экономические преимущества флотомашины Джеймсон, первым установкам для цветных металлов не хватало эксплуатационной надежности. Пилотные тестирования стабильно давали хорошие результаты, но ни конструкция промышленных флотомашин, ни методика их интеграции в цикл еще не были доведены до совершенства.

Наибольшая эксплуатационная проблема заключалась в том, что питание должно было иметь постоянный объем и давление, чтобы флотомашина Джеймсон могла выйти на устоявшийся режим работы. Первые установки не могли обеспечить это - питание было однопроходным или в замкнутом цикле с флотомашинами основной флотации. Расход питания был непостоянным, в результате чего аэратор работал нестабильно, и это сказывалось на металлургической эффективности и стабильности флотомашины. Если расход питания падал, отдельные аэраторы приходилось отключать, чтобы поддерживать постоянную скорость в остальных аэраторах. Это решение было непрактично. Кроме того, высокая интенсивность контакта с пузырьками в аэраторе означает не только быструю флотацию, но и меньше возможностей для образования контакта между пузырьком и частицей материала. В результате, флотомашина Джеймсон быстро производит высококачественный концентрат, но в одной машине невозможно добиться высокого извлечения (как правило, 50-80% за одну стадию при флотации руд цветных металлов). Чтобы получить высокое общее извлечение в цикле, необходима либо последовательная установка нескольких машин (что сокращает преимущества с точки зрения капитальных затрат), либо эксплуатация машины в замкнутом цикле с флотомашинами основной флотации.

Хотя в то время отсутствовал опыт правильного применения флотомашины Джеймсон для руд цветных металлов, в 1990-е годы она пользовалась большим успехом при флотации тонкой угольной фракции и при удалении органических примесей в технологии SXEW (Jameson, Goffinet, Hughes, 1991; Dawson, Jackson, 1995). Этот успех был связан со способностью флотомашины Джеймсон получать конечный продукт при очень высоком извлечении (для угля 95-98%) за одну стадию. Кроме того, высокая интенсивность флотомашины Джеймсон позволяла перерабатывать огромные объемы материала на небольшой по сравнению с другими технологиями установке. К настоящему времени флотомашина Джеймсон стала стандартным решением для таких применений.

Успех в угольной промышленности побудил обратиться к исследованию неудовлетворительных показателей эффективности при флотации руд цветных металлов. Разница в эффективности между флотацией угольных шламов и флотации полиметаллов заключалась в том, что если при флотации угля флотомашина Джеймсон могла обеспечить извлечение 95-98%, то при флотации полиметаллов извлечение было всего 50-80%. Флотация полиметаллов имеет более медленную дифференциальную кинетику и более высокий коэффициент обогащения по сравнению с углем. Различия в кинетике и коэффициенте обогащения требуют, чтобы при обогащении полиметаллов флотомашина Джеймсон работала в замкнутом цикле с другими флотомашинами (механическими или Джеймсон), если необходим высокий общий показатель извлечения в цикле.

СОВЕРШЕНСТВОВАНИЕ СХЕМЫ ЦИКЛА С ФЛОТОМАШИНОЙ ДЖЕЙМСОН

Применительно к рудам цветных металлов нужно было решить две основные задачи:

- Обеспечить постоянный объем питания флотомашины Джеймсон несмотря на обычное для промышленных установок непостоянство.
- Спроектировать малозатратный в целом цикл, учитывая низкое извлечение за один проход и используя все возможности, связанные с быстрой флотацией и высоким содержанием в концентрате.

Первая задача была решена рециркуляцией хвостов. Установки проектируются с расчетом на переработку объема, на 30-40% превышающего максимальную расчетную производительность. Питание флотомашины обеспечивает насос, работающий с постоянной скоростью и подающий постоянный объем и давление в аэраторы. Для поддержания постоянного объема производится рециркуляция хвостов в емкость питания. Даже если свежее питание отсутствует полностью, флотомашина продолжает работать на полностью рециркулируемом объеме хвостов с сохранением размера пузырьков (это также обеспечивает защиту насоса питания). Еще одним преимуществом этой системы является повышенное извлечение при первом проходе, поскольку, как правило, 40% питания имеет "второй шанс" в аэраторе. Столь простое решение существенно изменило технологичность флотомашин Джеймсон: они производят пузырьки постоянного размера, поддерживают постоянную эффективность и автоматически подстраиваются под условия цикла, не требуя постоянного вмешательства со стороны оператора. Механизм рециркуляции стал неотъемлемым элементом флотомашины Джеймсон и может быть внутренним (IRC), внешним (ERM) (рис.3) и внешним независимым (Cowburn, Stone, Bourke, Hill, 2005).

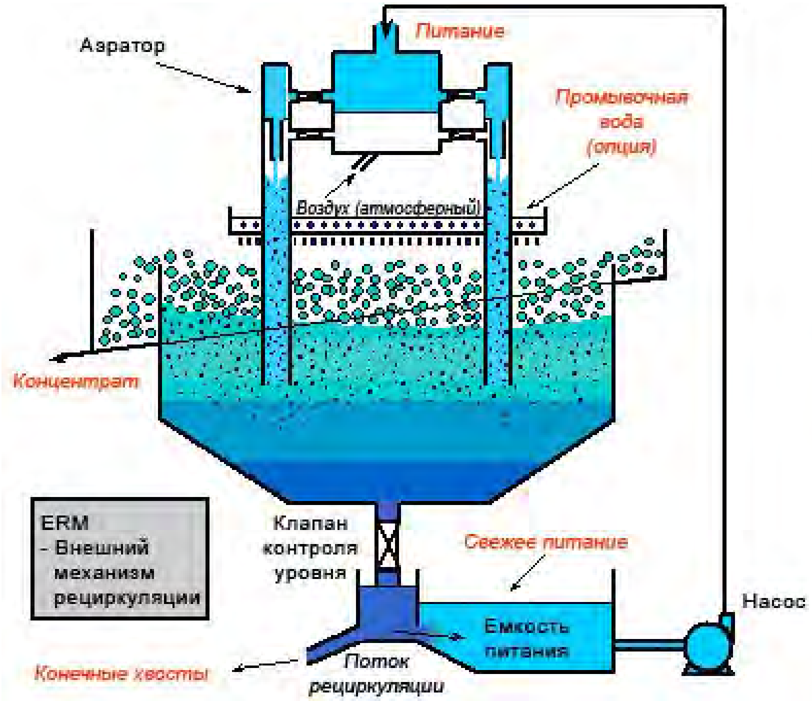


Рисунок 3. Флотомашина Джеймсон с внешним механизмом рециркуляции для стабилизации расхода питания аэратора.

К концу 1995 года на медной обогатительной фабрике Philex Mining Corporation весь цикл был оборудован флотомашинами Джеймсон. (Harbort, Murphy, Budod, 1997). Эта фабрика стала первым применением на рудах цветных металлов, на которой была использована система рециркуляции хвостов (ERM). Этот проект повлек разработку больших типоразмеров флотомашин Джеймсон, и значительное увеличение производительности цикла с флотомашиной Джеймсон. Система ERM получила дальнейшее развитие при проектировании обогатительной фабрики Алумбрера, на которой весь цикл перечистной флотации был основан на флотомашинах Джеймсон (Harbort, Murphy, Launder, Miranda, 2000). К 1999 году внутренний механизм рециркуляции (IRC) применялся на свинцово-цинковых (Young, Pease, Fisher, 2000) и медных обогатительных фабриках (Carr, Harbot, Lawson, 2003) компании Mount Isa Mines. Внутренняя система рециркуляции хвостов позволяла снизить стоимость монтажа, делая установки более компактными. (рис. 4)

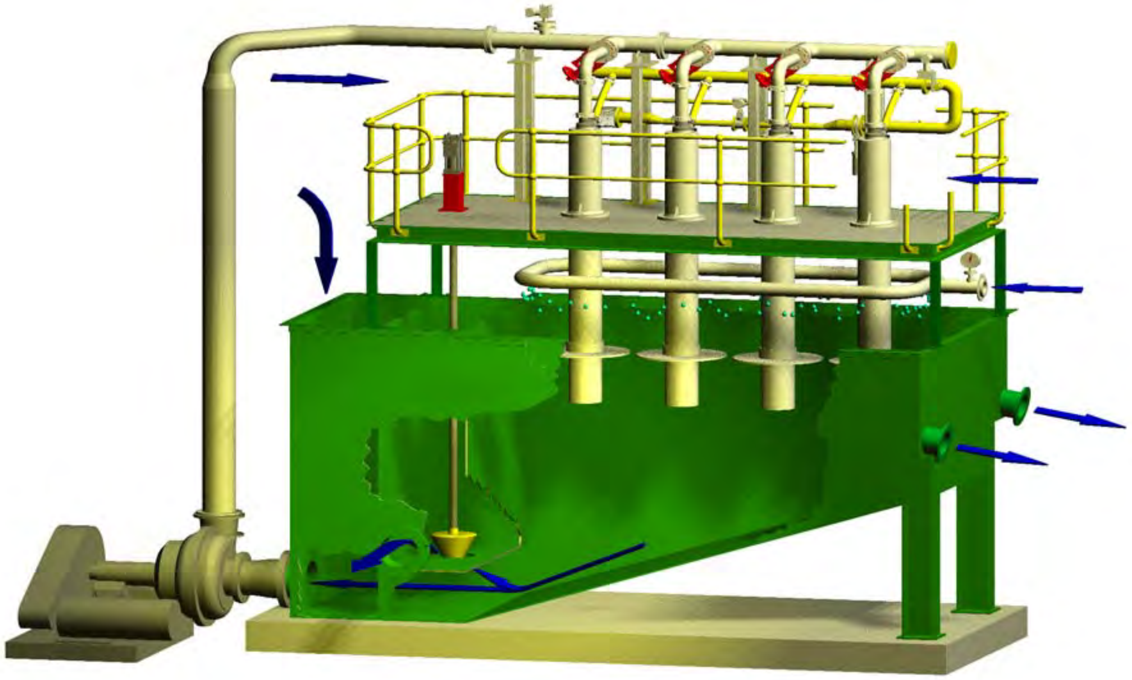


Рисунок 4. Флотомашина Джеймсон с внутренним механизмом рециркуляции (IRC) для стабилизации расхода питания аэратора.

Одновременно с внедрением рециркуляции хвостов для стабилизации питания флотомашины Джеймсон продолжались работы с целью повысить технологичность аэратора - "сердца" флотомашины Джеймсон. В 2000 году был представлен аэратор Mark 3 (рис. 5), в котором использовалось вдвое меньше деталей, благодаря чему все они могли быть расположены вне аэратора для удобства доступа.

Замена диафрагмы соплом пульпы увеличила срок службы деталей, а улучшенная конструкция подачи пульпы повысила коэффициент расхода и снизила потребление энергии насосом питания на 10-15%. Кроме того, расположение сопла пульпы по сравнению с диафрагмой увеличило полезную длину аэратора на 15%, что увеличило время пребывания в зоне смешивания и сделало возможной работу при более высоком отношении воздух-пульпа. Лабораторные тестирования показали, что увеличенная длина аэратора повышает вовлечение воздуха при том же значении вакуума (Cowburn et al, 2005).

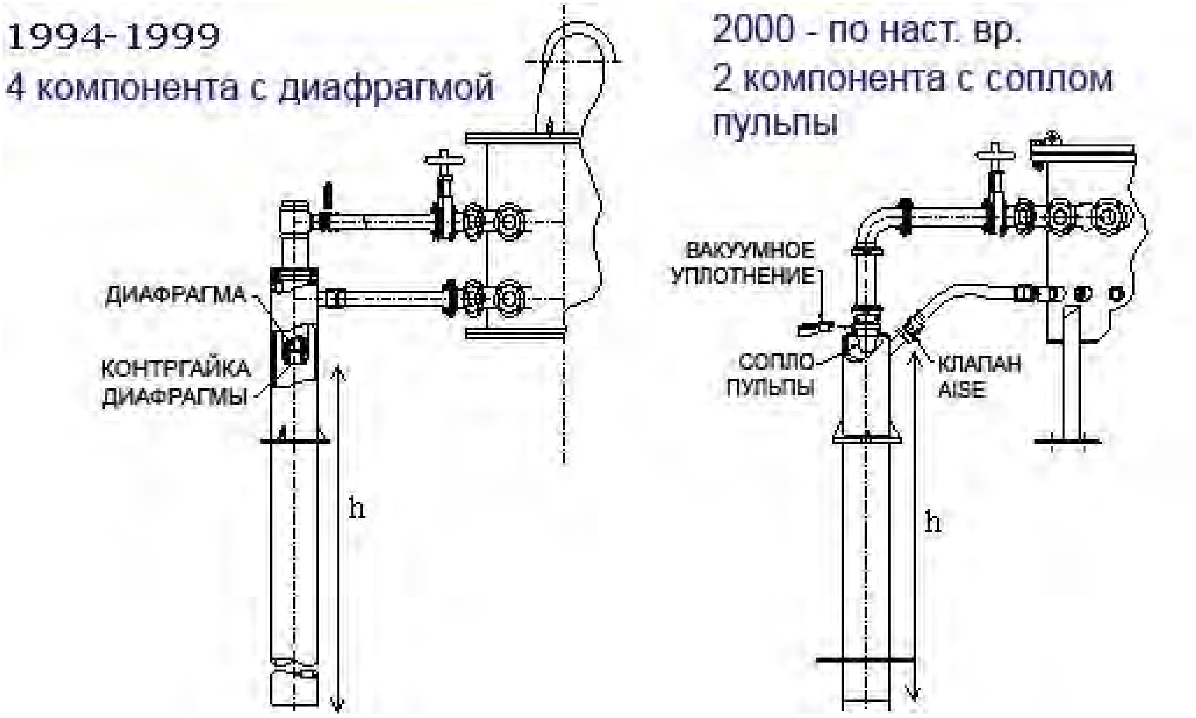


Рисунок 5. Аэратор старой и новой конструкции

Тенденция к увеличению производительности флотомашин повлекла разработку больших типоразмеров флотомашин Джеймсон: самая большая из установленных для руд цветных металлов флотомашин Джеймсон, B6500/22, перерабатывает до 22 000 т в сутки, а самая большая флотомашина Джеймсон, J7200/10, перерабатывает 40 000 т в сутки. Проектирование флотомашин Джеймсон большей производительности не представляет труда, поскольку диаметр флотомашины и количество аэраторов могут быть с легкостью увеличены, при этом общая высота установки остается неизменной.

Вторая задача, которую нужно было решить, чтобы флотомашины Джеймсон могли применяться для флотации руд цветных металлов, заключалась в проектировании установок, учитывая низкое извлечение за один проход, одновременно используя такие преимущества, как быстрая флотация и высокое содержание в концентрате, для эксплуатации в составе низкозатратного цикла. Для этого нужно было принять во внимание несколько принципиальных особенностей флотомашины Джеймсон. Во-первых, флотомашина Джеймсон производит высококачественный продукт. Это высокоинтенсивная флотомашина, использующая для флотации подсасываемый воздух. Высокая скорость сдвига образует мелкие пузырьки в аэраторе. Для образования мелких пузырьков флотомашина Джеймсон должна работать с низким или умеренным отношением воздух-пульпа (0,2 - 0,5) и умеренным или высоким вакуумом (15 - 30 кПа). При таких условиях эксплуатации происходит избирательное извлечение из пульпы минералов с высокой кинетикой и более четкое разделение минералов с высокой и низкой кинетикой

Флотомашина Джеймсон имеет систему промывки пены и умеренную глубину пены, что позволяет использовать промывку пены, чтобы минимизировать механический унос. Такая конструкция флотомашины позволяет получать высококачественный концентрат за одну стадию флотации. При эксплуатации коэффициент промывочной воды должен быть умеренно положительным (отношение добавляемой в пену промывочной воды к воде, извлекаемой с концентратом). Коэффициент на уровне 1,2 позволяет свести механический унос к минимуму, а более высокий коэффициент (до 1,5) может помочь минимизировать извлечение композитных частиц, имеющих слабую связь с пузырьками, что дает концентрат максимального содержания.

Хотя высокая интенсивность, небольшой размер пузырьков и промывка пены позволяют флотомашине Джеймсон производить высококачественный концентрат за одну стадию флотации, из-за этих особенностей извлечение за одну стадию является низким или умеренным. Низкое извлечение можно повысить за счет увеличения коэффициента рециркуляции, увеличив отношение воздуха к свежей пульпе, что опровергает распространенное заблуждение, будто извлечение можно повысить, уменьшив вакуум и таким образом увеличив расход воздуха, что в действительности приводит к увеличению размера пузырька и снижению эффективности. В промышленных условиях коэффициент рециркуляции легко увеличить, установив дополнительные аэраторы.

Принцип работы традиционных флотомашин и колонн отличается от принципа работы флотомашин Джеймсон, и эти различия должны быть учтены при проектировании цикла флотации руд цветных металлов. Если во флотомашинах Джеймсон для увеличения отношения свежей пульпы к воздуху увеличивают количество аэраторов, то в механических флотомашинах увеличивают расход сжатого воздуха. Хотя мелкие пузырьки образуются в лабораторных механических флотомашинах, в промышленных машинах интенсивность сдвига ниже и пузырьки крупнее. Флотомашина Джеймсон производит мелкие пузырьки как в лабораторных, так и в промышленных условиях, поскольку интенсивность сдвига не меняется.

Принципиальные различия между традиционными флотомашинами и флотомашинами Джеймсон означают, что технологическая схема должна быть составлена таким образом, чтобы использовать все преимущества соответствующих флотомашин, и оптимизировать реакцию руды с учетом кинетики и обогатимости.

Со строительной точки зрения масштабирование флотомашины Джеймсон намного проще масштабирования традиционной флотомашины. Для цикла, основанного на флотомашинах Джеймсон, достаточно большого типоразмера флотомашины с правильным количеством аэраторов, работающих при низком отношении воздуха к свежей пульпе, и не требуется дополнительных машин, как при использовании традиционных флотомашин.

ПРОЕКТИРОВАНИЕ ЦИКЛА ДЛЯ МАКСИМАЛЬНОЙ ЭФФЕКТИВНОСТИ ФЛОТАЦИИ РУД ЦВЕТНЫХ МЕТАЛЛОВ

При флотации руд цветных металлов флотомашины Джеймсон имеют целый ряд применений. Они могут быть установлены в новых технологических линиях или при модернизации действующих предприятий. Будучи машинами флотации, позволяющими получать конечный концентрат за одну стадию, они прекрасно подходят для установки в циклах перечистной флотации, однако их способность производить высококачественный концентрат при умеренном извлечении делает возможным их применение и в основной флотации.

При проектировании цикла флотации руд цветных металлов важно правильно подобрать флотомашину с учетом выполняемой задачи, поскольку флотомашина Джеймсон не сможет показать желаемую эффективность, если она будет использована для выполнения несвойственных ей задач. Машина должна соответствовать поставленной задаче.

Для каких же задач можно использовать флотомашину Джеймсон?

1. **Флотомашина Джеймсон в качестве машины основной перечистной флотации в замкнутом цикле с традиционным циклом основной и контрольной флотации.** Одним из первоначальных применений флотомашины Джеймсон была одностадийная перечистная флотация в замкнутом цикле с циклом основной и контрольной флотации (рис. 6). Эта технологическая схема успешно работает, производя концентрат с высоким содержанием, если быстрая кинетика флотации и умеренный коэффициент обогащения способствуют повышению содержания концентрата основной флотации. Для компенсации низкого извлечения при первом проходе может потребоваться значительная рециркулирующая наргрузка (до 300%), хотя хорошо высвобожденные минералы с высокой кинетикой могут обогащаться и при небольшой рециркулирующей нагрузке. Извлечение при флотации можно также повысить, установив вторую флотомашину Джеймсон последовательно.

   Этот цикл использовался на руднике Маунт Айза в цикле перечистки медных шлаков и в цикле перечистки перед флотацией.

Рисунок 6. Технологическая схема цикла основной перечистной флотации с использованием флотомашины Джеймсон

2. Флотомашина Джеймсон для перечистной скоростной флотации.

Применение флотомашины Джеймсон для перечистной скоростной флотации идеально при расширении и модернизации предприятий, на которых традиционный цикл перечистной флотации уже установлен (рис. 7). Этот цикл использует все преимущества обеих технологий: флотомашины Джеймсон собирают быстрофлотируемый материал для получения высококачественного тоннажного концентрата, а традиционные флотомашины обеспечивают конечное извлечение. Эти задачи выполняются более компактно и при меньших капитальных затратах, чем при использовании любой из технологий, взятой отдельно. Флотомашины Джеймсон являются недорогим решением при расширении для снижения нагрузки и позволяют повысить как общее содержание, так и извлечение. Промывка пены повышает содержание в концентрате за счет уменьшения уноса. За счет высокой производительности и компактности установки флотомашина Джеймсон позволяет экономично осуществлять промывку пены в отношении значительной доли концентрата. Более того, благодаря уменьшению объема питания последующих традиционных перечистных флотомашин, они работают при меньшей плотности и меньшем пенообразовании, что уменьшает механический унос. Поэтому относительно небольшая площадь промывки пены может значительно уменьшить общий механический унос в таком "гибридном" цикле. Среди изменений, внесенных в цикл, - дополнительные флотомашины Джеймсон, хотя затраты на перекачивание могут сделать затратной установку более чем двух флотомашин Джеймсон последовательно.

Дополнительная флотомашина Джеймсон для скоростной перечистной флотации на свинцово-цинковой обогатительной фабрике Маунт Айза была модернизирована, как описано в Примере 1.

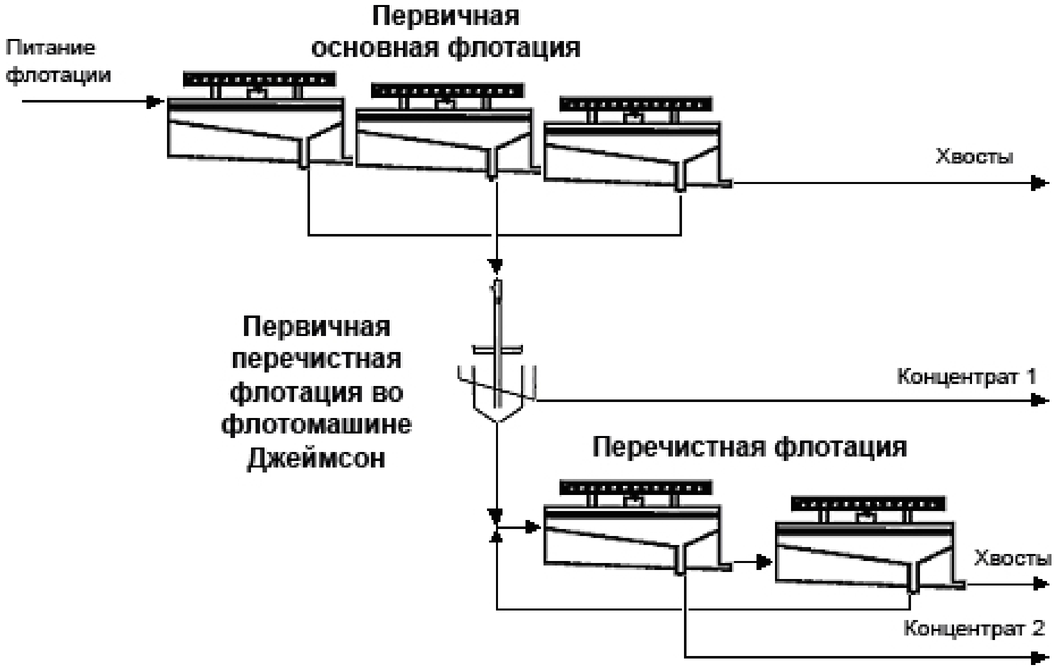


Рисунок 7. Технологическая схема цикла скоростной перечистной флотации с использованием флотомашины Джеймсон

3. Флотомашина Джеймсон в качестве машины скоростной основной флотации перед циклом традиционной основной и контрольной флотации

Использование флотомашин Джеймсон для скоростной основной флотации идеально подходит как для действующих предприятий, так и для проектов расширения. Флотомашины Джеймсон имеют высокую производительность и позволяют получать готовый высококачественный концентрат за одну стадию флотации (рис. 8). Благодаря их компактности, они могут быть с легкостью установлены на действующих предприятиях и являются идеальным решением при модернизации предприятий. Возможность установки двух флотомашин Джеймсон последовательно делает их особенно эффективными в этом качестве.

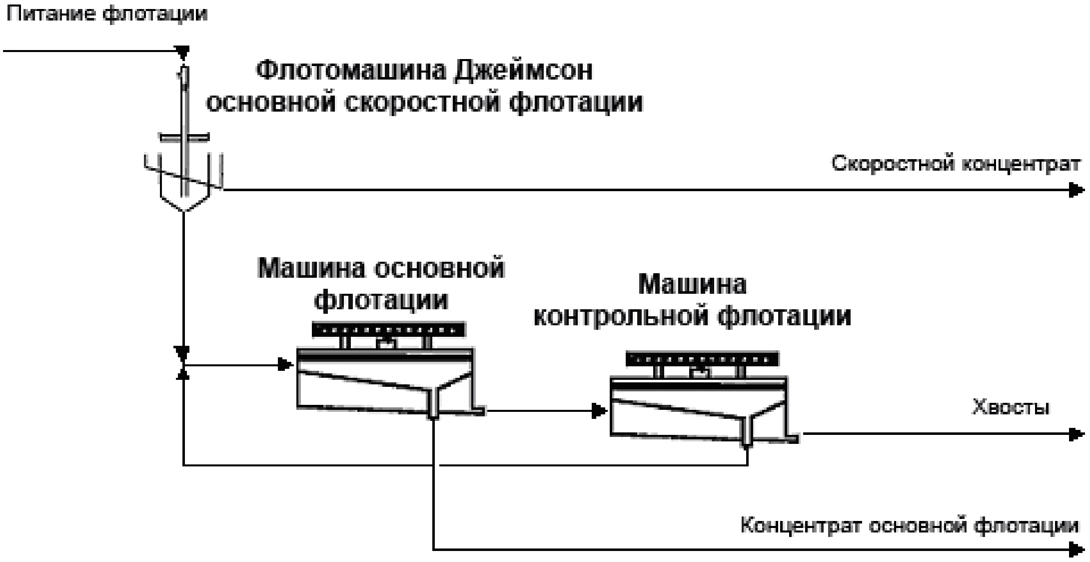


Рисунок 8. Технологическая схема цикла скоростной основной флотации с использованием флотомашины Джеймсон

Технологические схемы, перечисленные выше, описывают далеко не все возможные применения флотомашины Джеймсон и лишь демонстрируют широкий диапазон ее возможностей. Одновременное применение флотомашин Джеймсон в циклах основной и перечистной флотации позволяет еще более полно использовать их преимущества.

НЕКОТОРЫЕ ПРИМЕРЫ РЕАЛИЗОВАННЫХ ПРОЕКТОВ

Проект 1: Цикл перечистной флотации свинца, Маунт Айза

Для обогащения руды месторождения Джордж Фишер на свинцово-цинковой обогатительной фабрике рудника Маунт Айза требовалось увеличить производительность цикла перечистной флотации свинца. Задача была решена путем установки флотомашины Джеймсон в качестве машины перечистной флотации свинца для переработки тонкой (P80 = 12 мкм) фракции продукта доизмельчения IsaMill и получения высококачественного свинцового концентрата (Young et al, 2000). Продукт IsaMill подается непосредственно в перечистную флотомашину Джеймсон, как показано на рис. 9. Флотомашина Джеймсон производит концентрат с содержанием свинца 60% при извлечении свинца 35% и уменьшает циркуляционную нагрузку по сравнению с традиционными трехстадийными машинами перечистной флотации свинца в замкнутом цикле. В сочетании с традиционными машинами перечистной флотации (обеспечивающими содержание свинца 51% при извлечении 45%), флотомашины Джеймсон обеспечивают получение готового концентрата с содержанием свинца 55%.

Для решения этой задачи была выбрана флотомашина Джеймсон модели E1732/4 с IRC – внутренней системой рециркуляции хвостов. Она имеет площадь поверхности 5,2 м2 и объем 8,2 м3. По сравнению с традиционными машинами перечистной флотации, имеющими площадь поверхности 156 м2 и объем 200 м3 и обеспечивающими извлечение свинца 45%, флотомашина Джеймсон позволяет извлекать 35% свинца при на порядок меньшей площади поверхности и емкости машины. Высокая эффективность флотомашины Джеймсон является следствием высокой интенсивности, малого размера пузырьков и эффективной промывки пены.

Высокая производительность флотомашины Джеймсон при небольшой площади поверхности делает промывку пены экономически обоснованной благодаря более высокому общему содержанию в концентрате. В результате уменьшается питание традиционных машин перечистной флотации, что позволяет эксплуатировать их при меньшей плотности и загрузке пены, снижая механический унос и повышая эффективность. Хотя эффективность цикла повысилась, сложно определить, в какой мере это связано с установкой флотомашины Джеймсон, а в какой - с повышенной производительностью IsaMill (они были установлены одновременно). Независимо от этого, общим следствие является повышение содержания свинца в концентрате на 5% до 55% при увеличении извлечения свинца на 5%.

Циркуляционная нагрузка между циклом основной флотации и циклом перечистной флотации/флотомашиной Джеймсон невелика и нередко составляет всего 110% содержания свинца в питании нового цикла перечистной флотации. Таким образом, при использовании данного сырья, при данном высвобождении и схеме цикла можно добиться высокого общего извлечения по фабрике при низкой циркуляционной нагрузке.

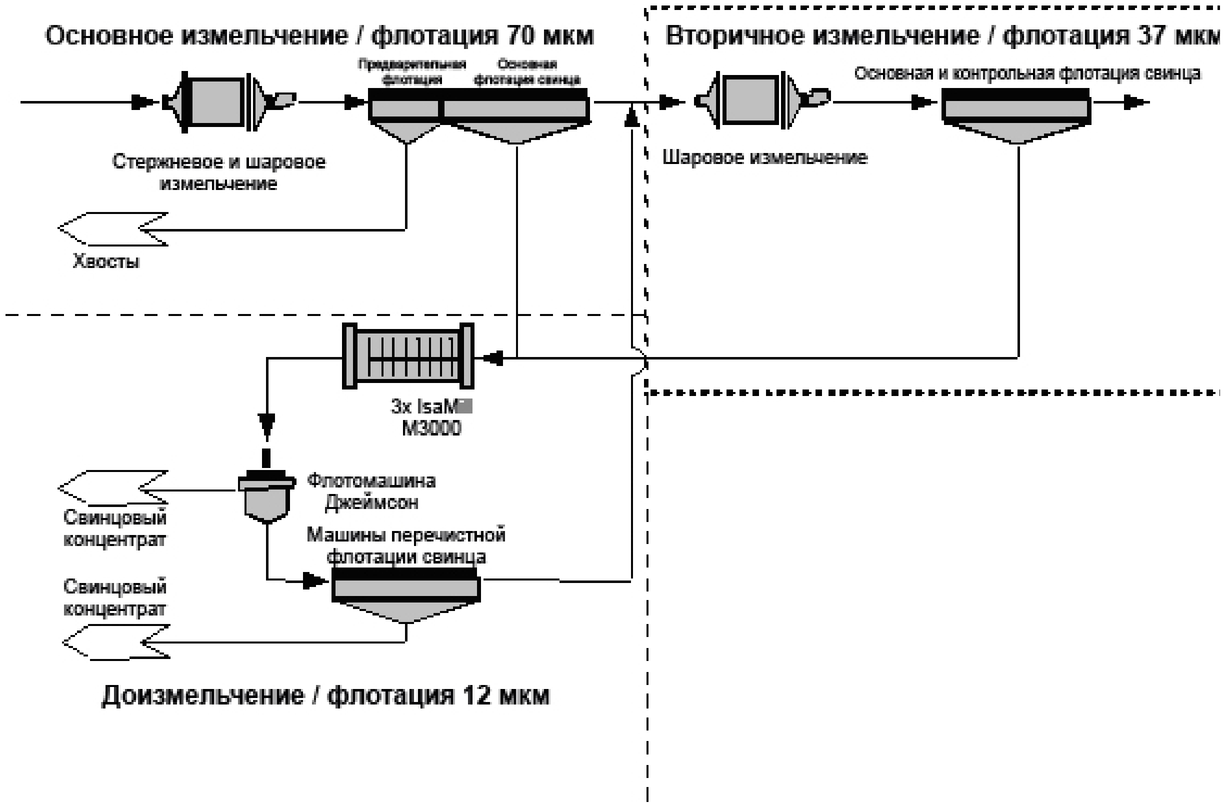


Рисунок 9. Технологическая схема цикла основной и перечистной флотации свинца, Маунт Айза

Флотомашина Джеймсон при использовании для перечистной флотации свинца очень хорошо извлекает мелкие, высвобожденные, быстро флотируемые частицы галенита. Как показано на рис. 10, 93% галенита представлено частицами крупностью менее 12 мкм, причем большинство из них не крупнее 6 мкм. Хотя флотомашина Джеймсон лучше извлекает такие тонкие частицы, традиционный цикл перечистной флотации лучше извлекает более крупные частицы (рис. 11). Флотомашина Джеймсон обеспечивает высокое извлечение крупной фракции при намного более высоком качестве концентрата (содержание свинца 60%) и более низком извлечении пустой породы по сравнению с традиционным циклом перечистной флотации (содержание свинца 51%). Одновременное применение флотомашины Джеймсон и традиционного цикла перечистной флотации позволяет эффективно извлекать как тонкий высвобожденный, так и крупный менее высвобожденный галенит, обеспечивая хорошее извлечение в конечный свинцовый концентрат при меньшем уносе тонких частиц.

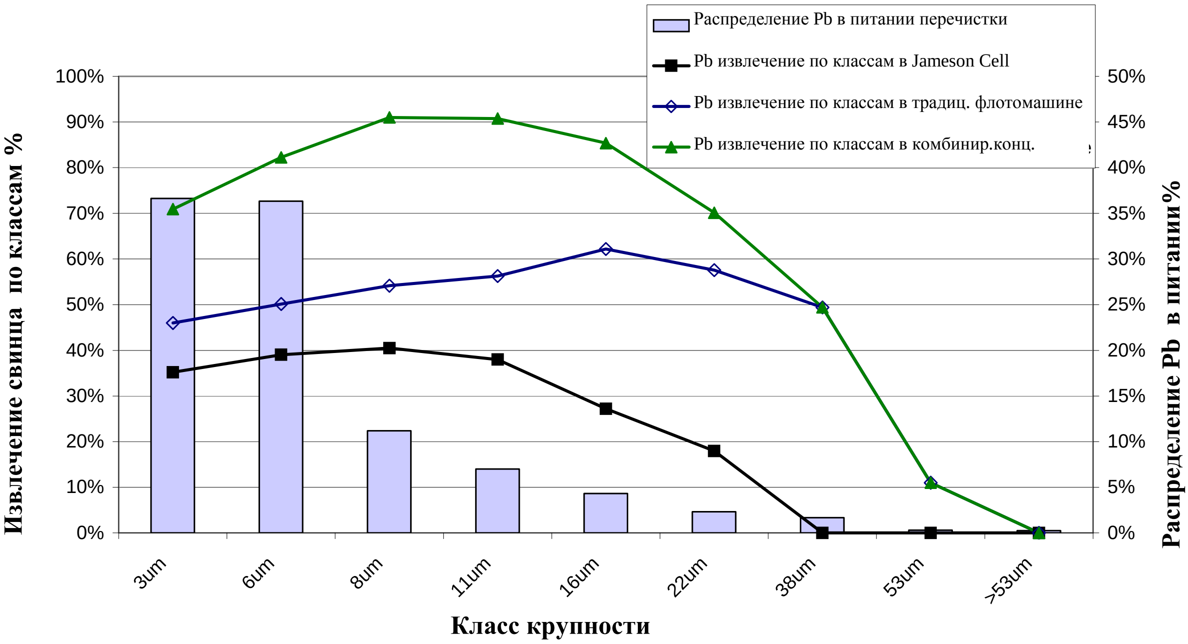


Рисунок 10. Зависимость извлечения от крупности, цикл перечистной флотации свинца на свинцово-цинковой обогатительной фабрике Маунт Айза

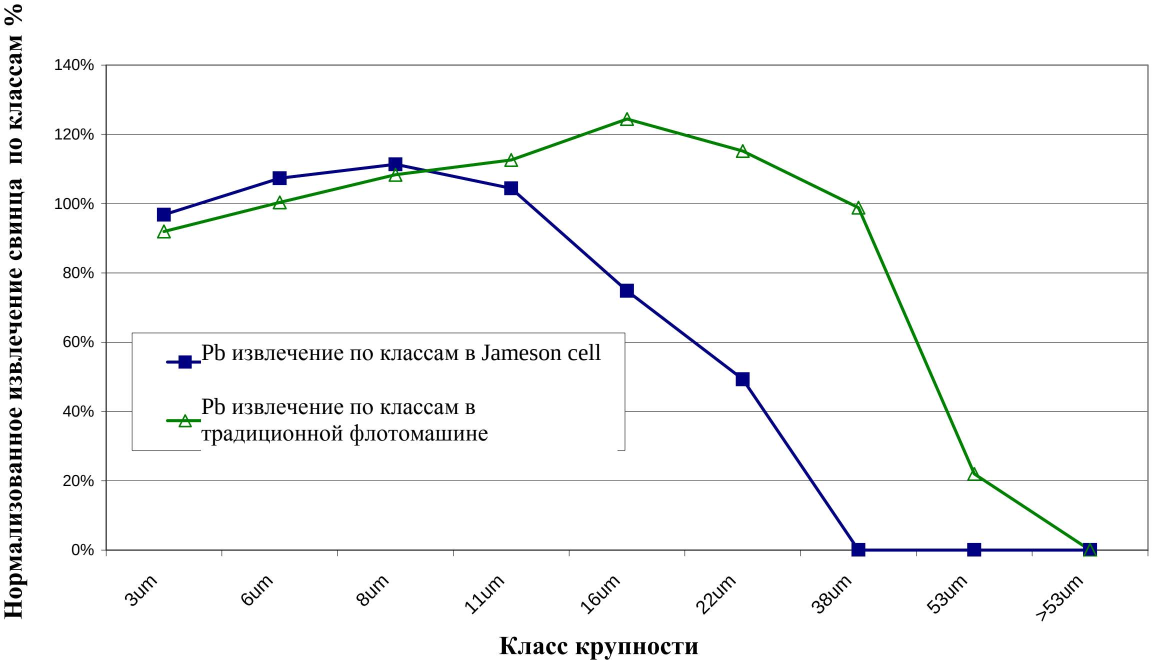


Рисунок 11. Нормализованная зависимость извлечения от крупности, цикл перечистной флотации свинца

До модернизации основным безрудным минералом, извлекаемом в цикле перечистной флотации свинца, был сфалерит, а основная часть потерь цинка в цикле флотации приходилась на бинарные частицы сфалерита и галенита, попадающие в свинцовый концентрат. Доизмельчение питания перечистной флотации свинца до 12 мкм высвободило значительную долю бинарных частиц, обеспечив лучшую сепарацию сфалерита в цинковый цикл (и более эффективную сепарацию пиритных примесей). На рис. 12 показана эффективная сепарация сфалерита из свинцового концентрата, в особенности во флотомашине Джеймсон. Очень хорошие показатели сепарации тонкого цинка и пирита во флотомашине Джеймсон по сравнению с традиционными флотомашинами указывают на то, что высокая интенсивность флотомашины Джеймсон позволяет оперативно извлекать быстро флотируемые минералы и эффективно отделять минералы с низкой скоростью флотации. По сравнению с традиционными флотомашинами, флотомашина Джеймсон увеличивает разницу в скорости флотации между быстро и медленно флотируемыми минералами.

Если бы установленные флотомашины Джеймсон имели большую производительность, сепарация сфалерита из свинцового концентрата была бы еще лучше.

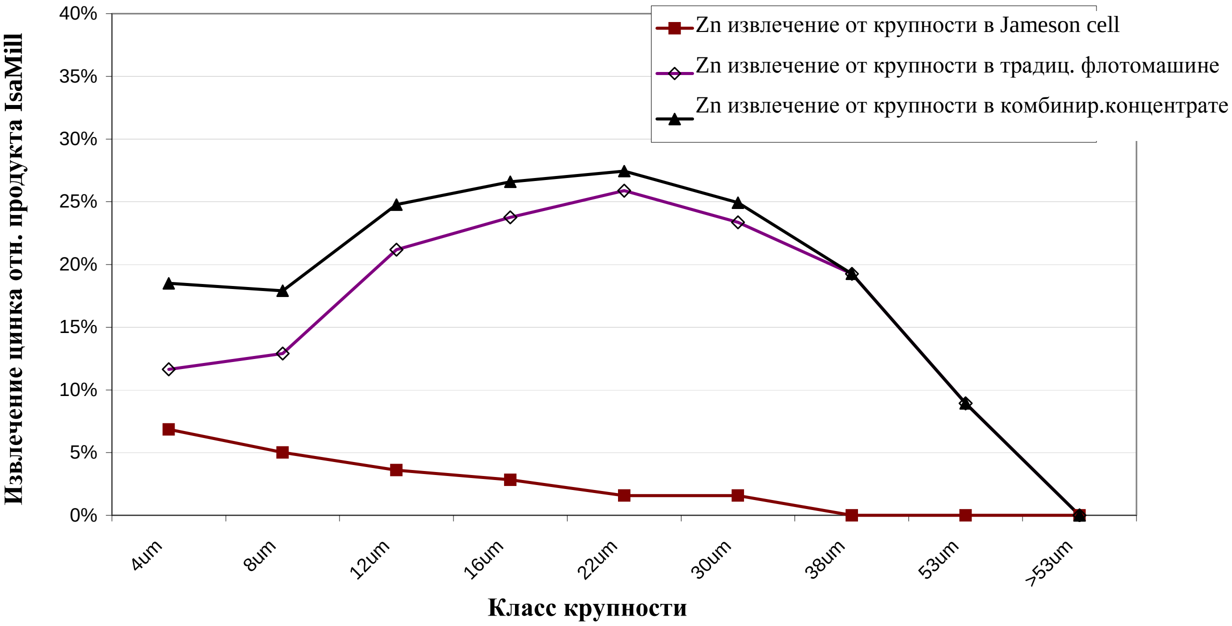


Рисунок 12. Извлечение цинка в зависимости от крупности в цикле перечистной флотации свинца

Проект 2: Эффективность флотомашины Джеймсон для предварительной флотации меди, Маунт Айза

Медные рудные тела на месторождении Маунт Айза содержат естественно флотируемый мелкозернистый углесодержащий пирит и естественно флотируемый тальк, загрязняющие медный концентрат. Примесь пирита снижает содержание медного концентрата, уменьшая производительность медеплавильного комбината и увеличивая потери меди в металлургические шлаки. Примесь талька увеличивает содержание магния в медном концентрате, что существенно снижает эксплуатационную готовность медеплавильного комбината и общий выход меди, повышая точку плавления шлаков, а соответственно и рабочую температуру (что, в свою очередь, ускоряет износ огнеупорной футеровки и сокращает срок службы плавильного цеха), а магний агрессивно воздействует на огнеупорную футеровку плавильных печей, также сокращая срок эксплуатации.

Первоначально переработка на медной обогатительной фабрике Маунт Айза состояла из флотации с последующей депрессией углесодержащего пирита и талька. Проведенные в 1990-е годы исследования показали, что предварительная флотация естественно гидрофобной пустой породы предпочтительна по сравнению с депрессией. Первый цикл предварительной флотации состоял из флотационных колонн. Эти колонны были переустановлены в 1996 году, однако увеличение количества талька и углесодержащего пирита в рудных телах обусловило необходимость возобновления цикла предварительной флотации. В конце 1990-х годов увеличение содержания примесей в концентрате и потерь меди в концентрат предварительной флотации стало причиной для установки цикла предварительной перечистной флотации с использованием флотомашины Джеймсон. Флотомашина Джеймсон (модель E2514/3 с IRC – внутренней системой рециркуляции хвостов) в качестве машины предварительной перечистной флотации была установлена согласно схеме, показанной на рис. 6, в рамках проекта расширения обогатительной фабрики в 2002 году (Carr, Harbort, Lawson, 2003).

Основным результатом установки флотомашины Джеймсон для предварительной перечистки стало уменьшение загрязнения медного концентрата естественно флотируемыми безрудными минералами. На рис. 13 показано, что примесь пирита в медном концентрате значительно уменьшилась после пуска флотомашины Джеймсон несмотря на увеличение содержания пирита в питании, произошедшее в этот период.

Потери меди в концентрат предварительной флотации также уменьшились. До установки флотомашины Джеймсон концентрат предварительной флотации содержал 2-2,5% меди, затем содержание снизилось до 1-1.5%. Количество естественно флотируемого материала в питании увеличилось, а выход в новом цикле удвоился в периоды высокого содержания естественно флотируемого безрудного материала в питании обогатительной фабрики при сохранении низких потерь меди.

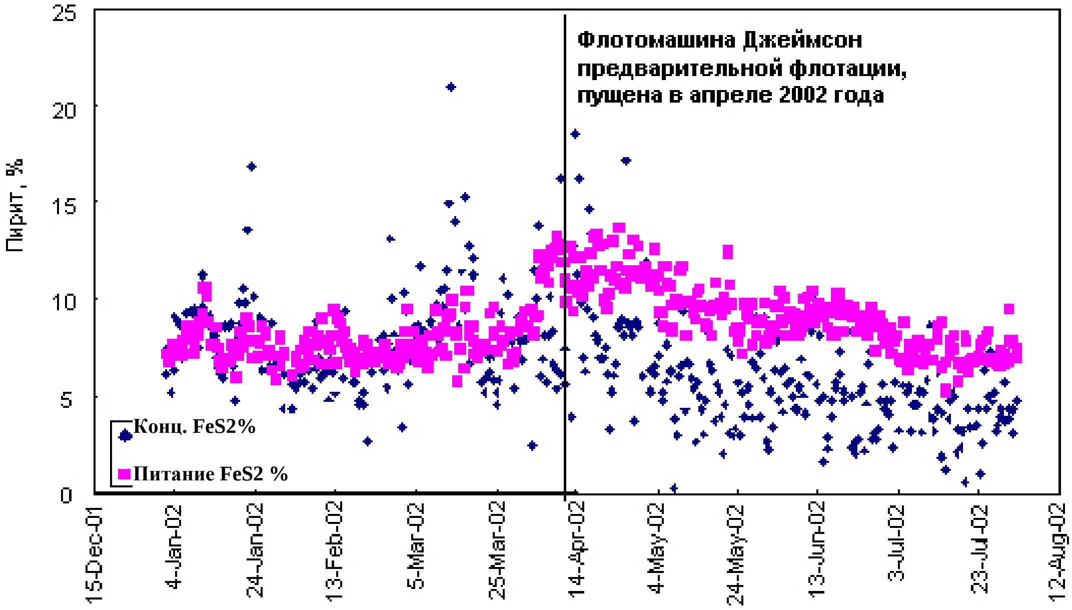


Рисунок 13. Загрязнение медного концентрата пиритом до и после пуска флотомашины Джеймсон

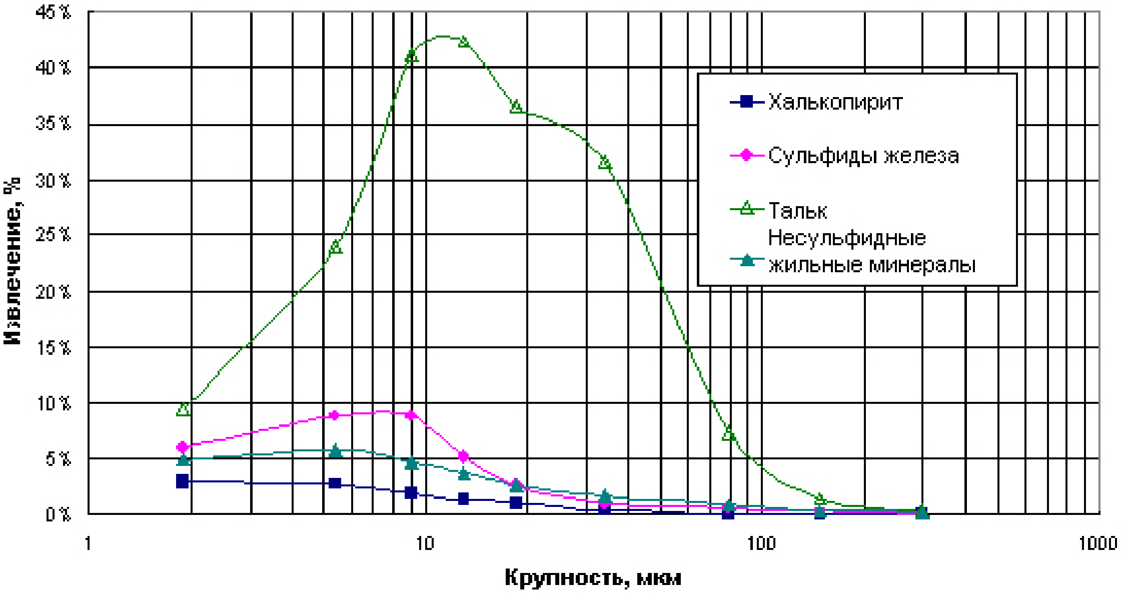


Рисунок 14. Извлечения в зависимости от крупности, цикл предварительной флотации

График зависимости извлечения от крупности в цикле предварительной флотации (рис. 14) указывает на хорошее отделение частиц талька средней крупности и естественно флотируемого углесодержащего пирита при минимальных потерях меди. Согласно результатам анализа концентрата предварительной флотации, содержание меди составило 1%. Необходимо отметить, что в настоящее время извлечение меди ниже, чем извлечение несульфидных безрудных минералов, что указывает на хорошую эффективность сепарации меди.

Проект 3: Эффективность цикла переработки медных шлаков, Маунт Айза

Медные металлургические шлаки перерабатываются на медной обогатительной фабрике партиями с использованием одной линии измельчения и основной флотации. Исторически для перечистки шлаков применялись традиционные машины перечистной флотации. После успешного опыта применения колонной флотации на обогатительной фабрике Хилтон цикл медных шлаков был переведен на колонную флотацию. В рамках модернизации медной обогатительной фабрики в 2002 году была установлена специальная перечистная флотомашина Джеймсон (модель E2532/6) для перечистки конвертерных шлаков и шлаков RHF (рис. 15) (Carr et al 2003).

Рисунок 15. Технологическая схема цикла переработки медных шлаков на обогатительной фабрике

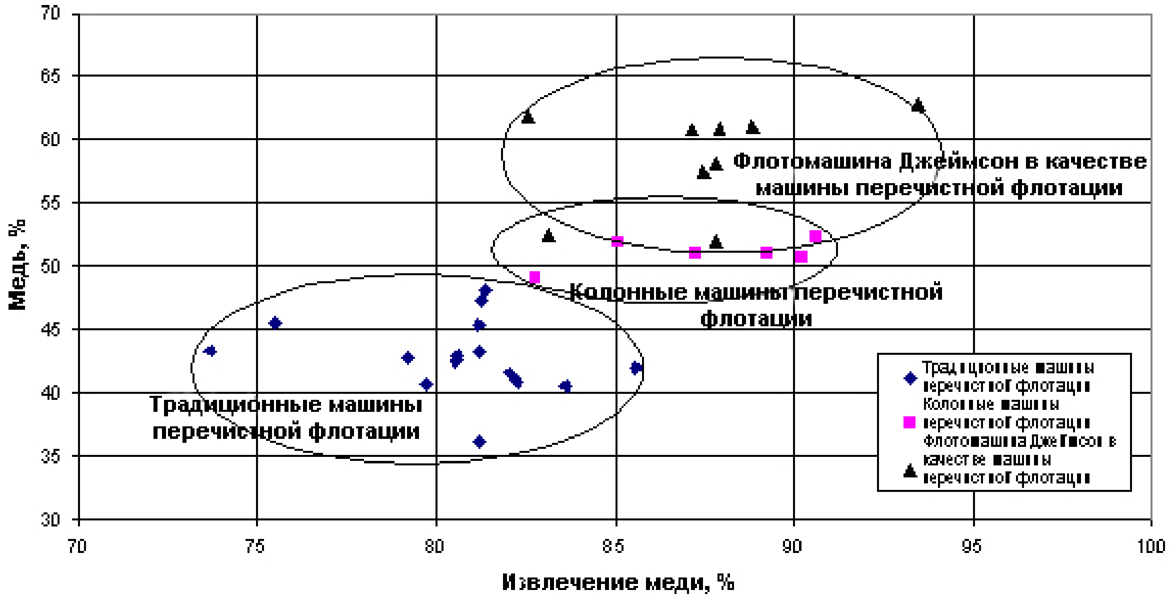


Рисунок 16. Эффективность цикла переработки медных шлаков в зависимости от типа перечистной флотомашины

Эффективность цикла переработки медных шлаков на протяжении нескольких лет и в зависимости от технологической схемы показана на рис. 16. Установка трех колонн диаметром 2,5 м и высотой 17 м для перечистки шлаков позволила повысить содержание и извлечение медного концентрата в цикле переработки медных шлаков, однако эффективность стала еще выше после замены колонн флотомашиной Джеймсон (E2532/6). В настоящее время концентрат шлака смешивается с халькопиритным медным концентратом для поддержания постоянного качества питания медеплавильного комбината.

Проект 4: Пилотный цикл основной скоростной флотации меди с использованием флотомашин Джеймсон, Маунт Айза

Пилотные испытания основной скоростной и основной флотации меди с использованием флотомашин Джеймсон (рис. 17) были проведены в составе программы испытаний в целях проекта модернизации флотации на медной обогатительной фабрике (Harbort, 2002). Данная технологическая схема будет реализована в составе второй очереди проекта модернизации.

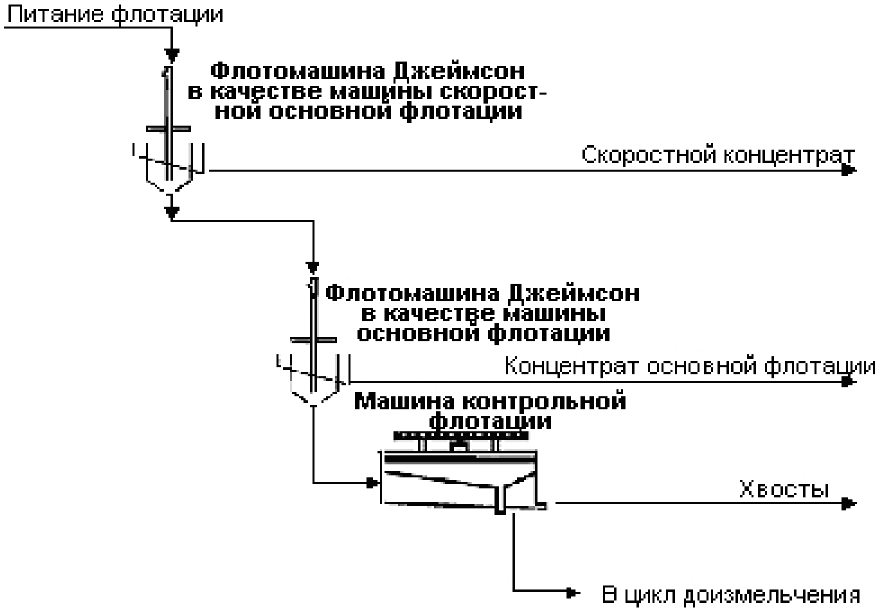


Рисунок 17. Цикл скоростной основной флотации и основной флотации с использованием флотомашин Джеймсон

Первый этап тестов скоростной основной флотации имел целью получить извлечение меди 60% при содержании меди в концентрате 30%. При испытаниях было получено среднее извлечение меди 63,37% при среднем содержании меди в концентрате 29,4% (рис. 18). 80% тестов показали результаты выше требуемого извлечения, хотя при немного более низком содержании в концентрате. Это означает, что концентрат скоростной основной флотации может направляться непосредственно в конечный концентрат со снижением нагрузки на циклы перечистной и повторной флотации.

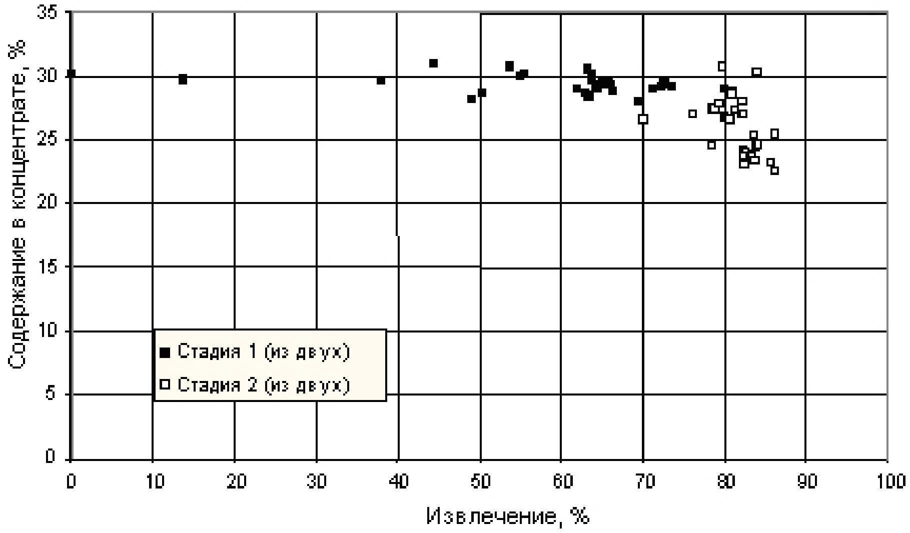


Рисунок 18. Испытания основной флотации меди в две стадии с использованием флотомашин Джеймсон

Использование флотомашины Джеймсон в основной флотации хвостов скоростной флотации повысило извлечение меди до 87% при общем содержании меди в концентрате основной флотации/скальпирования 25% (рис. 18). Это можно сравнить с действующими машинами основной флотации, обеспечивающим содержание меди 17% и извлечение 89%. Основная флотация в отдельно взятой флотомашине Джеймсон обеспечивает извлечение меди 60% при содержании 15%. Концентрат основной флотации будет направляться в цикл перечистной флотации для дальнейшей переработки.

Любопытно, что время пребывания во флотомашине Джеймсон, выполняющей данные задачи, составляет примерно 1 минуту на каждом этапе.

Согласно предлагаемой схеме, двухстадийный цикл основной флотации с использованием флотомашин Джеймсон заменит действующие 30-летние машины основной флотации. Хвосты основной флотации из флотомашин Джеймсон будут направляться на контрольную флотацию в 4x100м^3 Wemco, установленные при выполнении первой очереди модернизации. Концентрат контрольной флотации направляется в цикл доизмельчения, продукт которого поступает на перечистную флотацию.

Высокий выход является еще одним преимуществом флотомашины Джеймсон при скоростной основной флотации. При 1000 т/ч и содержании меди в питании 3,5% флотомашина Джеймсон обеспечиваете среднее извлечение меди 63,37% при среднем содержании меди в концентрате 29,4%, т.е. при использовании флотомашины Джеймсон за одну стадию флотации извлекается 75 т/ч конечного концентрата.

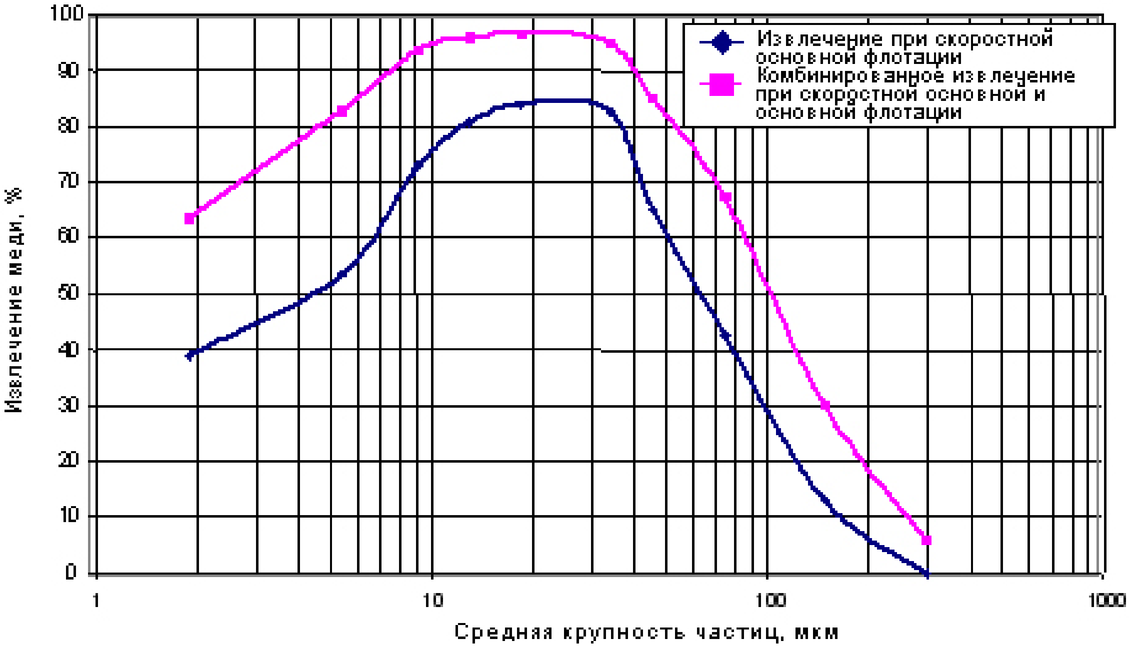


Рисунок 19. Извлечение в зависимости от крупности, флотомашина Джеймсон для основной флотации меди

График зависимости между крупностью и извлечением во флотомашине Джеймсон в качестве машины основной флотации меди (рис. 19) показывает очень высокое извлечение из фракций 9-38 мкм. Эти фракции хорошо высвобождаются и быстро флотируются и поэтому должны направляться непосредственно в конечный концентрат. Более крупные фракции недостаточно извлекаются, потому что они плохо высвобождаются и медленно флотируются. Их невозможно извлечь в высококачественный концентрат без загрязнения последнего и снижения его содержания. Более тонкие фракции (-6 мкм) извлекаются умеренно, поскольку они также медленно флотируются. Условия, создаваемые для сепарации крупных композитов и минимизации вовлечения тонкой пустой породы, способны обеспечить только умеренное извлечение этих медленно флотируемых фракций.

Плохо извлекаемые крупные композиты и медленно флотируемые тонкие фракции будут извлекаться в цикле контрольной флотации, и направляться на доизмельчение, где композиты будут высвобождены и переработаны в цикле перечистной флотации.

Проект 5: Цикл перечистной флотации, Minera Alumbrera

Компания Minera Alumbrera Ltd эксплуатирует обогатительную фабрику Алумбрера в Аргентине. Фабрика была пущена в 1997 года и перерабатывала 80 000 т/сутки, а после расширения в 2003 году перерабатывает 100 000 т/сутки золотосодержащей меднопорфировой руды. Эта фабрика является одной из немногих в мире, использующих исключительно флотомашины Джеймсон в цикле перечистной флотации.

Первоначальная технологическая схема фабрики Алумбрера (рис. 20) состояла из 2 параллельных циклов. Каждый цикл состоял из мельниц полусамоизмельчения мощностью 13,4 мВт, продукт которых поступал в 2 шаровые мельницы мощностью 6 мВт для измельчения до P80 150 мкм. Каждая линия флотации состояла из 8 флотомашин OK100 в основной флотации, после которых была установлена мельница доизмельчения для доведения питания перечистной флотации до класса P_{80} 37 мкм. Перечистная флотация происходит в двух параллельных циклах с использованием флотомашин Джеймсон.

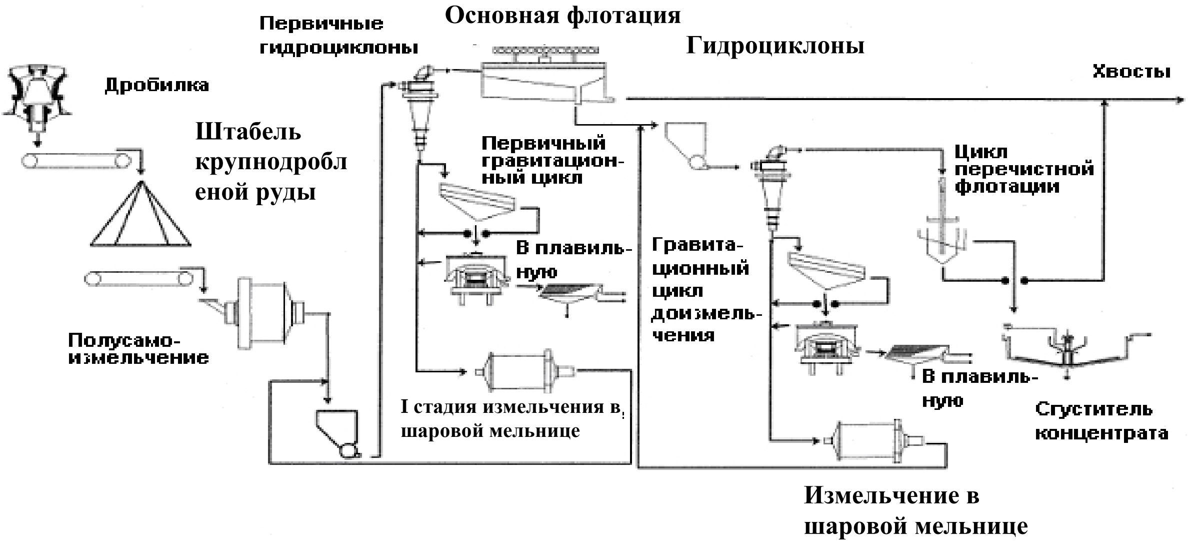


Рисунок 20. Технологическая схема Minera Alumbrera (Harbort *et al*, 2000)

Каждый цикл перечистки состоит из четырех машин первой перечистной флотации R5233/12, одной машины второй перечистной флотации R5245/12 и двух машин контрольной флотации R5233/12 (рис. 21). Концентрат перечистной флотации сгущается и перекачивается по трубопроводу длиной 312 км на фильтрующую установку в Сан Мигель де Тукуман, откуда продукт доставляется по железной дороге в порт для отгрузки.

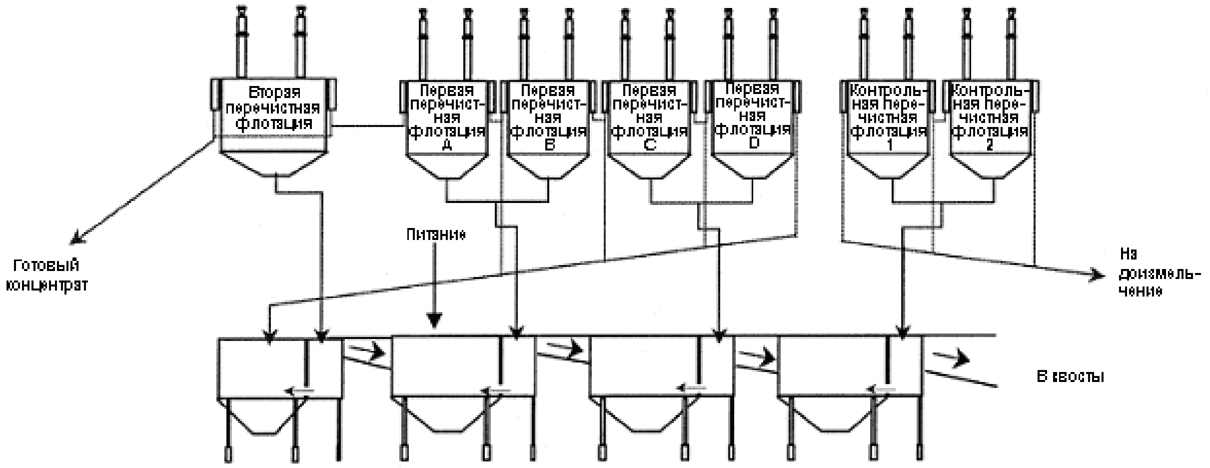


Рисунок 21. Схема цикла перечистной флотации, Minera Alumbrera (Harbort *et al*, 2000)

Данные флотации после пуска в эксплуатацию показывают, что извлечение меди в цикле перечистной флотации свыше 95% при средней производительности возможно, причем машина первой перечистной флотации А способна производить конечный концентрат при извлечении 70%. (Harbort *et al* 2000).

Характеристика извлечения в зависимости от крупности в цикле перечистной флотации меди показана на рис. 22. График подтверждает хорошее извлечение меди во всех классах крупности, особенно крупнозернистого хорошо высвобожденного халькопирита. В готовом концентрате халькопирит высвобожден на 90% благодаря крупнозернистой минерализации и тщательному доизмельчению в цикле.

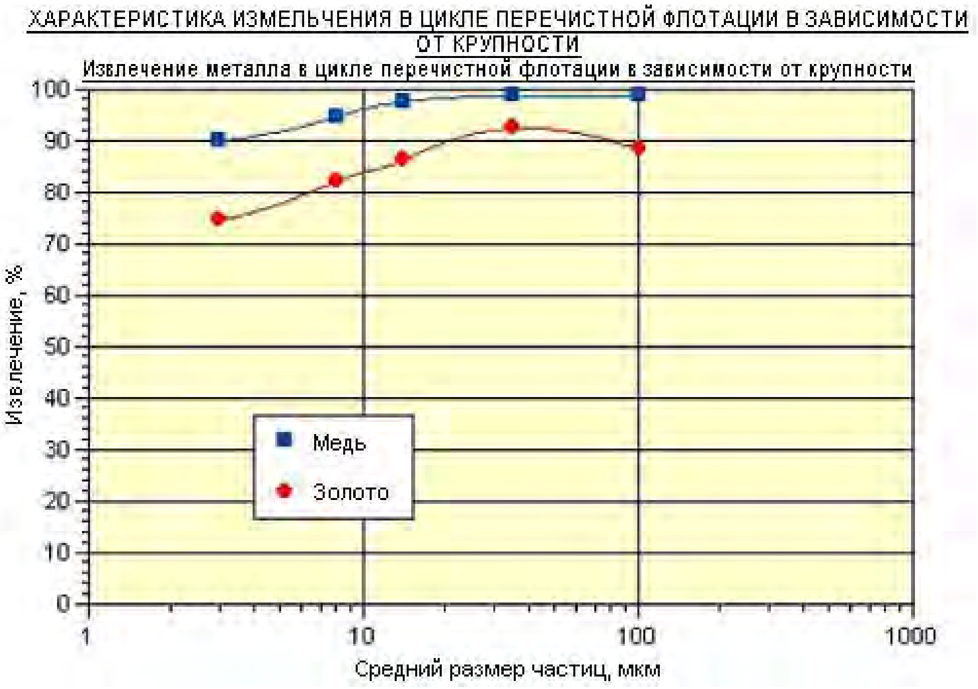


Рисунок 22. Извлечение в зависимости от крупности в цикле перечистной флотации, Minera Alumbrera (G&T, 2002)

После пуска в эксплуатацию производительность обогатительной фабрики Алумбрера выросла до более чем 100 000 т/сутки. Увеличению производительности способствовала установка третьей линии измельчения и увеличение на 50% мощностей основной и контрольной флотации. Эта модернизация была выполнена без увеличения производительности цикла перечистной флотации.

С момента пуска в эксплуатацию содержание меди в рядовой руде упало с 1% до 0,5%, а содержание в концентрате основной флотации - с 10% до 5%. Флотомашины Джеймсон обеспечивают вдвое более высокий коэффициент обогащения по сравнению с первоначальной схемой, что подтверждает способность флотомашины Джеймсон производить высококачественный концентрат и обеспечивать высокий коэффициент обогащения.

Текущие условия эксплуатации служат еще одним подтверждением надежности конструкции флотомашины Джеймсон, поскольку весь концентрат на фабрике Алумбрера производят первоначально установленные флотомашины Джеймсон. Однако это не значит, что возможности для дальнейшего совершенствования отсутствуют. Исходя из современного понимания флотомашин Джеймсон, а также сильных и слабых сторон традиционных флотомашин и флотомашин Джеймсон, авторы убеждены, что идеальным решением было бы добавить к флотомашинам Джеймсон несколько традиционных машин контрольной флотации, чтобы в полной мере использовать преимущества обеих конструкций и создать гибридный цикл с более низкой себестоимостью. В настоящее время предлагается новая схема цикла, позволяющая с использованием существующего оборудования уменьшить циркуляционную нагрузку и обеспечить высокий коэффициент обогащения для повышения качества конечного концентрата.

ЗАКЛЮЧЕНИЕ

С начала 1990-х годов флотомашины Джеймсон с большим успехом применяются в циклах обогащения угля и циклах SXEW, однако их применению для руд цветных металлов препятствовали особенности ранних моделей и отсутствие понимания наилучших методик их применения в циклах флотации руд цветных металлов. Конструктивные усовершенствования последнего десятилетия и более глубокое понимание сильных и слабых сторон и технологических аспектов флотомашины Джеймсон позволяют проектировать циклы с использованием флотомашины Джеймсон в качестве надежного решения при флотации руд цветных недрагоценных металлов.

Флотомашины Джеймсон способны обеспечить хорошие показатели захвата минералов с высокой кинетикой и получение высококачественного концентрата при минимальном уносе. Флотомашины Джеймсон являются машинами высокой интенсивности, в которых высокая скорость сдвига создает маленькие пузырьки, селективно собирающие минералы с высокой кинетикой. Компактность конструкции позволяет рационально применять промывку пены, уменьшая механический унос и получая высококачественный концентрат за один проход. Умеренное извлечение, наблюдаемое, как правило, при использовании одной флотомашины Джеймсон при флотации руд цветных металлов, может быть повышено за счет увеличения коэффициента рециркуляции, установки дополнительных флотомашин Джеймсон или дополнительных традиционных флотомашин.

Идеальным применением флотомашины Джеймсон часто является гибридный цикл с традиционными флотомашинами, обеспечивающий более высокую эффективность при меньшей стоимости и меньшей занимаемой площади по сравнению с любой из технологий, взятой в отдельности.

Хотя флотомашины Джеймсон идеально подходят для использования в новых технологических схемах, а благодаря их компактности и небольшой габаритной высоте они являются прекрасным решением при расширении и модернизации существующих производств. Флотомашины Джеймсон могут быть установлены для повышения эффективности флотации на действующем предприятии, а также являются экономичным решением для проектов расширения.

СПИСОК ИСПОЛЬЗОВАННОЙ ЛИТЕРАТУРЫ

Carr, D., Harbort, G., Lawson, V., 2003, Expansion of the Mount Isa Mines Copper Concentrator Phase One Cleaner Circuit Expansion, Eighth Mill Operators' Conference, AUSIMM, Townsville, QLD

Clayton, R., Jameson, G.J., Manlapig, E.V., 1991, The Development and Application of the Jameson Cell; Minerals Engineering ,July-Nov 1991

Cowburn, JA., Stone, R., Bourke S., Hill, B., 2005, Design Developments of the Jameson Cell, Centenary of Flotation Symposium, AUSIMM Brisbane

Dawson W.J., Jackson, B.R., 1995, Evolution of Jameson Cells for Solvent Extraction Applications, Copper Hydrometallurgy Forum, Brisbane

G&T Metallurgical Sevrices, 2002, Modal analysis of the Plant Process Streams, January, 2002, Minera Alumbrera, Argentina, Report KM1256 - Consultant Report

Harbort, G.J., Murphy, A.S., Budod, A, 1997, Jameson Cell Developments at Philex Mining Corporation, Sixth Mill Operators' Conference, AUSIMM, Madang, PNG

Harbort GJ, Lauder D, Murphy AS, Miranda J, 2000, Size by Size Analysis of Operating Characteristics of Jameson Cell Cleaners at the Bajo de Alumbrera Copper / Gold Concentrator, Seventh Mill Operators Conference, AUSIMM, Kalgoorlie

Harbort GJ, 2002, Pilot Plant Jameson Testwork at the Mount isa Copper Concentrator, MIM Holdings Limited - Internal Report.

Jameson, G.J., Harbort, G., Riches, N., 1991, The Development and Application of the Jameson Cell Fourth Mill Operator's Conference, AUSIMM, Burnie, Tasmania

Jameson, G.J., Goffinet, M., Hughes, D, 1991, Operating Experiences with Jameson Cell at Newlands Coal Pty Ltd, Queensland, 5th Australian Coal Preparation Conference

Jameson, G.J. and Manlapig, E.V., 1991 - Flotation cell design - experiences with the Jameson Cell, 5th AusIMM Extractive Metallurgy Conference.

Riches, N.J., 1991 - Jameson Cell testing of AG Mill discharge. MIM Holdings Limited - Internal Report.

Young, MF, J D Pease, JD, Fisher, KS., 2000, The George Fisher Project to Increase Recovery in the Mount Isa Lead/Zinc Concentrator, Seventh Mill Operators Conference, AUSIMM, Kalgoorlie

ОПЫТ МАСШТАБИРОВАНИЯ ФЛОТОМАШИНЫ ДЖЕЙМСОН НА УГЛЕОБОГАТИТЕЛЬНОЙ ФАБРИКЕ ГУНЬЕЛЛА КОМПАНИИ BHP COAL

Майкл Ф. Карретта (инженер-технолог), Джеймс Н. Грэм (инженер-технолог)
BHP Coal Pty Ltd, Goonyella/Riverside Mine, Moranbah, Qld, 4744, Australia

Уэнди Дж. Доусон (металлург, флотационная машина Джеймсон)
MIM Holdings Ltd, GPO Box 1433, Brisbane, Qld, 4001, Australia

Представлено на: Coal Prep 97
29 апреля - 1 мая
Лексингтон, Кентукки

Автореферат

В 1993 году компания BHP Coal выполнила оценку нескольких технологий флотации для использования в качестве замены устаревающих механических флотомашин на углеобогатительной фабрике Гуньелла. Исследование включало Проведение пилотного тестирования с использованием двух флотомашин Джеймсон диаметром 0,5 м с производительностью 7,5 м^3/ч. Результаты показали, что флотомашина способна обеспечить более высокие технологические показатели, нежели механические флотомашины. В следующем году BHP Coal приняла решение заменить механические флотомашины флотомашинами Джеймсон с использованием двухступенчатой схемы (основная флотация - контрольная флотация).

В 1995 году BHP Coal предприняла проект с целью проверить возможность масштабирования флотомашин Джеймсон с пилотных до промышленных. Цель исследования заключалась в том, чтобы установить, позволит ли аэратор промышленного размера получить результаты, смоделированные на основании тестирования с использованием аэратора меньшего размера. Для проверки возможностей масштабирования компания BHP Coal закупила две демонстрационные флотомашины Джеймсон диаметром 1,25 м каждая с с аэратором промышленного размера производительностью 60 м^3/ч. Объем емкости демонстрационной флотомашины был рассчитан с использованием параметров, полученных в ходе пилотного тестирования. Полученные результаты демонстрационных испытаний соответствовали данным, полученным в ходе пилотного тестирования.

После доказанного прямого масштабирования пилотной флотомашины Джеймсон до демонстрационной с использованием аэраторов промышленного размера, BHP Australia Coal установила восемь флотомашин модели 5000/16 (диаметр 5 м, 16 аэраторов) с использованием последовательной двухступенчатой схемы. В этих промышленных флотомашинах Джеймсон производительностью 960 м^3/ч каждая установлены аэраторы такого же размера, как и в демонстрационных машинах. Установка флотомашин Джеймсон на фабрике Гуньелла была частью проекта модернизации линии обогащения мелкой фракции.

Модернизированная линия обогащения мелкой фракции была пущена за две недели и быстро достигла максимальной производительности. Оптимизационные данные промышленной флотомашины Джеймсон практически полностью соответствуют данным пилотного тестирования. Они подтверждают, что пилотные испытания позволяют точно прогнозировать эксплуатационные показатели флотомашин Джеймсон промышленного масштаба.

Эксплуатационные данные по флотомашинам Джеймсон, собранные за несколько месяцев, показывают, что промышленная установка уверенно обеспечивает показатели, спрогнозированные по результатам пилотного тестирования. Также доказано, что показатели эффективности флотомашины Джеймсон намного более постоянны, чем показатели механических флотомашин, применявшихся ранее.

Стабильное получение продукта с низкой зольностью во флотомашине Джеймсон способствовало приросту извлечения в цикле обогащения крупной

фракции повышению выхода по фабрике на 3,5%, установив рекордные производственные показатели BHP Coal.

ВВЕДЕНИЕ

Разрез Гуньелла / Риверсайд является одним из семи рудников, эксплуатируемых BHP Australia Coal Pty Ltd (ранее - BHP Australia Coal Pty Ltd) в бассейне Боуэн в центральной части штата Квинсленд. Хотя разрез Гуньелла (владелец - Central Queensland Coal Associates) и соседний разрез Риверсайд (владелец - BHP Mitsui Coal) были объединены в 1989 году, они эксплуатируются на основании разных договоров аренды и имеют собственные углеобогатительные фабрики. В совокупности в этом финансовом году на обоих предприятиях будет добыто примерно 10,7 млн. тонн угля, из которых 7,2 млн. тонн - на разрезе Гуньелла (Gosnell, 1996).

Разрез и углеобогатительная фабрика Гуньелла были пущены в эксплуатацию 25 лет назад и ежегодно производят 4,4 млн. тонн угля. Согласно первоначальной технологической схеме, исходный материал делился на два технологических потока:

- уголь крупной фракции (щелевой грохот +0,5 мм), составляющий 80% питания фабрики и перерабатываемый в шести модулях тяжелосредных циклонов; и

- уголь мелкой фракции (щелевой грохот -0,5 мм), перерабатываемый в стандартном цикле флотации, состоящем из 4-х батарей механических флотомашин Wemco емкостью 8,5$м^3$ (300 куб.фт.). Ультратонкая фракция удалялась.

При последующих модернизациях была добавлена секция флотации шламов (1979 год) и винтовые сепараторы перечистки хвостов (1985 год), а в 1987 году производство было расширено на треть для достижения производительности 6,7 млн. тонн в год.

К августу 1993 года, когда Гуньелла продала свою стомиллионную тонну угля, углеобогатительная фабрика перерабатывала до 1 800 т сырья в час и состояла из восьми модулей тяжелосредных циклонов, секции раздельной флотации, состоящей из четырех батарей по 4 флотомашины Wemco объемом 8,5$м^3$ (300 куб.фт.), четырех батарей по 4 флотомашины Wemco объемом 8,5$м^3$ (300 куб.фт.) для флотации шламов и винтовых сепараторов перечистки хвостов. Продукт флотации, как правило, имел зольность порядка 9-10% с извлечением горючей массы 71-83%горючей массы. Однако флотационные машины для крупной фракции имели сильную коррозию и нуждались в замене. В целях их замены были изучены новые технологии, которые смогли бы повысить выход по фабрике за счет получения концентрата флотации с меньшим уровнем зольности.

В 1993 году BHP Coal приступила к интенсивным испытаниям на предприятии с целью изучить преимущества, которые углеобогатительная фабрика Гуньелла могла получить благодаря применению новых технологий флотации повышенной интенсивности, в частности:

- флотационная машина Джеймсон (MIM Holdings Ltd),
- Microcel (Bulk Materials Coal Handling) и
- Машина интенсивной флотации Centrifloat (Century Oils).

Машина Джеймсон представляет собой пневматическую флотационную установку, не имеющую подвижных частей и не нуждающуюся в подаче сжатого воздуха и устройствах барботирования. Техническое описание флотационной машины можно найти в других источниках (Evans *et al*, 1995).

Испытания

Пилотные тестирования

Пилотные установки Джеймсон, Microcel и Centrifloat были испытаны в открытом цикле с подачей следующего питания непосредственно с углеобогатительной фабрики:

- питание секции флотации, щелевой грохот 0,5 мм x 0 (см. табл. 1),
- питание флотации шламов, 75 микрон x 0 (минус 200 меш),
- пески сгустителя хвостов.

Таблица 1. Характеристики питания для флотации

Фракция Меш (номинальный размер ячейки)	Масса (%)	Зольность (%)
меш +16 (+1 мм)	0.5	19.0
меш -16 +30 (-1 +0.5 мм)	12.1	14.0
меш -30 +60 (-0.5 +0.25 мм)	21.1	14.0
меш -60 +100 (-0.25 +0.15 мм)	13.7	15.2
меш -100 +250 (-0.15 +0.063 мм)	13.2	18.2
меш -250 (-0.063 мм)	39.4	39.0

Питание флотации, как правило, имело зольность 18-25%.

Для испытаний на фабрике Гуньелла была использована мобильная пилотная флотомашина Джеймсон. Пилотная установка состояла из двух флотомашин Джеймсон модели JC500/1 (диаметр 500 мм) с одним аэратором 100 мм. Каждая из флотомашин имела производительность 7,5 м^3/ч и могла работать в одну стадию, две стадии или одну стадию с частичной рециркуляцией хвостов. Различные конфигурации флотомашины Джеймсон показаны на рис. 1.

Демонстрационные флотомашины Джеймсон

Тестирование флотации на флотомашинах Джеймсон проводилось на демонстрационных установках, чтобы убедиться в том, что аэраторы промышленного размера (диаметр 300 мм) производительностью 60 м^3/ч позволят воспроизвести результаты, полученные в результате пилотного тестирования. Демонстрационные машины Джеймсон имели диаметр 1,25 м и частично закрытый порог, чтобы весовая скорость продукта в пересчете на длину порога был равна аналогичному показателю в полноразмерной машине, которую предполагалось установить на фабрике. Две машины были смонтированы на раме с целью эксплуатации в составе двухступенчатой схемы. Из-за проблем, связанных с реагентами, двухступенчатая схема оказалась сложной в эксплуатации. Поэтому демонстрационные испытания были проведены с использованием только одной машины в режиме частичной рециркуляции хвостов. Пилотные испытания показали, что эксплуатационные показатели одной флотомашины в режиме частичной рециркуляции были близки к показателям двухступенчатой схемы.

Четырехдневные демонстрационные испытания флотомашины были проведены в то время, когда, по словам персонала предприятия, в качестве питания подавался окисленный уголь, а следовательно, флотация происходила не в оптимальном режиме. Предположительно, исходный уголь был перегрет. Кроме того, содержание твердой фракции в питании флотации было выше обычного (14-18% против 10-12%).

Полноразмерные флотомашины Джеймсон

Восемь флотомашин Джеймсон модели 5000/16 (диаметр 5 м, по 16 аэраторов диаметром 30,5 см) были установлены в две стадии (четыре первичные машины и четыре вторичные, как показано на рис. 2). Питание - пульпа 3 800 $м^3$/ч с содержанием твердой фракции 9,2% (по весу).

Масштабирование флотационной машины Джеймсон было произведено с использованием простых алгоритмов и значений основных параметров, установленных в ходе испытаний. Флотомашина Джеймсон модели 5000/16 была спроектирована с непосредственным использованием оптимизированных параметров пилотного тестирования. Методика эксплуатации и расход реагентов были также определены на основании данных пилотного тестирования. Такие эксплуатационные переменные, как расход воздуха, глубина пены и подача промывочной воды, могут корректироваться в соответствии с требуемыми технологическими показателями.

Установка флотомашин Джеймсон стала одним из этапов модернизации линии обогащения мелкой фракции, наряду с проектированием и сооружением сгустителя, расширением существующего здания фабрики для размещения флотомашин Джеймсон, модернизацией системы управления и изменением расположения ряда единиц технологического оборудования.

Рисунок 1. Технологическая схема флотации мелкой угольной фракции с с использованием флотомашин Джеймсон

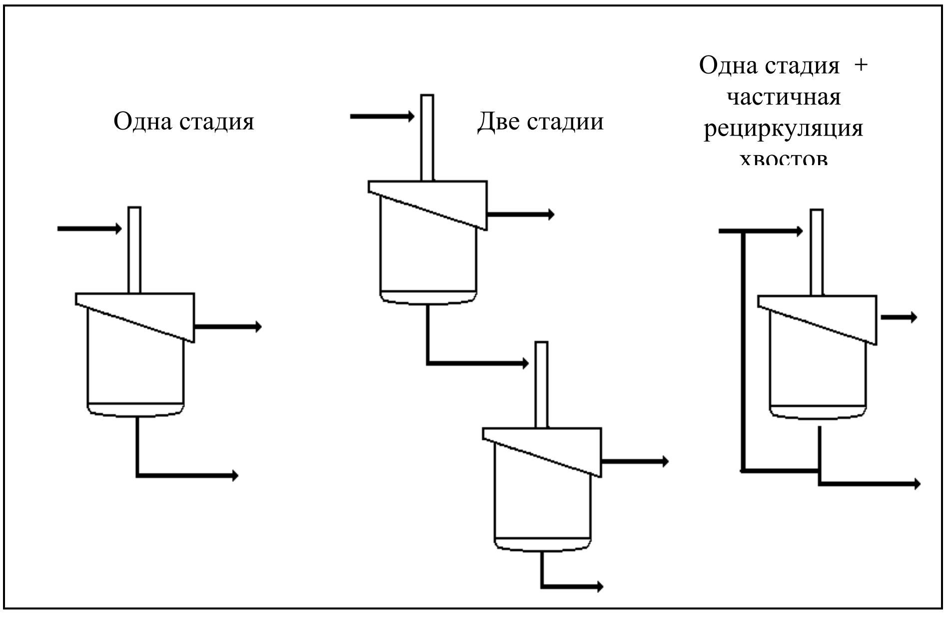
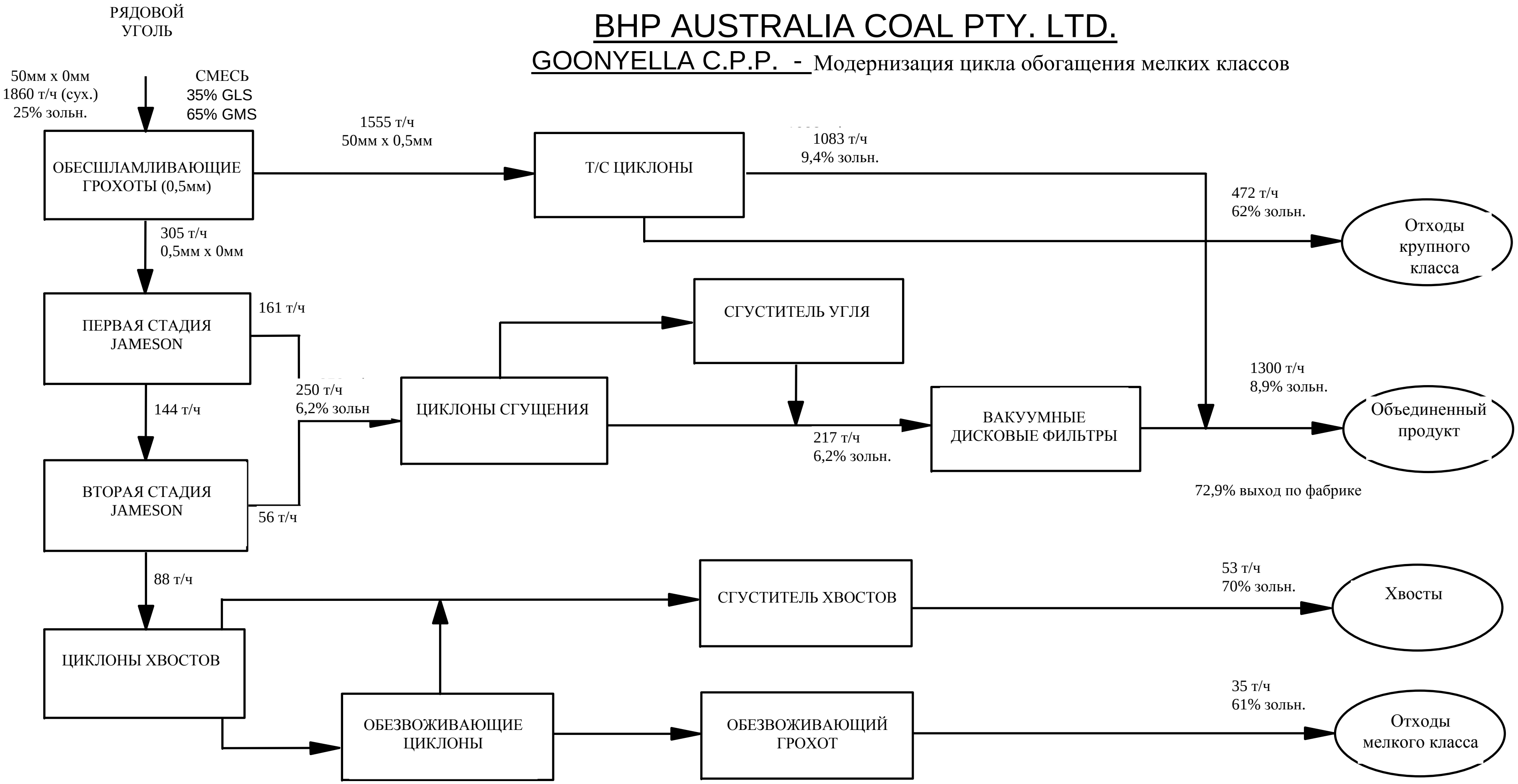


РИСУНОК 2

BHP AUSTRALIA COAL PTY. LTD.

GOONYELLA C.P.P. - Модернизация цикла обогащения мелких классов



Результаты и анализ

Пилотные тестирования и оценка результатов

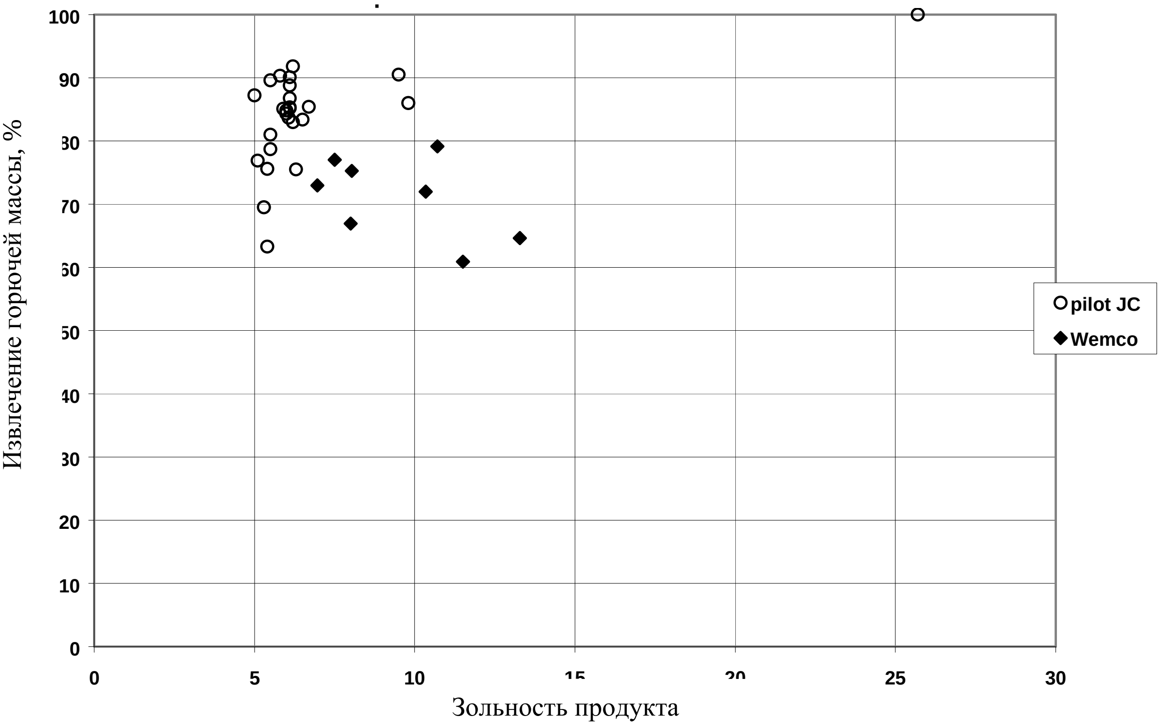
Пилотные флотационные испытания показали, что флотомашина Джеймсон способна обеспечить значительно меньшую зольность продукта по сравнению с существующими флотомашинами Wemco при сохранении или увеличении извлечения горючей массы. На рис. 3 показаны эксплуатационные характеристики пилотной флотомашины Джеймсон при переработке питания флотации в сравнении с характеристиками линии, состоящей из механических флотомашин Wemco и машин для флотации шламов. Данные по флотомашинам Wemco были получены в результате независимых флотационных испытаний.

В ходе пилотного тестирования было установлено, что результаты двухступенчатой флотации на флотомашинах Джеймсон очень близки к результатам испытаний одноступенчатой флотации с частичной рециркуляцией хвостов. Это характерный результат, который может позволить добиться существенной экономии на оборудовании при проектировании полноразмерных установок.

Результаты испытаний трех новых технологий флотации и эксплуатационные данные механических флотомашин были подвергнуты комплексной оценке капитальных и эксплуатационных затрат, прогнозируемого прироста прибыли, эксплуатационной годности и с точки зрения технических рисков.

Рисунок 3

Извлечение горючей массы – Зольность продукта
Пилотное тестирование Jameson Cell – Wemco



Проведя тщательный анализ, компания BHP Coal приняла решение установить флотомашины Джеймсон на линию переработки всего питания флотации. Машины Джеймсон были установлены в две стадии с подачей хвостов после первой стадии на переработку во второй стадии. Хотя одностадийная схема с флотомашинами Джеймсон с частичной рециркуляцией хвостов показала результаты, сравнимые с показателями после двух стадий, предпочтение было отдано последней схеме, поскольку она была ранее успешно установлена и эксплуатировалась на руднике Блэкуотер (BHP Coal) и на предприятии Ньюлэндс (MIM Holdings) (Manlapig *et al*, 1993). На данном этапе одноступенчатая схема с частичной рециркуляцией хвостов была сопряжена с определенным риском, поскольку пилотные тестирования на разрезе Гуньелла стали первым тщательным исследованием данного режима эксплуатации и промышленные машины Джеймсон, установленные с использованием данной схемы, отсутствовали. Тем не менее, данная схема была впоследствии успешно применена на ряде углеобогатительных фабрик (Dawson *et al*, 1996; Le Page *et al*, 1995).

Технологии Джеймсон было отдано предпочтение перед конкурирующими технологиями по следующим причинам (Carretta, 1996):

- меньшие капитальные затраты на промышленную установку;
- меньшие эксплуатационные затраты (энергия, реагенты) и затраты на обслуживание;
- прямое масштабирование результатов пилотных испытаний было доказано на руднике Блэкуотер (BHP Coal);
- флотомашина Джеймсон обеспечивала такое же извлечение горючей массы и зольность продукта, как и другие технологии; и
- лабораторная установка Джеймсон оказалась наиболее простой в эксплуатации и наименее восприимчивой к колебаниям содержания твердого в питании и расхода питания.

Получение продукта с низкой зольностью во флотомашине Джеймсон означало, что можно было повысить извлечение цикла переработки крупной фракции (при немного более высокой зольности), что, в свою очередь, позволяло увеличить производительность фабрики при соблюдении требований по зольности продукта.

Демонстрационные испытания

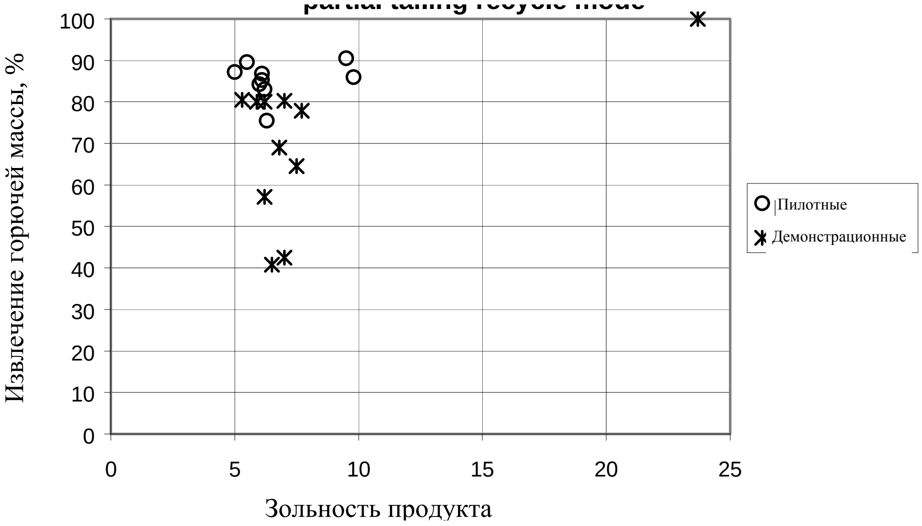
Показатели извлечения горючей массы и зольности продукта по результатам демонстрационных флотационных испытаний в целом соответствовали данным пилотного тестирования, хотя извлечение горючей массы было немного ниже (рис. 4). Показатели установленных на фабрике флотомашин Wemco при переработке питания более низкого качества в ходе демонстрационных испытаний также были хуже обычных.

Демонстрационные испытания были прекращены досрочно, когда на установку стал подаваться в качестве исходного материала частично окисленный уголь среднего пласта Риверсидс.

Высокое содержание твердого в исходном материале, как известно, отрицательно сказывается на эксплуатационных показателях флотомашины Джеймсон, особенно при приближении к максимальной несущей способности (Espinosa-Gomez *et al*, 1988). Поэтому повышенное содержание твердого вещества могло снизить извлечение горючей массы. Учитывая неидеальные условия проведения демонстрационных испытаний, результаты были признаны обнадеживающими.

Рисунок 4

Извлечение горючей массы – Зольность продукта
Пилотные и демонстрационные тесты Jameson Cell – с частичной рециркуляцией хвостов



Промышленный масштаб

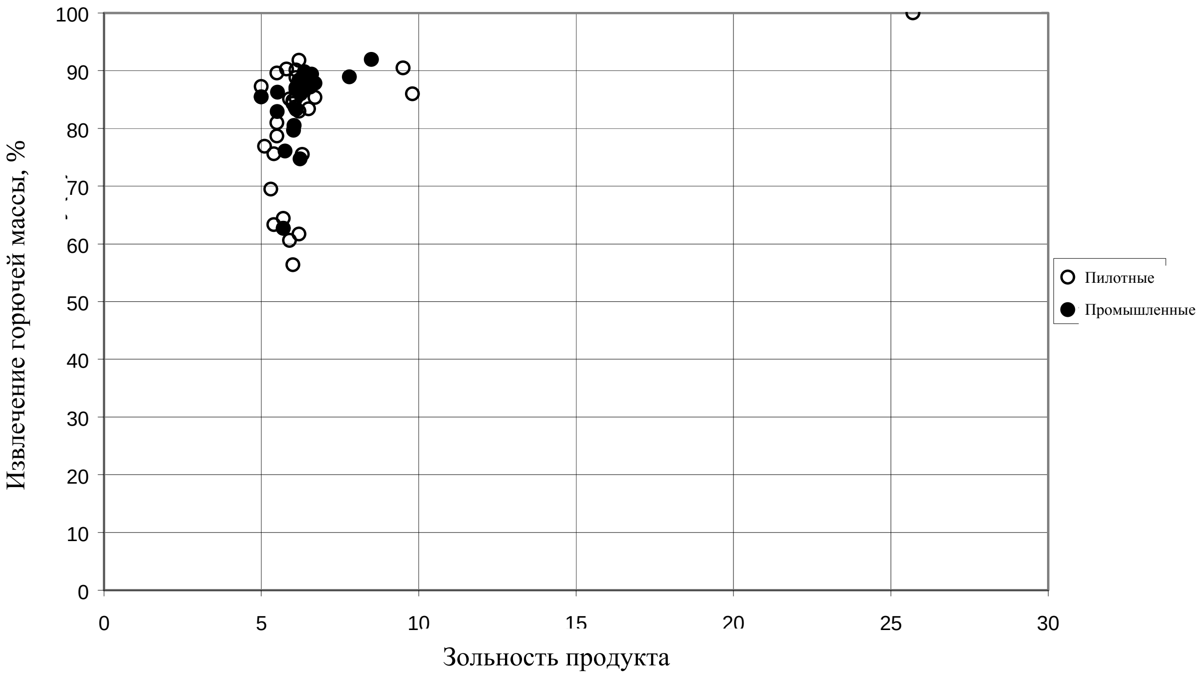
Модернизированная линия переработки мелкой фракции была пущена в эксплуатацию в декабре 1995 года. Пусконаладочные работы были выполнены за две недели, причем основные проблемы были связаны с недостатками обслуживания и эксплуатации циклонов классификации хвостов (диаметр 1 м), сгустителя угля и новой системы подачи реагентов. Установка вышла на максимальную производительность 2000 т/ч (до модернизации - 1800 т/ч) через шесть недель после повторного пуска линии переработки мелкой фракции.

На рис. 5 показано извлечение горючей массы и зольность продукта по результатам оптимизационных испытаний промышленных флотомашин Джеймсон, которые были проведены после пуска в эксплуатацию, относительно результатов пилотного тестирования. Это подтверждает способность пилотного тестирования точно предсказать характеристики промышленной флотомашины Джеймсон.

Эксплуатационные параметры промышленной флотомашины Джеймсон, такие как подача реагентов и глубина пены, были определены по результатам пилотного тестирования и использованы непосредственно в полноразмерной установке.

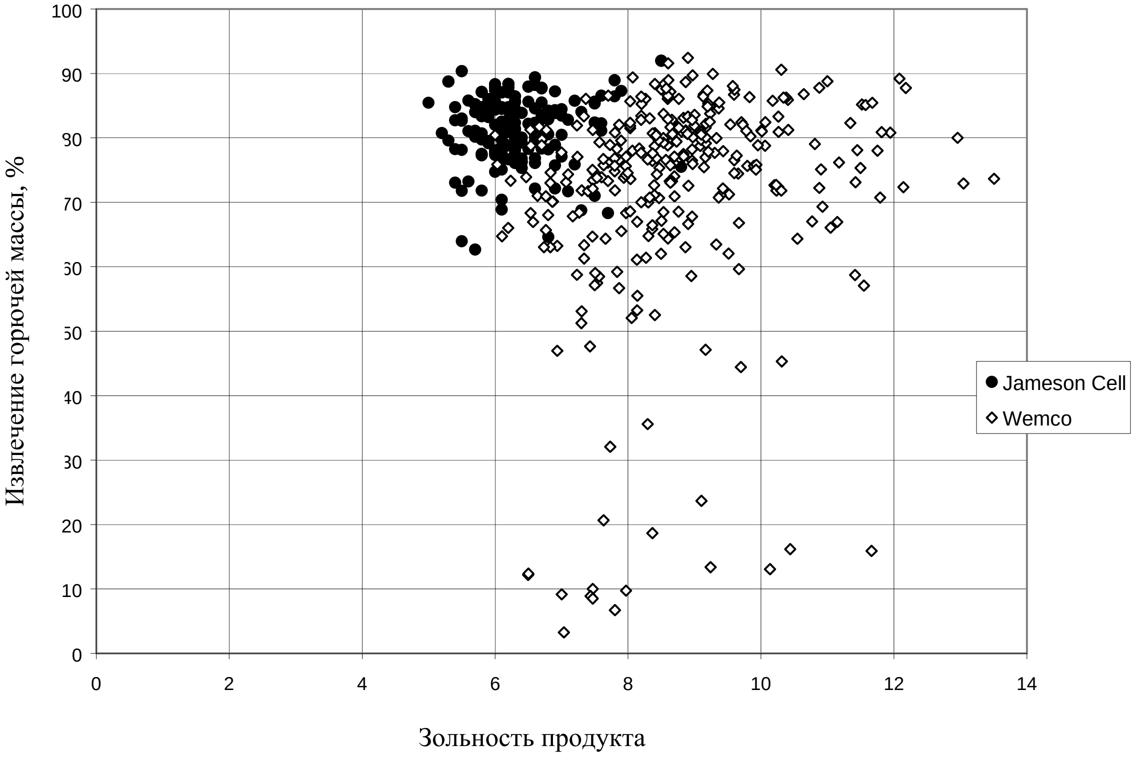
Стабильные показатели промышленных флотомашин Джеймсон показаны на рис. 6, где сравниваются эксплуатационные данные за год работы флотомашин Wemco и примерно за шесть месяцев работы флотомашин Джеймсон. Флотомашины Джеймсон уверенно обеспечивают низкую зольность продукта, несмотря на колебания в характеристиках питания.

Рисунок 5



Извлечение горючей массы – Зольность продукта
Пилотное тестирование и Промышленная эксплуатация Jameson Cell

Рисунок 6



Извлечение горючей массы – Зольность продукта
Технологические показатели Jameson Cell – Wemco

На рис. 7 и 8 показано распределение горючей массы в питании между продуктом и хвостами при флотации во флотомашине Джеймсон и флотомашине Wemco, соответственно. Флотомашина Джеймсон обеспечивает высокое извлечение горючей массы в широком диапазоне крупностей питания флотации. Механические флотомашины имеют более низкие показатели извлечения горючей массы из наиболее мелкой и наиболее крупной фракций, причем наиболее мелкая фракция имеет определяющее значение для общего показателя извлечения.

Рисунок 7

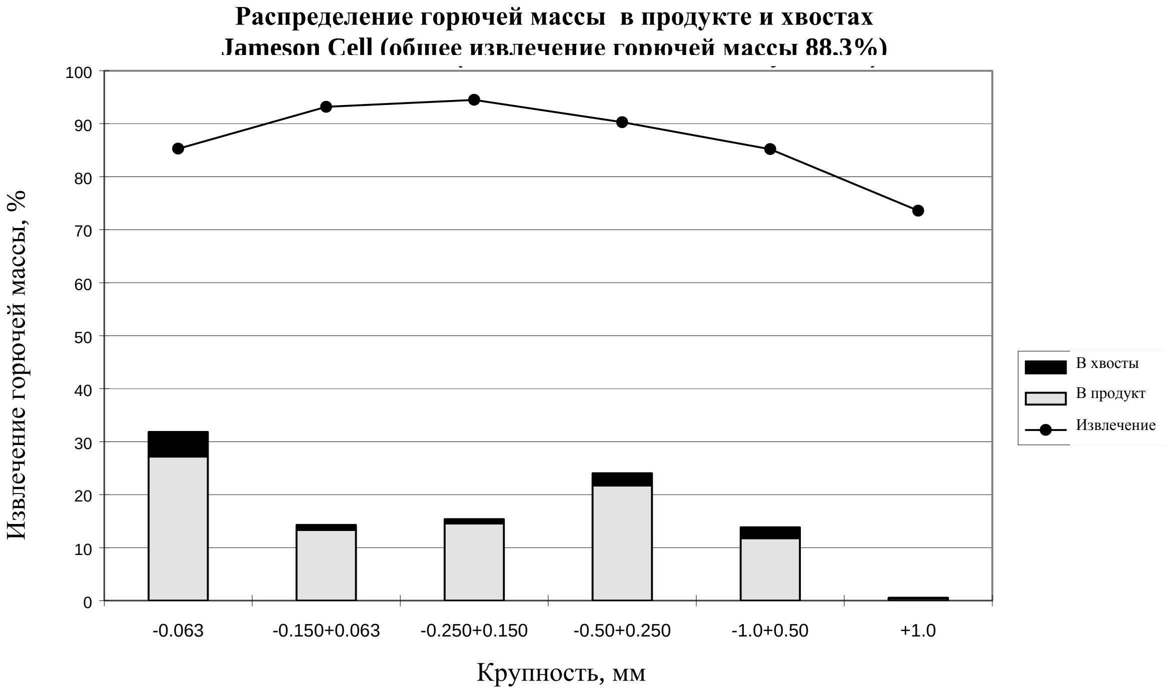
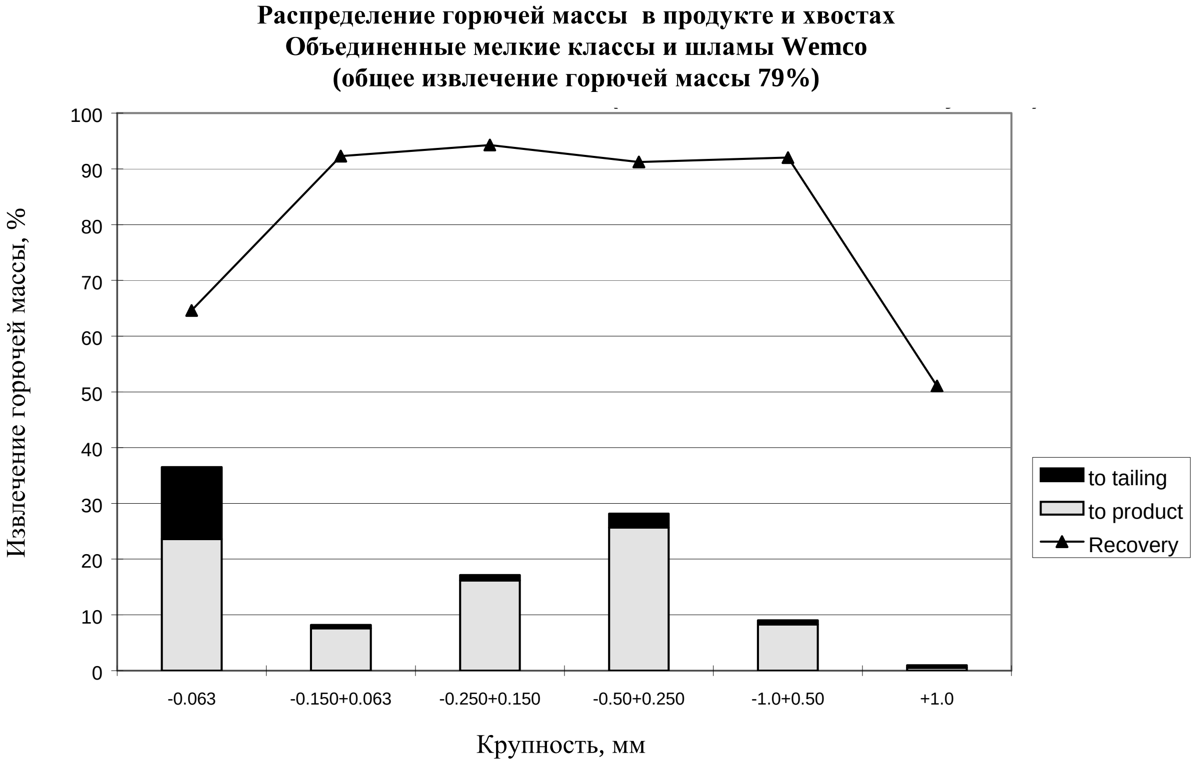

Распределение горючей массы в продукте и хвостах
Jameson Cell (общее извлечение горючей массы 88.3%)

Рисунок 8


Распределение горючей массы в продукте и хвостах
Объединенные мелкие классы и шламы Wemco
(общее извлечение горючей массы 79%)

Технологические показатели за период после пусконаладки флотомашин Джеймсон подтверждают, что общий выход по фабрике вырос на 3,5%. Фактический прирост оказался значительно выше прогнозного (2,1%). В результате срок окупаемости проекта модернизации линии переработки мелкой фракции сократился с первоначальных 28 месяцев до 17 месяцев.

Успех проекта установки флотомашин Джеймсон на углеобогатительной фабрике Гуньелла способствовал рекордно высоким производственным показателям предприятия. В марте 1996 года фабрика Гуньелла превысила рекордный показатель недельного объема производства угля в BHP Coal (ранее поставленный на близлежащем разрезе Пик Даунз), произведя 198 645[1] тонн угля. Впоследствии фабрика Гуньелла добилась новых рекордных показателей недельного объема производства в мае 1996 года (200 163 тонны) и в январе 1997 года (210 476 тонн).

Заключение

1. Пилотные тестирования флотомашин Джеймсон доказали превосходство этой технологии по отношению к применяемым механическим флотомашинам с получением продукта меньшей зольности и более высоким извлечением горючей массы.

2. В ходе демонстрационных испытаний флотомашин Джеймсон были получены аналогичные пилотным показатели зольности, хотя показатели извлечения горючей массы не достигли уровня, отмеченного в результате пилотного тестирования. Пониженное извлечение горючей массы было, вероятно, связано с худшим качеством исходного угля.

3. Технологические показатели промышленной установки Джеймсон очень близки к результатам пилотного тестирования. Они подтверждают, что пилотные тестирования позволяют точно прогнозировать эффективность флотомашин Джеймсон промышленного масштаба.

4. Флотомашина Джеймсон обеспечивает более высокое извлечение горючей массы из наиболее мелкой (-0,063 мм) и наиболее крупной (+1,0 мм) фракций и благодаря этому позволяет повысить общее извлечение.

5. Стабильное получение продукта низкой зольности при высоком извлечении во флотомашине Джеймсон позволило повысить выход по фабрике на 3,5%, что, в свою очередь, позволило поставить новые производственные рекорды.

Благодарности

Авторы выражают благодарность BHP Coal Pty Ltd и MIM Holdings Ltd за разрешение опубликовать данные, содержащиеся в этом докладе.

---

[1] Тоннаж приведен к стандартному для BHP Coal содержанию влаги 8%.

Список использованной литературы

Carretta, M. F., 1996. Goonyella/Riverside CPP1 fine circuit upgrade, *Technology Review Symposium - Coal Flotation, A Review of current Practice*, Brisbane.

Dawson, W. J., Yannoulis, G.F., Atkinson, B. W., and Jameson, G. J., 1996. Applications of the Jameson Cell in the Australian coal industry. *Column '96 - International Symposium on Column Flotation*, Quebec, August 26-28, pp. 233-246.

Espinosa-Gomez, R., Finch, L. A., Yianatos, J. B., and Dobby, G. S., (1988). Technical note: Flotation column carrying capacity: particle size and density effects, *Minerals Engineering*, Vol. 1, No. 1, pp. 77-79.

Evans, G. M., Atkinson, B. W., and Jameson, G. J., 1995. The Jameson Cell, *Flotation Science and Technology*, Matis, K. A., (ed.), Marcel Dekker Inc., New York, pp. 331-363.

Gosnell, P., 1996. Latest upgrade will see Goonyella/Riverside into 21st century, *Australian Mining*, Vol. 88, No. 11, December, pp.14-15.

Le Page, M., Goodman, P., and Halpin, B., 1995. Gordonstone coal preparation plant upgrade process design, in Smitham, J. (ed.), Proceedings of the Seventh Australian Coal Preparation Conference, Paper A4, pp. 46-70.

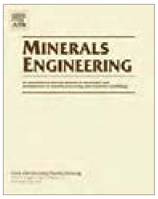
Manlapig, E. V., Jackson, B. R. Harbort, G. J. and Cheng, C. Y., 1993. Jameson Cell coal flotation, *Coal Prep '93: 10th International Coal Preparation Exhibition and Conference*, Lexington, pp. 204-219.

Minerals Engineering 27–28 (2012) 52–59

Contents lists available at SciVerse ScienceDirect

Minerals Engineering

journal homepage: www.elsevier.com/locate/mineng



The beneficiation of ultrafine phosphate

A.J. Teague [*], M.C. Lollback

Legend International Holdings Inc., Level 8, 580 St. Kilda Road, Melbourne, Vic. 3004, Australia

ARTICLE INFO

Article history:
Received 20 October 2011
Accepted 12 December 2011
Available online 29 December 2011

Keywords:
Industrial minerals
Fine particle processing
Flotation machines
Flotation depressants

ABSTRACT

This paper discusses a novel method of beneficiation of ultrafine phosphate which allows the recovery of phosphate particles that are less than 20 μm in size and have long been considered in the industry to be unrecoverable. It has been standard practice over many years in the phosphate industry to separate and discard the ultrafine particles, due to unacceptable processing difficulties such as excessively high viscosity and/or poor flotation recovery. In contrast with the established methods of beneficiation of phosphate where classification by hydrocyclone is mainly used to remove ultrafines as tailings, the method described in this paper allows a high proportion of ultrafines to be recovered via flotation without prior separation.

A number of variables and their effect on the flotation recovery of ultrafine phosphate are investigated including the pulp density and water quality during conditioning and flotation, type of flotation machine and reagents used to depress Fe_2O_3 and Al_2O_3. Some excellent results were achieved using samples containing up to 75 wt% −20 μm particles, including for example 91.2% P_2O_5 recovery to a concentrate grade of 34.7% P_2O_5 from a low feed grade of 6.46% P_2O_5 and 92.4% P_2O_5 recovery to a concentrate grade of 30.2% P_2O_5 from 10.6% P_2O_5 feed. Guar gum was found to be the most effective depressant for Fe_2O_3, whilst the Al_2O_3 was determined to be hydrophilic, resulting in low amounts being recovered to the concentrate, regardless of whether a depressant was used or not. The results of this work led to the development of the method described herein which is designed to recover phosphate from ores containing particles up to 80 wt% passing 20 μm, by flotation using a Jameson cell.

1. Introduction

Phosphate beneficiation operations around the world which use flotation as the principal mechanism to concentrate the phosphate bearing minerals, typically discard the “ultrafine” phosphate bearing particles, where for this study “ultrafine” is defined as particles being smaller than 20 μm. It is standard practice in the phosphate industry to separate the ultrafines by scrubbing and hydrocycloning before the remainder of the raw feed is transferred to the flotation plant for concentration (Kogel et al., 2006). Phosphate beneficiation has been carried out this way for many years because of the known poor floatability of the ultrafines and because in the past it has been determined that the larger phosphate bearing particles float and concentrate more efficiently in the absence of ultrafines. After the ultrafines have been separated they are typically discarded and stored in large slimes ponds as tailings, effectively becoming lost revenue to the mining operation.

Legend International Holdings Inc. major phosphate deposits Paradise South and DTree are known to average between 20 wt% and 60 wt% ultrafine particles, containing up to 55 wt% of the phosphate (Teague and Lollback, 2010). After grinding, the proportion of ultrafines can increase to up to 80 wt% of the total feed mass, containing up to 75 wt% of the phosphate. At the beginning of the development of Legend’s beneficiation process, it became clear that if standard practice was followed by desliming and discarding the ultrafines, the project would be potentially uneconomical due to the removal of a large portion of the phosphate values. Hence the major aim of this work was to develop a process that could successfully concentrate ore from either of these two major phosphate deposits without discarding any ultrafines.

Bench scale flotation tests were initially carried out in a conventional Denver cell to confirm the poor floatability of the ultrafine phosphate and to investigate the effect of pulp density and water quality during conditioning and flotation, on phosphate recovery. Some bench scale comparison tests between Denver cells and Jameson cells were carried out, which effectively finalised the design of Legend’s beneficiation process. Lastly, the entire beneficiation process was set-up at laboratory pilot scale to test diamond drill core and RC chip samples, to enable Legend to calculate a reserve for the Paradise South and a pending reserve for the DTree ore deposits. These tests confirmed the novel process was successful using samples with highly variable phosphate feed grades and mineralogy and also established guar gum as the best depressant for Fe_2O_3.

---

[*] Corresponding author.
E-mail address: adamt@axisc.com.au (A.J. Teague).

0892-6875/$ - see front matter Crown Copyright © 2011 Published by Elsevier Ltd. All rights reserved.
doi:10.1016/j.mineng.2011.12.007

2. Theory

Examples of methods of desliming and their importance to effective recovery are discussed by Ahmed (2007). The use of screens or hydrocyclones to deslime the crushed ore dramatically improves the efficiency of flotation but in many cases the loss of phosphate fines is over 15% of the total phosphate. A technical report by Luttrell (2004) explains that beneficiation plants in Florida, one of the major areas for phosphate beneficiation in the USA, typically wash and deslime ore matrix at 150 mesh (105 μm) so that finer than 150 mesh particles are considered tailings and pumped to settling ponds so that approximately 30% of phosphate contained in the original ore is lost to tailings ponds. Thus, while flotation is a very effective method for the beneficiation of phosphate, there remains a need for a process which can recover a portion of fines normally discarded as tailings.

There is some reference in the literature to the flotation of ultrafine phosphate with column cells such as that practised at the Barreiro concentrator in Brazil (Wyslouzil, 2009; Wyslouzil et al., 2010) however there are some drawbacks to this process. The feed to the ultrafine circuit has a particle size of 100% passing 30 μm and is deslimed again at approximately 5–10 μm where 60 wt% of the P_2O_5 is still discarded. Also, of the 40 wt% of the P_2O_5 that is fed to the CPT flotation columns, only 7 wt% is recovered to a marketable concentrate grade of 33.5% P_2O_5 and 33 wt% is discarded as the flotation tailing. It can be seen that this process still has major limitations of; not being able to treat and recover ultrafine particles in the range of 0–10 μm; and having to install additional desliming and flotation circuits for a low additional mass recovery of P_2O_5. The method discussed in the current paper does not have these limitations.

Jameson cell flotation machines have a number of attributes which have made them a popular choice for facilitating the flotation of fine particles in base metal and coal operations. These have been well described and documented in the past by Jameson et al. (1988) and Young et al. (2006) and others. The average air bubble size is 300 μm compared to 1000 μm for conventional flotation machines and the total surface area of 1 mm^3 of air is 20 mm^2 compared with 6 mm^2 for conventional machines. This larger bubble surface area generated by the Jameson cell increases the probability of particle-bubble collision and attachment when floating fine particles. It has been documented in the past that fine and ultrafine particles exhibit improved flotation characteristics when floated with small bubbles (Nguyen and Schulze, 2003; Fillipov, 1998), since the collision efficiency increases and there is an increased chance of particle-bubble attachment.

Although conventional and column cells have been used before in phosphate flotation, there is no reference in the literature to Jameson cells ever being used. This is surprising, given they have been used for some time in potash processing (Eisner, 2010) and they are considered to be better suited to the flotation of fine particles than both conventional and column cells since they generate smaller bubbles. With the development of Legend's process to float the ultrafine phosphate, it was thought there would be merit in trialling the Jameson cell, because of its known success in the minerals industry and its ability to generate small bubbles and recover fine particles.

3. Materials and methods

3.1. Bench scale conventional flotation cell tests

Bench scale studies were carried out on two of Legend's phosphate deposits, Paradise South and DTree, to determine the effect of ultrafines removal, pulp density and water quality during conditioning and flotation and depressants for Fe_2O_3 and Al_2O_3 rejection. Prior to flotation, 1.5 kg of dry solid was wet-ground for 2 min in a 7 l laboratory rod mill at 63 wt% solids using distilled water, until a particle size distribution of 80 wt% passing 150 μm was achieved. The slurry was then wet screened at 150 μm and 20 μm, so that the +150 μm oversize and −20 μm ultrafines were removed respectively. This left 1 kg of solid (−150 μm + 20 μm fraction)∗∗∗ available for flotation. The slurry was dewatered to 75 wt% solids for conditioning, using a pressure filter. High intensity conditioning was carried out by vigorously agitating the solids and reagents in a 1 l stainless steel flask, at 1100 rpm, using a drill press with impellor. Soda ash (Na_2CO_3) was added to the slurry at a dose of 0.8 kg/t and stirred for 2 min until a pH of between 9.3 and 9.5 was achieved. Tall oil fatty acid and diesel were then added together as an emulsion at a dose of 1.3 kg/t (ratio of 1:1) and the slurry stirred for a further 6 min. When the amount of Fe_2O_3 in the feed was known to be 4% by weight or greater, guar gum depressant was added to the slurry as a 1 wt% solution and stirred for a further 2 min. After conditioning, each sample was diluted to 30 wt% solids with distilled water and transferred to the flotation cell.

Batch flotation was conducted in an aerated 2.5 l Denver flotation cell, with a constant impeller speed of 1000 rpm. Froth was scraped at a constant rate during rougher/scavenger flotation until the froth was barren; on average, this took about 4 min. The combined rougher/scavenger concentrate was then re-floated in a cleaning stage using the same flotation cell, without any further reagent or water addition. The flotation products were dried, weighed and assayed for P_2O_5, Fe_2O_3, Al_2O_3, CaO, MgO and SiO_2 by ICP and XRF so that flotation recoveries could be calculated.

3.2. Laboratory pilot plant tests

Continuous flotation tests using drill core and RC chip samples with highly variable mineralogy from the Paradise South and DTree deposits were carried out at laboratory pilot scale. For these tests, the −20 μm ultrafine fraction was not removed from the ore samples prior to conditioning and flotation. Samples of between 30 kg and 50 kg each were tumbled in a tumble mill for 15 min and screened at 25 mm to remove oversize silica. Each sample was wet-ground in a 95 l rod mill with de-ionised water at 63 wt% solids to achieve a particle size of 80 wt% passing 150 μm, thickened to 75% solids by filtration and conditioned using the same reagent scheme as described in Section 3.1.

After conditioning, the slurry was transferred to a baffled 300 l drum and diluted to between 15 and 20 wt% solids using de-ionised water for flotation in a L150 (6 m high) Jameson cell. The slurry was continuously fed to the Jameson cell, at a constant flow rate of 0.35 m^3/h, for rougher flotation. Air to the cell was adjusted to create a vacuum of between 15 and 20 kPa, which produced a feed pressure of between 120 and 150 kPa. When the operation was stable at the set process conditions, a survey was conducted by collecting 5 l feed, concentrate and tailing slurry samples. When the rougher stage was completed, the tailing was collected and fed continuously to the Jameson cell at 0.35 m^3/h, to simulate scavenger flotation. Surveys were again conducted upon reaching stable operation. After the scavenger stage, the Jameson cell was drained and cleaned, and the rougher and scavenger concentrates were collected and fed though the cell to simulate cleaner flotation. The enrichment ratios for each flotation stage were found to vary with feed grade; typically they averaged between 1.8 and 2.7 for the rougher, 3 and 5 for the scavenger and between 1 and 1.5 for cleaner flotation. Similar to the bench scale tests, no further reagent additions were made to the scavenger and cleaner flotation stages. The flotation products were each dried, weighed and assayed for P_2O_5, Fe_2O_3, Al_2O_3, CaO, MgO and SiO_2 by ICP and XRF so that flotation recoveries could be calculated.

During operation of the Jameson cell, only one problem was encountered with short-circuiting. Occasionally the feed pressure to the cell would drop due to airlocks in the pump caused by excess froth in the slurry. When this occurred, it was necessary to stop-start the pump so that the feed pressure and vacuum in the downcomer were restored. Although this problem was encountered occasionally, it was quickly managed and did not affect the flotation results.

3.3. Mineralogy

Phosphorite at Legend's DTree deposit occurs predominantly as mudstone and silty mudstone phosphorite (also known as microsphorite) with some minor occurrences of peloidal grainstone phosphorite (also known as pelletal phosphorite). At Paradise South the phosphorite is made up of pelletal phosphorites almost completely comprised of collophane (carbonate fluorapatite), mudstone phosphorite and replacement phosphorite, which is a more indurated, porcellanous phosphorite (Hough, 2010). Mineralogical studies of DTree and Paradise South ore by QEMSCAN (Jones, 2011) and quantitative XRD (Johnson, 2009) determined the major mineral species to be apatite, quartz, kaolinite and goethite. Average head assays taken from the test samples that were used in this study for DTree and Paradise South ore are given in Table 1.

4. Results and discussion

4.1. Ultrafines removal – conventional flotation cell tests

Table 2 shows results that were obtained from bench scale flotation tests on DTree ore using a conventional Denver cell, with and without ultrafines. These tests were designed to investigate and confirm the effect ultrafine particles have on P_2O_5 concentrate grade and recovery when floated at different pulp densities.

It can be seen from the results that leaving the ultrafines in the ore has a negative impact on the amount of P_2O_5 recovered as well as the concentrate grade, when flotation is carried out in a conventional Denver cell. The difference in P_2O_5 recovery is more pronounced at 20 wt% solids (51.1%) compared to 30 wt% and 40 wt% solids (25.2% and 31.6%), nevertheless the difference is significant at each pulp density. Given the phosphate industry has used conventional flotation cells for many years, the poor recoveries in these tests demonstrate why the ultrafines are commonly separated from the ore and discarded before flotation. The results also show that in this case, 30 wt% solids is the optimum pulp density to achieve the highest P_2O_5 recovery and grade for DTree ore, marginally better recovery (6.7%) than 40 wt% solids with ultrafines and the same recovery (94%) with slightly higher grade (2.3%) without ultrafines.

4.2. Effect of pulp density during conditioning

Bench scale tests using Paradise South ore, were designed to investigate the effect of pulp density during conditioning on the flotation recovery of P_2O_5. It has been previously documented that conditioning at high wt% solids promotes collector adsorption onto phosphate surfaces and reduces the activation of silica (Gruber et al., 1995). Table 3 shows the results. It can be seen that conditioning at low wt% solids yields poor P_2O_5 recoveries and grades. The P_2O_5 recoveries increase from a low 10.3% at 30 wt% solids to a high of 97.8% when the slurry is conditioned at 75 wt% solids. There is not much difference in concentrate grade when the slurry is conditioned at 51 or 75 wt% solids but the P_2O_5 recovery is over 50% greater when conditioned at 75 wt% solids indicating that the hydrophobicity of the P_2O_5 in the feed has increased markedly under these conditions, resulting in improved floatability. Gruber et al. (1995) explain that the benefits of conditioning at high wt% solids are twofold; firstly, it increases the concentration (moles per litre) of the collector and therefore the force driving chemisorption of collector onto phosphate is increased and secondly, the effects of dissolved cations such as Ca^{2+} and Mg^{2+} are reduced when the quantity of water is reduced so that less reagent is lost due to bulk precipitation and there are fewer cations available to activate quartz. The results in Table 3 confirm this hypothesis.

4.3. Effect of water quality

Further to the discussion in the previous Section 4.2, the study undertaken by Jacobs Engineering (Gruber et al., 1995) clearly showed that the use of distilled or de-ionised water in phosphate conditioning and flotation, instead of plant or tap (hard) water, increased the recovery of phosphate and reduced the recovery of quartz. They suggested that the Ca^{2+} ions in the hard water activated the quartz and residual oleate ions from the collector were physically adsorbed onto the quartz in the flotation cell. In the current study, bench scale tests were carried out using DTree ore to confirm this effect and the results are given in Table 4, whilst Fig. 1 shows a plot of water hardness as $CaCO_3$ equivalent versus P_2O_5 recovery. The results demonstrate that the flotation recovery of P_2O_5 is inversely proportional to the hardness of the water that is used in grinding, conditioning and flotation, for instance 97.6% P_2O_5 recovery for distilled water with <1 mg/L hardness and only 59.9% P_2O_5 recovery for fresh site water with 410 mg/L hardness. Interestingly, the result for Test 4 of 97.1% P_2O_5 recovery using IX softened site water, confirmed that the ion-exchange apparatus successfully removed hard cations to a level <0.1 mg/L hardness, making it similar in quality to distilled water. The P_2O_5 grade of the concentrates was found to increase from 26.6% P_2O_5 to 30.9% P_2O_5 as the water hardness decreased, although this did not appear to be as linear as the relationship between water hardness and P_2O_5 recovery as illustrated in Fig. 1. When hard water was used during conditioning, a pH change in the slurry was observed, although more soda ash was needed to raise the pH from approximately eight to the target of between 9.3 and 9.5 and hence neutralise the concentration of Ca^{2+} in the water. Elevated levels of water hardness resulted in more silica reporting to the phosphate concentrate which agrees with the investigations carried out with Gruber et al. (1995) and their hypothesis of Ca^{2+} activating silica as discussed previously in this section. The water hardness may have also contributed to the lower phosphate recovery and concentrate grade by the reaction between fatty acid collector and Ca^{2+} and Mg^{2+}, reducing the efficiency of the collector as found by Gruber et al. (1995).

In a full scale operation it is expected that all the recycled process water would be collected into a common collection tank which would then pass through a filter to remove fine particulate

Table 1
Average head assays for DTree and Paradise South ore.

Ore type	P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	MgO (%)	CaO (%)	SiO_2 (%)
DTree	15.52	5.45	6.13	0.35	21.43	47.28
Paradise South	15.58	6.29	5.40	0.70	22.91	39.92

Table 2
DTree conventional flotation cell tests with and without ultrafines.

Pulp wt% solids	DTree with ultrafines		DTree without ultrafines	
	Concentrate grade P_2O_5 (%)	P_2O_5 recovery to concentrate (%)	Concentrate grade P_2O_5 (%)	P_2O_5 recovery to concentrate (%)
20	29.3	37.1	34.2	88.2
30	27.4	69.2	36.4	94.4
40	26.5	62.5	34.1	94.1

Table 3
Effect of pulp density during conditioning on flotation of Paradise South ore.

Test	Pulp wt% solids in conditioning	Feed grade P_2O_5 (%)	Concentrate grade P_2O_5 (%)	P_2O_5 recovery to concentrate (%)
1	30	17.63	20.76	62.45
2	30	17.63	30.09	10.30
3	51	19.50	31.50	45.50
4	51	19.30	32.50	41.40
5	75	19.80	30.80	97.80
6	75	19.60	32.00	97.40

solids and ion exchange system to remove Ca^{2+} and Mg^{2+}, such as that facilitated by industrial water providers.

4.4. Conventional flotation cell versus Jameson cell

Table 5 shows results for the flotation of DTree ore with ultrafines in a conventional Denver cell versus a L150 Jameson cell at a pulp density of 20 wt% solids. From these results it can be seen that even though the cell types were tested using DTree samples of varying P_2O_5 feed grades, the difference in performance of the Denver cell and Jameson cell in terms of P_2O_5 recovery (30–50%) is obvious and significant. The Jameson cell tests were found to be very encouraging since they were the first trials ever attempted using phosphate ore and as such the process conditions were far from optimised, yet they still gave an outstanding result on material with 75 wt% ultrafines. The high P_2O_5 recoveries of 82.9% and 90% for Tests 3 and 4 at grades of 27.5% P_2O_5 and 29.7% P_2O_5 respectively, demonstrate that on a grade/recovery curve, the recovery could potentially be reduced to improve the grade to >30% P_2O_5. The results demonstrate that the use of the Jameson cell has a significant impact on the recovery of ultrafine phosphate particles compared to the conventional Denver cell, most likely due to the presence of small bubbles in the mixing zone which increase the probability of particle-bubble collision and attachment. Even though the average bubble size was not measured in this study, over the last twenty years it has been a major focus in Jameson cell research; Evans et al. (1992, 1994) measured and determined the average bubble size to be between 360 μm and 950 μm compared to conventional column cells that had an average bubble size between 2 mm and 3 mm. This was proven for full scale and pilot cells of various sizes. In the current study, it has been assumed that the bubble size generated during operation of the L150 pilot cell is not significantly different to that measured in other Jameson cells.

4.5. Depression of Fe_2O_3 and Al_2O_3

Bench scale flotation tests were carried out on Paradise South ore using a Denver cell to compare a number of reagents known to depress Fe_2O_3 and Al_2O_3 during phosphate flotation (Finch, 2009). These included corn starch, potato starch and guar gum. It was necessary for Legend to find a suitable depressant for these species so that the rock concentrate would be more amenable to downstream manufacture of the fertilizers mono-ammonium-phosphate (MAP) and di-ammonium-phosphate (DAP). On average, Paradise South and DTree run of mine (ROM) ore contain approximately 5–6% each of Fe_2O_3 and Al_2O_3 by weight, so the aim of these tests was to find a reagent that could reduce these contaminants to as low a level as possible in the flotation concentrate. From the results in Table 6 it can be seen that the best depression of Fe_2O_3 was achieved in Test 7, using 500 g/t of guar gum. This test reduced the grade of Fe_2O_3 by the greatest margin, compared to the other depressants, from 4.40% to 2.89%, whilst the Al_2O_3 grade was reduced from 0.93% to 0.74%. The next best results were Tests 4 and 5, using 1500 g/t of corn starch and 500 g/t of potato starch respectively. In Test 4, the corn starch produced the lowest concentrate grade of Fe_2O_3 and Al_2O_3 out of all the tests (2.78% Fe_2O_3 and 0.72% Al_2O_3) and in Test 5, the potato starch gave a similar margin of Fe_2O_3 grade reduction to Test 4 (1.47% compared to 1.48%). The P_2O_5 recoveries to the concentrate are all consistently between 97% and 98% which is a good result, given that starch and guar gum are known to also adsorb onto phosphate surfaces and cause flotation depression if added in excessive quantities. From these tests, guar gum was selected as the overall best depressant because it performed the best in reducing the concentrate grade of Fe_2O_3 and Al_2O_3 at a minimal dosage of between 250 g/t and 500 g/t, compared to the other reagents which used between 500 g/t and 1500 g/t.

4.6. Laboratory pilot plant

The results from the bench scale work led to the refinement and finalisation of the process flowsheet. This flowsheet was then constructed at laboratory pilot plant scale at Amdel laboratories in Adelaide, Australia to carry out beneficiation tests on Paradise South and DTree ore, to enable Legend to calculate a mineable reserve for these deposits. All tests were completed without the removal of any −20 μm particles. The results of this work proved the novel process worked effectively using diamond drill core and RC chip samples with highly variable feed mineralogy and wt% proportions of −20 μm particles.

4.6.1. Paradise South

Table 7 shows a selection of flotation results from the Paradise South reserve determination tests whilst Figs. 2 and 3 illustrate the relationship between the grade of Fe_2O_3 and Al_2O_3 in the feed versus concentrate respectively.

The results in Table 7 indicate excellent P_2O_5 recoveries of between 76% and 97.3% from a variety of feed grades ranging from a low 6.46% P_2O_5 to 23.8% P_2O_5. The result for Test 7, giving a concentrate grade of 34.7% P_2O_5 from 6.46% P_2O_5 feed whilst

Table 4
The effect of water quality on flotation of DTree ore.

Test	Water type	Hardness as $CaCO_3$ equivalent (mg/L)	Feed grade P_2O_5 (%)	Concentrate grade P_2O_5 (%)	Concentrate grade SiO_2 (%)	P_2O_5 recovery to concentrate (%)
1	Distilled	<0.1	17.1	30.9	17.6	97.6
2	IX softened site	<0.1	17.1	29.8	18.2	97.1
3	Recycled softened site	110	18.7	22.3	28.0	78.8
4	Brisbane tap	120	20.2	29.5	22.3	76.2
5	Fresh site	410	17.5	26.6	20.9	59.0

recovering 91.2% $P_2O_{5.}$ is outstanding. Tests 4 and 6 also gave excellent results from low grade feeds of about 10% P_2O_5; they recovered between 85% and 86% P_2O_5 to concentrate grades between 33% and 34% P_2O_5 from feed samples that consisted of between 47 wt% and 54 wt% −20 μm particles. Most phosphate operations around the world would consider the feed grades of Tests 4, 6 and 7 to be too low to process using established methods, so these results really highlight the benefits of the using the new beneficiation process described in this paper. These results indicate that this method of high wt% solids conditioning, de-ionised water and flotation with Jameson cells is very efficient and selective in recovering phosphate particles, even when the feed consists of a high wt% proportion of −20 μm particles.

With regard to Fe_2O_3 depression, the results show that 250 g/t guar gum performed well when it was used in 5 out of the 7 Tests, reducing the Fe_2O_3 grade by margins between 17% and 70%. The best result was achieved in Test 5 which reduced the Fe_2O_3 grade from 7.98% in the feed to 2.38% in the concentrate. Fig. 2 illustrates the relationship between Fe_2O_3 in the feed versus Fe_2O_3 in the concentrate for the entire campaign of tests using Paradise South ore. It can be seen that there is a straight line relationship between the Fe_2O_3 grade in the feed and concentrate, indicating consistent depression of Fe_2O_3 by the guar gum.

In terms of Al_2O_3 depression, it can be seen from the feed and concentrate grades in Table 7, that the process is extremely efficient in rejecting Al_2O_3 regardless of the feed grade and whether guar gum depressant has been used or not. The Al_2O_3 grade is consistently reduced by between 82% and 87% to levels ranging from 0.66% Al_2O_3 to 1.14% Al_2O_3 with the exception of Test 7 which reduced the grade from 5.34% Al_2O_3 to 2.23% Al_2O_3. The best individual result was in Test 4 where 7.22% Al_2O_3 in the feed was reduced to 0.94% Al_2O_3 in the concentrate. Fig. 3 shows a scatter plot of Al_2O_3 grade in the feed versus Al_2O_3 grade in the concentrate. The amount of scatter in the data indicates that there is no relationship between the Al_2O_3 in the feed and concentrate, because regardless of the feed grade, only minimal Al_2O_3 is recovered to the concentrate. The Al_2O_3 appears to behave differently to the Fe_2O_3 during flotation; it is postulated that it does not adsorb fatty acid collector and become hydrophobic like the Fe_2O_3, hence the guar gum has no effect on its depression, since it is already hydrophilic. It is also thought that the unique operating mechanism of the Jameson cell has added to the excellent Al_2O_3 rejection because of its small bubble generation and froth washing, resulting in only minimal entrainment of Al_2O_3 into the froth.

4.6.2. DTree

Table 8 shows a selection of flotation results from the entire campaign of DTree pending reserve determination tests, whilst Figs. 4 and 5 illustrate the relationship between the grade of Fe_2O_3 and Al_2O_3 in the feed versus concentrate respectively.

Similar to the results discussed in Section 4.6.1 for Paradise South, the DTree results shown in Table 8 demonstrate excellent P_2O_5 recoveries ranging from 76.7% to 95.6% from samples of

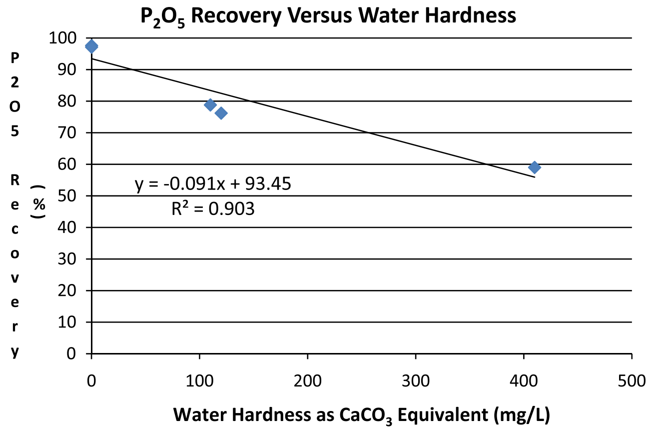


Fig. 1. P_2O_5 recovery versus water hardness.

variable feed grades (10.6% P_2O_5–19.1% P_2O_5) and wt% proportions of −20 μm particles (40.7–65.2%). In general the wt% proportions of −20 μm particles in the milled DTree samples were higher than for the Paradise South samples, nevertheless, the beneficiation method still produced excellent flotation recoveries and concentrate grades. Some of the best results included Test 7 which achieved the highest P_2O_5 recovery of 95.6% at a concentrate grade of 33.3% P_2O_5 from a feed grade of 16.9% P_2O_5 with 53.4 wt% −20 μm particles and Test 5 which gave 92.4% P_2O_5 recovery at a concentrate grade of 30.2% P_2O_5 from a low feed grade of 10.6% P_2O_5 with 49.7 wt% −20 μm particles.

The Fe_2O_3 grades in the feed and concentrate for Tests 2, 3, 5, 6 and 8 indicate that the guar gum at 250 g/t worked effectively as a depressant. The reductions in Fe_2O_3 grade range from 6.6% to 67%, although on average the reductions are less than they were for the Paradise South tests, which may have been due to the presence of fine goethite needing further liberation. The best individual result was Test 6 which reduced the Fe_2O_3 feed grade of 8.36% to a concentrate grade of 2.74%.

Fig. 4 illustrates the relationship between Fe_2O_3 in the feed versus Fe_2O_3 in the concentrate and similar to Fig. 2, gives a straight line relationship. This result suggests that the Fe_2O_3 mineralogy in both DTree and Paradise South is similar and responds to the guar gum depressant in the same way, although the R^2 value for DTree is slightly lower than for Paradise South (0.61 compared to 0.73) which could mean some of the goethite in the DTree samples was not as well liberated as it was for the Paradise South tests.

From the results in Table 8, it can be seen that similar to the Paradise South results, the process rejects Al_2O_3 from the phosphate concentrate extremely well, regardless of the feed grade and whether guar gum has been used or not. The Al_2O_3 grade is consistently reduced by between 66% and 88% to grades ranging from 0.64% Al_2O_3 to 1.67% Al_2O_3. The best individual result was achieved in Test 8 which gave a concentrate grade of 0.64% Al_2O_3 from a feed grade of 7.75% Al_2O_3. Fig. 5 shows a scatter plot of Al_2O_3 grade in

Table 5
Flotation of DTree ore using a Denver cell and Jameson cell.

Test	Cell type	Wt% passing 20 μm	Feed grade P_2O_5 (%)	Concentrate grades			Recovery
				P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	P_2O_5 (%)
1	Denver	72	15.8	29.3	7.3	2.7	37.1
2	Denver	52	16.4	33.8	2.9	1.2	52.0
3	Jameson	75	21.4	27.5	1.8	1.4	82.9
4	Jameson	75	21.7	29.7	1.6	1.7	90.0

Table 6
Flotation of Paradise South ore comparing depressants.

Test	Reagent	Addition rate (g/t)	Head grade			Final grade			Recovery
			P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	P_2O_5 (%)
1	Baseline	–	19.8	4.42	1.03	30.8	3.49	0.94	97.8
2	Corn starch	500	19.6	4.35	0.94	31.3	2.91	0.79	97.4
3	Corn starch	1000	19.7	4.35	0.95	31.1	2.89	0.80	97.7
4	Corn starch	1500	19.9	4.25	0.91	32.4	2.78	0.72	98.0
5	Potato starch	500	19.5	4.42	0.97	32.0	2.94	0.75	97.6
6	Guar gum	250	19.7	4.31	0.96	30.5	2.97	0.82	97.7
7	Guar gum	500	19.5	4.40	0.93	31.4	2.89	0.74	97.3

Table 7
Paradise South laboratory pilot plant flotation results.

Test	Wt% passing 20 μm	Head grade				Concentrate grade				Recovery
		P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	SiO_2 (%)	P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	SiO_2 (%)	P_2O_5 (%)
1	46.7	20.9	5.63	5.77	29.8	36.4	2.19	0.86	8.30	96.4
2*	48.6	23.8	3.55	3.65	30.9	36.7	2.13	0.66	4.30	97.3
3	41.3	16.2	5.35	7.01	38.4	33.0	3.33	1.14	6.30	94.1
4	47.2	10.1	2.75	7.22	56.2	34.3	2.05	0.94	6.10	85.2
5	55.3	10.2	7.98	4.47	51.3	30.4	2.38	0.91	11.0	76.0
6	54.3	10.4	4.80	4.48	59.5	33.0	3.99	0.98	10.7	86.5
7*	35.5	6.46	0.92	5.34	75.1	34.7	1.17	2.23	9.13	91.2

* Indicates tests where no guar gum was added during conditioning.

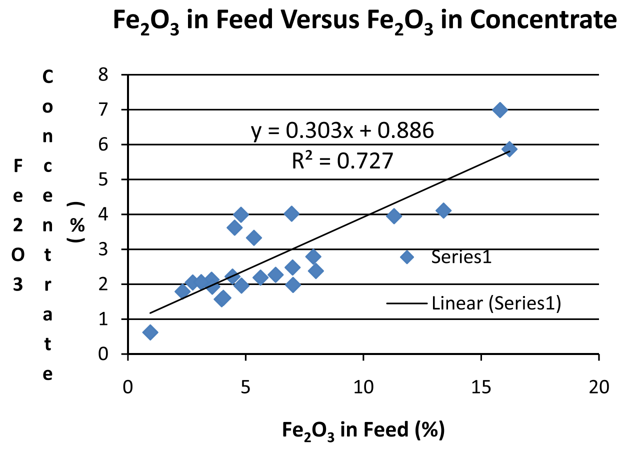


Fig. 2. Fe_2O_3 grade in feed versus Fe_2O_3 grade in concentrate for Paradise South laboratory pilot plant tests.

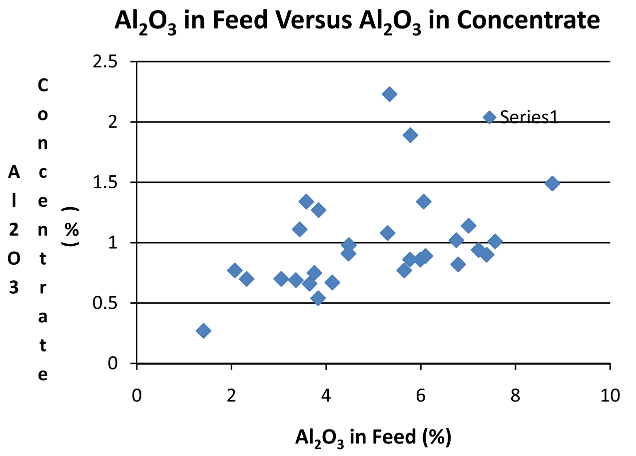


Fig. 3. Al_2O_3 grade in feed versus Al_2O_3 grade in concentrate for Paradise South laboratory pilot plant tests.

the feed versus Al_2O_3 grade in the concentrate. Again similar to the Paradise South tests, the amount of scatter in the data demonstrates there is no relationship between the Al_2O_3 in the feed and concentrate, because the beneficiation process has performed so well it has rejected Al_2O_3 regardless of the feed grade. Overall, the similarity of the P_2O_5, Fe_2O_3 and Al_2O_3 flotation results between DTree and Paradise South ore indicate the excellent repeatability and robust nature of the designed beneficiation process.

4.6.3. Final process method

The results from all of the test work discussed in this paper culminated in the design of Legend's beneficiation process which is summarised here and is currently in patent pending status following the submission of an international patent application (Teague, 2011). This method allows the successful flotation recovery of phosphate from ore with a particle size distribution of up to 80 wt% passing 20 μm.

Table 8
DTree laboratory pilot plant flotation results.

Test	Wt% Passing 20 μm	Head grade				Concentrate grade				Recovery
		P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	SiO_2 (%)	P_2O_5 (%)	Fe_2O_3 (%)	Al_2O_3 (%)	SiO_2 (%)	P_2O_5 (%)
1*	52.4	19.1	2.99	5.84	37.1	33.8	2.76	1.65	8.70	83.7
2	40.7	17.6	4.20	4.93	43.1	30.6	3.68	1.67	15.5	92.4
3	44.9	17.3	5.18	5.09	40.6	32.0	4.62	1.64	11.2	88.4
4*	54.3	18.7	2.41	6.10	39.1	31.8	2.25	1.15	13.7	92.1
5	49.7	10.6	5.84	6.05	54.9	30.2	3.82	1.11	15.2	92.4
6	56.5	15.3	8.36	7.12	40.8	35.1	2.74	0.84	6.40	79.4
7*	53.4	16.9	3.45	5.54	45.6	33.3	3.74	0.91	12.2	95.6
8	65.2	16.3	3.24	7.75	43.3	36.2	1.09	0.64	8.10	76.7

* Indicates tests where no guar gum was added during conditioning.

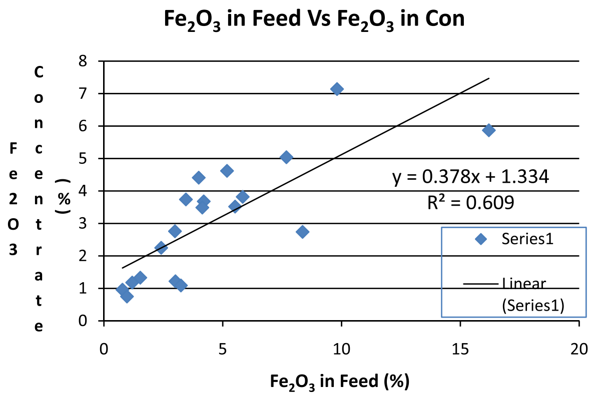


Fig. 4. Fe_2O_3 grade in feed versus Fe_2O_3 grade in concentrate for DTree laboratory pilot plant tests.

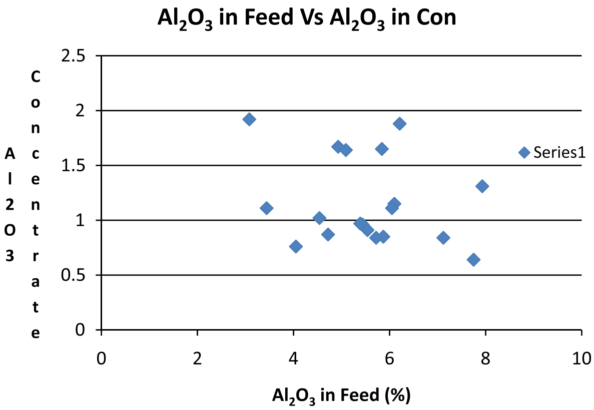


Fig. 5. Al_2O_3 grade in feed versus Al_2O_3 grade in concentrate for DTree laboratory pilot plant tests.

The slurry is formed by milling the ore in a rod mill, followed by classification using a hydrocyclone to provide a size fraction of at least 80 wt% passing 150 μm, then it is dewatered to at least 70 wt% solids using a belt filter. Conditioning is carried out by contacting the slurry at 70 wt% solids with at least one collecting agent selected from fatty acids and salts thereof such as tall oil fatty acid (C12–C36) and at least one hydrocarbon such as diesel. Typically the dose of collector and hydrocarbon are in the range of 0.5–3 kg per tonne of dry solid contained in the slurry. The collector and hydrocarbon may be added to the aqueous slurry together or separately. They can be added neat or in the presence of diluents such as alkaline aqueous diluents if desired. In cases where the content of Fe_2O_3 and Al_2O_3 is relatively high such as at least 4% each by weight of the solids content of the slurry, guar gum can be used to depress these constituents. The guar gum may be added to the slurry at a dosage of 0.05–1 kg/t.

A pH adjusting agent such as soda ash (Na_2CO_3) may be added at the milling stage or up to or including the conditioning tank so that a pH of between 9.3 and 10 is achieved. The soda ash may be added and mixed with the slurry in the conditioning tank, for example for 2 min and the fatty acid collector and diesel together or separately added and mixed with the slurry for a further period of, for example, 6 min. Where used, the guar gum may then be added and mixed with the slurry for a period (for example 2 min). Typically when it is used, the guar gum is mixed with the slurry after the phosphate particles have been made hydrophobic by the collector and hydrocarbon.

The method comprises a step of diluting the conditioned aqueous slurry to provide a solids content of no more than 35% by weight and preferably 20% solids by weight. The diluted conditioned slurry is introduced to the one or more downcomers of a Jameson flotation cell. Rougher, scavenger and cleaner flotation is carried out using Jameson cells to obtain a 32% P_2O_5 concentrate with at least 80% P_2O_5 recovery.

The water used in the method may be fresh water, recycled water or a mixture. It is particularly preferred that the water used in the method has a concentration of no more than 10 mg/L combined Ca^{2+} and Mg^{2+} and most preferably no more than 1 mg/L. The water can be treated using an ion exchange unit to reduce the concentration of Ca^{2+} and Mg^{2+} ions in the water, for example by exchange with Na^+ ions. The reduction in the combined concentration of calcium and magnesium significantly improves adsorption of the collector onto the phosphate during conditioning and this improves flotation performance.

5. Conclusions

A process has been designed which successfully beneficiates phosphate ore with highly variable mineralogy and ultrafine (−20 μm) size fractions of up to 80 wt% of the feed. The process uses conditioning with reagents at high wt% solids (at least 70 wt%) and flotation with Jameson cells in a rougher, scavenger, cleaner configuration to recover at least 80% P_2O_5 at a grade of 32% P_2O_5 or greater. The Jameson cell was found to have an advantage over conventional flotation cells when treating ultrafine particles, due to their intense mixing zone and propensity to form small bubbles. The use of de-ionised water in the process was also found to be important to minimize the concentration of hard cations that could activate silica and hence adsorb collector, thereby interfering with collector adsorption onto phosphate, decreasing its floatability.

Acknowledgements

Thanks to Legend International Holdings Inc. for funding the research and providing the ore samples. Lyndon Ryan and Le Huynh from Xstrata Technology are thanked for their technical assistance and supply of the Jameson cells. Jim Thorpe, Adam Wilkinson and staff at Amdel Ltd are thanked for their assistance with carrying out the test work.

References

Ahmed, 2007. Optimisation of desliming prior to phosphate ore upgrading by flotation. Physicochemical Problems of Mineral Processing 41, 79–88.

Eisner, M., 2010. Is Canada Ready for the Jameson Cell. CIM Bulletin 5 (2).

Evans, G.M., Jameson, G.J., Atkinson, B.W., 1992. Prediction of the bubble size generated by a plunging liquid jet bubble column. Chemical Engineering Science 47 (13/14), 3265–3272.

Evans, G.M., Atkinson, B.W., Jameson, G.J., 1994. In: Matis, K.A. (Ed.), The jameson cell. flotation science and technology. Marcel Dekker Inc, New York.

Fillipov, Y.M., 1998. Flotation of Fine Particles and Intergrowths by Small Air Bubbles. Journal of Mining Science 34 (5).

Finch, E. 2009. Summary of Lady Annie Bulk Samples Preliminary Testing. Confidential report to Legend International Holdings Inc, Florida.

Gruber, G.A., Moudgill, B.M., Somasundaran, P. 1995. Understanding the Basics of Anionic Conditioning in Phosphate Flotation. Jacobs Engineering Group, Inc., Florida Institute of Phosphate Research, Bartow, Florida.

Hough, M., 2010. Geology of Samples. Internal Report. Legend International Holdings Inc., Melbourne.

Jameson, G.J., Belk, M., Johnson, N.W., Espinosa-Gomez, R., Andreaditis, J.P., 1988. Mineral flotation in a high intensity column. Chemeca 88. In: 16th Australian Conference on Chemical Engineering. Sydney. pp. 507–510.

Johnson, R. 2009. Lady Annie and D-Tree Iron Mineral XRD Report 1. Confidential report for Legend International Holdings Inc. Crescent Technology Inc. Louisiana.

Jones, K., 2011. Min623 Legend Mineralogy Data – Summary. Confidential report for Legend International Holdings Inc. Ammtec, Perth.

Kogel, J.E., Trivedi, N.C., Barker, J.M. and Krukowski, S.T., 2006. Industrial Minerals and Rocks: Commodities, Markets and Uses. In: 7th Ed. Society for Mining, Metallurgy and Exploration (U.S.). pp. 714–720.

Luttrell, G.H. 2004. Phase I technical report on "In Plant Testing of High –Efficiency Hydraulic Separators, Virginia Polytechnic Institute & State University (Virginia Tech) pp. 67–102.

Nguyen, A.V., Schulze, H.J., 2003. Colloidal Science of Flotation. CRC Press.

Teague, A.J. 2011. Method of Beneficiation of Phosphate. International patent application PCT/AU2011/000651. Legend International Holdings Inc. Melbourne, Australia.

Teague, A.J., Lollback, M.C. 2010. Paradise South Reserve Beneficiation Test Work Report – Campaigns 1 and 2. Legend International Holdings Inc. Internal Report. Melbourne.

Wyslouzil, H.E., 2009. The Use of Column Flotation for the Recovery of Ultra-Fine Phosphates. http://en-ca.eriez.com/Products/Markets/mineralflotation.

Wyslouzil, H.E., Kohmeunch, J., Christodoulou, L., Fan, M. 2010. Coarse and Fine Particle Flotation. In: Proceedings of the 48 mining engineering congress.

Young, M.F., Barnes, K.E., Pease, J.D., 2006. Jameson Cell: The "Comeback" in Base Metals Applications Using Improved Design and Flow Sheets. In: 38th Annual Meeting of the Canadian Mineral Processors. Ottawa, Ontario, Canada.

Separation efficiency improvement of a low grade copper-gold flotation circuit

David SEAMAN / Peter MANTON / Paul GRIFFIN
Newcrest Mining, Australia

Abstract

Newcrest's Telfer gold mine, located in the South Western area of the Great Sandy Desert in the Paterson Province of Western Australia, processes low grade copper-gold ores from open pit and underground sources in two parallel processing trains. The copper flotation circuits upgrade copper content by a factor of 80-160 times from feed to a saleable copper concentrate containing approximately 50% of the contained gold in feed (remaining gold is recovered both as a gravity product within the primary grinding circuit and also via leaching of a gold-containing pyrite concentrate).

The copper concentrate grades achieved from the flotation circuit at Telfer have typically fallen well short of the theoretical achievable grades despite the majority of copper being hosted in high copper bearing minerals: chalcopyrite, chalcocite, bornite and others. The recovery of gold-containing pyrite to the copper concentrate is encouraged to achieve a higher total gold recovery provided that a saleable copper concentrate grade is still produced.

Mineralogical diagnosis of feed and product streams from the copper flotation circuit have shown that the copper bearing minerals are well liberated and thus further liberation of the copper minerals was not selected as a method of improving Telfer's concentrate grade or recovery from current main dome ore sources. Furthermore, repeated mineralogical analyses have demonstrated that there is significant dilution in the copper concentrate resulting from the presence of fine, well liberated non-sulphide gangue minerals.

This paper presents a diagnosis of the flotation circuit that resulted in a proposal for a circuit modification that afterwards was indeed implemented, aiming at improving the existing flotation circuit with respect to the rejection of non-sulphide gangue, which is achieved by improved separation efficiency of the copper flotation circuit. The techniques used for diagnosis include: quantitative mineralogical evaluation of key streams, flotation circuit surveys, floatability component model fitting and simulation, and the application of dilute batch flotation tests.

INTRODUCTION

Telfer is a gold/copper operation located in the Pilbara region of Western Australia. Open pit mining (Main Dome) recommenced in 2003, followed by an underground mine (Telfer Deeps) in mid-2006. Ore mineralogy varies significantly from mainly chalcocite in the open pit ore to predominately chalcopyrite in the underground ore. The ore processing plant consists of two parallel trains, currently treating a total of 21 million tonnes of ore per annum including approximately six million tonnes from the underground mine. Train 1 is receiving a blend of the underground and open pit ores, while Train 2 is treating open pit ore alone. Details of the mine geological and ore mineralogical information, the initial process plant design criteria and operating strategies and a summary of the commissioning phase, can be found in previous publications by Goulsbra et al. (2003) and Benson et al. (2007).

Ore is processed through both trains in a variety of configurations. The predominant configuration is sequential flotation, where copper bearing minerals are recovered to a saleable copper concentrate, followed by re-activation and flotation of the pyrite which is leached with cyanide to recover gold.

Both trains produce a copper concentrate with significant quantities of non-sulphide gangue minerals. The target concentrate grade is approximately 16% Cu, with copper minerals, chalcopyrite and chalcocite, being the most abundant copper bearing minerals in the feed to the plant. Stoechiometrically, these minerals contain 34.6 and 79.9% Cu respectively (up to five times the copper grade produced at Telfer).

A reduction in the non-sulphide gangue content of the concentrate will result in an improvement of the copper concentrate grade and of the ability of the plant to recover additional pyrite (containing gold) to the copper concentrate, thereby improving gold recovery from the operation. Zheng, Crawford & Manton (2009) presented details on how a reconfiguration of Train 1 was completed to assist rejection of some of this gangue. While this modification was successful, further improvements are still possible.

This paper presents diagnostic investigations carried out to determine the quantity and nature of this non-sulphide gangue reporting to the copper concentrate. Subsequently, a modification to the cleaning circuit is proposed, that aims at rejecting a significant quantity of the non-sulphide gangue.

MINERALOGY OF COPPER CONCENTRATE

Since processing recommenced at Telfer, there have been several size and mineralogical quantifications of monthly composites from both operating trains. **Figure 1** below shows a summary of the mineral abundance in the copper concentrate streams from all trains. These data are taken from a recent (October 2010) quantification that was completed by G&T Metallurgical Services Ltd. (Ma & Johnston, 2011). It is evident in the mineral abundance that approximately 40% (on a mass basis) of the copper concentrate produced is made up of non-sulphide gangue. Although not presented here, similar mineral abundances have been found in other mineralogical campaigns carried out over the last two years at Telfer.

The greatest source of gangue in the copper concentrate is the re-cleaner concentrate streams and the Train 1 rougher concentrate A.

Figure 2 shows the size distribution and liberation characteristics of the non-sulphide gangue contained in the two re-cleaner streams. Note that liberated gangue is defined as those particles containing more than 98% of the gangue surface area (measured in 2D) as measured by G&T in a QEMSEM (Ma & Johnston, 2011). Firstly, it is evident that the non-sulphide gangue present in the copper recleaner concentrate streams is very well liberated in most size fractions. Secondly, the gangue is finely distributed with 80% of the total gangue from these streams containing particles of less than 30 μm in size.

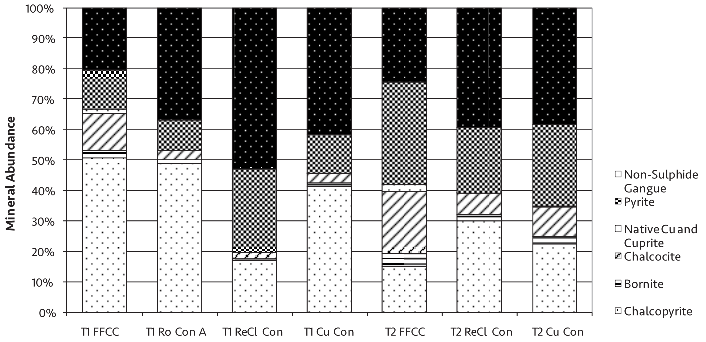


Figure 1 Mineral abundance in concentrate streams from the Telfer plant in October 2010[1]

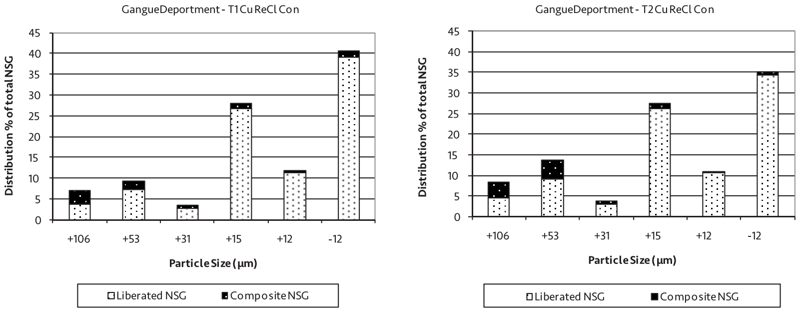


Figure 2 Non-sulphide gangue liberation and size distribution, October 2010 (Ma & Johnston, 2011)

Analysis of the mineralogical data reveals that theoretically, a final concentrate assaying about 20 to 25% copper could be achieved in both trains by rejecting half of the liberated gangue currently reporting to final concentrate. This theoretical improvement would be equivalent to an improvement in overall copper and gold recovery by moving further down the grade-recovery curve of the operation.

The mineralogical deportment and size of the gangue presented in these two graphs suggest that the non-sulphide gangue is being recovered to the flotation concentrate by the mechanism of entrainment. The entrainment process is size dependant, with more recovery by entrainment occurring at finer particle sizes due to their lower settling velocities in flotation froth. Johnson, McKee & Lynch (1975) and Savassi et al. (1998)

---

[1] In this figure, FFCC refers to the flash flotation cleaner concentrate; Ro Con A refers to the first two rougher cell concentrates which are currently sent to final concentrate – as described in Zheng, Crawford & Manton (2009); ReCl Con refers the re-cleaner concentrate; Cu Con refers to the final combined copper concentrate from each respective train. T1 and T2 refer to the two operating trains – Train 1 and Train 2.

provide further details on the entrainment mechanism. A common expression used to describe the entrainment process is the use of an entrainment classification function, ENT, which is dependant on the size distribution of the particles in the slurry phase:

$$M_{ent,\boldsymbol{i}} = ENT_i \cdot R_w \cdot M_{tail,i} \tag{1}$$

Where, ENT_i is the entrainment classification number for particles of size class i; $M_{ent,i}$ is the mass flow rate of particles in the i size class collected in the concentrate by the entrainment mechanism; $M_{tail,i}$ is the mass flow rate of particles in the i size class in the tailings stream; and R_w is the water recovery of the flotation cell. Values of ENT_i for different size classes can be found in Johnson, McKee & Lynch (1974) or these can be determined from a non-floating fully liberate tracer mineral in the ore body. ENT_i typically ranges from zero for particles greater than 53 μm and approaches a value of one for ultra-fine particles.

Once the ore has been ground and a size distribution of particles is presented to a flotation cell, the only parameter that can be manipulated to reduce the entrainment is water recovery. If additional water is added to dilute the contents of the flotation cell, and the flow rate of water to the concentrate remains the same, the water recovery will be reduced. The downside of diluting flotation banks is the resulting loss of residence time which can have a negative effect on valuable mineral recovery.

FLOTATION CIRCUIT DIAGNOSTICS

Surveys of each flotation train were conducted to investigate the potential for diluting the cleaner circuits further, with the objective to assist in entrainment rejection and to provide a circuit balance. This balance is used to build a flotation model that is used to investigate further configuration changes to the circuit. Dilution cleaning tests were also carried out with the aim of confirming whether the non-sulphide gangue in the concentrate streams was being recovered by entrainment or was hydrophobic and therefore recovered by true flotation.

Mass balance results

A survey and mass balance was completed on each flotation train. The %solids (w/w) and calculated residence times of each flotation bank are shown for both trains in the table below.

Table 1 Estimated densities and residence times of flotation banks

	Train 1		**Train 2**	
	%solids	**τ (min)**	**%solids**	**τ (min)**
Cu Ro	22.0	10.2	34.3	15.0
Cu Cl	13.0	10.9	10.5	13.6
Cu Cl Scav	9.0	14.9	9.2	14.4
Cu ReCl	13.4	17.2	10.1	20.7

As can be seen in **Table 1**, the cleaner and recleaner banks are being operated with sufficient residence time to achieve flotation exhaustion of floatable minerals. In order to improve cleaning efficiency by further dilution of the cleaner and recleaner banks, additional flotation capacity would be required in the cleaning circuit.

Dilute cleaning batch flotation tests

Plant samples were collected, diluted and floated immediately to assess the NSG rejection potential of these streams. In all tests, a pulp density in the batch flotation test of ~5% was targeted to minimise the amount of non-sulphide gangue (NSG) reporting to the flotation concentrates by entrainment. Raw water was used as make-up dilution water for these tests, and only plant frother (DSF004) was added as necessary to maintain the froth in the batch flotation cell.

Figure 3 below shows the resulting copper-mass selectivity of these dilute batch flotation tests conducted on various concentrate streams around the plant.

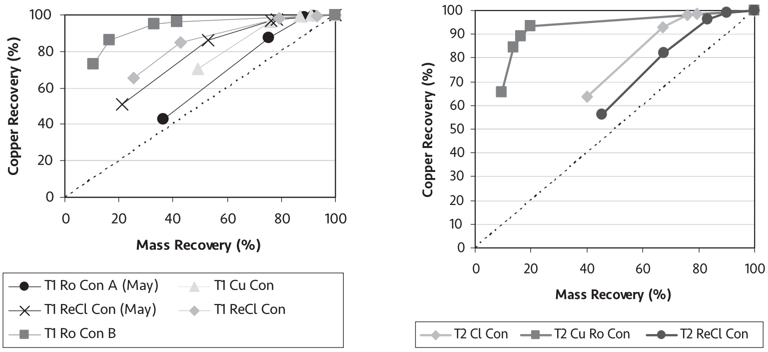


Figure 3 Copper recovery as a function of mass recovery - dilution cleaning tests

It is evident that with all streams re-floated, there is a significant upgrading achieved with an additional dilute cleaning stage, and in all cases, high copper recoveries were achieved after four concentrates with a cumulative flotation time of 9-10 minutes used in each test. For all the streams tested, it was possible to reject greater than 20% of the mass of the concentrates while recovering 90-99% of the valuable minerals (copper and gold). It can be concluded from these tests that the non-sulphide gangue present in the concentrate streams is predominantly non-floating and is being recovered by entrainment. Sulphur recovery was higher, similar to the copper result shown above, confirming that the improved selectivity was a result of non-sulphide gangue rejection.

FLOTATION PLANT RECONFIGURATION

Several alternative configurations were explored to improve the gangue rejection of the cleaning circuits at Telfer. As this is a Brownfield application, practical issues related to the installation of additional cleaning capacity and reconfiguration of the existing equipment were important in selecting the most effective improvements to the cleaner circuit performance at Telfer.

Barns, Colbert & Munro (2009) demonstrated the application of a Jameson cell at Prominent Hill, where a Jameson cell was installed as a 'cleaner scalper' unit, in order to reduce the load on the mechanical cleaning circuit and improve fine grained fluorine entrainment.

Jameson cells described by Evans, Atkinson & Jameson (1995) are highly efficient flotation machines that require a smaller footprint than conventional mechanical flotation cells and enable the efficient use of froth washing to improve gangue rejection. A schematic figure of the latest Jameson cell technology presented by Young et al. (2006) is shown below. Typically these cells achieve more than 50% recovery of valuable mineral per stage, with negligible gangue entrainment due to the froth wash water that is operated to achieve a net downward water flow in the froth.

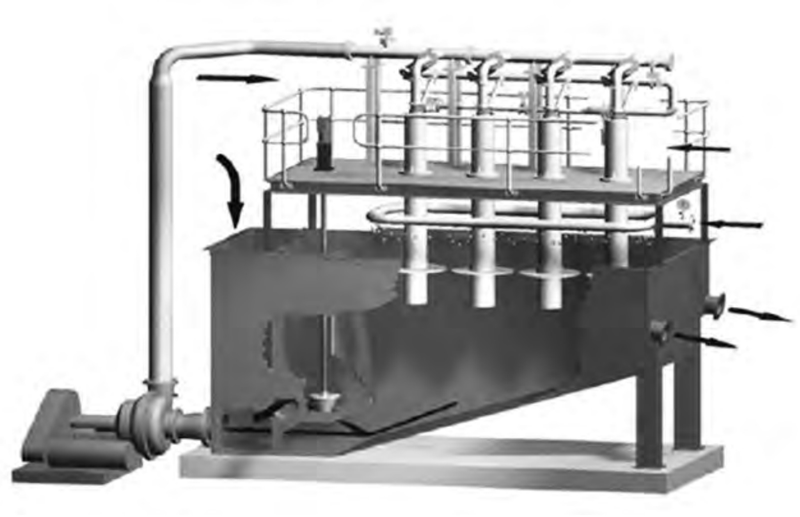


Figure 4 The Jameson cell with internal recycle to stabilise downcomer feed rate, after Young et al. (2006)

The proposed circuit reconfiguration for improving NSG rejection at Telfer is shown in **Figure 5** – a Jameson cell is to be installed as a scalper unit ahead of the cleaning circuit, targeting the removal of 50% of the valuable (i.e., copper and gold containing) minerals at high concentrate grades (through effective froth washing and high intensity bubble-particle contact). The cleaner-scavenger concentrate is also to be redirected to the re-cleaner, allowing the cleaner bank to be operated at a higher dilution (5–10% solids w/w) without compromising cleaner bank residence time. In the case of Train 1, the first two rougher cell concentrates are to be redirected through the new Jameson cells to assist in gangue rejection of this stream. In Train 2, all the copper rougher concentrate currently reports to the cleaning circuit.

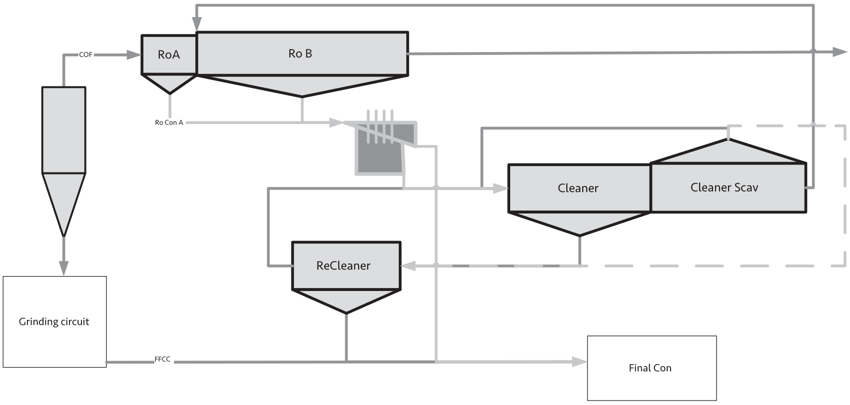


Figure 5 Jameson cell configurations with concentrate reporting to final concentrate

In order to estimate the potential improvement in the operating grade-recovery curve of the Telfer operation, a flotation model was calibrated with recent plant survey data, and used to predict the performance of the modified circuit shown above.

Model

A flotation model was developed for both trains based on the AMIRA P9 flotation modelling methodology as presented by Harris et al. (2002). In this methodology the feed is split into a number of floatability fractions that describe the observed distributed flotation rate. The recovery of a single floatability fraction across a single perfectly mixed cell is described by:

$$R = \frac{P \cdot S_b \cdot Rf \cdot \tau(1 - R_w) + ENT \cdot R_w}{1 + P \cdot S_b \cdot Rf \cdot \tau(1 - R_w) + ENT \cdot R_w} \quad \textbf{(2)}$$

where R is the recovery of fraction, P, the floatability of the fraction, S_b, the superficial gas velocity of the cell, Rf, the froth recovery of the floatability fraction in the cell, ENT, entrainment parameter (size dependent) and Rw, the water recovery across the cell.

A total of three components were used to describe the flotation of each mineral species: fast floating, slow floating and non-floating.

In order to simplify the model, the terms $P^*S_b{}^*R_f$ were lumped into a single first order rate constant k^*C where k is a first order rate constant conserved around the entire Cu circuit (for each floatability class) and C being a scale-up number assigned to each physical cell or flotation bank. This C parameter is actually a measure of how hard the cell is pulled by changing froth depth or aeration rate. The model adequately describes the entrainment mechanism, critical to evaluating cleaning circuits through the use of the entrainment classification number, ENT, which was estimated based on the size distribution of the tailing streams from each bank.

The feed stream to each bank was split into three floatability fractions: fast, slow and non-floating fractions for each of the elements assayed for (Au, Cu, Fe and S), and model parameters were fitted to the plant data.

Simulation results

Monte Carlo simulations were conducted of the as-is flotation circuit models and with the modifications shown in **Figure 5**. The Monte Carlo simulations were conducted by sampling the flotation cell scale-up parameters, randomly, from a normally distributed scale-up number with standard deviation of 10% and 20% for the rougher and cleaner/re-cleaner cells respectively, and simulating the resulting parameter set. This was repeated 100 times for each scenario that was modelled. This manipulation of the cell scale-up parameter is akin to operators changing the level and the air flow rate set-points randomly in the plant.

The impact of redirecting the cleaner scavenger concentrate from the cleaner feed to the recleaner feed and diluting the cleaner bank from 10% solids to 7% solids, without installation of the Jameson cell, is shown for Train 2 in the figure below (the copper recovery is shown as a change from current recovery).

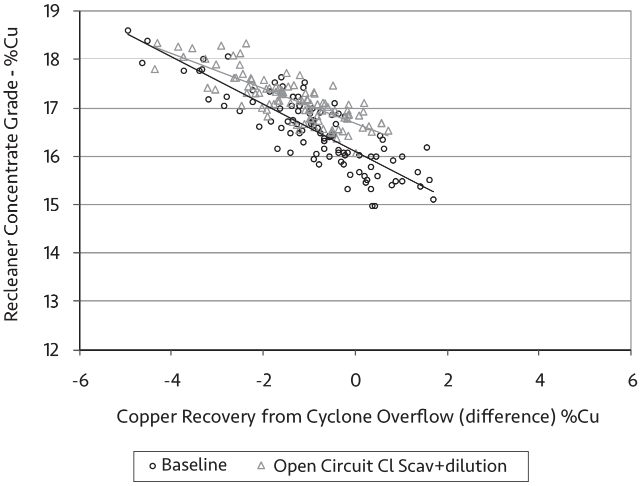


Figure 6 Modelled effect of open-circuiting cleaner scavenger concentrate and dilution of the cleaner bank from 10% solids to 7% solids on Train 2 at Telfer

This modification shows some improvement in gangue rejection, however further benefit is possible combining this change with the additional installation of Jameson cells as shown in **Figure 5.**

The Jameson cells were modelled using the selectivity obtained in the dilute cleaning tests on the rougher concentrate streams, as shown in **Figure 3**, and assuming that at least 50% copper and gold recovery can be achieved in the Jameson cell. This recovery value was based on similar installations of Jameson cells in similar duties, such as Prominent Hill (Barns, Colbert & Munro, 2009) and Phu Kham (Young & Crnkovic, 2011). The results of the Monte Carlo simulation results are shown in **Figure 7**.

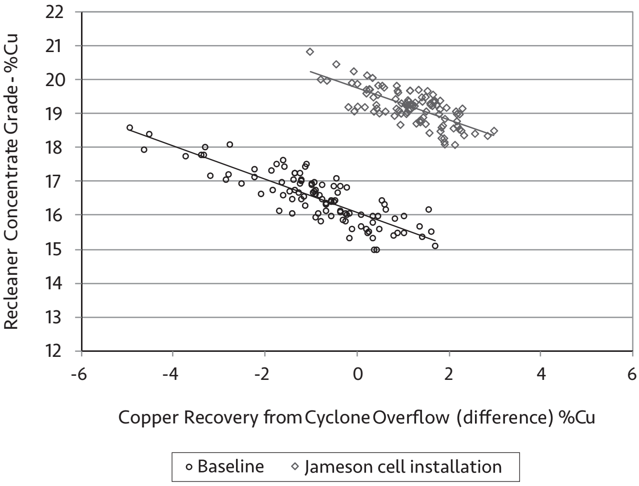


Figure 7 Model results for Train 2 Jameson cell installation

It can be seen in **Figure 7** that the resulting improvement in copper grade and recovery is consistent with the mineralogical predictions that the concentrate grade could be raised to 20-25% Cu, by rejecting half of the liberated gangue species that are currently reporting to the final concentrate. Although not shown, gold grade and recovery showed improvements similar to those of copper.

Implementation

On the basis of the mineralogical circuit analysis, dilute cleaning tests and modelling outputs, it was decided that piloting of Jameson cells was unnecessary and capital expenditure for the installation of one Jameson cell per train has been approved. The Jameson cells have been sized to treat a maximum feed rate of 60 tph (solids) at a solids content of 15% w/w and a feed grade of 3-7% Cu. The recycle of the cell has been designed at 50% and the Jameson Cell was designed with an eight downcomer configuration – E3432/8.

The cleaner scavenger concentrate will be re-routed to allow for the stream to be directed to cleaner feed or recleaner feed, and additional water addition and density control will be installed to allow for the optimisation of the cleaning circuit post Jameson cell installation.

Installation and commissioning of these cells are expected to be completed by the end of December 2011.

CONCLUSIONS

It has been shown that there is a significant portion of non-sulphide gangue reporting to the final concentrate from both trains (~40%) at Telfer. In both trains, this non-sulphide gangue has been shown to be well liberated and contained in relative fine fractions suggesting its deportment to final concentrate by the mechanism of entrainment. This was confirmed by completing dilute batch flotation tests on various concentrate streams where 80% mass rejection was achieved with minimal loss of valuable (copper and gold) species.

Flotation modelling was completed on both trains to simulate the benefit of operating the cleaning circuit at a lower density – by diverting the cleaner scavenger concentrate to the recleaner to avoid loss of residence time in the cleaning circuit – and with the installation of a Jameson cell at the front of each cleaning circuit. The estimated benefit of this modification to the circuit is consistent with theoretical predictions that are based on rejection of half of the liberated gangue from the final concentrate producing a copper concentrate grade of 20 to 25% Cu.

The use of mineralogical circuit data, together with the application of a flotation model, allowed for the redesign of the cleaning circuit as well as demonstrating the need of capital expenditure required to implement the modification at Telfer.

REFERENCES

Barns, K.E., Colbert, P.J. & Munro, P.D. (2009) 'Designing the optimal flotation circuit – The Prominent Hill case', *AUSIMM 10th Mill Operators Conference*, *Adelaide, 12-14 October 2009*, pp. 173–182.

Benson, M.A., Headley, C.R., Hille, S.O. & Jacob, N.P. (2007) 'Differential flotation commissioning at Telfer', *Proceedings Ninth Mill Operators' Conference*, The Australasian Institute of Mining and Metallurgy, Melbourne, pp 17–24.

Evans, G.M., Atkinson, B.W. & Jameson, G.J. (1995) 'The Jameson Cell', *Flotation Science and Technology*, K.A. Matis (ed), Marcel Dekker Inc., New York, pp. 331–363.

Goulsbra, A., Dune, R., Lane, G., Dreisinger, D. & Hart, S. (2003) 'Telfer project process plant design', *Proceedings Eighth Mill Operators' Conference*, The Australasian Institute of Mining and Metallurgy, Melbourne, pp. 103–114.

Harris, M.C., Runge, K.C., Whiten, W.J. & Morrison, R.D. (2002) 'JKSimFloat as a practical too for flotation process design and optimization', *Proceedings Mineral Processing Plant Design, Practice and Control Conference*, Society for Mining, Metallurgy, and Exploration, Littleton, pp. 461–478.

Johnson, N.W., McKee, D.J. & Lynch, A.J. (1974) 'Flotation rates of non sulfide minerals in chalcopyrite flotation processes', *Transactions of Society of Mining Engineers*, AIME, vol. 256, pp. 204–209.

Ma, W. & Johnston, H. (2011) *Mineralogical Assessment of Telfer Process Streams – KM2883 October 2010*, Newcrest confidential report, G&T Metallurgical Services Ltd.

Savassi, O.N., Alexander, D.J., Franzidis, J-P. & Manlapig, E.V. (1998) 'An empirical model for entrainment in industrial flotation plants', *Minerals Engineering*, vol. 11, no. 3, pp. 243–256.

Young, M., Barns, K., Anderson, G. & Pease, J. (2006) 'Jameson Cell: The 'comeback' in base metals applications using improved design and flowsheet', *38th Annual Meeting of the Canadian Mineral Processors*.

Young, M.F. & Crnkovic, I. (2011) 'Development of an Innovative Copper Flowsheet at Phu Kham', *AUSTMINE 2011, Brisbane 17-18 May*.

Zheng, X., Crawford, A. & Manton, P. (2009) 'Telfer Train 1 Copper Flotation Circuit Reconfiguration', *AUSIMM 10th Mill Operators Conference, Adelaide, 12-14 October 2009*, pp. 233–243.

Separation efficiency improvement of a low grade copper-gold flotation circuit

Newcrest Mining Limited

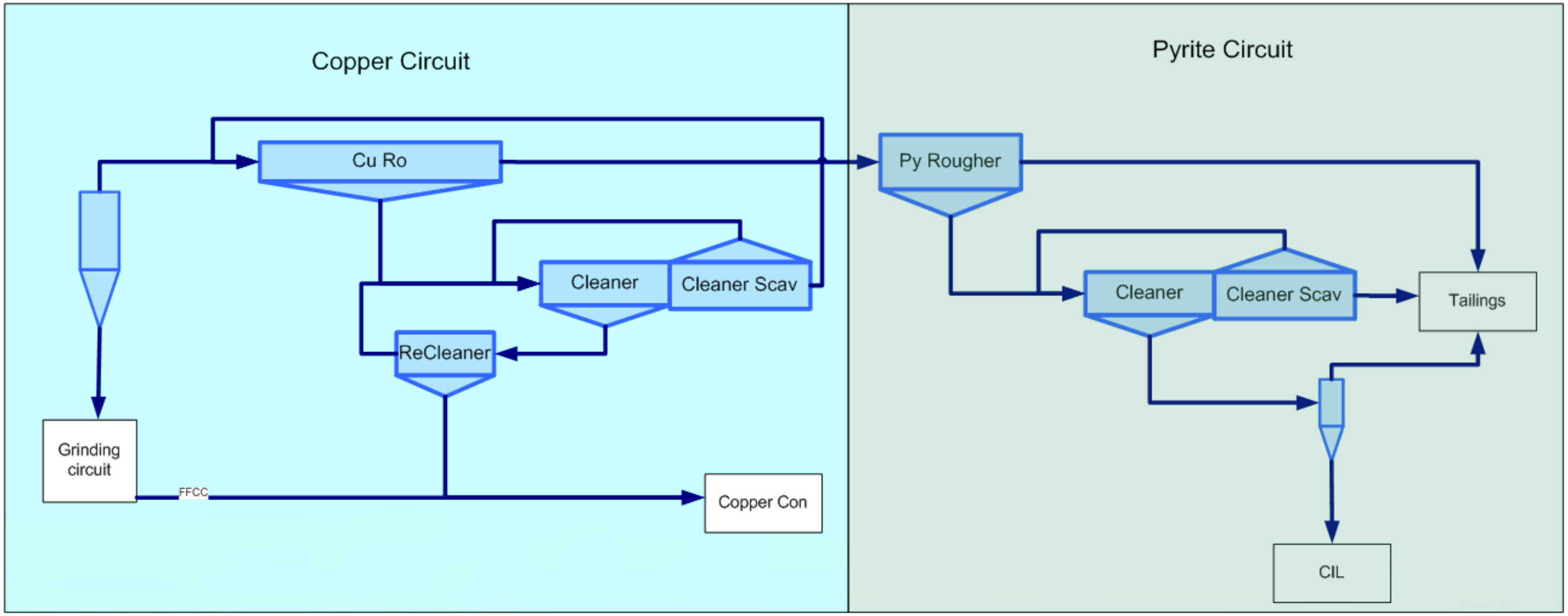
David Seaman, Peter Manton & Paul Griffin

Contents

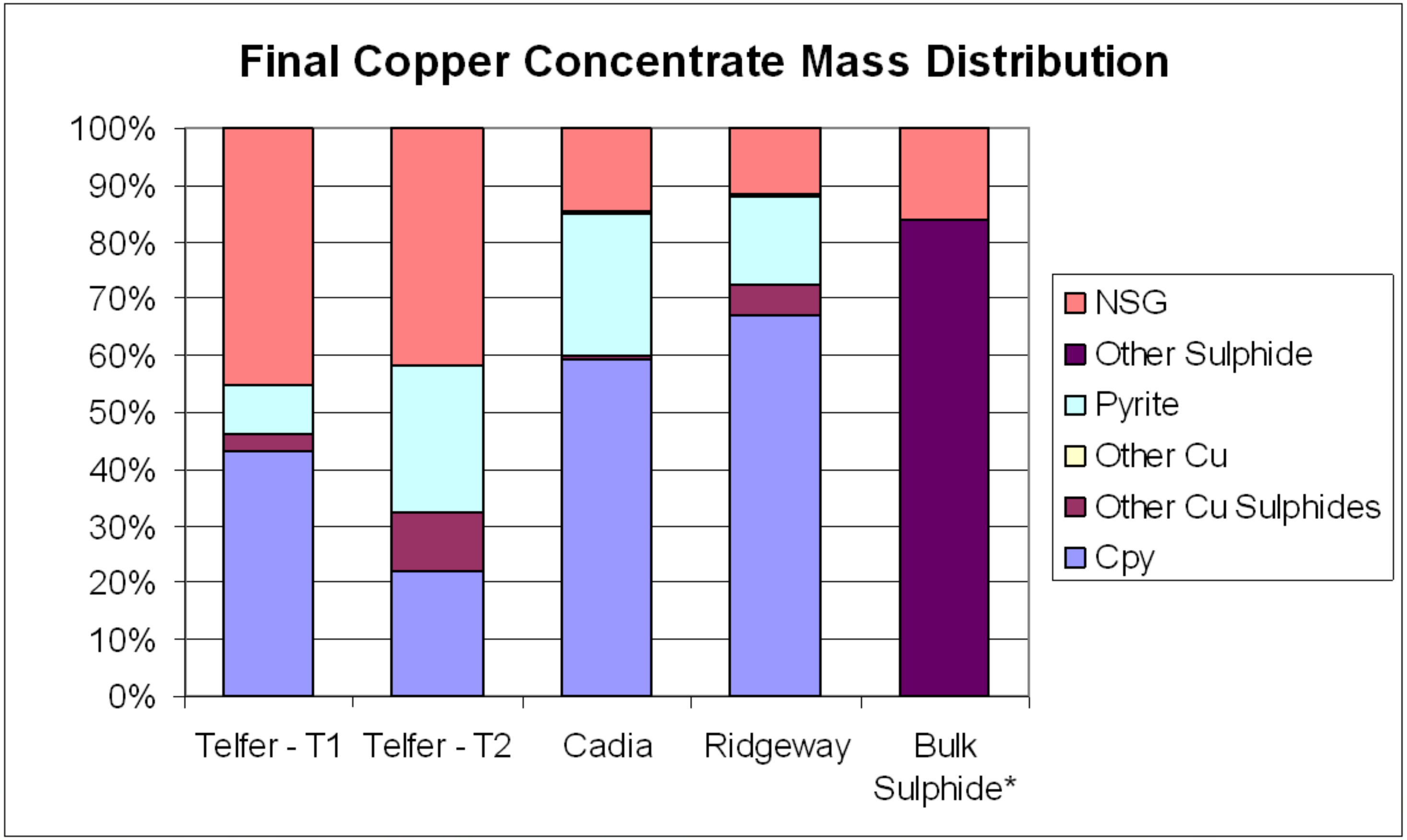
- Process Overview
- Mineralogy
- Circuit Modification
- Installation Details

Telfer Flotation Overview

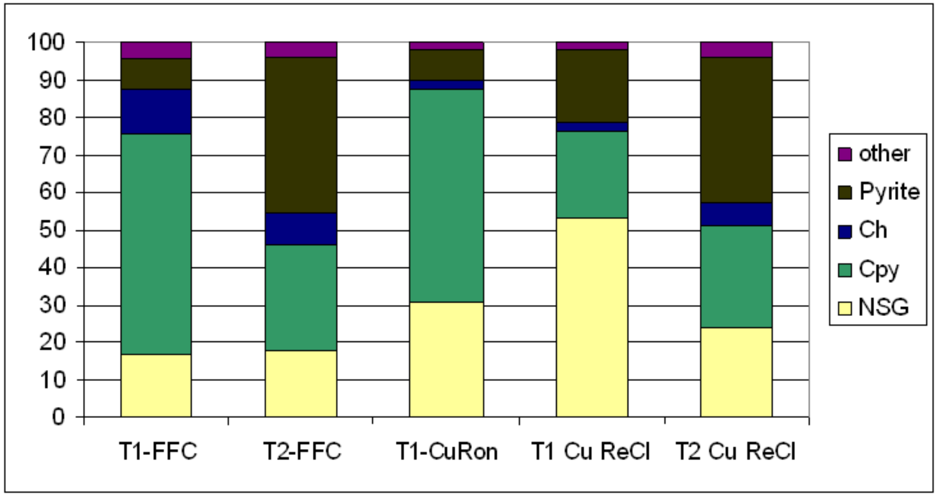
- Dual train (underground and open pit) ~ 23Mtpa
- Typical head grades – 0.1-0.2 % Cu; 1-2g/t Au; 1-5% Sulphur
- Gold distribution – free gold/copper minerals/Pyrite (~5g/t)
- Copper minerals – Predominantly Chalcopyrite and Chalcocite
- Copper concentrate produced - ~16% Copper

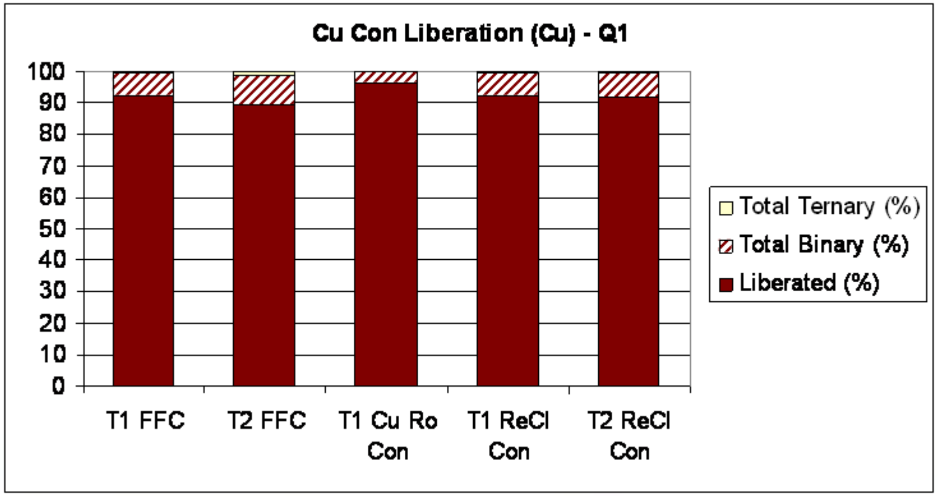


Copper Concentrate Mineralogy



Copper Concentrate Mineralogy

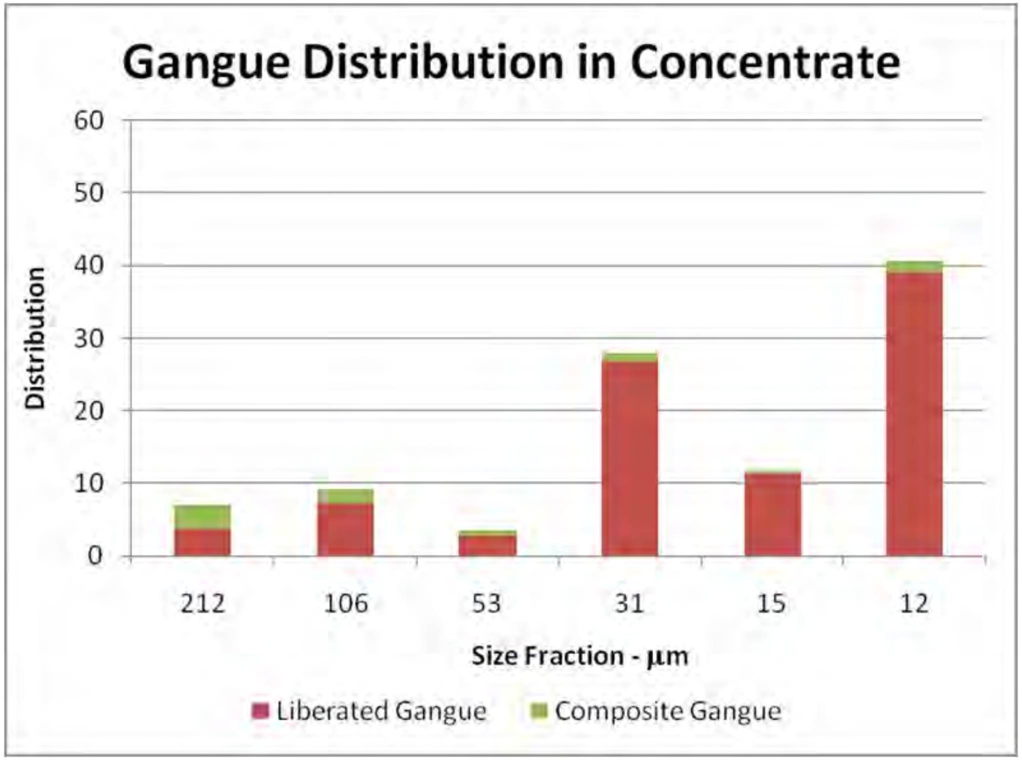



Cu Con Liberation (Cu) - Q1

- Significant quantity of liberated NSG diluting copper concentrate
- Copper minerals well liberated
- Opportunity to replace NSG with Pyrite
  - –improve gold recovery
  - –Increase copper concentrate grade

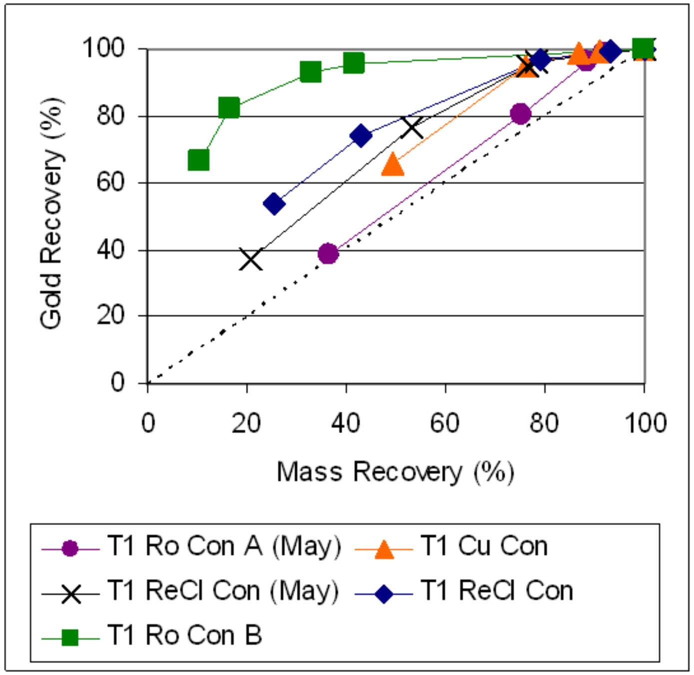
Gangue Characteristics – Size

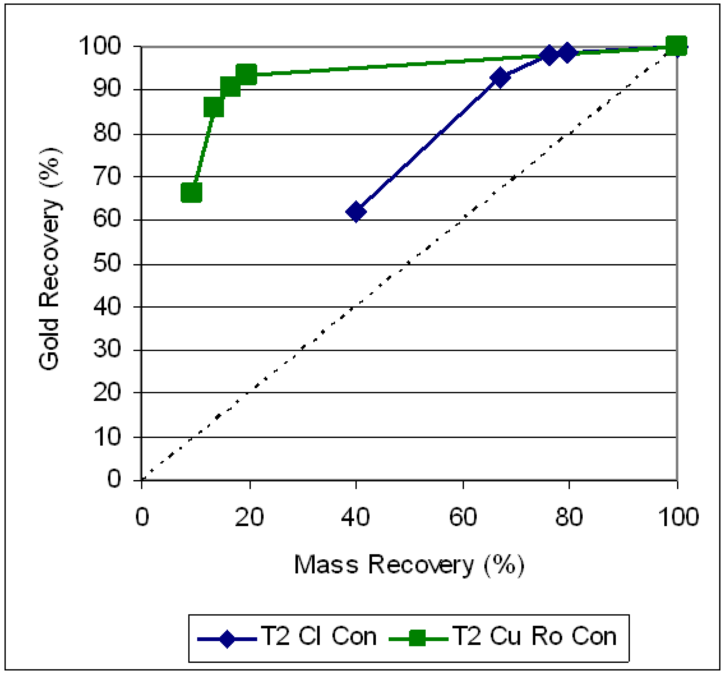
- Predominantly fine and well liberated
- Mostly Quartz & Feldspars



Mechanism of Gangue Dilution

- Dilute “Hot” Flotation tests - ~5% solids





- 20% of Mass (NSG) can be rejected by further cleaning with negligible change to overall recovery
- NSG is not being recovered by true flotation

Cleaning Circuit limitations

- Cleaner cells heavily loaded
- Cells have insufficient launder capacity
- Residence time constraints do not allow for further dilution in cleaners to reduce entrainment

	%solids	τ (min)
Cu Ro	34.3	15.0
Cu Cl	10.5	13.6
Cu Cl Scav	9.2	14.4
Cu ReCl	10.1	20.7

Cu Recleaner Cells

Reconfiguration of the Cleaning Circuit

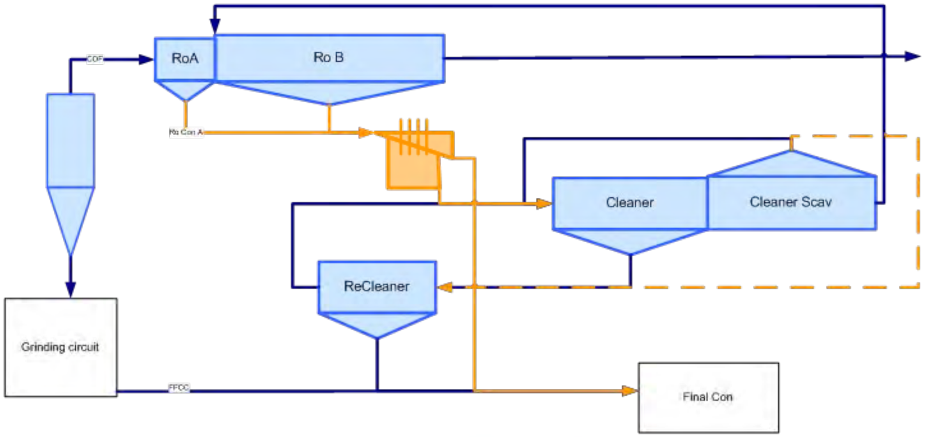
- Simplified AMIRA P9 model:

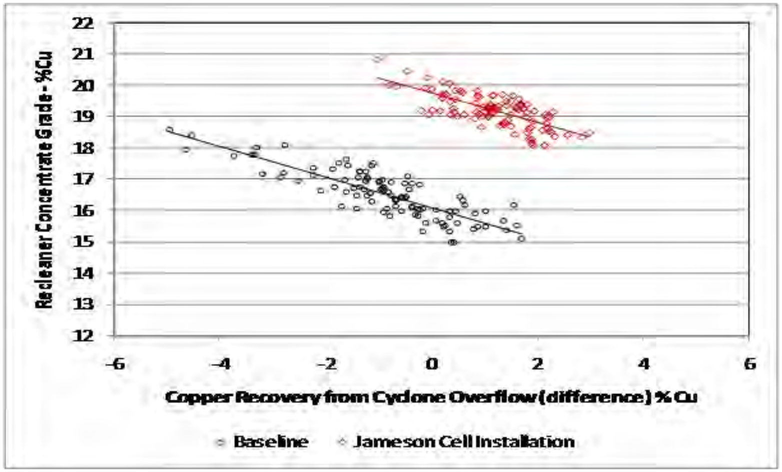
$$R_{ij} = \frac{k_{ij}\,\tau(1-R_w)+ENT\,R_w}{(1+k_{ij}\,\tau)(1-R_w)+ENT\,R_w}$$

- Various reconfigurations simulated using Monte-Carlo
- Minor improvements could be made using existing equipment:
  - open-circuiting the Cl Scav Con (to Recleaners)
  - diluting the Cl/Cl Scav bank to reduce Rw
- Additional capacity required to make a significant improvement
- Jameson cell chosen as new capacity
  - published performance data (>50% stage recovery at high grades)
  - ability to reject entrainment by froth washing
  - Incorporated in model using the dilute batch float selectivity

Modified Cleaner Flowsheet

- Installation of one Jameson cell per Train fed by Cu Ro Con
  - Similar approach to recent installations at Phu Kam and Prominent Hill
- Open circuit Cl Scav Con to allow additional dilution of Cleaners
- Modelled improvement consistent with mineralogical characterisation

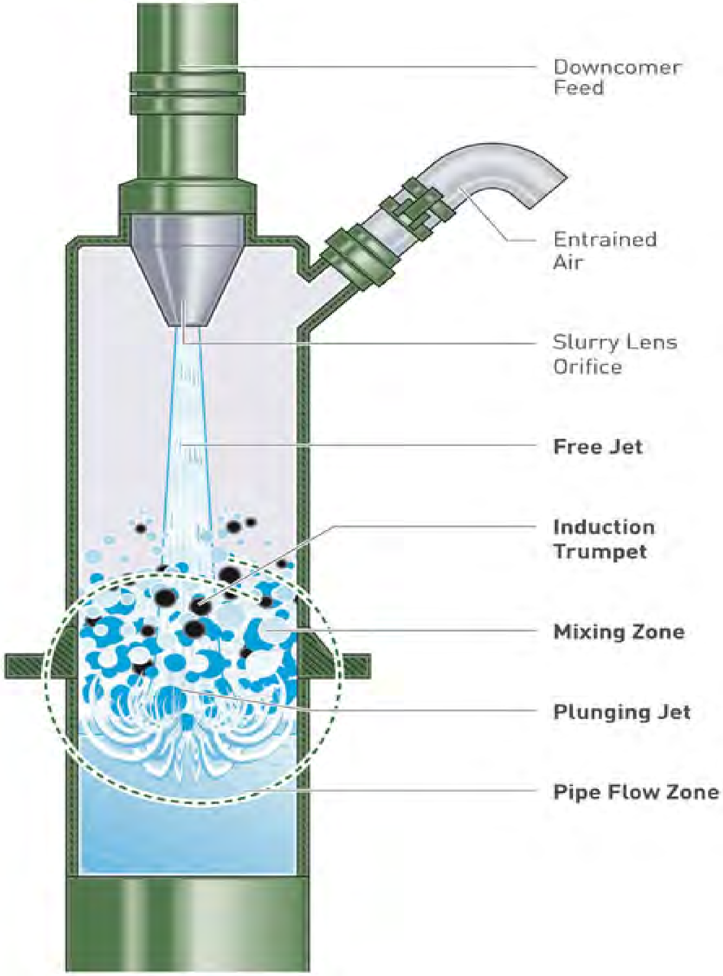




Jameson Cell Technology

The Downcomer

- Slurry is pumped through a slurry lens orifice at high pressure (150kPa)
- Air is entrained into the downcomer
- Mixing zone is formed in the downcomer where particles and bubbles contact
- Internal recycle of cell content allowing the use of a fixed speed pump, designed for approximately 2:1 (downcomer flowrate: fresh feed)
- Cells are fitted with wash-water trays to allow effective froth washing to reduce entrainment

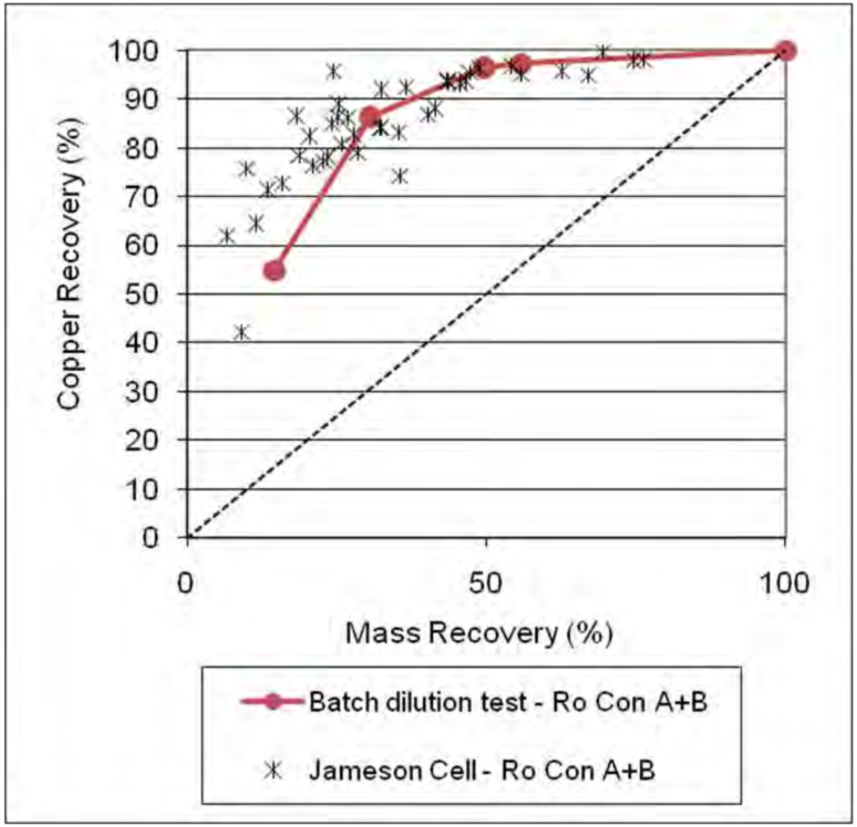


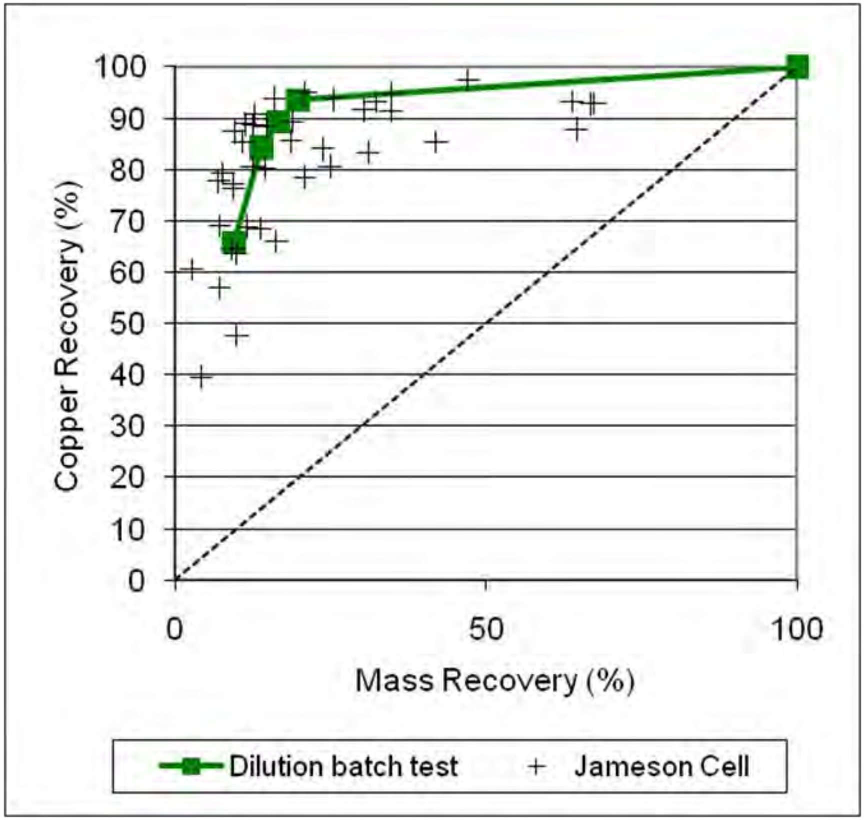
Telfer Jameson Cell Installation

Installation Details

- 2 x E3432/8 Jameson Cells
- Designed to treat up to 60tph fresh feed each (350m^3/h slurry)
- Concentrate and tailings flow by gravity
- Each cell driven by a 75kW Warman 10/8 pump (700m^3/h)
- Wash water utilisation ~ 30-100m^3/h
- Major equipment direct purchased by Newcrest, installation carried out under a lump sump EPC - reduced project delivery time
- Timing
  - Oct 2010 – Project initiated to address NSG in concentrate
  - May 2011 – Order placed for major equipment
  - November 2011 – New cleaning circuit commissioned

Jameson Cell Actual vs Predicted

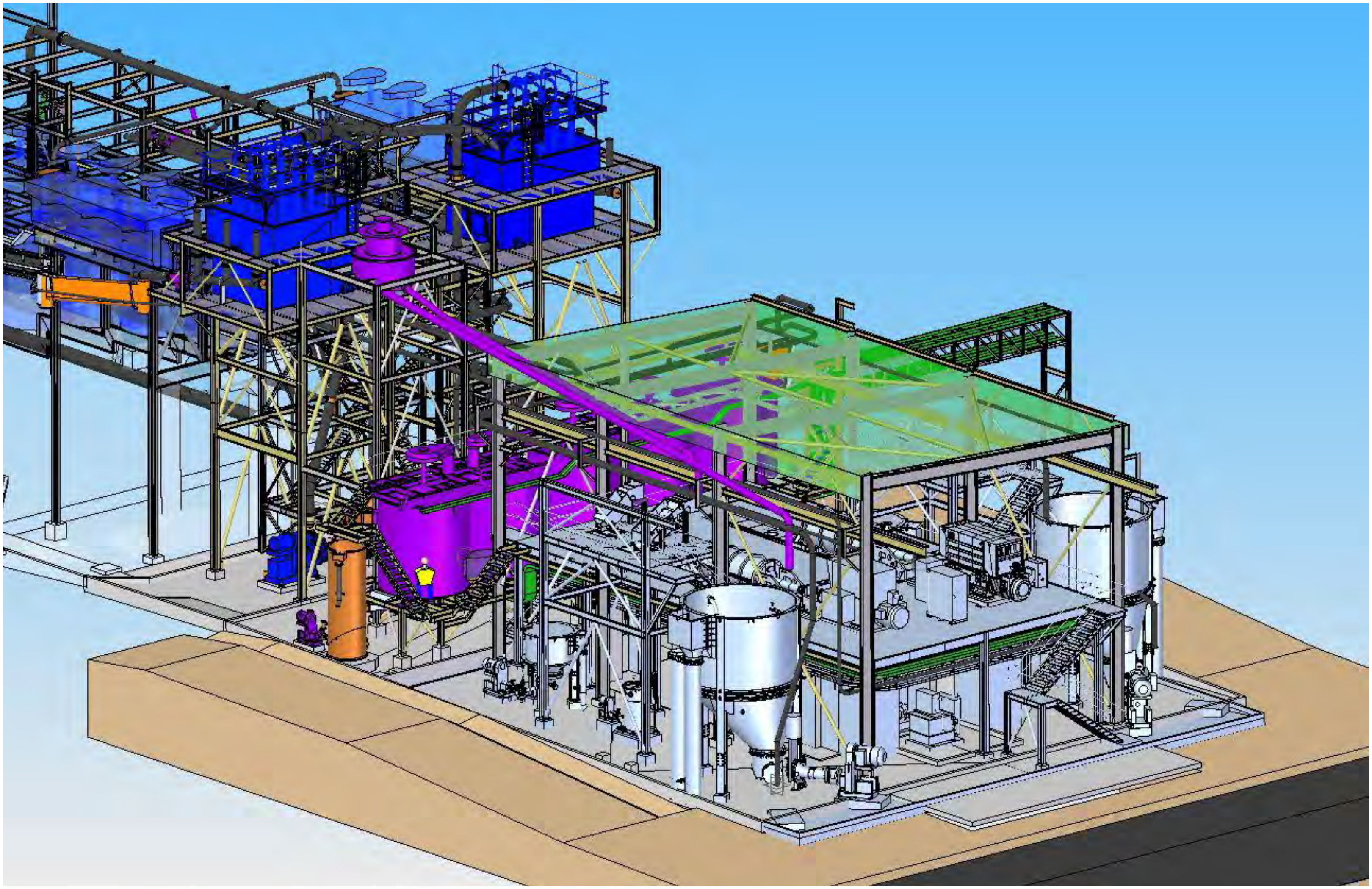




Similar selectivity observed to date, high stage recoveries achieved (>80%) for Cu and Au

Further Circuit Modifications in 2012

- Copper regrind mill for improved liberation (Cu/Pyrite) of a complex ore body to be processed
  - ISAMill M3000 – 1.2MW
- Pyrite regrind to liberate gold and copper from Pyrite and recover by flotation prior to leaching
  - ISAMill M5000 – 1.5 MW
  - 5xOutotec TC 30s



Development of an Innovative Copper Flowsheet at Phu Kham

M F Young and I Crnkovic

Phu Kham Location in Laos

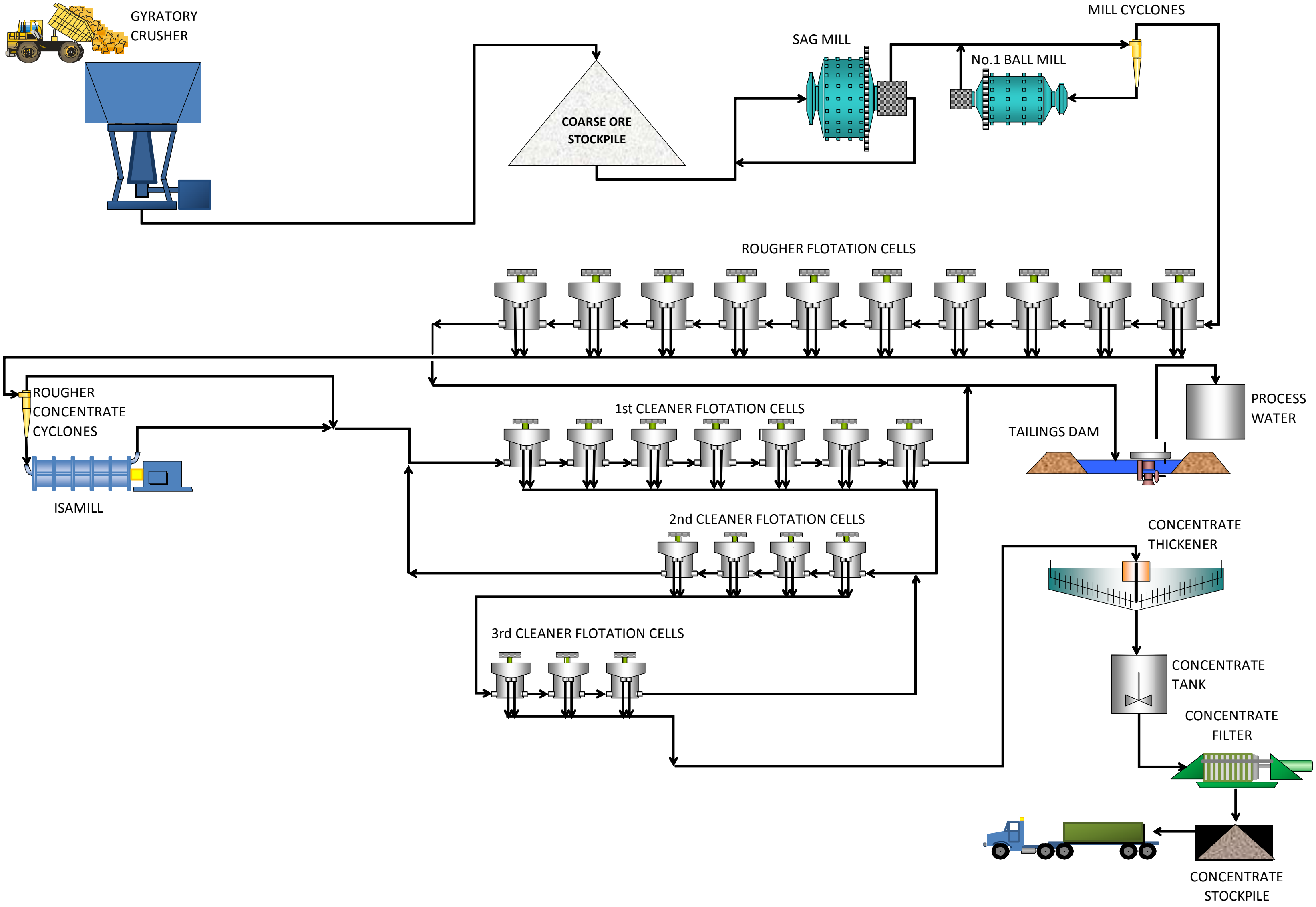
Phu Kham Plant Layout

Phu Kham process plant at night

Plant Layout showing

- Primary Grinding
- Rougher Flotation
- IsaMill Regrinding

Phu Kham Flowsheet



Phu Kham Flowsheet

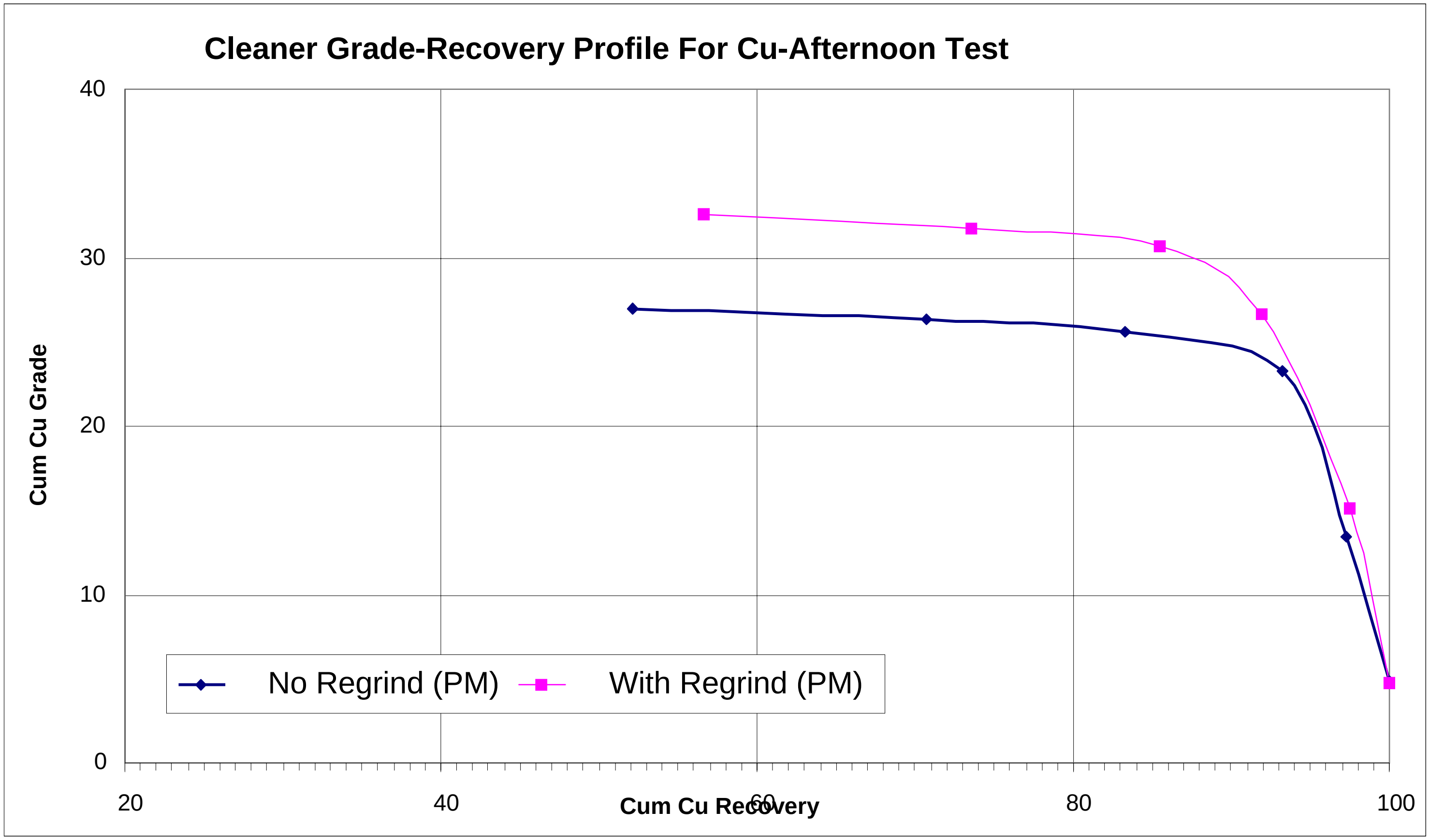
- 12.8 Mt/a of copper-gold bearing ore from the open pit
- Plant Feed Grades are 0.75% Cu and 0.33g/t of Au and 3.8g/t of Ag
- Concentrate quality is +25% Cu, 7 g/t of Au and 60g/t of Ag

- Primary Grinding Circuit contains
  - 34 ft × 18 ft (13MW) SAG mill
  - 24 ft × 39 ft (13 MW) ball mill
- Regrinding Circuit contains
  - M10,000 (2.6MW) IsaMill

- Ore has fine locking of copper and gangue minerals
- Ore requires regrinding rougher concentrate to 38 microns to make good quality final copper concentrate

Cleaner Feed Performance

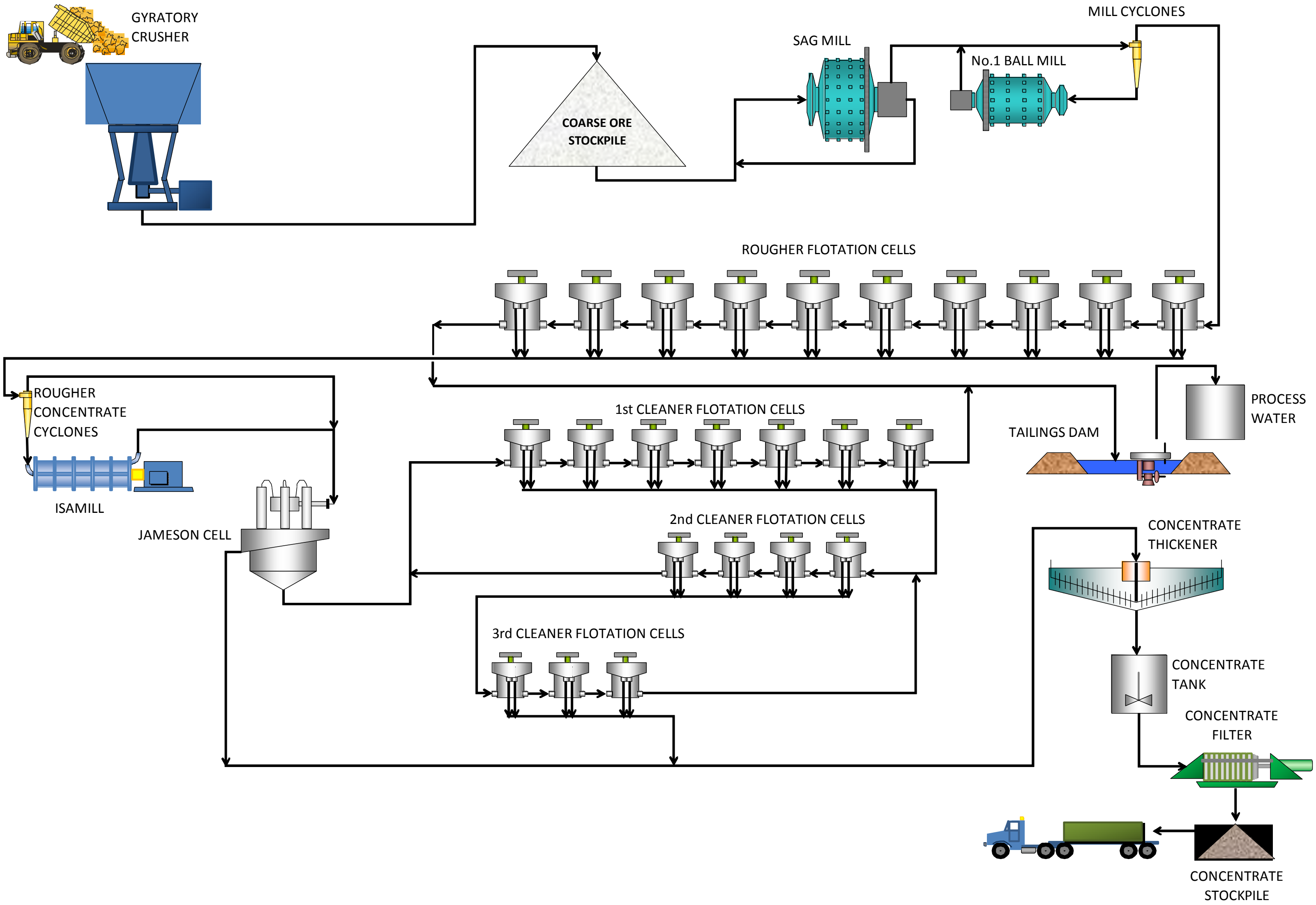
- IsaMill currently grinds from 90 microns to 38 microns
- Cleaner feed performance at 38 microns and 25 microns
- Laboratory Flotation Tests



New Phu Kham Flowsheet

- Cleaning Circuit was overloaded at times
- Fine grinding to improve concentrate quality, slowed flotation rate and increase frothing issues in the cleaner circuit, decreasing the performance of the cleaner circuit
- Jameson Cell with froth washing was installed at the head of the cleaner circuit to increase cleaning capacity
- This new circuit allowed final concentrate to be produced from the IsaMill regrind product in one flotation cell and remove more than half the load from the existing cleaning circuit

New Phu Kham Flowsheet

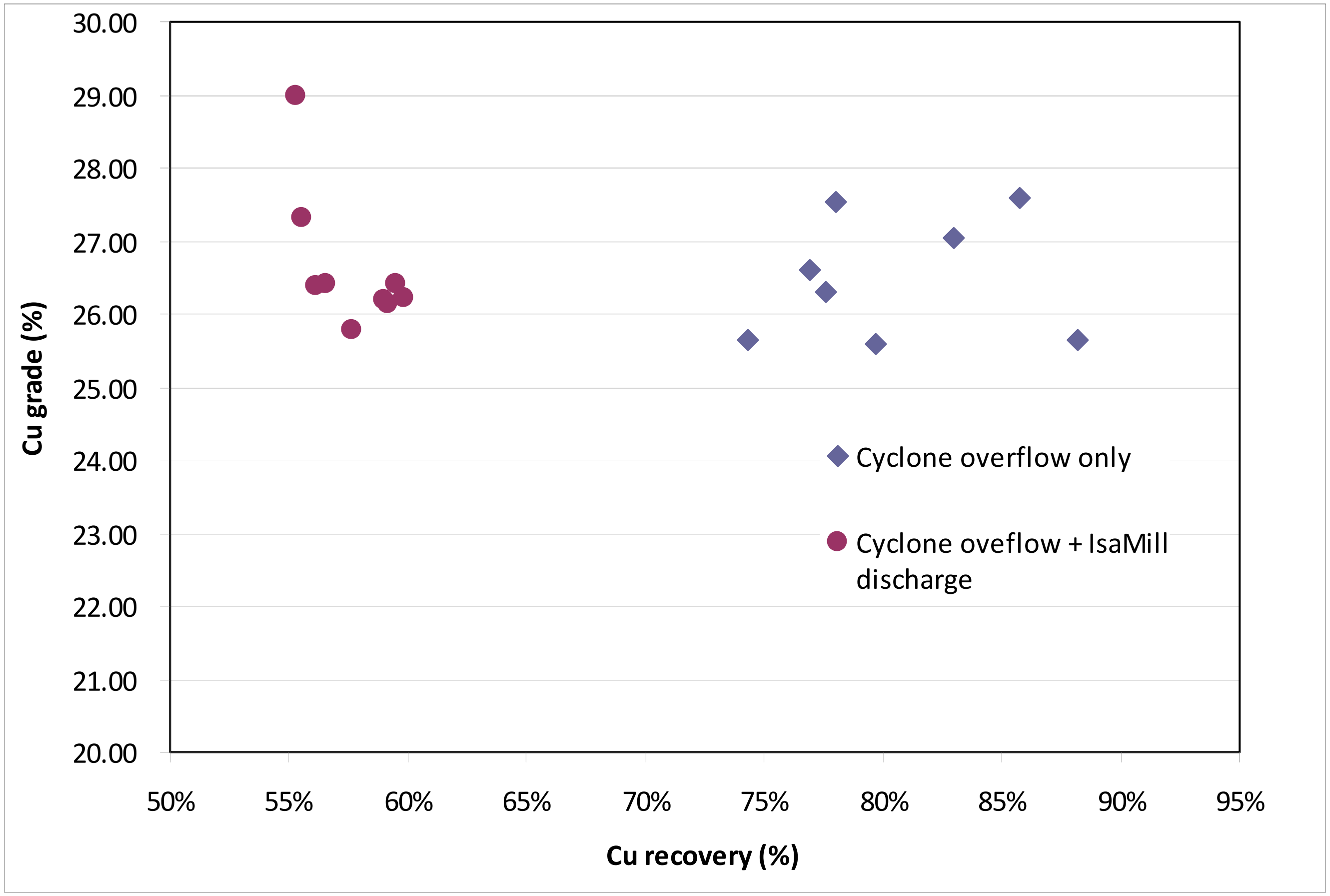


Layout of IsaMill and Jameson Cell

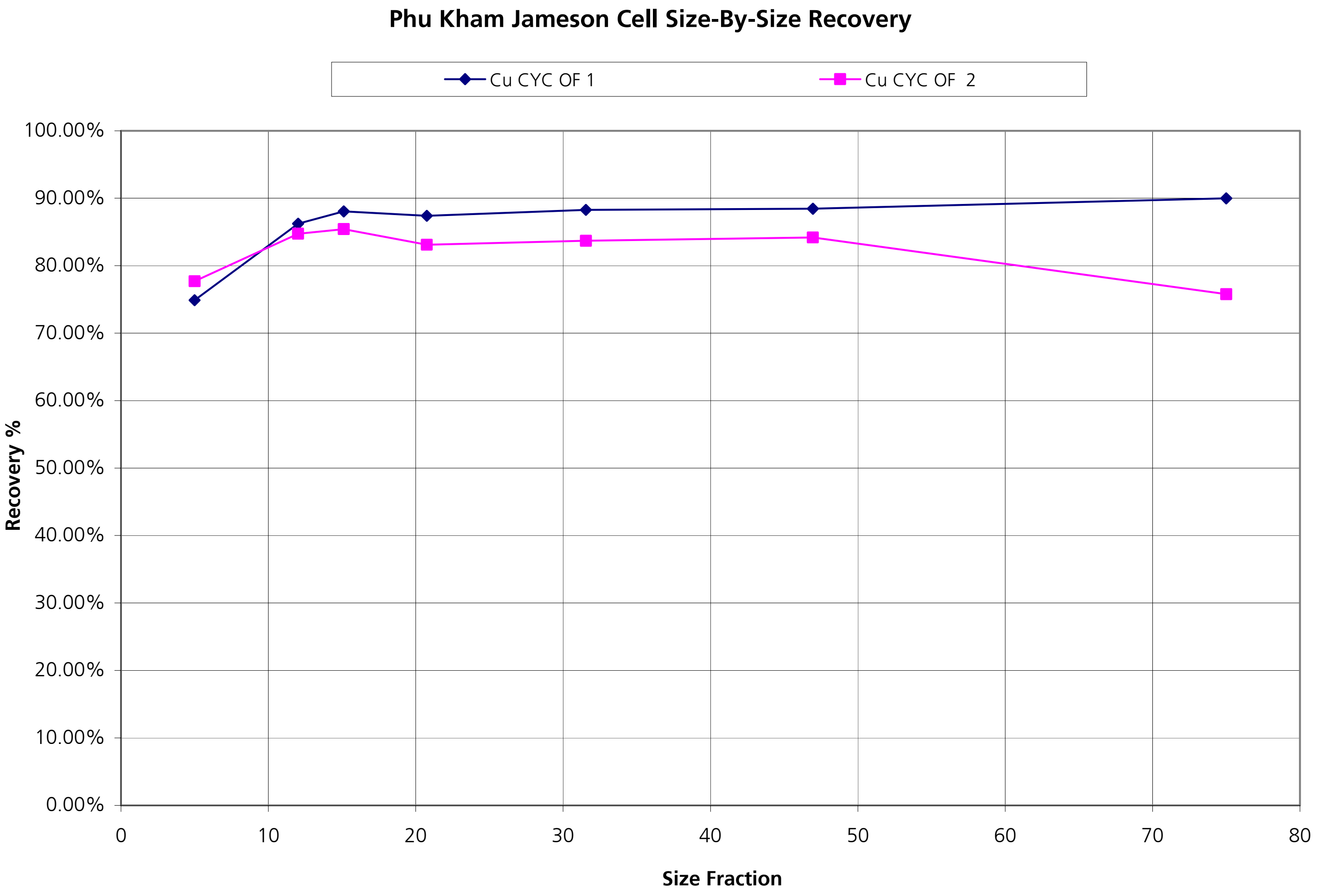
Jameson Cell Flotation at Phu Kham

Phu Kham Jameson Cell
Grade-Recovery Curve Performance

- IsaMill Cyclone Overflow P80=20 microns
- IsaMill Discharge P80=38 microns
- Feed (Rougher Conc) Grade = 3-5% Cu



Jameson Cell Size by Size Recovery on IsaMill Cyclone Overflow Stream

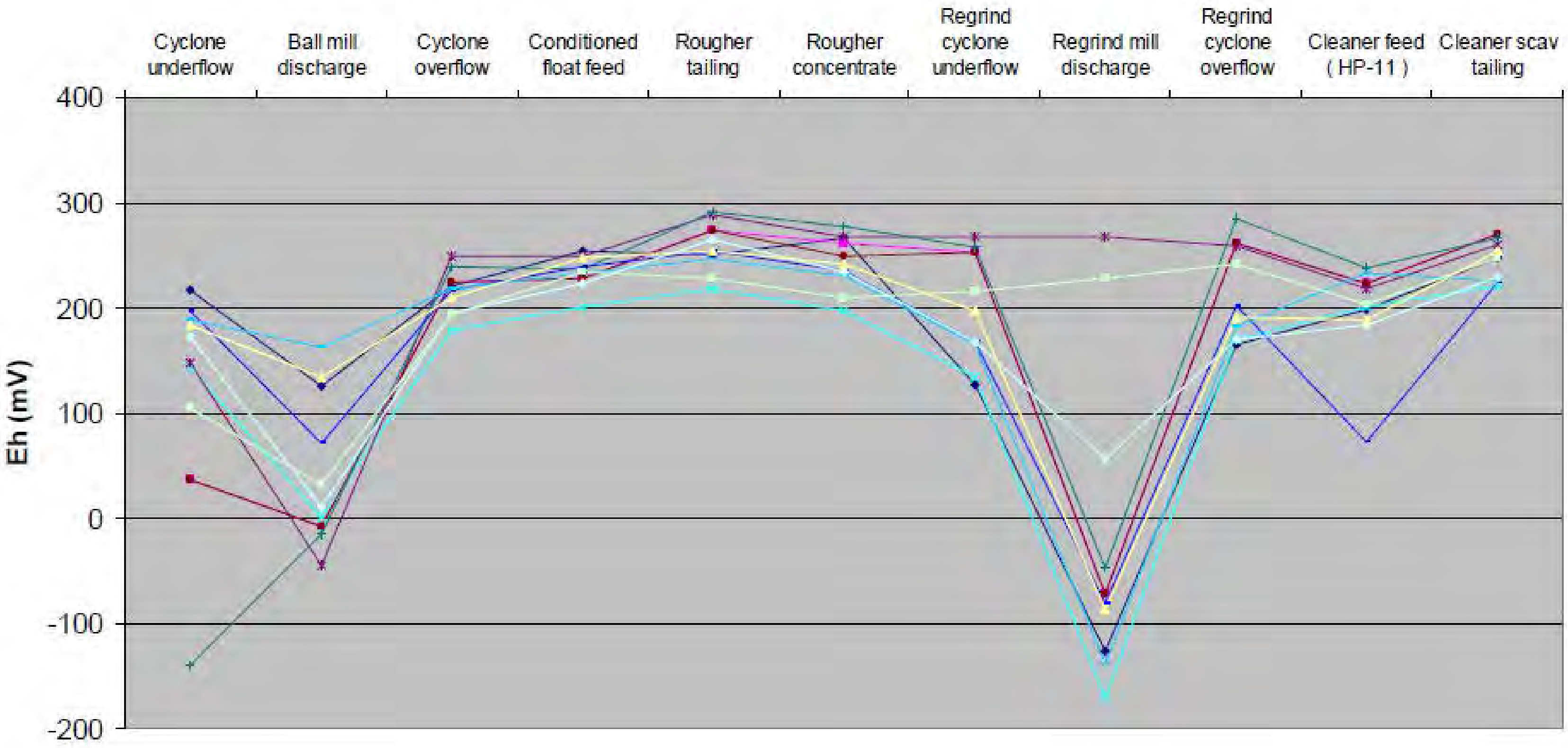


New Cleaning Circuit Results

- Jameson Cell was initially commissioned on the regrind cyclone overflow
- Cyclone Overflow was fine (P80=20 microns) and well liberated
- Jameson Cell produced 25.5% to 27.5% Cu grade at 75% to 90% Cu recovery on this stream
- Jameson Cell produced high recovery in all size fractions for this stream

- Jameson Cell then treated cyclone overflow plus IsaMill discharge
- Jameson Cell still produced high grade copper concentrate at 26% to 29% Cu grade at 55% to 60% copper recovery
- The recovery of the coarser size fraction was lower due to the locking and lower liberation in the size fractions

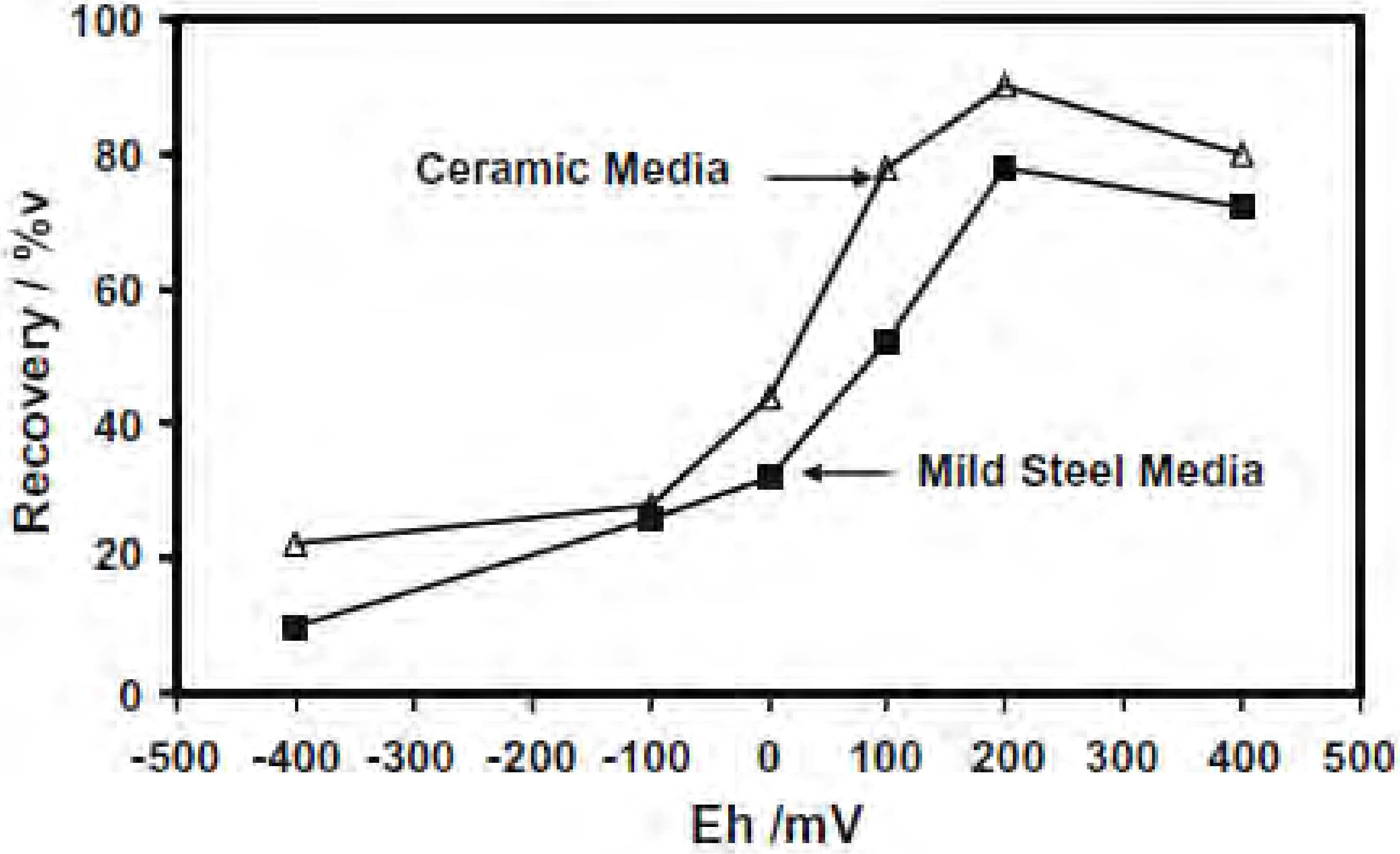
Redox Measurements in Plant



Low Eh reduces the copper mineral flotation performance

Impact of Inert Grinding - Chalcopyrite

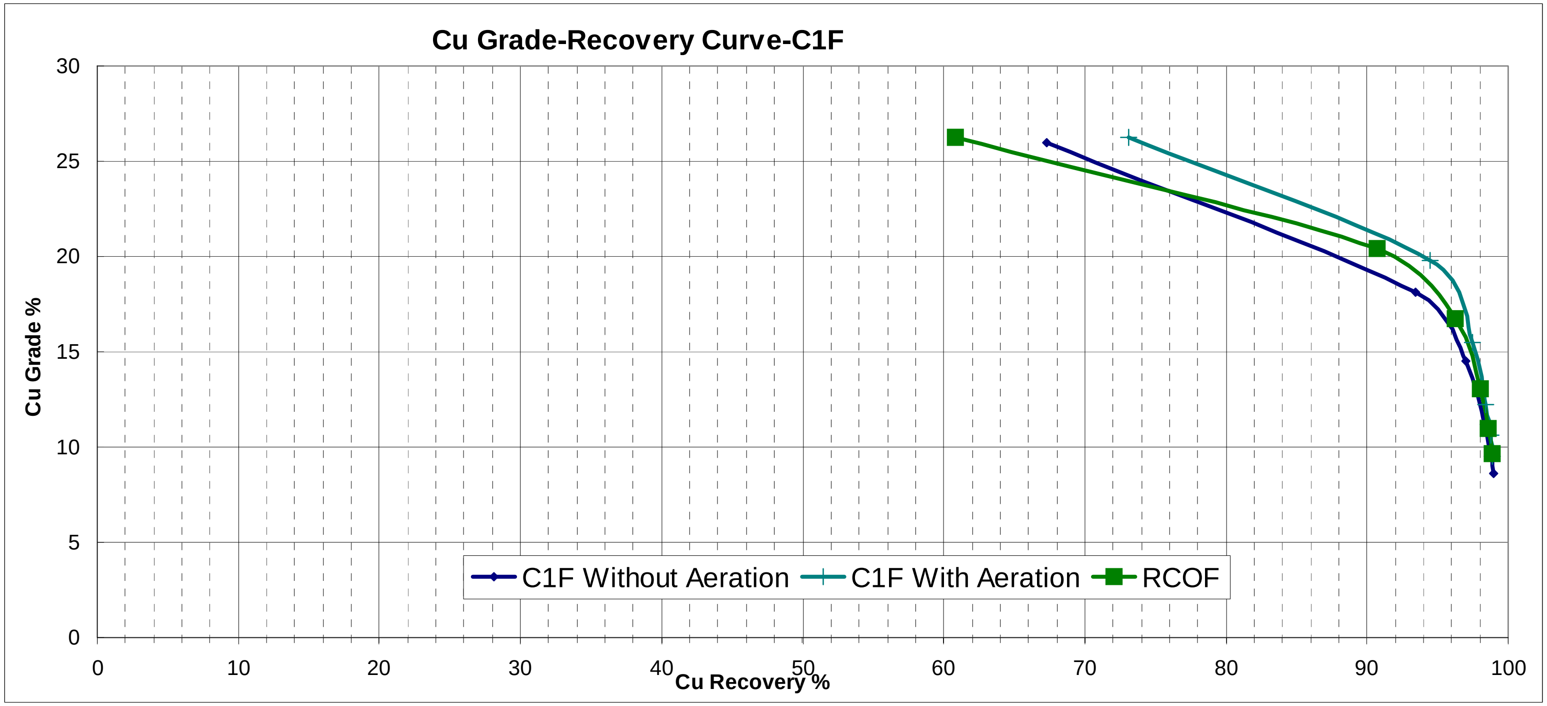
The Effect of Eh – Mt Isa Cu Ore



The critical importance of the grinding environment on fine particle recovery in flotation - Stephen Grano - 2009

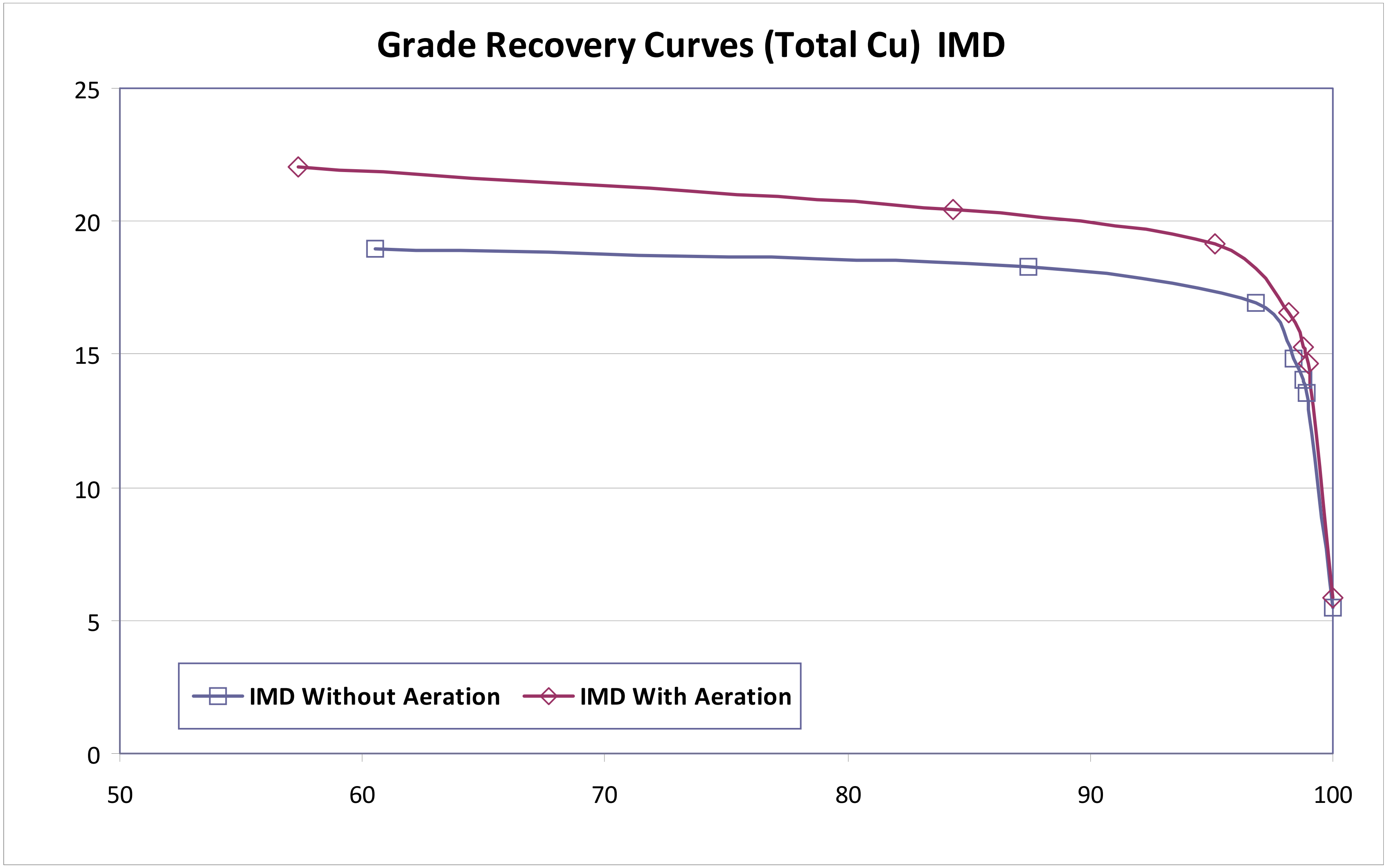
Cleaner Feed Performance

The Effect of Aeration in Laboratory Float



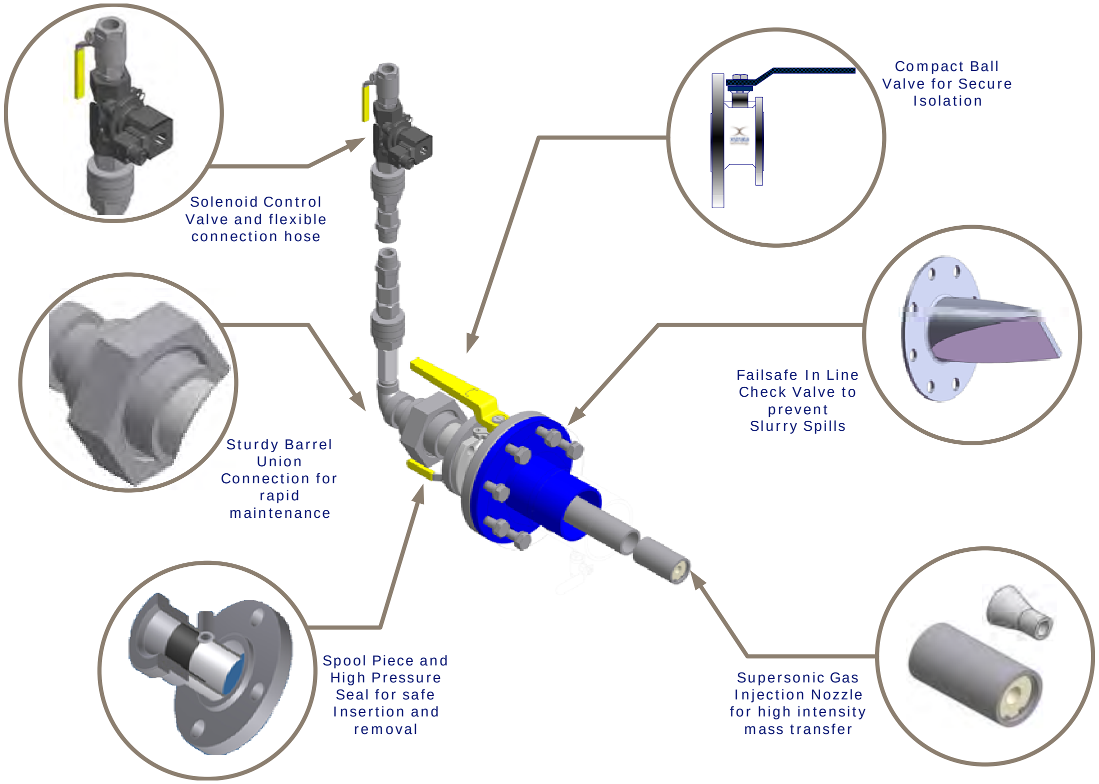
IsaMill Discharge Performance

The Effect of Aeration in Laboratory Float



HYPERSPARGE

Supersonic Gas Injection



Solenoid Control Valve and flexible connection hose

Compact Ball Valve for Secure Isolation

Sturdy Barrel Union Connection for rapid maintenance

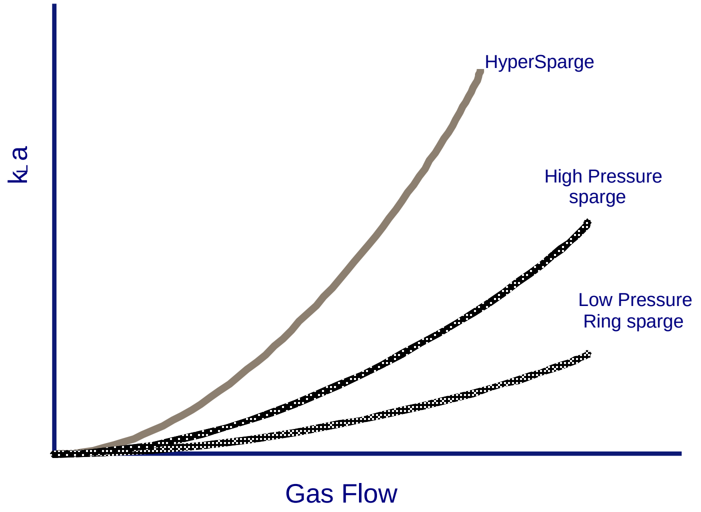
Failsafe In Line Check Valve to prevent Slurry Spills

Spool Piece and High Pressure Seal for safe Insertion and removal

Supersonic Gas Injection Nozzle for high intensity mass transfer

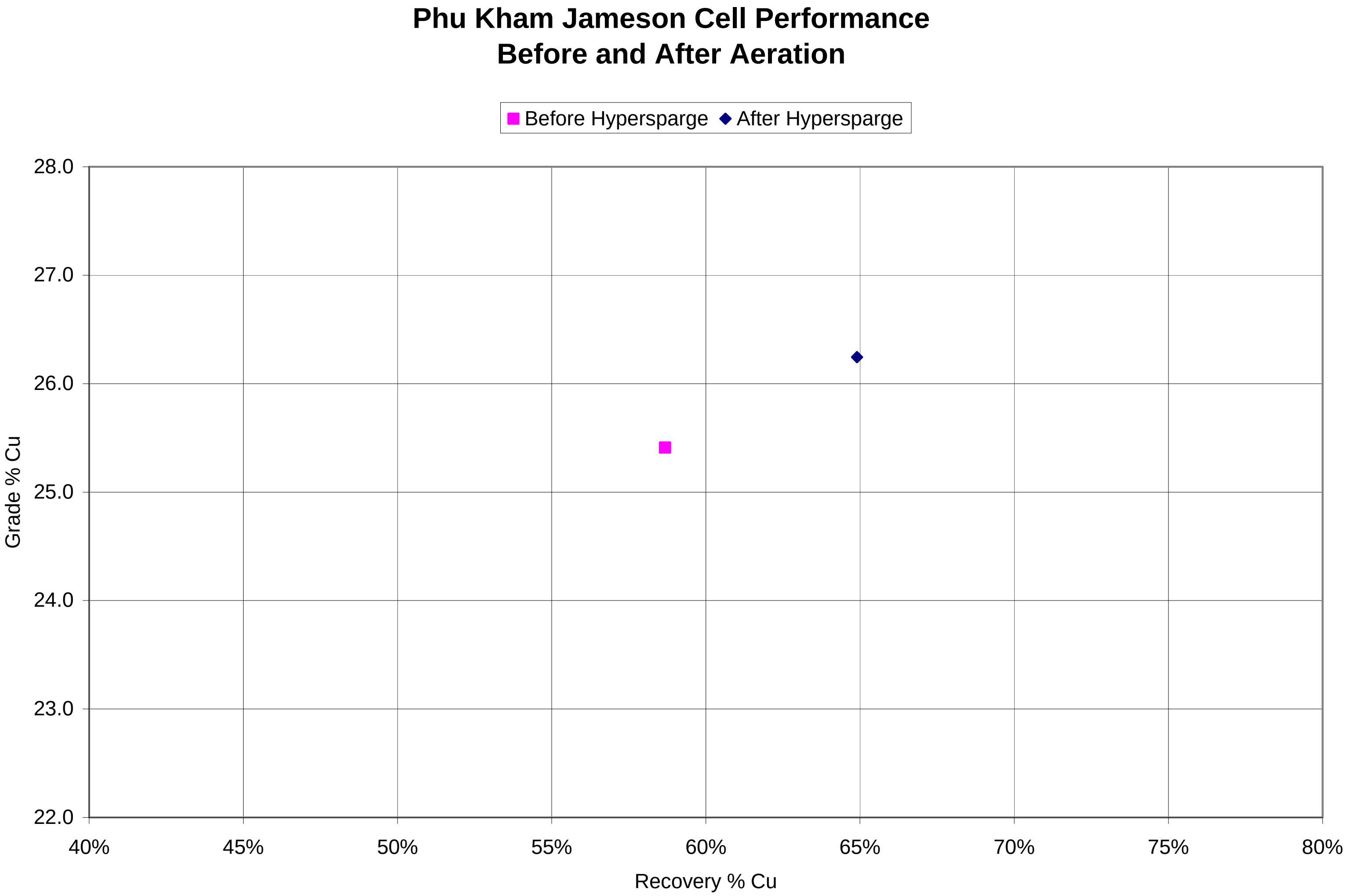
Fine Bubble Generation

Able to generate much finer bubbles for fast gas transfer rates

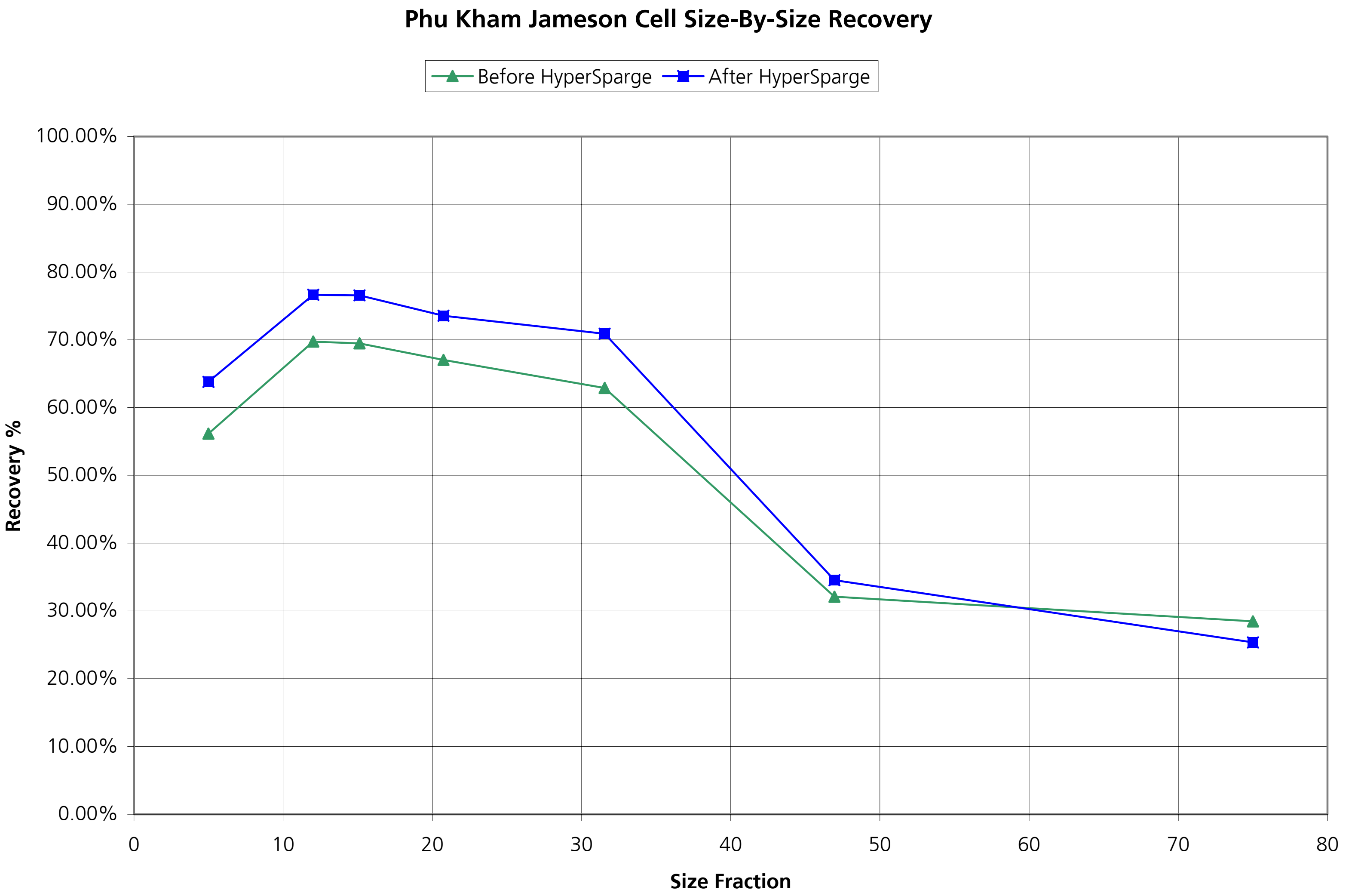


HyperSparge installed in IsaMill Discharge line

Phu Kham – Effect of Aeration Plant Surveys



Phu Kham – Effect of Aeration Plant Surveys



Summary

- IsaMill regrinds rougher concentrate to 38 micron to improve the final copper concentrate quality

- Cleaner feed performance at 38 microns is 25-28% Cu in final concentrate
- and regrinding 25 microns can produce +30% Cu in final concentrate

- Jameson Cell installation was an innovative flowsheet change to removed load from cleaning circuit and increased plant capacity and decrease cleaner circuit frothing issues

- Aeration of Cleaner feed has improve performance and fines recovery and selectivity

- Jameson Cell has allowed the IsaMill to operate at Increase power to improve the circuit performance.

PAPER 21

Removal of Organic Carbon with a Jameson Cell at Red Dog Mine

Todd Smith[1], Project Metallurgist
David Lin[2], Senior Research Engineer
Brigitte Lacouture[1], Chief Metallurgist
Greg Anderson[3], Manager - Mineral Processing

[1]Teck Cominco Alaska Incorporated
Red Dog Mine
3105 Lakeshore Dr, Building A Suite 101
Anchorage, Alaska, 99517
PH: (907) 426-9206
Email: Todd.Smith@teckcominco.com
Email: Brigitte.Lacouture@teckcominco.com

[2]Teck Cominco Metals Ltd
Applied Research and Technology
Email: David.Lin@teckcominco.com

[3]Xstrata Technology – Canada
PH: (905)-281-9063
Email: GAnderson@xstratatech.com.au

Key Words: flotation, organic carbon, Jameson Cell

40th Annual Meeting of the
Canadian Mineral Processors

January 22 to 24, 2008
Ottawa, Ontario, Canada

ABSTRACT

The lead and zinc flotation circuits at Red Dog Mine are preceded by a preflotation circuit which recovers naturally-floating organic carbon contained in the ore. Organic carbon is a potential contaminant in the lead concentrate and is detrimental to lead-zinc selectivity. Prior to 2007, the Red Dog preflotation circuit consisted of a single stage bank of tank cells which floated the organic carbon. The preflotation concentrate was discarded directly to tailings, however, this stream contained significant amounts of lead and zinc. Zinc loss to preflotation ranged from 2% to 6%, relative to the zinc in the mill feed and was proportional to the amount of organic carbon in the ore. The main mechanism of the zinc loss was entrainment.

Pilot plant and modeling testwork were conducted to examine the use of a Jameson Cell to clean the preflotation concentrate. Results indicated that up to 90% of the zinc deporting to preflotation concentrate could be returned to the flotation circuit for recovery. The installation of a 5.4 m Ø Jameson Cell with 18 downcomers was initiated in June 2006. The new preflotation cleaning circuit was commissioned in March 2007. Conservatively, preflotation cleaning has resulted in zinc and lead absolute recovery gains of 1.0% and 1.5%, respectively. The $9.1M project has a payback period of approximately 1 year at current metal prices.

INTRODUCTION

Red Dog Preflotation Circuit

Red Dog Mine is located above the Arctic Circle in northwest Alaska. It is the world’s largest zinc concentrate producer with a production rate of more than 1.0M t/yr. Red Dog is operated by Teck Cominco Alaska Incorporated under an operating agreement with the Northwest Alaska Native Association (NANA). Due to the remote location of the mine, it is accessible only by air, or seasonally, by ocean-going barges.

The Red Dog deposit is a rich sedimentary exhalative (sedex) zinc-lead-silver deposit. The major sulphides, in decreasing order of abundance, are sphalerite, pyrite, galena and marcasite. Due to the complex nature of the ore body, run-of-mine ore is blended in stockpiles before being crushed and processed through the concentrator. The average mill feed grade for 2006 was 6.1% lead, 20.6% zinc, 8.0% iron and 0.67 % organic carbon.

When Red Dog Mine began production in November 1989, the lead and zinc flotation circuits were not preceded by preflotation. The initial mill feed was highly weathered and contained high levels of elemental sulphur and organic carbon both of which are naturally hydrophobic. Without preflotation, these contaminants reported directly to the lead concentrate and resulted in poor concentrate quality. As a makeshift measure, the first two lead rougher cells were converted to preflotation duty. In 1991, two SK-50 Outokumpu Skimair® cells were installed ahead of the lead circuit for preflotation. The Skimair® cells along with the first two lead rougher cells were operated in preflotation duty until the depletion of the weathered ore in 1995. The rougher cells were then returned to lead flotation duty. Unlike elemental sulphur, which was found predominantly in the surface ore, organic carbon is present throughout the orebody. Organic carbon occurs as disseminated organic matter in the black shale host rock.

The preflotation circuit was subsequently modified during major circuit expansion projects in 1998 and 2000. In 2001, the preflotation circuit consisted of six tank cells as shown in Figure 1. The first two cells were Maxwell MX14 cells which were retrofitted with OK-50 Outokumpu agitators. The other four cells were OK-50 Outokumpu cells arranged in two parallel banks. Concentrate from these preflotation rougher cells reported directly to the final tailings stream even though it contained significant amounts of lead and zinc.

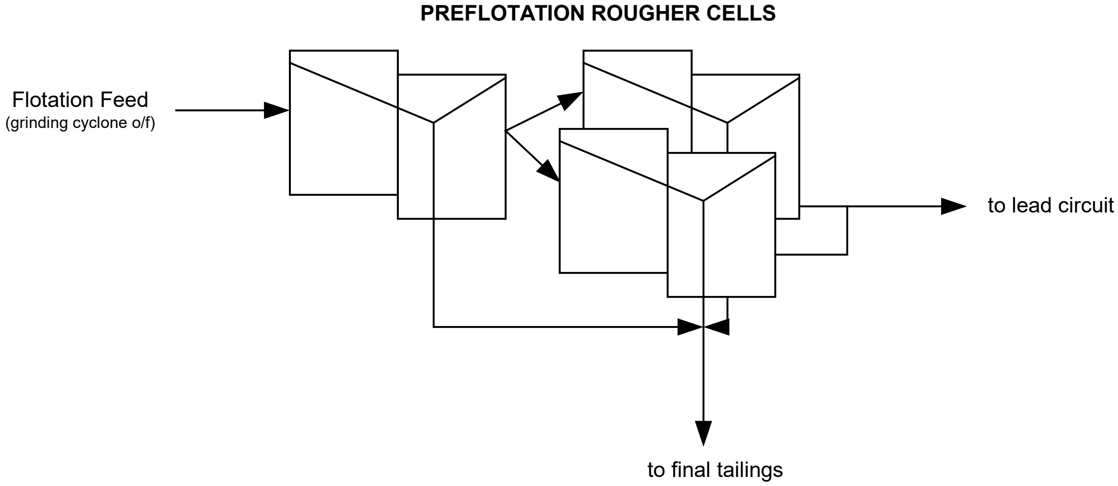


Figure 1: Red Dog preflotation circuit from January 2001 to March 2007.

Preflotation Concentrate Cleaning

Plant surveys identified that the deportment of zinc to the preflotation concentrate accounted for 2% to 6% of the total zinc loss, relative to the zinc in the mill feed. Mass balance data indicated that zinc and lead losses are proportional to the amount of total organic carbon (TOC) in the mill feed as illustrated in Figure 2. The zinc loss to the preflotation concentrate was as high as 8.3% for one stockpile in 2005. The majority of the zinc loss to the preflotation concentrate is due to entrainment. Zinc and lead losses to preflotation have been increasing as Red Dog mines higher TOC areas of the deposit. TOC in the mill feed is expected to remain high (>0.5% TOC) until 2012.

In 2000, laboratory batch flotation tests were conducted which demonstrated that it was possible to recover zinc from the preflotation concentrate stream by cleaning it. The testwork also showed that an acceptable zinc rougher concentrate could be subsequently generated from the preflotation cleaner tailings. A preflotation cleaning plant trial was conducted in 2002 utilizing a 4.1 m Ø flotation column with the column tailings sent to the zinc circuit. The preflotation cleaner tailings could not be sent to the lead circuit since lead flotation circuit capacity limited throughput at that time. Unfortunately, this trial was unsuccessful since significant short-circuiting in the column resulted in low unit TOC recovery making the column tailings unsuitable for the zinc circuit.

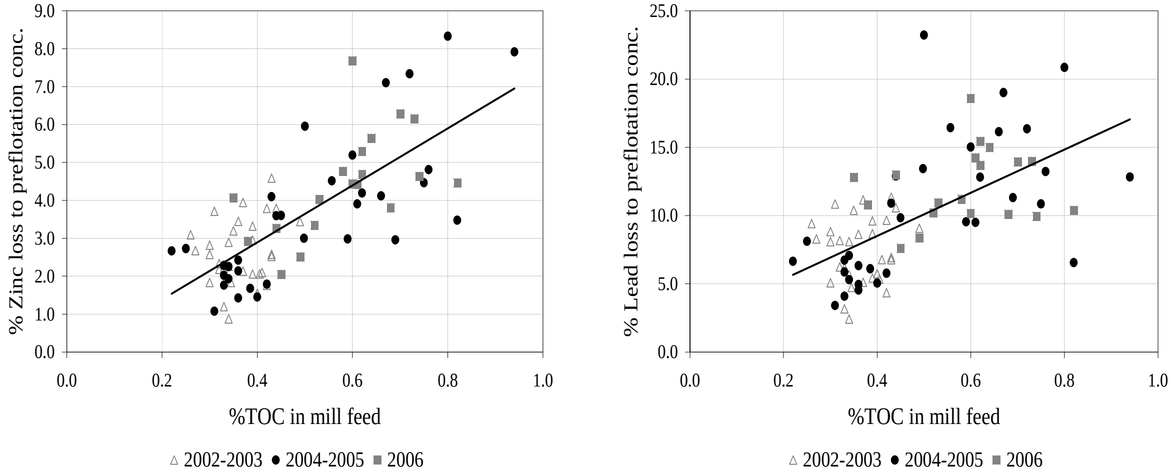


Figure 2: Relationship between mill feed TOC and zinc and lead losses to preflotation concentrate.

Additional lead flotation capacity was added in 2005 which allowed the consideration of other circuit configuration options for preflotation cleaning. It was proposed that preflotation cleaner tailings be re-circulated to the head of the flotation circuit instead of introduced to the zinc circuit. Re-circulating the preflotation cleaner tailings to the head of the flotation circuit provided an opportunity to recover both lead and zinc while rejecting TOC. A Jameson Cell was considered for preflotation cleaning duty.

The Jameson Cell

The Jameson Cell operating fundamentals have been described by numerous authors including Harbort et al (2003). A schematic of the Jameson Cell is shown in Figure 3. The most notable feature of the Jameson Cell is the downcomer where a plunging jet of slurry entrains air and bubble/particle contact occurs. The Jameson Cell typically operates with tailings recycle via the external recycle mechanism (ERM). A portion of the tailings stream is directed back to the feed pumpbox where it mixes with the fresh feed stream and is pumped to the Jameson Cell downcomers. The tailings recycle maintains a steady feed rate to the downcomers despite fluctuations in the fresh feed rate and increases unit cell recovery without adversely affecting concentrate grade. The circuit design and the benefits of a Jameson Cell in base metal applications have been described by Young et al (2006).

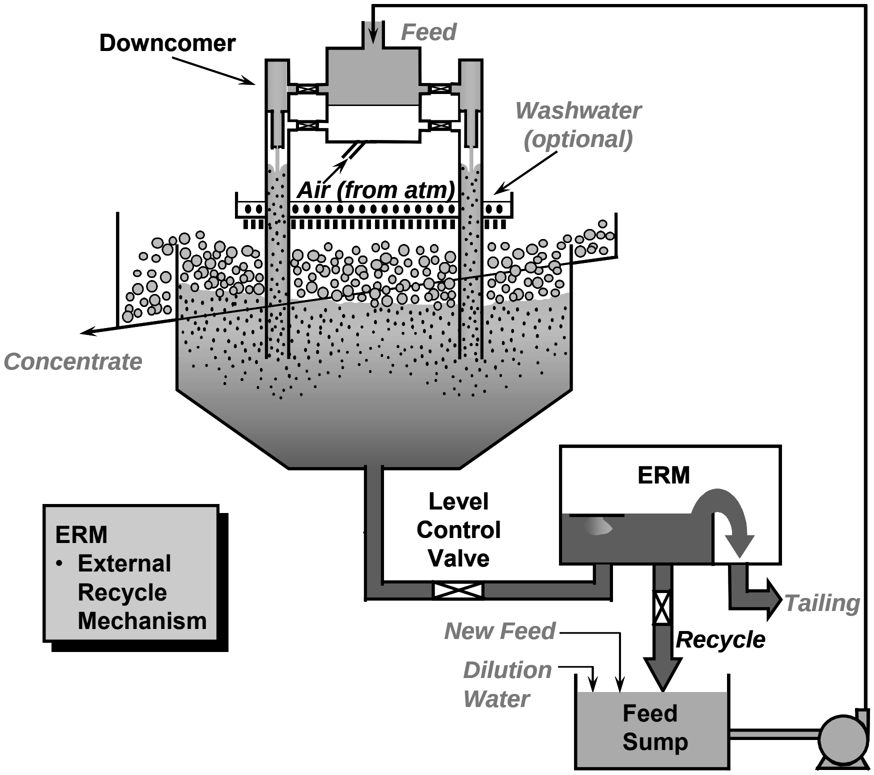


Figure 3: Jameson Cell schematic with external recycle mechanism (ERM).

RESULTS AND DISCUSSION

Pilot Plant Testwork

In April 2005, Xstrata Technology was invited to Red Dog to conduct pilot plant tests on preflotation concentrate cleaning. A 300 mm Ø Jameson Cell pilot rig was installed and tests were conducted in open circuit configuration (i.e., the Jameson Cell tailings were not returned to the preflotation circuit). It was not possible to simulate full scale closed circuit operation due to the small capacity of the pilot cell.

The testwork program was performed in a two week period. The aim was to minimize zinc loss to the preflotation concentrate while maximizing TOC removal. Feed for the Jameson Cell was bled from the discharge of the preflotation concentrate disposal pump. The mill feed TOC assays ranged from 0.7% to 0.8% during the course of the pilot plant testwork.

A summary of the pilot plant test results is shown in Figure 4. The Jameson Cell was able to achieve a maximum TOC unit recovery of 70%. Under optimized conditions, the zinc unit recovery was as low as 10% (i.e., 90% zinc rejection).

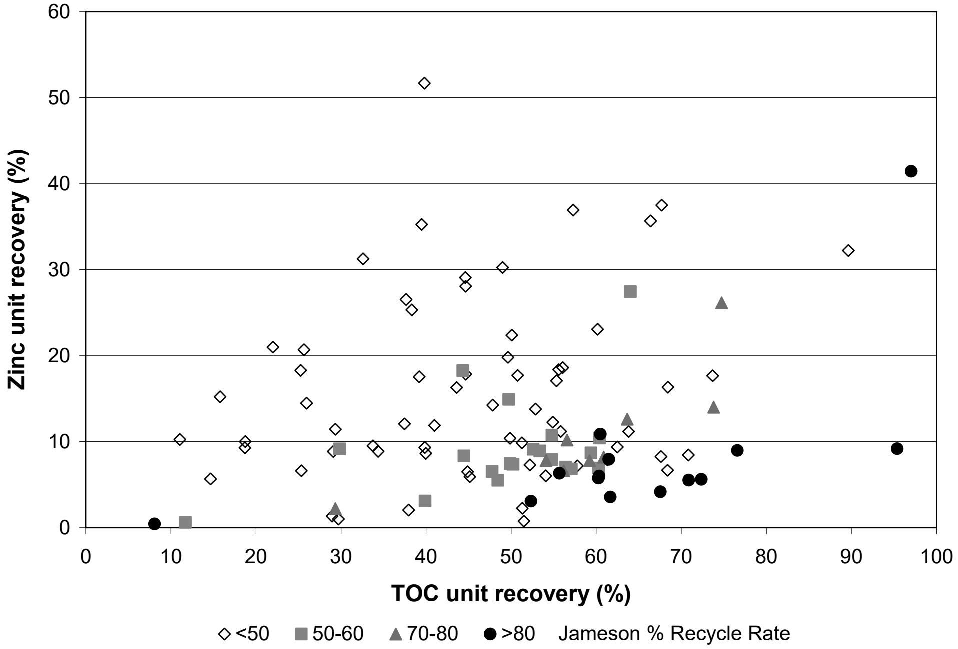


Figure 4: Relationship between TOC and zinc unit recoveries for the Jameson pilot plant.

The test program evaluated the following parameters: recycle rate, froth depth, jet velocity, feed density, air to pulp ratio (APR), superficial gas velocity (Jg), frother addition and wash water. Recycle rate and feed density were found to have the most significant effect on selectivity as seen with other Jameson Cell preflotation cleaner applications (Pokrajcic et al, 2005).

As a result of this testwork, Xstrata outlined the following design criteria for a Jameson Cell preflotation cleaner at Red Dog:

- Minimum of 80% cell recycle rate. The data indicated that for a given TOC recovery, improved selectivity of TOC over zinc was achieved at higher recycle rates.
- Jameson Cell feed density < 16% solids. Dilute feed improves dilution cleaning.
- Jet velocity at 15 m/s. Test data showed no significant impact on cell performance across the range of jet velocities tested.
- APR of up to 1.5. This is higher than normal for cleaning operation, but was selected to assist TOC recovery.
- Wash water addition was not required.

Batch kinetic flotation tests were also performed on each stream of the pilot plant. This permitted the construction of a mathematical model of the preflotation circuit with a Jameson Cell in preflotation cleaning duty using the principles of the AMIRA P9 project (Harris, 2002). Simulation results showed that at a maximum mill feed rate of 455 tph and a high TOC feed grade of 0.8%, the Jameson Cell would be required to handle 75 tph. The simulations also concluded that closed circuit Jameson Cell cleaning could reduce preflotation concentrate zinc loss by up to 80% in high TOC stockpiles.

Based on the above data, Xstrata Technology determined that a 5.4 m Ø Jameson Cell with 18 downcomers (B5400/18) should be selected for preflotation cleaning. The B5400/18 met Red Dog's requirement for a nominal feed rate of 50 tph and would be capable of handling up to 80 tph with an 80% minimum recycle rate.

Construction of Preflotation Cleaning Circuit

Engineering and procurement for the addition of a B5400/18 Jameson Cell in the preflotation circuit began in February 2006. Major equipment and construction materials were brought to site during the 2006 summer sealift with additional materials arriving by air freight. Due to limited space within the existing Red Dog concentrator, a new module was constructed to house the Jameson Cell. Construction began in June 2006. The project also required upgrading the preflotation rougher concentrate pumpbox and pumps within the existing module. Photos from construction of the new module and the completed Jameson Cell are shown in Figures 5 and 6.

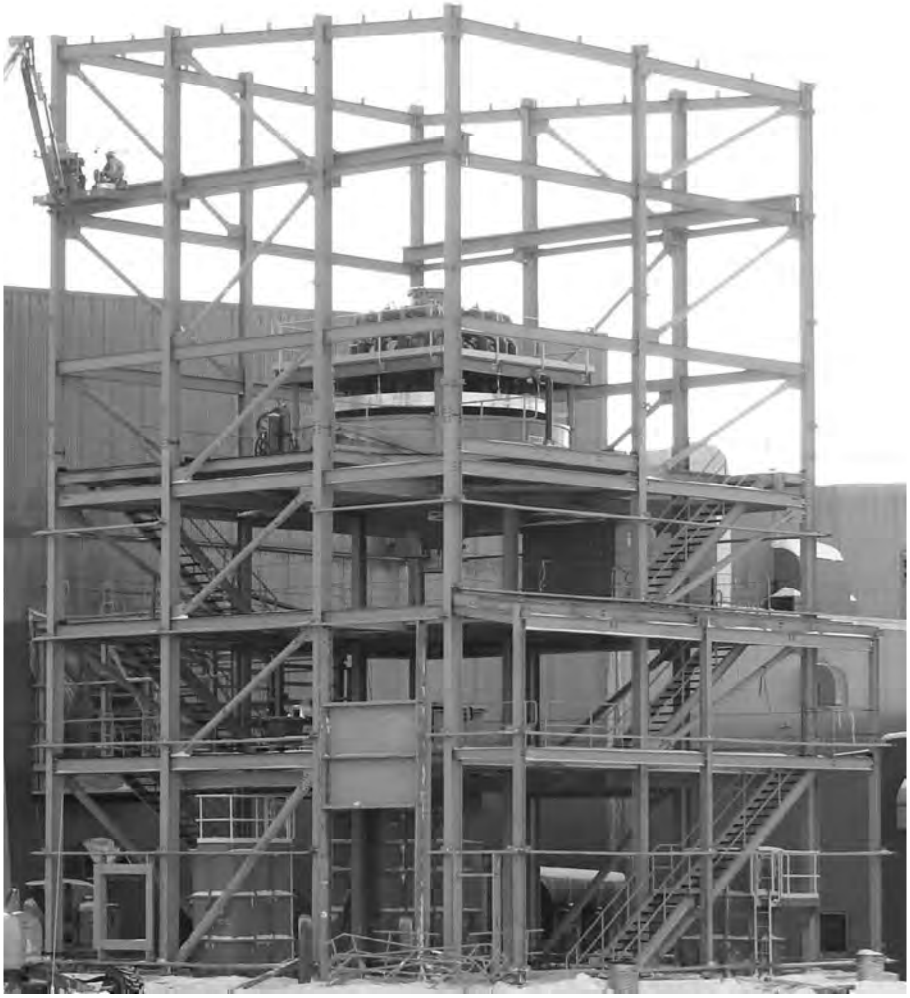


Figure 5: Construction of the new module for the Jameson Cell.

Figure 6: The preflotation Jameson Cell during commissioning at Red Dog Mine.

Commissioning Surveys

The preflotation cleaning circuit was commissioned in mid-March 2007. A schematic of the circuit is shown in Figure 7. Commissioning surveys of the new preflotation circuit were conducted over the next few months. During the first month, the mill treated an unusually low TOC stockpile; mill feed assays ranged from 0.2% to 0.3% TOC compared to the expected 2007 average of 0.6% TOC. Additional surveys were later conducted when the mill feed was 0.5% to 0.7% TOC.

The Jameson Cell was operated under different air rates, froth depths and pulp densities during the commissioning period. Since the pilot plant work indicated wash water was not necessary, all but three tests were conducted without wash water. Feed density to the Jameson Cell ranged from 12% to 19% solids. Recycle rates were >80% as per design.

For comparison, surveys of the preflotation circuit were conducted with and without the Jameson Cell in operation. A summary of these results are shown in Figures 8 and 9. These figures show TOC selectivity against zinc and lead for both configurations. Preflotation cleaning was found to be effective at reducing zinc and lead losses to the preflotation concentrate stream. There is a statistically significant difference between the regression curves with and without the Jameson Cell on-line. At 35% TOC recovery, zinc and lead losses were reduced by an absolute of 1.5%

and 2.5%, respectively. The magnitude of the reduction is dependant upon the mill feed TOC grade and the preflotation recovery target. Since the Jameson Cell tailings returns to the head of the circuit, zinc and lead rejected by the Jameson Cell has the opportunity to be recovered in their appropriate flotation circuit.

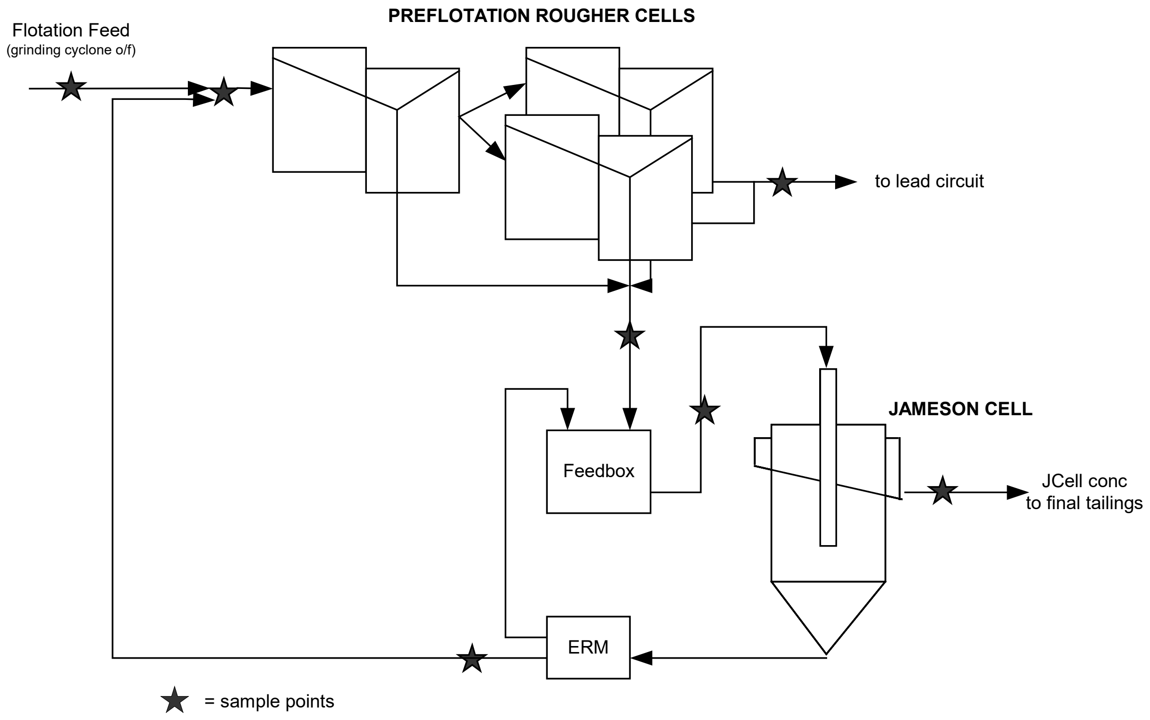


Figure 7: Red Dog preflotation circuit with the Jameson Cell preflotation cleaner.

It should be noted that prior to the Jameson Cell commissioning, Red Dog implemented the addition of dextrin (an organic depressant) to the lead cleaner circuit which allows for lower preflotation TOC recovery. Regardless of whether the Jameson Cell is on-line or not, lower TOC recovery helps minimize zinc and lead losses via entrainment to the preflotation concentrate.

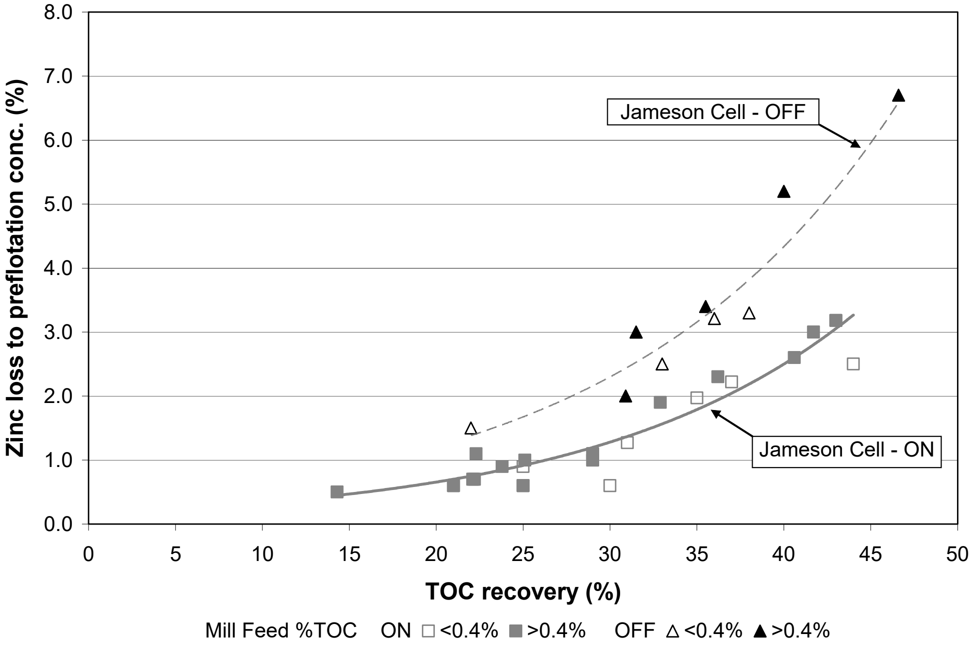


Figure 8: Overall TOC recovery versus zinc loss for the preflotation circuit.

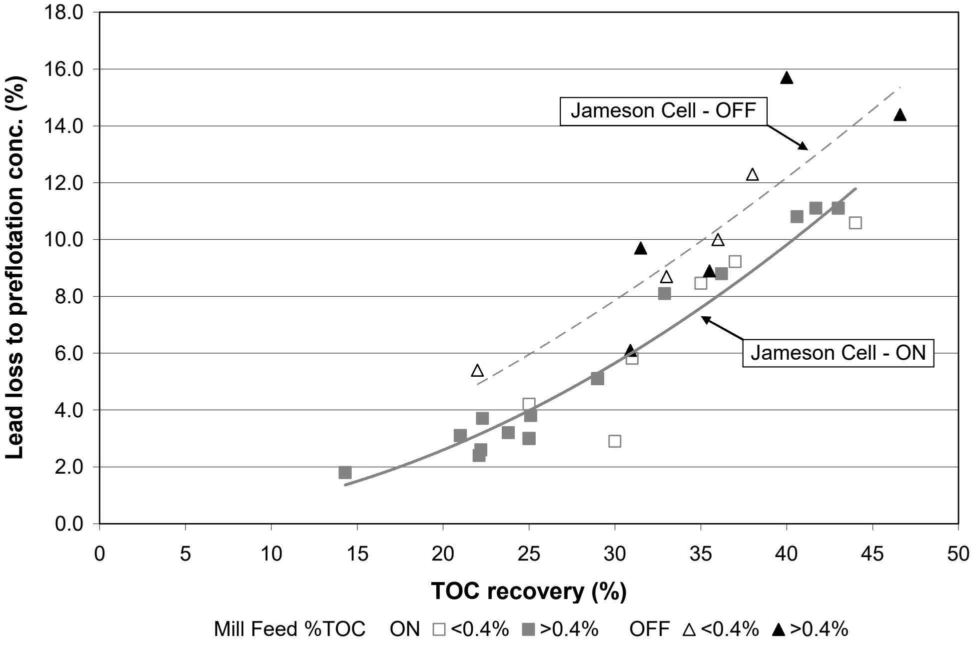


Figure 9: Overall TOC recovery versus lead loss for the preflotation circuit.

Figure 10 shows the unit TOC and zinc recoveries for the Jameson Cell during these surveys. At 70% TOC unit recovery, 34% of the zinc was recovered (i.e., 66% zinc rejected to Jameson Cell tailings). Future optimization of the Jameson Cell is expected to improve performance.

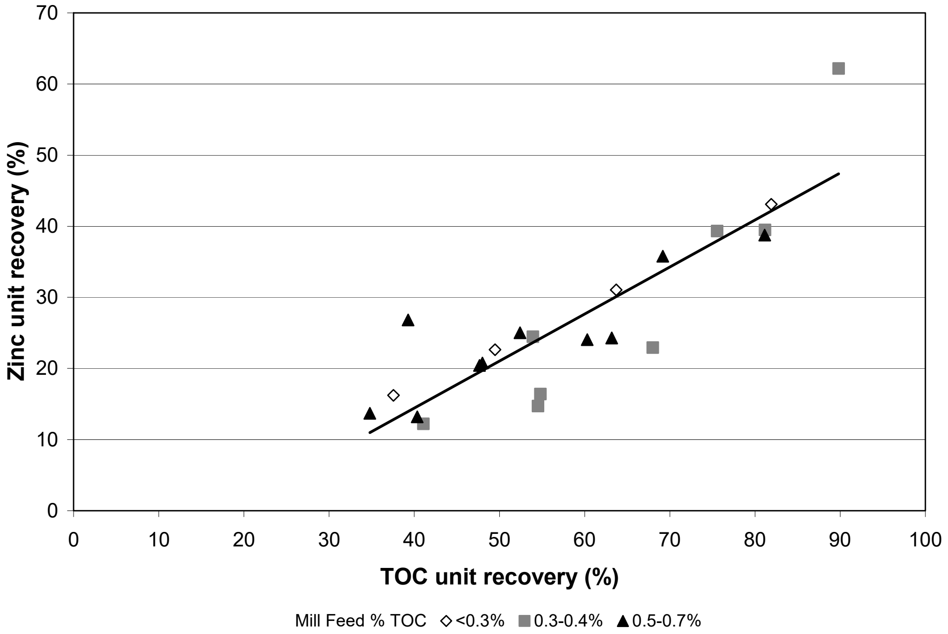


Figure 10: TOC unit recovery versus zinc unit recovery for the Jameson Cell.

Recent changes have been made to the Jameson Cell:

- The 40 mm slurry lenses on the downcomers were replaced with 44 mm lenses per the original Xstrata design criteria. This has increased the Jameson Cell recycle rate.
- A frother addition point was added to the Jameson Cell feed for use when the TOC in the feed is >0.7%. The additional frother increases the amount of air entrained in the downcomers and increases unit recovery.

Future testwork will be conducted to evaluate these changes. Wash water addition will also be evaluated.

Recoverability of Jameson Cell Tailings

To quantify how much lead and zinc rejected by the Jameson Cell reports to their appropriate final concentrate, laboratory batch cleaner flotation testwork was conducted on Jameson Cell tailings. The flotation performance of the Jameson Cell tailings was expected to be significantly worse than that of fresh feed for the following two reasons:

- Zn/Pb, Zn/Fe and Zn/TOC ratios are significantly lower than typical feed grade.
- The majority of Jameson Cell tailings is fine (<15 µm) and thus presents a potential challenge in lead and zinc flotation.

Results from the batch flotation tests are summarized in Figures 11 and 12. This testwork was performed only on the initial low TOC stockpile.

In Figure 11, one stage of lead cleaning yielded concentrates that graded from 39% to 54% lead with lead recoveries ranging from 47% to 59%. In Figure 12, four stages of open circuit zinc cleaning yielded concentrates that graded from 47% to 52% zinc at recoveries ranging from 54% to 61%. Cleaning circuits in the plant operate in closed circuit, thus plant recoveries will be higher than in open circuit batch tests. It is expected that at a minimum, 60% of the lead rejected by the Jameson Cell will be recovered to lead final concentrate and 70% of the zinc will be recovered to the zinc final concentrate.

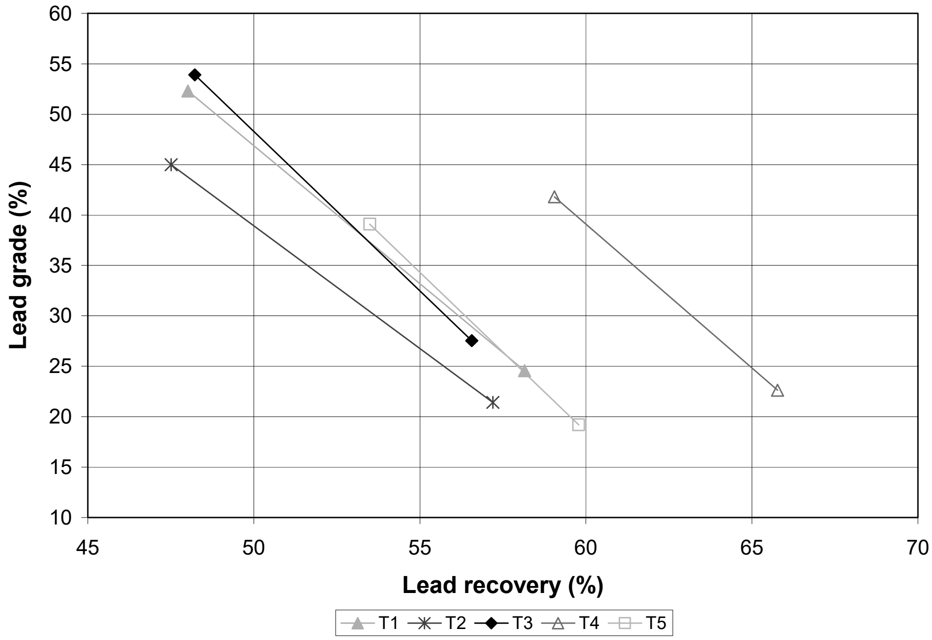


Figure 11: Lead grade-recovery relationship for the flotation of Jameson Cell tailings.

Project Economics

At a zinc price of \$1.30/lb and a lead price of \$1.20/lb, this \$9.1M project has an approximate payback period of 1 year based on the following:

- A 1.5% absolute reduction in the zinc loss to preflotation concentrate equates to a 1.0% increase in overall zinc recovery (assuming 70% zinc recovery of the rejected material).
- A 2.5% absolute reduction in the lead loss to preflotation concentrate equates to a 1.5% increase in overall lead recovery (assuming 60% lead recovery of the rejected material).

Figure 13 shows the daily zinc and lead flotation circuit recoveries before and after the Jameson Cell was commissioned. On average, zinc recovery increased 2.1% and lead recovery increased 3.0% since March 2007. The increase in flotation circuit recoveries has been attributed to the synergy between the Jameson Cell preflotation cleaner and the use of dextrin in the lead circuit.

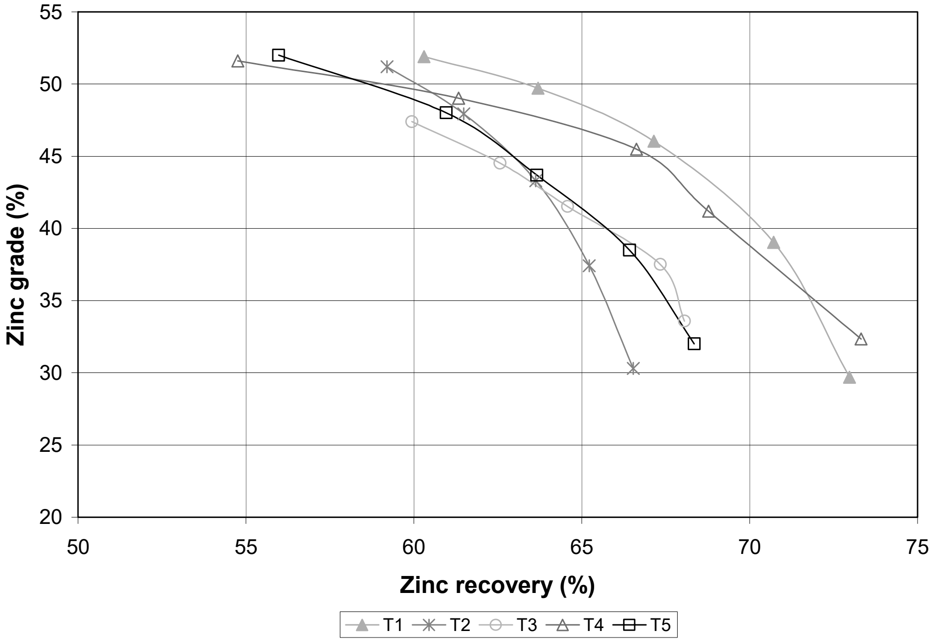


Figure 12: Zinc grade-recovery relationship for the flotation of Jameson Cell tailings.

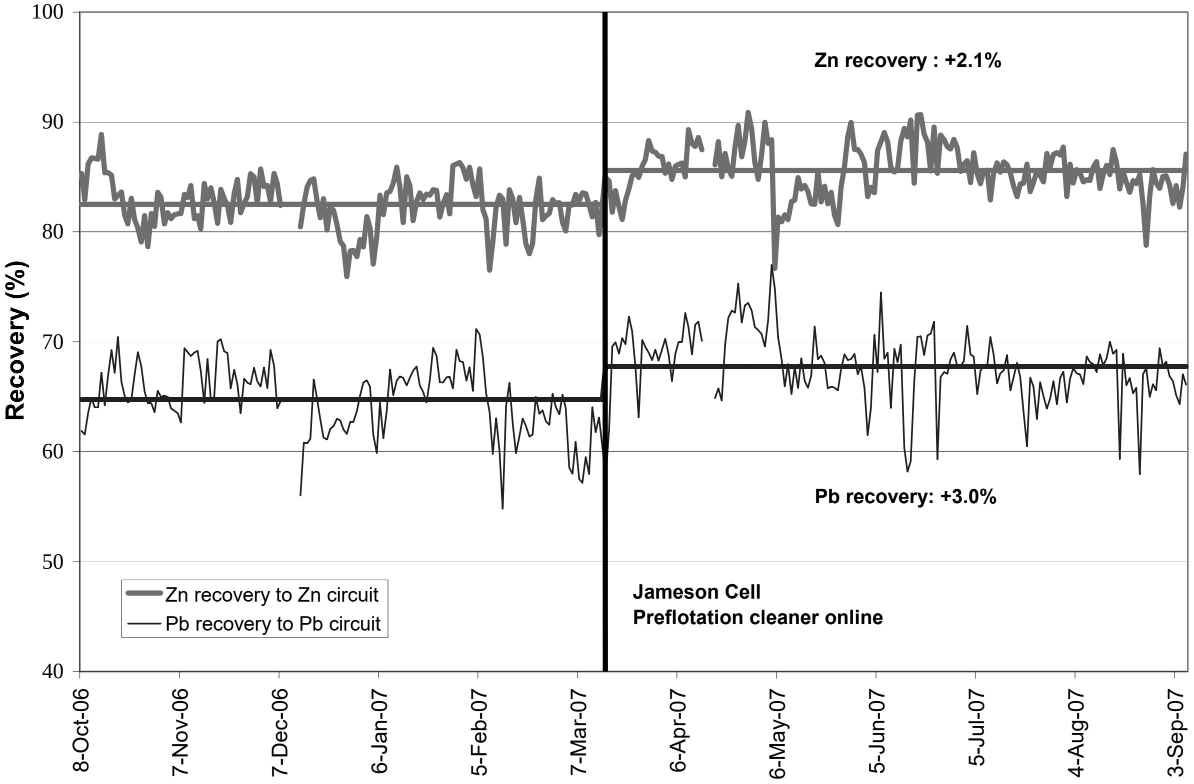


Figure 13: Zinc and lead circuit recoveries pre/post Jameson Cell.

CONCLUSIONS

Survey and operating data have shown that the installation of a Jameson Cell for preflotation concentrate cleaning has improved flotation circuit performance at Red Dog. Conservatively, preflotation cleaning has resulted in zinc and lead absolute recovery gains of 1.0% and 1.5%, respectively. The $9.1M project has a payback period of approximately 1 year at current metal prices.

ACKNOWLEDGEMENTS

The authors gratefully acknowledge the metallurgical teams at both Red Dog Mine and Teck Cominco Applied Research and Technology for their assistance and support on this project. We thank Xstrata Technology for their contributions and support with project testwork. We also wish to thank Teck Cominco for permission to publish this work.

REFERENCES

Habort, G., De Bono, S., Carr, D. Lawson, V., 2003. “Jameson Cell fundamentals – a revised perspective”. *Minerals Engineering*. 16 (2003), 1091-1101.

Harris, M.C., Runge, K.C., Whiten, W.J. and Morrison, R.D., 2002. “JKSimFloat as a practical tool for flotation process design and optimization”. *SME Mineral Processing Plant Design, Practice and Control Conference*. Vancouver, Canada, October, 461-478.

Pokrajcic, Z., Harbort, G.J., Lawson, V., Reemeyer, L., 2005. “Applications of the Jameson Cell at the Head of Base Metal Flotation Circuits”. *Centenary of Flotation Symposium, Brisbane QLD*, 165-170.

Young, M., Barns, K., Anderson, G., Pease, J., 2006. “Jameson Cell: The “comeback” in base metals applications, using improved design and flowsheet”. *38th Annual CMP proceedings. Ottawa, ON*, 311-332.

DESIGN DEVELOPMENTS OF THE JAMESON CELL

Cowburn, J (1), **Stone, R**(2), **Bourke, S**(3), **Hill, B**(4)

(1) Xstrata Technology

(2) Sedgman

(3) Rio Tinto Coal, Hail Creek Mine

(4) BHPB Mitsubishi Alliance, Riverside Mine

Centenary of Flotation 2005 Symposium, Brisbane. June 5-9 2005

Joan Ann Cowburn

Xstrata Technology, Level 2, 87 Wickham Terrace, Brisbane, QLD, 4000

Telephone (07) 3833 8569

Fax (07) 3833 8555

jcowburn@xstrata.com.au

DESIGN DEVELOPMENTS OF THE JAMESON CELL

Cowburn, J (1), Stone, R (2), Bourke, S (3), Hill, B (4)

(1) Xstrata Technology

Level 2, 87 Wickham Terrace, Brisbane, QLD 4000

(2) Sedgman Pty. Ltd.

40 Station Road, PO Box 163, Indooroopilly, QLD 4068

(3) Hail Creek Mine, Rio Tinto Coal

PO Box 212, Nebo, QLD 4742

(4) Goonyella Riverside Mine, BMA Coal

Moranbah, QLD 4744

Keywords:

Flotation, Coal, Jameson Cell, Hail Creek, Riverside

ABSTRACT

Since Jameson Cell flotation technology was first installed in a coal flotation application in 1988/89 at Newlands Coal Handling and Preparation Plant (CHPP) there have been fundamental improvements to its design which has led to its current status as the leading technology in Australia for fine coal flotation. These improvements, which include key hardware modifications, consist of the re-development of the orifice plate to produce considerable increase in wear life and reduction in power consumption for a given air entrainment and a significant increase in the amount of air that is entrained. Modifications to the downcomer allow maximisation of residence time and air entrainment while changes to the feed and air distribution system allows a reduction in installation cost and a reduction of solids ingress into the air line. Additionally, operational improvements such as the use of recycle to maintain constant flow to the cell feed are discussed. The recent installation at Hail Creek in the Bowen Basin, which consists of three B6000/20 Jameson Cells (6m diameter bottom-fed distributor with 20 downcomers) will be reviewed as a case study. The latest design and operability of the cells are summarised and compared against one of the original designs at Riverside CHPP also in the Bowen Basin.

JAMESON CELL OPERATION

The Jameson Cell is a high intensity flotation device, which utilises induced air as the medium for froth flotation. It was developed jointly by Mount Isa Mines and Prof. G J Jameson of the University of Newcastle in the 1980's initially treating lead slimes in the lead zinc concentrator in Mount Isa. To date there are 228 installed Jameson Cells, 94 being in coal flotation applications across the globe, 77 of which are in Australia. The principles of Jameson Cell operation have been discussed by numerous authors including Jameson et al (1988) and Evans et al (1995) and recently by Harbort et al (2003) and Harbort et al (2004) and so will not be discussed at length in this paper.

The slurry is fed to the Jameson cell slurry distributor at elevated pressure and is evenly split between the downcomers. The jet created in each downcomer by the slurry passing through the orifice promotes the inducement of air (Figure 1). The shearing action of the jet on the column of slurry within the downcomer generates fine bubbles and transports them through the mixing zone. Particles and the bubbles collide and attach to each other and subsequently travel down the downcomer through the pipe flow zone. Bubbles are removed by hydrostatic pressure from the downcomer creating a vacuum for further air entrainment. The aerated slurry exits the bottom of the downcomer and the buoyancy of the bubble/particle aggregates cause them to rise towards the froth zone. The pulp zone is principally a region of disengagement although some further collection can occur. The froth zone is the where entrained materials are removed from the froth by froth drainage and/or froth washing (Harbort et al, 2004)

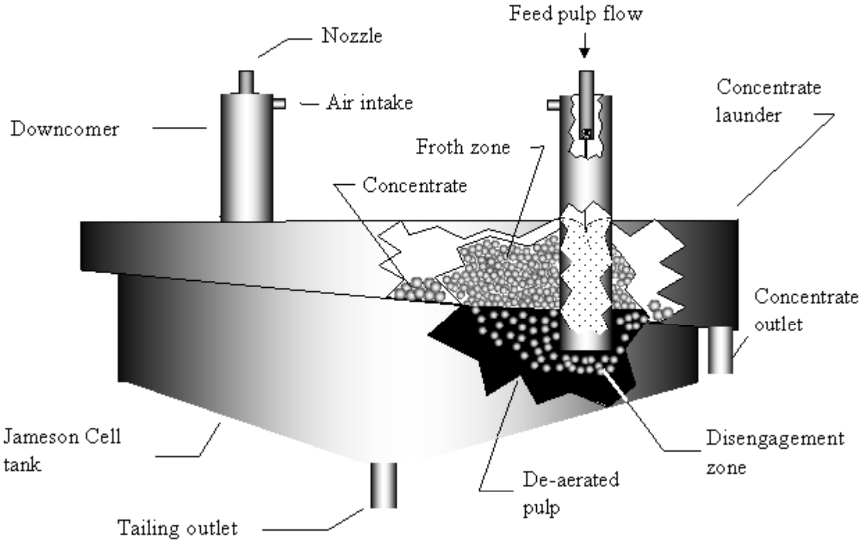


Figure 1. Schematic of Jameson Cell

JAMESON CELL INCEPTION AND DEVELOPMENT 1985-1990

In 1985 Mount Isa Mines commissioned G J Jameson to commence a project to improve the sparger design in the column cleaners in the zinc circuit. Following extensive research the concept of the downcomer was developed. The notion of co-current air and slurry direction and naturally aspirating air under a vacuum was a seldom used concept and was initially thought of as a new sparger design for a tall column (Harbort, 1992). However, further investigations showed that most bubble particle interactions took place within the high void fraction environment in the downcomer and so the collection zone of a column was unnecessary. This lead to the development of the short tank design now realised as the Jameson Cell.

In 1986 a provisional patent was lodged by G J Jameson, which was later assigned to TUNRA Ltd, University of Newcastle. Research into the technology continued with a small 2 tph pilot cell with a 100 mm downcomer and approximately 13mm orifice plate being tested at Mt Isa in the lead/zinc concentrator. In 1988 Mount Isa Mines (MIM) decided to increase the capacity of their heavy medium plant slimes flotation circuit to improve lead recovery. Investigations were undertaken into mechanical, column and

Jameson flotation cells with the latter giving the highest recoveries, which was attributed to the combination of a mineral whose hydrophobicity decreases with time and the short residence time of slurry within the Jameson Cell.

Orders were secured in 1989 for two full-scale Jameson cells for Mt Isa Pb/Zn concentrator (Harbort, 1992) and two units for the Hilton Pb/Zn concentrator.

Also in 1989, testwork was conducted at Peko Wallsend concentrator in Tennant Creek in a copper cleaning duty in an attempt to increase final concentrate grade. The test cell had a diameter of 530mm with a 102mm downcomer treating 2 tph of solids (Jameson et al, 1991). Following the test program two 1.4m diameter cells were installed in December 1989 (Harbort, 1992).

In parallel to the investigations occurring in metalliferous operations, MIM, in conjunction with G J Jameson, began trials on the settling cone overflow at Newlands Coal. The objective being to recover very fine coal: 100% passing 100 μm and a d_{50} 35 μm. Six flotation cells were installed in a two-stage arrangement. The primary units had seven 200mm downcomers while the secondary units had six downcomers. The flotation tanks were rectangular in design, each 1.5 x 3.5 m, allowing the cells to be incorporated in existing plant, and the full-scale plant was commissioned in 1988/89.

By 1990 the standard orifice plate diameter had increased to 28mm, allowing a throughput of 30 m^3/hr per downcomer. Table 1 details the modifications in downcomer diameter, orifice diameter and the consequent flow per downcomer since the inception of the Jameson Cell.

In April 1989 MIM Holdings Ltd acquired world rights from TUNRA Ltd/University of Newcastle for metallurgical purposes with TUNRA pursuing applications in wastewater treatment.

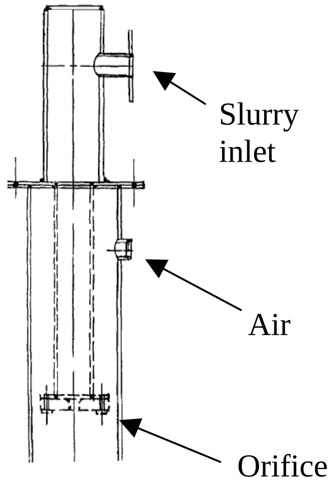
Table 1. Standard Downcomer Diameters and orifice sizes

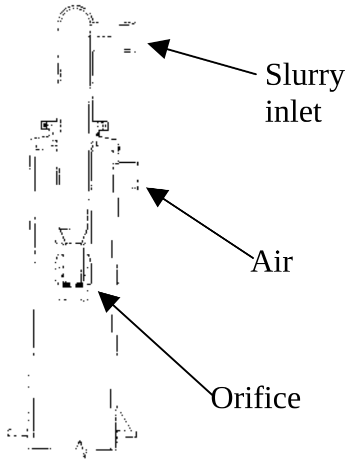
Year	**Downcomer Diameter (mm)**	**Orifice Diameter (mm)**	**Flow per Downcomer (m3/hr)**
1989	200	18	14
1990	200	28	30
1993	280	34	50
1997	280	38	60
1999	280	42[*]	75

[*] 42mm refers to a slurry lens design compared to an orifice plate design

DEVELOPMENT 1991-1993

The principal development during this stage of the Jameson cell early life was the downcomer, figure 2. From initial fabrication in polyurethane lined steel the design migrated into a HDPE construction with seven elements. Although this addressed the issue of weight, concerns with wear of the downcomer were experienced and so both designs were relatively short lived. Materials of construction of the orifice plate were investigated in 1991, including high chromium hardened steel and various ceramics (Harbort et al, 1994). High density Alumina was deemed to have excellent wear properties, and became the standard. The maximum Jameson Cell diameter for this period was 3.5m with 200mm diameter downcomers allowing a surface area of 8.2m^2.





DEVELOPMENT 1994-1999

During this phase many developments occurred in Jameson Cell technology, namely:

- Increase of tank diameter, downcomer diameter and so distance between downcomers
- Increase depth of tank from bottom of downcomer
- Introduction of both internal and external recycle in place of downcomer isolation
- Optimisation of bubble diffusers
- Mark 3 downcomer incorporating AISE valve and Slurry lens
- Design of the Z Cell

A few of the above will now be discussed in depth below:

Tank Diameter and Depth, Downcomer Diameter and Downcomer Distances

Design of the Jameson Cell has been driven by users directing machine development. Added to this has been very practically oriented research into fundamental behaviour by three of Australia’s premier research groups, the University of Newcastle, University of Queensland/JKMRC and the Commonwealth Scientific and Industrial Research

Organisation (CSIRO). With this input, a key improvement was the increase in the maximum cell diameter, from 3.5m in 1993 to 6.5m in 2000. Parallel to this, as referred to previously in table 1, the diameter of the downcomer and the orifice diameter were also increased. Consequently, the distance between the downcomers was able to be increased thereby reducing the interaction of aerated slurry exiting neighbouring downcomers. This interaction would cause increased pulp phase turbulence that could affect overall cell recovery by causing particles recovered in the downcomer to become detached.

In terms of operation within the Jameson Cell, tank void fraction measurements show that bubble patterns in general form a central, air swept cone, as described by Taggart in 1945. The Jameson Cell tank contains areas of high, localised air void throughout the pulp zone. The rising swarm of bubbles is governed by a number of factors including recirculating patterns within the tank, pulp flow volumes and air flow volumes. Increasing the volumetric flowrate per downcomer by the above mentioned design changes may result in secondary bubble recirculation patterns within the tank, potentially resulting in attached particles becoming detached from the bubble, mineral laden bubbles being drawn into the tailing and possibly erosion of the lower portion tank structure. Resulting from a period of intense investigation the redesign of the Jameson cell took into account the increase volumetric throughput per downcomer ensuring the aerated pulp would not impact on the cell floor.

Recycle

During early Jameson Cell development it was understood that for optimum operation and plunging jet formation the device should be operated at a fixed volumetric feed rate. Due to the nature of most concentrators and coal preparation plants this was an unrealistic request and so initially in periods of reduced flow downcomers were isolated. This requires quite complex instrumentation or manual intervention and results in uneven flow across Jameson cell. This would also result in variable feed pressure and implicitly jet velocity that would affect air entrainment and reduce overall flotation performance. Consequently recycle was instigated. The Jameson cell and feed system are designed to operate at a higher volumetric throughput than the nominal fresh feed flow, with approximately 30% to 40% of the cell feed being made up of recycled tailing. In addition to stable flow to the flotation cell, this can improve recovery while not affecting concentrate quality by providing mineral misreported to tailing with another opportunity to attach to a bubble and be recovered to the concentrate launder. Additionally if cell fresh feed is lost completely the device can operate in a 100% recycle mode to act as pump protection. Recycle has now become an integral part of the Jameson cell with internal, external and detached external options available.

Bubble Diffusers

Underneath the downcomer is an area of significant turbulence and so bubble diffusers have become a feature of the Jameson cell. The design of diffuser plates has been optimised through the continuous development of the flotation device. Significant amounts of testwork has been conducted to optimise the shape, location, and porosity of the bubble diffuser. Diffusers allow uniform bubble rise velocities across the surface of the cell by slowing the superficial gas velocity in the high void faction area immediately around the downcomer. Diffusers also act to ensure even bubble dispersion thereby reducing entrainment in the froth (Harbort, 1997, internal report). The current design reduces turbulence by 69% compared to a standard downcomer with no diffuser.

Mark 3 Downcomer

The downcomer is the heart of the Jameson Cell and its design and operability are keys to the performance of the technology. Although various designs have been used for different applications and improvements introduced, the basic design remained the same for some time. A fresh approach to downcomer design has resulted in a dramatically different design that reduces the number of parts by over half and further increases the simplicity of operation of the equipment. The Mark 3 downcomer allows all parts to be located outside the downcomer, with access greatly simplified. Additionally, with the location of the slurry lens compared to the orifice plate, the effective length of the downcomer has been increased by 15%, thereby improving residence time in the mixing zone and allowing operation at higher Air-to-Pulp ratios. Laboratory scale test work has shown that the longer length in downcomer allows increased air entrainment for a given vacuum (Figure 3).

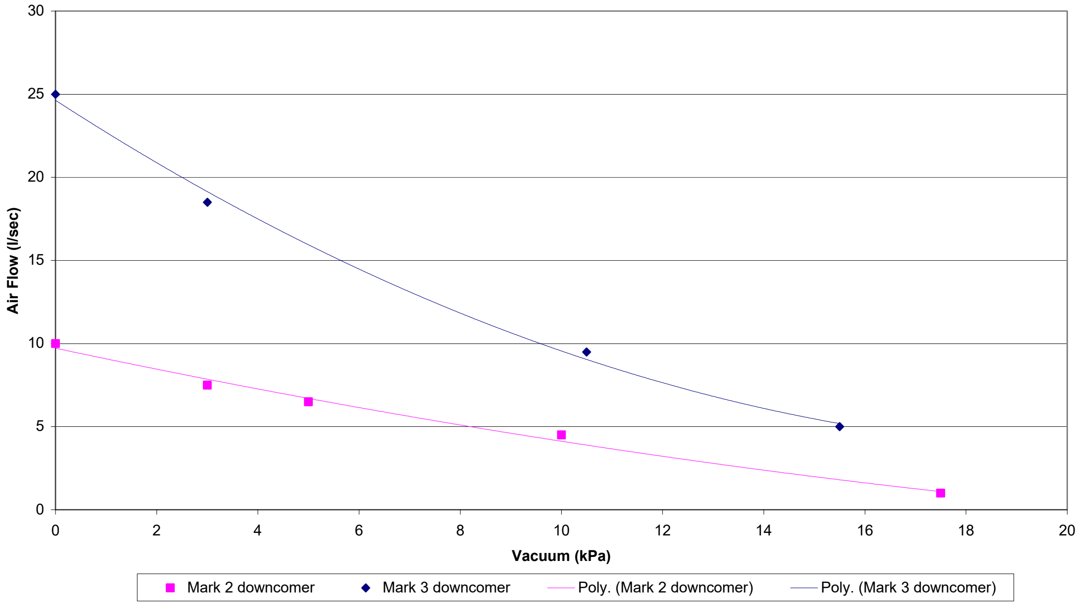


Figure 2. Relationship between Downcomer length and Induced Air Flowrate

Slurry Lens

1999 saw the replacement of the orifice plate used in the downcomer to form the plunging jet with the Slurry Lens. This was a staged development with the initial prototype being conceived in 1996 and the current design being settled in 1999. The key feature of the design is the smooth shallow entry angle (Xstrata Technology website). This ensures an optimum flow regime over the ceramic for maximum wear life. The ceramic is backed by polyurethane to cushion the impact of large heavy objects such as bolts. Even if the ceramic is damaged, the polyurethane serves to keep the ceramic serviceable.

Further benefits of the profile include:

- The ability to pass rod-like objects with greater ease, minimising blockages.

- An increase in the discharge coefficient of the orifice, decreasing power consumption by as much as 10%.
- Better jet formation with less splashing resulting in improved vacuum and air entrainment.

AISE Valve

Under normal conditions small fluctuations occur in the operation of the downcomer that results in slight changes in the vacuum. As all downcomers are connected through a common air distribution manifold, this can result in slight, momentary, movements of slurry from the downcomer into the entrance of the air system. The Air Isolating Slurry Eliminating (AISE) valve prevents this occurrence and also automatically isolates the downcomer from the air distributor when the downcomer slurry flow is isolated (figure 4). The AISE valve is a non-return check valve using the concept of a rubber curtain closing against a flat seal. The design is optimised so that minimal impedance to air flow is created whilst ensuring a rapid response to any alteration in air flow direction. The AISE valve is installed just prior to the air's entry into the downcomer at a 45° angle to allow slurry to drain back into the downcomer (Murphy et al, 2000). The use of soft rubber materials also ensures that should particles of solids be present around the valve, an effective seal will still be achieved.

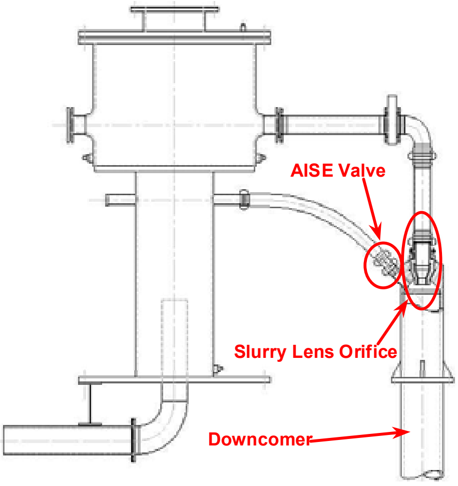


Figure 3. Schematic of Jameson cell Mark 3 Downcomer

Z-Cell

All the above mentioned features have been incorporated in the Z-Cell design, which integrates the flotation tank, feed sump and tailing recycle mechanism into a single unit that can be located on one level. Feed fluctuations are compensated internally in the cell by the hydraulic head difference between the feed and tailing boxes with no instrumentation being required. The Z-cell design is currently operating in several solvent extraction/electrowin circuits in Mexico and as a scalper in a gold operation in South Africa (Smith, 2005), see figure 5 below.

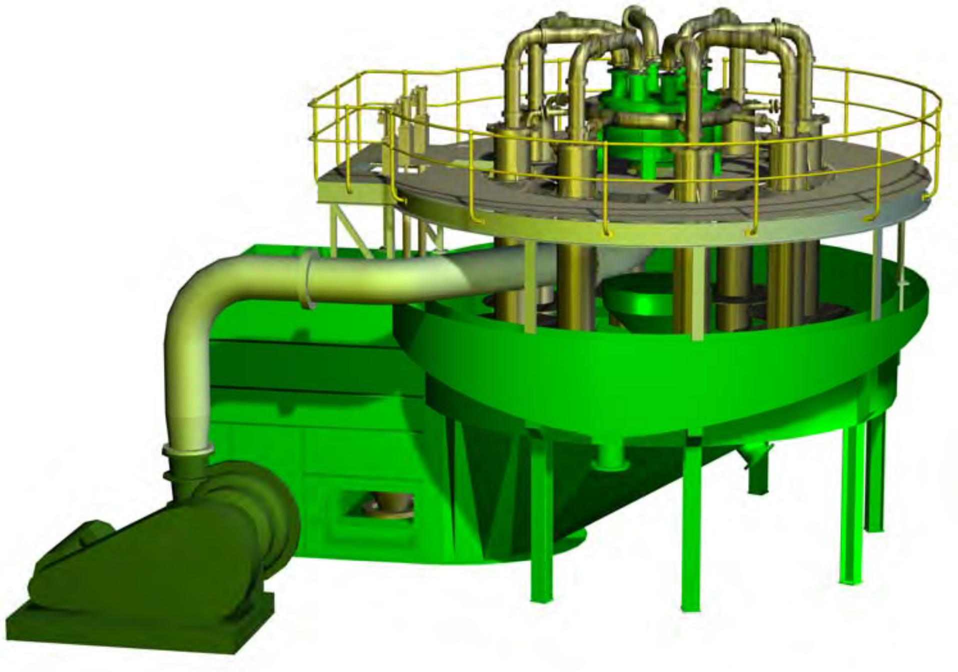


Figure 4. Schematic of Z Cell

DEVELOPMENT 2000-TO DATE

Air and Slurry Distribution

About the turn of the century design developments turned to the distribution methods of slurry and air from the single point entry to the Mark 3 downcomer. In conjunction with a leading cyclone manufacturer, Krebs, a design was developed for extruded radial slurry feed branches from the distributor to each downcomer. This modification reduced wear and also assisted in the applications of wear linings inside the slurry distributor. Also quick-release fittings were incorporated to ease any maintenance and ensure correct installation of the AISE valve.

Wash Water

For many flotation applications requiring a clean concentrate grade, wash water is an invaluable tool. Two distinct methods of wash water addition have been realised namely, above froth and in-froth. The latest design in wash water systems has been installed and operated at numerous coal preparation plants. It consists of stainless steel circular rings attached to a manual lifting system. Holes are drilled into the side of the rings to allow wash water to flow. The system can be easily located in three positions in the froth or completely above the froth. When determining the location of wash water addition (above-froth vs in-froth) the following items should be considered:

In-froth washing produces a drier concentrate, assisting in downstream filtration processes. Washing occurs closer to the froth-pulp interface allowing increased time for bubble drainage in the froth phase.

In-froth washing generally increases washing efficiency. The steady coalescence and drainage of bubbles in the froth phase leads to a wider size distribution of bubbles at the top of the froth. This can lead to channelling of water that is introduced above the froth, leading to inefficiency of froth washing. Introduction of wash water lower in the froth zone reduces this channelling.

Above froth washing results in more froth being exposed to wash water. This increased water in the froth phase acts as a lubricant to the froth, increasing mobility and decreasing bubble coalescence. These factors act together to increase froth recovery but sometimes at the cost of froth grade.

Above froth washing can lead to some froth breakage due to the impact shock of the water stream hitting the bubbles. This can lead to a decrease in froth recovery, particularly at high wash water flowrates (required for high concentrate grade operations).

Frothermiser

For decades aerosol addition of reagents has been a point of interest in flotation research although there has been little quantitative work performed until recently. Various people such as Wada et al (1968) and Flint et al (1988) published that aerosol addition of frother reduced bubble size and could lead to a reduction in frother consumption. More recently the Energy Technology division of the CSIRO compared the air and slurry phase addition methods for a pilot scale generic flotation column and a pilot scale Jameson cell (Ofori et al, 2003).

In December 2001 MIM Process Technology applied for a patent for the technology later to be marketed as the Frothermiser. The Frothermiser is an in-line device and adds aerosol frother, normally Methyl IsoButyl Carbinol, to the naturally aspirated air as it is drawn into the Jameson cell. Compressed air impacts on the liquid frother within an atomising nozzle and forms it into a mist, which is drawn into the air distributor and dispersed into the downcomers. There are two commercial installations of the Frothermiser in coal preparation plant in Australia, one in the Bowen Basin, the other in the Hunter Valley (Cowburn et al, 2005). At the time of writing no information was available on the Hunter Valley installation whereas an independent report on the Bowen Basin installation shows an increased ash on the tailing thickener underflow by 7% (Pokrajcic et al, 2004).

CASE STUDY 1- RIVERSIDE MINE

Goonyella Riverside is located 30km north of Moranbah township and 190km south west of the Hay Point port facilities. Riverside mine commenced operations in 1983 and has the capacity to produce approximately 4.0 million tonnes per annum of prime hard coking coal with Riverside coal being produced from the BHP Mitsui coal leases. This is predominantly a combination of Riverside Coke, produced from the Goonyella Lower

Seam, and Goonyella Coke, a 60/40 feed blend of Goonyella Middle Seam and Goonyella Lower Seam.

In 1989, Goonyella merged operationally with the adjoining Riverside mine, and the combined operation is known as Goonyella Riverside Mine. Following the amalgamation of the Goonyella and Riverside mines, coal sources from similar regions could be concurrently processed through both Goonyella and Riverside CPPs. Riverside underwent a significant upgrade in 1996. This upgrade saw the replacement of the conventional flotation circuit with six J5000/14 Jameson cells, arranged in single stage operation with tailings recycle, and the installation of two horizontal belt filters. These cells showed significant performance improvements over the incumbent Wemco mechanical cells and delivered yield increase in the order of 7%.

The Riverside plant processes coal utilising dense medium cyclones (DMCs) for coarse coal (nominally –50mm + 0.5mm) and Jameson Cell froth flotation for fine coal (-0.5mm w/w). The total CPP is configured as two largely independent half plants nominally treating 800tph per half plant. The half plants are further subdivided into three individual modules (i.e. six modules in total). Each module consists of a DMC circuit and a single J5000/14 Jameson Cell (Wex et al, 2004). Generally 20% to 25% of the total plant feed reports to flotation.

Within each module the fine coal feed reports to the Jameson Cell feed sump where diesel collector is added to increase hydrophobicity. The coal slurry, containing approximately 7 to 10% solids is pumped to the Jameson Cell, with MIBC as frother being added to the pump suction. The Jameson Cells are generally operated to maximise coal recovery to product whilst still maintaining a relatively low ash concentrate stream of between 5% and 6% ash.

The cells operate at feed pressures of 150kPa that equate to a jet velocity of 17.7 cm/sec. The cells operate with a recycle of between 30% and 40% of the total downcomer feed volume with the proportion of recycle being controlled by an actuated butterfly valve on an external recycle box. The level in the feed sump controls the valve position, with additional slurry overflowing the launder in the recycle box and gravitating to the tailing sump.

Reagent dosages are in the range of 8 to 14ppm of fresh feed for MIBC and 160-250 g/t diesel dependent on coal type (Wex et al, 2004). The cells have above froth washing operating at a wash water ratio of 0.8 to remove entrained gangue. Froth depths are typically run at 200mm although this may vary depending on filtration constraints as froth depth is seen to have a great affect on concentrate moisture, which has implications in the filtration circuit.

Recently further work has been progressing on the flotation cells:

In 2002 larger air intakes were fitted allowing significantly higher operating air-to-pulp ratios and also increased vacuums. This increased the superficial gas velocity (Jg) from 0.8 to around 1.5 whilst the vacuum pressure was able to be decreased from –5kPa to –10kPa. The cells now operate at air-to-pulp ratios of the order of 1.2, which has increased recovery of coal in all size fractions.

In late 2004 one of the Jameson cells was retrofitted with mark 3 downcomers. As mentioned previously the effective length of the downcomer has been increased so modifications to the length of the slurry feed pipe from the pump were made and the slurry distributor was raised (figure 6).

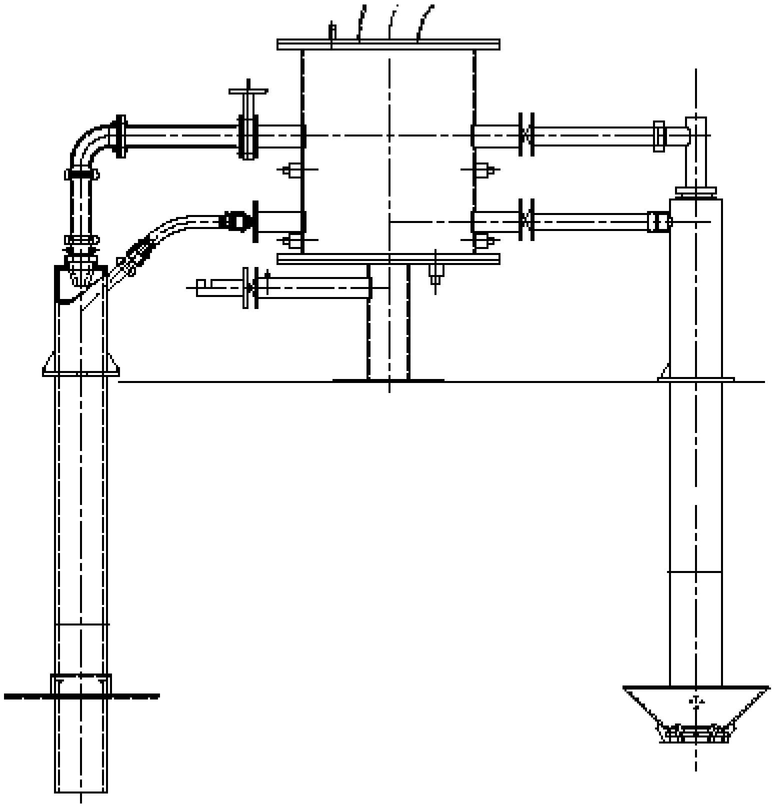


Figure 5. Schematic of Downcomer retrofit at Riverside

CASE STUDY 2- HAIL CREEK MINE

Introduction

Located 100 kilometres west of Mackay and 35 kilometres North-East of Nebo, Queensland, the Hail Creek operation produces coal from one of the world's largest coking coal deposits. Rio Tinto Coal Australia manages the operation on behalf of the joint venture interests - Rio Tinto Coal Australia (92%), Marubeni Coal (5.33%) and Sumisho Coal Development (2.67%). Hail Creek is recognised as a high quality, large-scale coal resource of some 1.2 billion tonnes with proven open cut mineable reserves in excess of 200 million tonnes. The Hail Creek coal mine has the capacity to produce 5.5 million tonnes of prime hard coking coal annually.

Plant Design

Initial testing was carried on large bore samples as this was a greenfield development. These tests confirmed that flotation would achieve targeted recovery at laboratory scale. Flowsheet development was carried out and the throughput determined from which it was decided that three cells, each six-metres in diameter with twenty downcomers would be required (Figure 7).

To simplify plant layout it was decided to use a single pump to feed the three cells. One drawback to this decision was that a non-standard impellor was required for the chosen duty. This was required to increase efficiency to bring the power draw down to match the 500 kW drive (Proud et al, 2004). During layout of the plant the sump design was changed to a tube style sump to decrease floor areas required in the plant. This led to a balance pipe configuration from the tailings deaeration sump to the flotation feed sump being used for recycle.

To counteract the migration of frother in plant water to the rest of the plant a split water system was developed. This allowed clarified water from the thickener overflow that is rich in frother to recirculate to the flotation circuit without mixing with other plant circuits.

Concentrate from the three Jameson cells was collected in open launders and flows to a distributor that feeds two horizontal belt filters. The launders were replaced with pipes when the operating level of the distributor increased above the top of the launders.

Commissioning and Optimisation

Commissioning coal for the plant was of lower quality than expected and this impacted on the initial set up of the cells. The coal was finer than expected and contained large amounts of shale. Consequently, cell operation was adjusted to handle this material and performance testing delayed until better quality coal was fed to the plant. Commissioning of the flotation feed pump was hampered by the fact the wrong impellor was supplied with the pump that caused continual overloads. When the problems were finally resolved, the required pressure was achieved at the feed distributor to the cells.

Cell optimisation was based upon achieving a concentrate that meets expected yield, although other factors such as ash content and percent solids were considered. Test work was not a reliable predictor of actual plant reagent dosage. This was due, in part, to the recirculation of frother in the plant water supply, another source of variation was conditioning time with collector. Table 2 below shows operating parameters for the three cells installed at Hail Creek.

Table 2. Operating Parameters for Hail Creek

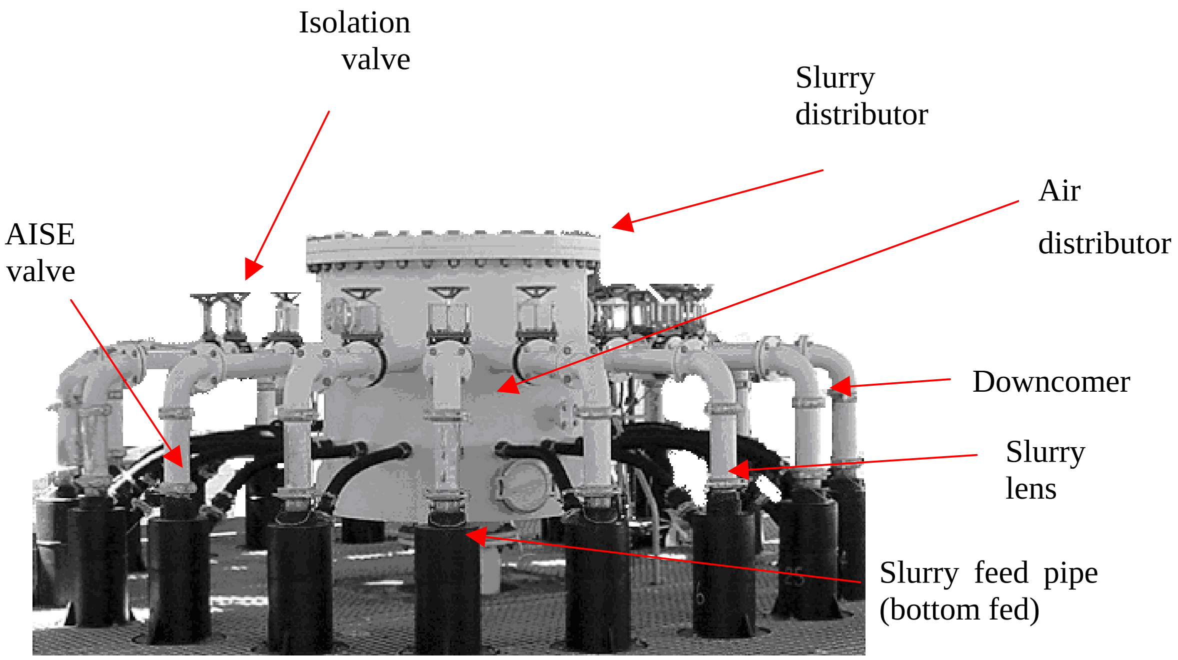
Parameter	Operating Value
Feed Pressure	150 kPa
Vacuum	-8 kPa
Air Flow	1100 m^3/hr
Froth Depth	300mm
Wash Water	78 m^3/hr
Frother (MIBC)	15.2 ppm
Collector (Diesel)	0.6 l/min

Varying levels of wash water was used to remove entrained ash from the froth, dependent on required concentrate ash. The cells at Hail Creek were designed with stainless steel concentric rings to achieve even distribution of wash water into the rising froth for submerged froth washing. Experience has shown the optimum location for the rings was just above the pulp to froth transition zone. Submerged froth washing gave good flow of clean water in the transition zone where bubbles were coalescing (Stone, 2004).

As clarified water was used as wash water it contains ultra fine particles and flocculant and had a tendency to settle in pipes that were low in velocity. In the wash water ring design, the flow around the rings had to be controlled to avoid high velocity jets from exiting the rings and disrupting the froth. Consequently, the rings contain areas of low velocity and some settling occurred. The problem was accentuated in parts by the presence of burrs from the drilling process in the rings. This settled material led to eventual blocking of large portions of the rings. The solution was to remove the burrs from the inside of the wash water rings and to provide flushing points. High pressure water was injected into the wash water manifold and opening flushing valves attached to the rings. This stream of high pressure water was sufficient to break up the settled material within about 30 minutes. The operators were instructed to repeat this operation as required (nominally weekly) with the measured flow rate to the wash water rings being used as the indicator for the timing of the flushing operation.

Water content in the froth was controlled by cell level or implicitly froth depth. A deeper froth depth will allow more time for drainage of the froth as well as removing some of the entrained high ash slimes.

As with all flotation this optimisation was not as simple as described above as there are other factors that will influence the operation of the cell. This required operators who were trained to respond to the needs of the flotation circuit. Additionally, cameras were installed in the plant allow a visual check of the operation of the cells at all times.



CONCLUSIONS

Jameson Cell development has improved performance in a number of specific areas. These include design changes to minimise wear, maximise aeration and optimise grade and recovery.

High density alumina has been incorporated into orifice plate manufacture increasing operational life. This was further enhanced with the development of the slurry lens, whose shallow entry profile has increased wear life significantly.

Modifying the feed arrangement to each downcomer, in addition to simplifying access to key components, has resulted in a 15% increase in downcomer residence time. Subsequently, for the same vacuum a higher volume of air can be induced into the downcomer.

Operation at optimal grade and recovery has been enhanced with the inclusion of components which optimise superficial gas velocity within the flotation tank.

ACKNOWLEDGMENTS

The authors would like to thank Xstrata Technology, Rio Tinto Coal Australia, Sedgman and BHP Mitsui and BMA for permission to publish this paper. The historical and technical knowledge and assistance of Greg Harbort was gratefully received.

REFERENCES

Cowburn, J, A, Pokrajcic, Z, Harbort, G, J, Manlapig, E, V, 2005. Improving coal flotation using a new method of frother addition, in Proceeding Centenary of Flotation 2005 Symposium -in press. (The Australasian Institute of Mining and Metallurgy: Melbourne).

Evans, G, M, Atkinson, B, W, Jameson G, J, 1995. The Jameson Cell. Flotation Sci. Eng, (Ed: K A Matis), pp331-363 (Marcel Dekker: New York).

Flint, I, M, MacPhail, P, Dobby, G, S, 1988. Aerosol frother addition in column flotation. CIM Bulletin, May, pp 81-84

Harbort, G, J, 1992. Jameson Cell development and applications. Presented at AMMTEC Flotation Colloquium, Perth (Unpublished).

Harbort, G, J, Jackson B, R, Manlapig, E V, 1994. Recent Advances in Jameson Flotation Technology. Minerals Engineering Vol 7. Nos 2/3 pp 319-332.

Harbort, G, J, 1997. Internal report for MIM Process Technologies.

Harbort, G J, Manlapig, E V, DeBono, S K, 2002. A discussion of particle collection within the Jameson Cell downcomer. Trans *IMM (Section C: Mineral Process. Extr. Metall*), 111 / *Proc. Australas. Inst. Min. Metall.*, 307, January/April, 2002, ppC1-C10.

Harbort, G J, Cowburn, J, A, Manlapig, E V, 2004. Recovery Interactions between the Froth Zone, Pulp Zone and Downcomer within a Jameson Cell. *Proceedings of Tenth Australian Coal Preparation Conference, ACPS, Pokolbin New South Wales,* 2004. Membrey, WB, ed. Paper C8.

Jameson, G, J, Harbort, G, Riches, N, 1991. The Development and Application of the Jameson Cell. *4th Mill Operators' Conference, Burnie, 10-14 March, AusIMM, Melbourne, pp45-50.*

Jameson, GJ, 1988. A new concept in flotation column design. Column '88 – *Proceedings of an International Symposium on Column Flotation, SME, Phoenix Arizona,* 1988. Sastry, KV, ed. pp 281-289.

Ofori, P, Firth, B, Taylor, A, 2003. Aerosol Frother Addition in Coal Flotation. *ACARP Report C11007.*

Pokrajcic, Z, Manlapig, E, S, V, Harbort, G, J, Smith, S, and Jameson, G, J, 2003. Reagent Supply apparatus for Supplying Reagent to Froth Flotation Cell e.g. Jameson Cell. *Provisional patent PO9354.*

Murphy, A, S, Honaker, R, Manlapig, E, S, V, Lee, D, J, Harbort, G, J. Breaking the Boundaries of Jameson Cell Capacity. *Proceedings of Eighth Australian Coal Preparation Conference, ACPS, Nelson Bay, New South Wales, 2000*. (Ed: Membrey, WB)

Proud, D, Lamb, R, Shooter, C, 2004. Design, Construction and Commissioning of the Hail Creek Coal Handling and Preparation Plant. *Proceedings of Tenth Australian Coal Preparation Conference, ACPS, Pokolbin New South Wales, 2004*. (Ed: Membrey, WB) Paper A1.

Smith, S, R, du Preez, P, Irons, J, Curry, D, 2005. New Frontiers for the Jameson Cell in Africa, in Proceeding Centenary of Flotation 2005 Symposium -in press. (The Australasian Institute of Mining and Metallurgy: Melbourne).

Taggart, A, F, 1945. Handbook of Mineral Dressing 12 (p120-130). (John Wiley and Sons, New York)

Wada, M, Ishchin, G, Kano, S, Nadatani, A, Suzuki, H, 1968. Experimental study on aerosol flotation. *8th Intl'n Mineral Processing Congress*. Leningrad. Paper D-8.

Wex, T, Hill, B, Meyers, A, D, Clark, L, 2004. Successes Achieved Through Incorporating Recent Developments in Jameson Cell Technology into an Existing Jameson Cell Circuit. *Proceedings of Tenth Australian Coal Preparation Conference, ACPS, Pokolbin New South Wales, 2004*. (Ed: Membrey, WB). Paper D11.

Xstrata Technology Website, www.xstratatech.com/current/t_jameson_cell_advantages _advantages

BENEFITS OF HIGH INTENSITY FLOTATION AT THE HEAD OF BASE METAL FLOTATION CIRCUITS

Z Pokrajcic[(1)], G J Harbort[(2)], V Lawson[(3)], L Reemeyer[(4)]
(1) Xstrata Technology
(2) Julius Kruttschnitt Mineral Research Centre
(3) Inco
(4) Zinifex

Greg Harbort,
Julius Kruttschnitt Mineral Research Center.
Isle Road Indoorooopilly. Brisbane QLD 4068.
Phone 07 3365 5888, Fax 07 3365 5999
mailto: g.harbort@jktech.com.au

ABSTRACT

Keywords; Jameson Cell, Prefloat, Zinc, Copper, Scalping

The Jameson Cell is a high intensity flotation device, which utilises induced air from atmosphere. It was developed jointly by Mount Isa Mines and Prof. GJ Jameson of the University of Newcastle in the 1980's. It is proven to generate fine bubbles, in the order of 300 to 500µm, in a high intensity, high shear and compact zone contained in the downcomer. This aerated mixture exits the downcomer into the pulp zone, which is the quiescent mineral and gangue seperation zone.

Preflotation is the process where some of the readily floatable gangue material is recovered directly to tailing with minimal valuable mineral contamination. This can be done in a reverse flotation stage where the valuable mineral is suppressed and the naturally hydrophobic gangue material is floated off. Eliminating a portion of the gangue before it enters the downstream circuit reduces required flotation capacity and improves ease of valuable mineral recovery. It is a very effective and low cost method of expanding flotation capacity and improving overall circuit performance.

Scalping is the process where a final grade concentrate is produced in the roughing circuit. The scalper concentrate typically consists of well liberated and fast floating minerals. Valuable mineral which is not recovered in scalping is treated in the remainder of the circuit where it is floated as a lower grade concentrate and upgraded and recovered in the cleaner circuit. The removal of the liberated mineral prior to a regrinding stage allows the regrind and cleaning circuits to be designed and operated more appropriately to the middling material.

A number of Australian base metal flotation circuits feature a reverse flotation stage at the head of the circuit. Test work and plant operating data has shown that the use of a Jameson Cell in the prefloat cleaner application has further improved prefloat gangue recovery and selectivity. Operation of a Jameson Cell in a carbonaceous/pyrite prefloat cleaner duty at the Mt Isa copper concentrator increased copper recovery and reduced pyrite in the copper concentrate. Testwork at Zinifex Century Zinc Mine showed a decrease in zinc losses by the utilisation of Jameson Cell prefloat cleaner. Appraisal of a Jameson Cell in a scalping role within the Mt Isa Copper Concentrated indicated significant benefits could be achieved.

INTRODUCTION

The Jameson Cell

The principles of Jameson Cell operation have been discussed by numerous authors including Jameson (1998), Jameson et al (1998) and Evans et al (1995). Recent developments have been reviewed by Harbort et al (2003) and Carr et al (2003). Operation can be described with reference to Figure 1.

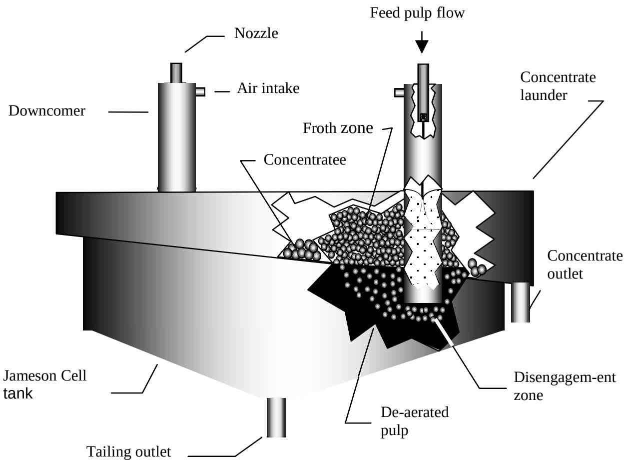


Figure 1. Jameson Cell Schematic

The operation of the Jameson Cell can be divided into three main zones:

The downcomer where primary contacting of bubbles and particles occurs. Feed slurry is pumped into the downcomer through an orifice plate, creating a high-pressure jet. The plunging jet of liquid shears and then entrains air, which has been naturally aspirated. Due to a high mixing velocity and a large interfacial area there is rapid contact and collection of particles.

The tank pulp zone where secondary contacting of bubbles and particles occurs and bubbles disengage from the pulp. The aerated mixture exits the downcomer and enters the pulp zone of the flotation tank. The velocity of the mixture and large density differential between it and the remainder of pulp in the tank results in recirculating fluid patterns, keeping particles in suspension without the need for mechanical agitation.

In the froth zone, gangue material that is entrained in the froth is removed by froth drainage and/or froth washing.

Applications at the head of base metal flotation circuits

Preflotation

Preflotation is the process where the bulk of the readily floatable gangue material is recovered directly to tailing with minimal valuable mineral contamination. Typically, the hydrophobic gangue material is floated and the entrained valuable mineral is removed by froth washing Eliminating a portion of the gangue before it enters the downstream circuit reduces required flotation capacity and improves ease of valuable mineral recovery. It is a very effective and low cost method of expanding flotation capacity and improving overall circuit performance.

Periodically amounts of naturally floatable species of carbonaceous pyrite and talc occur in ore bodies of the northwest Mining Province of Queensland. Their presence if untreated results in contamination of final concentrate. Talc in particular has detrimental down stream affects, causing an increase in smelter slag viscosity and higher metal losses to slag. Historically, treatment has consisted of flotation, followed by depression, often with ambiguous results Hoffman et al (1965) and Lyon et al (1971). Investigation by Grano et al (1990) identified the natural floatability of carbonaceous pyrite.

Also present in northwest mining area of Queensland is a host rock of carbonenous shale. A carbon preflotation stage is required to remove the organic carbon from downstream flotation stages.

Use of preflotation prior to roughing is increasingly considered a viable alternative to the flotation/depression route.

Scalping

Scalping involves producing a final grade concentrate in the roughing circuit. Although not a new concept in flotation, the high selectivity and rapid collection kinetics of the Jameson Cell enables the separation to be achieved with enhanced economics Gray et al (1998). Following scalping the valuable mineral remaining in the circuit is floated as a lower grade concentrate that is then upgraded the cleaner circuit. The removal of the liberated mineral prior to a regrinding stage allows the regrind and cleaning circuits to be designed and operated more appropriately to the middling material. This allows a greater efficiency of separation of composite particles. Minimising the quantity of regrinding decreases slimes generation and reduces the losses that inevitably result from their presence. Also, the shorter residence time of mineral in the flotation circuit means the less likely the oxidation of the mineral surface.

Case Study Sites

The Mount Isa Mines Copper Concentrator and the Zinifex Century Zinc Concentrator were used to evaluate the benefits of the Jameson Cell at the head of the flotation circuit. A brief description of the operations is given below.

Mount Isa Mines Copper Concentrator

Chalcopyrite ore and converter slag are treated to produce copper concentrate at the Mount Isa Mines Copper Concentrator. The concentrator was commissioned in 1973, replacing the original No. 1 lead-zinc concentrator, for the processing of all the chalcopyrite ore coming from Mount Isa mine Lumsdaine et al (1980).

At present production ore is sourced from the southern 1100, 3000 and 3500 orebodies. Currently 50% of mill feed is from the 3000 and 3500 orebodies. Chalcopyrite is the only significant copper mineral and occurs as a replacement deposit in a silica-dolomite host rock. Sulphide gangue consists of pyrite (FeS_2) and minor amounts of pyrrhotite and cobaltite. The ore averages 8% sulphur. The current flowsheet is shown in Figure 2.

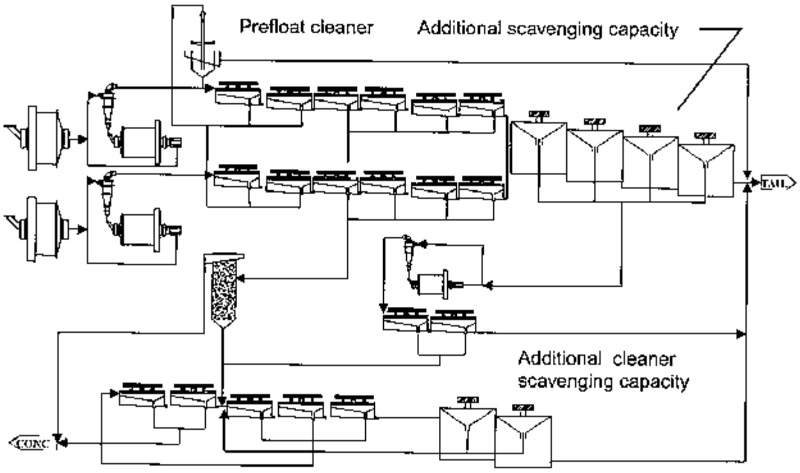


Figure 2. Mount Isa Mines Copper Concentrator Flowsheet

Future ore production will draw increasingly larger tonnages from the 3000 and 3500 orebodies. The host rock is very similar to that of the 1100 orebody but is slightly more siliceous in nature. Silica assays range from 60 to 70% SiO_2 and the ore is both abrasive and hard with a current Bond Work Index of approximately 22kWh/t. Copper mineralisation is disseminated through the silica-dolomite. Mineralogical examination indicates liberation of chalcopyrite is about 75% at a P_{80} of 150µm. Flotation feed sizing is normally in the range 80% passing 150µm, depending on milling rate and ore type.

Over time there has been an increase in the amount of naturally floatable species of carbonaceous pyrite and talc in ore. This would normally decrease concentrate grade for a given recovery; however at the same time the concentrator has been required to increase concentrate grade from 25% to 27% Cu. A preflotation circuit to remove naturally floating species has been necessary to achieve this, Carr et al (2003).

Zinifex Century Mine

The concentrator at Zinifex Century Mine produces a zinc and lead concentrate. The flotation circuit consists of carbon prefloat, lead flotation, zinc primary flotation, ultra fine milling, zinc ultra fine flotation Burgess et al (2003). Figure 3 shows the concentrator flowsheet at Century.

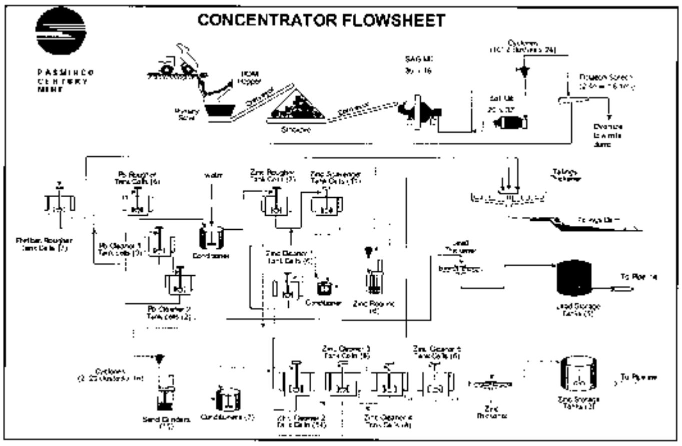


Figure 3. Zinifex Century Zinc Concentrator Flowsheet

Currently the carbon prefloat stage consists of a bank of roughers. The function of the carbon prefloat roughers is to remove a portion of carbon from the feed before it enters the remainder of the circuit, the aim being to prevent downstream contamination of concentrates and reduce the demand for flotation reagents. Prefloat rougher concentrate is pumped to final tailing. The prefloat rougher tailing gravitates to the lead circuit. Recently zinc losses to the throwaway prefloat concentrate has been identified as an area for improvement. Research into the preflotation of circuit has shown that zinc losses are in the form of fine entrained sphalerite that is carried with water into the froth.

Discussion

Prefloat duty at Mount Isa Mines

Work on a pilot scale Jameson Cell for one stage preflotation and for preflotation cleaning had been performed over a number of years. A series of pilot tests were conducted in 2000 to confirm parameters for engineering design. A full scale Jameson Cell was installed in 2002 in the copper concentrator to upgrade the rougher prefloat concentrate and minimise copper losses from the circuit Carr et al (2003).

The optimum Jameson Cell prefloat cleaning operating conditions were found to be:

- Wash water ratio between and 0.5 and 1.0
- Feed percent solids 20%
- Jg - 1.25cm/sec
- Air-to-pulp ratio between 0.44 and 0.82
- Froth depth – 150mm to 400mm

Figure 4 shows total talc recovery versus copper rejection from concentrate for the commissioning period, the original test work and for bench scale tests taken during commissioning, respectively. Target performance was to achieve a copper rejection of 90%, at a talc recovery of 70%. The upper envelope for commissioning surveys indicates a minor shortfall from this point, with an interpolated copper rejection of 90% at a talc recovery of approximately 67%. At 90% copper rejection talc recoveries of 50% and 47% were achieved in the bench scale tests and original test work respectively.

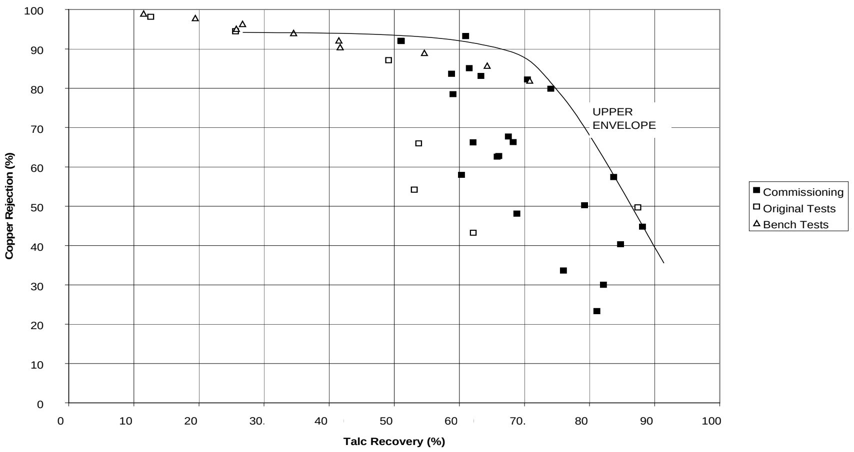


Figure 4. Talc Recovery verses Copper Rejection at Mount Isa Mines Copper Concentrator

Figure 5 shows total FeS_2 recovery versus copper rejection from concentrate for the commissioning period, the original test work and for bench scale tests taken during commissioning, respectively. Targeted performance was to achieve a copper rejection of 90%, at an FeS_2 recovery of 50%. The commissioning surveys indicate a shortfall from this point, with a copper rejection of 90% at an FeS_2 recovery of only 17%. At 90% copper rejection FeS_2 recoveries of 30% and 28% were achieved in the bench scale tests and original tests respectively. The commissioning results show a strong linear relationship between copper rejection and FeS_2 recovery indicating some degree of carbonaceous chalcopyrite being present and also FeS_2 recovery gains being due to entrainment rather than true flotation. There is some initial evidence that carbonaceous pyrite that did not float within the Jameson Cell has not been refloated in the prefloat roughers and does not report to final concentrate.

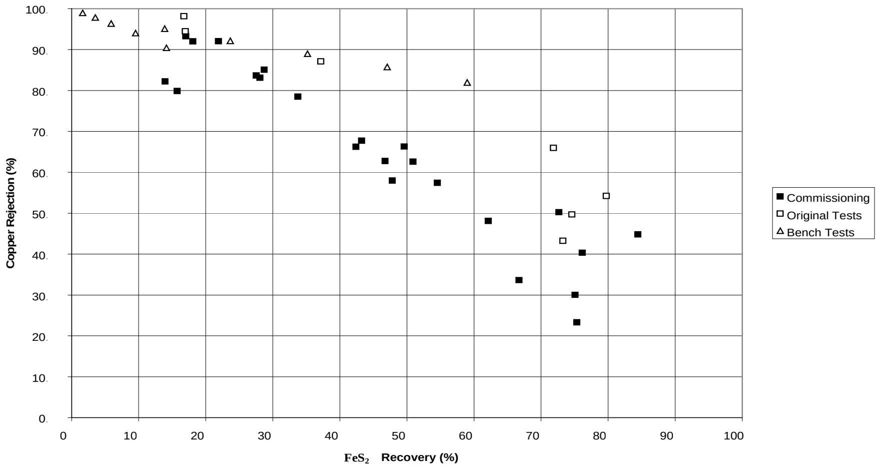


Figure 5. Pyrite Recovery verses Copper Rejection at Mine Isa Mines Copper Concentrator

A review of earlier analysis by Grano et al (1990) indicates that significant amounts of pre-aeration in bench tests results in a decrease in carbonaceous recovery and in an increase in chalcopyrite recovery. This same phenomenon is seen to be occurring within the Jameson Cell, although with the high aeration and turbulence within the downcomer it is happening in a much shorter time.

The most obvious explanation for the lack of pyrite recovery in the prefloat cleaner is surface passivation by iron hydroxides. Should this be occurring it is possible that the pyrite will not be refloated in the remainder of the circuit. A review of pyrite in feed versus pyrite in concentrate, Figure 6, indicates that even with the lower Jameson Cell pyrite recovery the amount of pyrite reporting to final concentrate has decreased following the Jameson Cell installation. This suggests that pyrite did not float in the remainder of the circuit either.

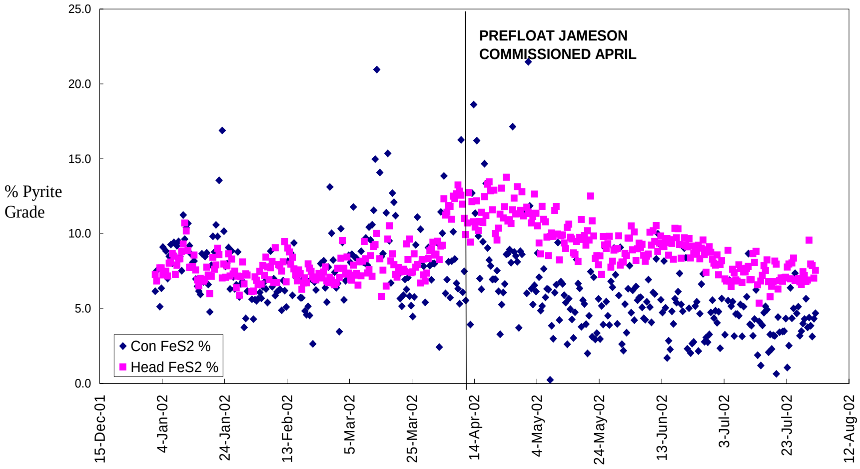


Figure 6. Pyrite Levels in Feed and Concentrate pre and post Jameson Cell installation

Preflotation duty at Zinifex Century Zinc Mine

In October 2002 a L500 Jameson Cell pilot plant rig was installed in the concentrator at Zinifex Century Zinc Mine, Pokrajcic (2003). The aim of the test work was to minimise zinc losses to carbon prefloat concentrate while maximising carbon removal by using the Jameson Cell as a prefloat cleaner. The majority of zinc losses are in the form of fine sphalerite entrained in the preflotation rougher concentrate.

Periodic test work conducted from October 2002 to January 2003, concentrated on appraising the Jameson Cell in a carbon prefloat cleaning duty.

Optimum Jameson Cell operating conditions in a prefloat cleaner duty were found to be:

- Feed percent solids between 7% and 8.5%
- Tailing Recycle at 55%
- Air to pulp ratio (APR) between 0.58 and 0.85
- Superficial air velocity (Jg) between 0.65 and 0.9 cm/s
- Jet Velocity of 13.0 m/s
- Zero wash water

When optimised the Jameson Cell in a prefloat cleaning duty reduced the overall zinc recovery to final tailing from approximately 4.5% to less than 2.0% while maintaining a carbon recovery to final tailing close to 10%. Figure 7 shows all the data generated in the test program, highlighted is the data showing best performance.

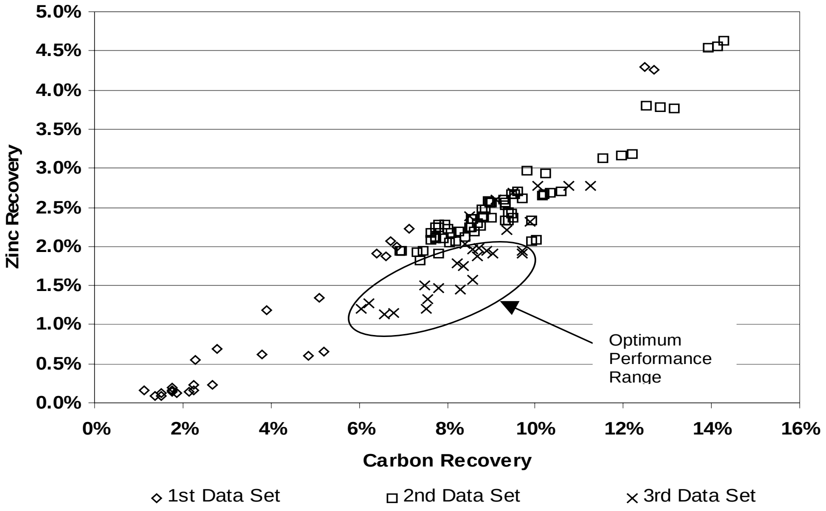


Figure 7. Overall Carbon and Zinc Recovery for the carbon Prefloat Circuit when using a Jameson Cell as prefloat cleaner.

Reducing the feed density from approximately 18% solids about 8% solids and increasing the tailing recycle had the most significant effect on selectivity without compromising rejection. This is due to the effect of reducing entrained sphalerite losses. Hence a significant reduction in zinc losses to the throwaway concentrate can be achieved with a small reduction in carbon recovery by the addition of a Jameson Cell prefloat cleaner.

Scalper duty at Mount Isa Mines

In February 2000 an L500 Jameson Cell pilot plant rig was installed in the copper concentrator at Mount Isa Mines to assess the Jameson Cell performance in prefloat and slag cleaning duties. Results from the slag cleaning work on converter slag produced high mass recovery per flotation surface area and indicated that some potential existed for the Jameson Cell in a chalcopyrite roughing role.

In July 2000 a series of sighter tests which were not optimised were conducted on rougher feed material to determine if similar production rates were achievable with a chalcopyrite ore feed. The chalcopyrite tests were extended to produce a comprehensive Jameson Cell grade recovery curve. This work showed that the potential did exist to operate the Jameson Cell as a scalper, producing final concentrate from rougher feed. In late 2002 the pilot unit was continuously operated at optimised conditions to check stability of operation and reproducibility of results under variable feed conditions. A second Jameson Cell was installed in series as a scavenger to determine whether scalping would have a detrimental effect at the higher recovery section of the grade recovery curve.

Optimum operating conditions for the Jameson Cell in a scapler duty were found to be:

- Wash water ratio between and 0.5 and 1.5
- Feed percent solids between 35% and 45%
- Jg between 0.7 and 0.8cm/sec
- Air-to-pulp ratio between 0.32 and 0.38
- Froth depth – 150mm

An interesting point to note is that the residence time in the Jameson is approximately 1 minute.

An overview of the scalping grade recovery curve is shown in Figure 8. These results include all tests undertaken, including non-optimised tests. Operation of the Jameson Cell in a chalcopyrite scalping role produced a concentrate grade of 29.6%Cu at up to 80% copper recovery.

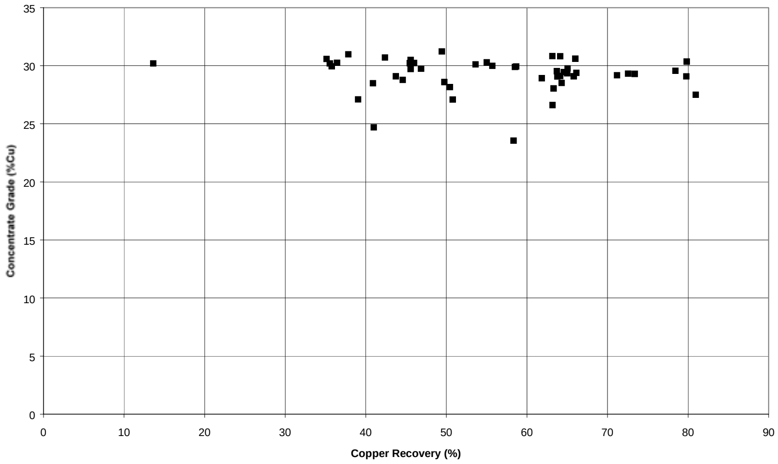


Figure 8. Scalper Testwork at Mount Isa Mines Copper Recovery verses Copper Grade

For the reproducibility tests a copper recovery of 60% with a concentrate grade of 30%Cu was considered an acceptable target. The reproducibility tests achieved an average copper recovery of 63.37%, with an average concentrate grade of 29.4%Cu. Eighty percent of tests achieved results better than the targeted recovery although at a slightly lower concentrate grade.

High mass recovery is another benefit of the Jameson Cell in a scalper duty. At 800tph at a feed grade of 3.5%Cu as mentioned previously the Jameson Cell produced an average copper recovery of 63.37%, with an average concentrate grade of 29.4%Cu, meaning 56tph of final concentrate is recovered in one stage of flotation using a Jameson Cell.

Conclusion

Preflotation is the process where the bulk of the readily floatable gangue material is recovered directly to tailing with minimal valuable mineral contamination.

Scalping is the process where a final grade concentrate is produced in the roughing circuit.

Both of these steps exploit the fast kinetics and well liberated properties of the gangue or valuable mineral. The Jameson Cell is particularly suited to such a duties because of it high intensity mixing of bubbles and slurry and relatively low tank residence time.

The addition of a prefloat and/or scalping stage is a very effective and low cost method of expanding flotation capacity and improving overall flotation circuit performance. For instance:

1. Eliminating a portion of the gangue before it enters the remainder of the downstream circuit reduces required flotation capacity.

2. Removal of liberated mineral prior to regrind allows the regrind and cleaning circuits to be designed and operated more appropriately for the middling material.

3. Reducing the amount of material going to regrind minimises slimes generation and the losses that result from their presence. It also decreases the mineral residence time in the flotation circuit, thereby decreasing the likelihood of mineral oxidation.

Test work and plant operating data has shown that the use of a Jameson Cell in a prefloat and/or scalper duty has improved the performance and efficiency of the flotation circuit. At Mount Isa Mines a Jameson Cell in a prefloat cleaner application increased copper recovery and reduced pyrite in the copper concentrate. Test work at Zinifex Century Zinc Mine showed a decrease in zinc losses by the application of a Jameson Cell as a prefloat cleaner.

Acknowledgments

The authors would like to thank Xstrata Technology, Xstrata Copper and Zinifex Century Mine for their kind permission to publish this paper.

References

Carr, D, Harbort, G, Lawson, V, 2003. Expansion of the Mount Isa Mines Copper Concentrator Phase One Cleaner Circuit Expansion, *The AusIMM 8th Mill Operators' Conference.*

Carr, D, DeBono, S, Harbort, G, J, Lawson, V, 2003. Jameson Cell fundamentals: a revised perspective. Presented at *Flotation 03*.

Evans, G M, Atkinson, B W, Jameson G J, 1995. The Jameson Cell. *Flotation Sci. Eng*, page 331-363

Grano, S, Ralston, J, Smart, R, St, C, 1990. Influence of electrochemical environment on the flotation behaviour of Mt. Isa copper and lead-zinc ore. *International Journal of Mineral Processing*, 30 page 67-97

Gray, M, P, Harbort, G, J, Murphy, A, S, 1998. Flotation circuit design utilising the Jameson Cell. *Proceedings of Metallurgical Plant Design and Operating Strategies*.

Harbort, G ,J, Manlapig, E, V, DeBono, S,K, 2003. A discussion of particle collection within the Jameson Cell downcomer. *Trans IMM (Section C: Mineral Process. Extr. Metall*), 111 / *Proc. Australas. Inst. Min. Metall*., 307, page C1-C10

Hoffmann, C, W, 1965. A review of developments in flotation treatment of Mount Isa chalcopyrite ores. *Proceedings of 8th Commonwealth Mining and Metallurgy Congress Australasia*, Paper no 103, pp 1133-1149

Jameson, G,J, 1988. A new concept in flotation column design. Column '88 –
Proceedings of an International Symposium on Column Flotation, SME, Phoenix Az, Sastry, K V, ed. page 281-289

Jameson, G, J, Belk, M, Johnson, N, W, Espinosa-Gomez, R, Andreaditis, J P, 1988. Mineral flotation in a high intensity column. *Chemeca 88, 16th Australian Conference on Chemical Engineering*. Sydney, page 507-510

Lumsdaine, I, O'Hare, S,M, 1980. Copper concentrator practice at Mount Isa Mines Limited, Mount Isa, Qld. *Mining and Metallurgical Practices in Australasia* (Ed:JT Woodcock), page 649-653, (AusIMM: Melbourne)

Lyon, G, Fewings, J, 1971. Control of the flotation rate of the carbonaceous material in Mount Isa chalcopyrite ore. *The AusIMM Proceedings*, No 237

Pokrajcic, Z, 2003, Report for pilot plant Jameson Cell flotation at Century Zinc Mine for carbon prefloat cleaning. Xstrata Technology report – unpublished.

RECOVERY INTERACTIONS BETWEEN THE FROTH ZONE, PULP ZONE AND DOWNCOMER WITHIN A JAMESON CELL

G Harbort(1), J Cowburn(2), E Manalpig(3)

(1) University of Queensland

(2) Xstrata Technology

(3) Julius Kruttschnitt Mineral Research Centre

ABSTRACT

The work reported in this paper shows that the primary area of coal recovery within the Jameson Cell is the downcomer, where the air and pulp are dispersed into a dense foam of fine bubbles. This creates an intense collection zone for bubble-particle collision and coal collection.

Historically, coal recovery in the Jameson Cell has not differentiated between recovery in the downcomer, pulp zone or the froth zone. As such, attempts to optimise the recovery of coal in one zone can result in non-optimum performance of another. A major flotation programme has been undertaken to measure coal recovery in the three zones of the Jameson Cell, separate from each other. The findings from this work should allow, upon completion, recovery in each zone to be individually optimised to give maximum overall coal recovery, from a size by size perspective.

INTRODUCTION

The Jameson Cell was a joint development between Mount Isa Mines and Prof. Graeme Jameson of the University of Newcastle (Jameson, 1988). Since its invention in 1986 there have been 94 Jameson Cells installed in the coal industry, both in Australia and overseas.

Jameson Cell operation

The principles of Jameson Cell operation have been discussed by numerous authors, including Jameson et al (1988), Evans et al (1995) and recently by Harbort et al (2002). The Jameson Cell can be divided into three main zones, as described with reference to Fig. 1.

1. The downcomer is where primary contacting of bubbles and particles occurs. Feed pulp is pumped into the downcomer through an orifice plate, creating a high-pressure jet. The plunging jet of liquid shears and then entrains air, which has been naturally aspirated. Due to a high mixing velocity and a large interfacial area there is rapid contact and collection of particles.
2. The tank pulp zone is where secondary contacting of bubbles and particles occurs and bubbles disengage from the pulp. The aerated mixture exits the downcomer and enters the pulp zone of the flotation tank. The velocity of the mixture and large differential between it and the remainder of the pulp in the tank results in recirculating fluid patterns, keeping particles in suspension without the need for mechanical agitation.
3. The froth zone is where entrained materials are removed from the froth by froth drainage and/or froth washing.

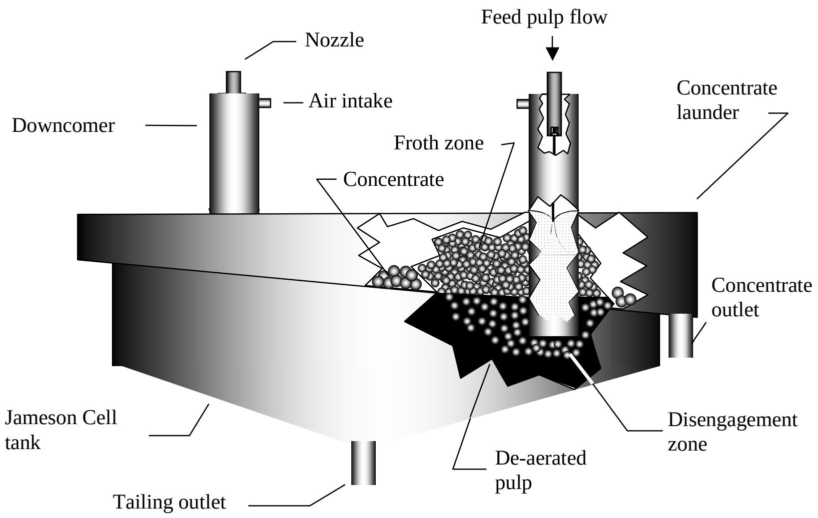


Figure 1.
Jameson Cell operation

Recovery interactions

Although a number of studies have been conducted on the effect of operating variables on the Jameson Cell (eg, Mohanty and Honaker, 1999) they have reported total Jameson Cell recovery, rather than the recovery in the three specific zones of the Jameson Cell.

The total recovery in a Jameson Cell is a function of the recovery gain in the downcomer, recovery loss or gain in the pulp zone and recovery losses in the froth zone.

Froth zone recovery

It is generally recognised (Vera, 1999) that recovery within the froth zone of any flotation machine is a function of the froth zone residence time τ, which in turn is determined by the aeration rate, Q_a, concentrate pulp flow rate, Q_c, the cell cross sectional area, A, and the froth depth, h. As such,

$$R_f = f.\tau$$
$$= f.A.h/(Q_a + Q_c)$$

Where f is a frothability factor effected by reagents and particle size.

Pulp zone recovery

Flotation equipment such as mechanical flotation cells and flotation columns are commonly designed to provide even dispersion of bubbles within the pulp zone of the tank. This dispersion results in pulp zone recovery becoming primarily a function of the residence time any one particle has in the pulp zone. In terms of operation within the Jameson Cell, tank void fraction measurements show that bubble patterns in general form a central, air swept cone surrounding each downcomer (Harbort et al, 2003). The Jameson Cell tank contains areas of high, localised air void throughout the pulp zone. The rising swarm of bubbles is governed by a number of factors including recirculating patterns within the tank, pulp flow volumes and air flow volumes, all of which affect the pulp zone recovery.

Downcomer recovery

The recovery that occurs within the downcomer is an area that is still under active investigation. Downcomer recovery is thought to be governed by a number of factors including the air-to-pulp ratio, turbulence, residence time and the amount of the mixing zone contained within the pipe.

Experimental procedure and equipment

The study was conducted using the Jameson Cell continuous recycle procedure as developed in 1992 by Cheng and associates (Manlapig et al, 1993). Two versions of this procedure are used by industry, these being the Simple Test, which approximates AS4156.2.1and the Coal Characterisation Test, which approximates AS4156.2.2. This test work used the Simple Test.

An experimental rig was used at the University of Queensland, which included:

- a 150mm diameter Jameson Cell, with a 25mm I.D. downcomer, fitted with a 3.8mm orifice plate
- a 100 litre capacity sump with stirrer
- a variable speed pump
- a Magnahelic flow meter measuring the Jameson Cell feed stream
- a pressure gauge on the feed line
- an air flow rotameter and a vacuum gauge on the Jameson Cell air line

As supplied coal contained material up to five millimetres in size. To prevent orifice plate blockage, coal greater than one millimetre was screened from the sample. The size distribution of the flotation feed coal is shown in Table 1, together with size fraction ash. At an overall ash of 15.6% the sample was considered a relatively clean flotation feed.

Table 1.
The size distribution and ash content in size fractions for the flotation feed.

Size range (µm)	Retained (%)	Cum. Retained (%)	Fractional Ash (%)	Cum. Ash (%)
-63 + 0 micron	52.01	52.01	18.84	18.84
+63 - 125 micron	13.30	65.31	11.98	17.44
+125 - 250 micron	10.33	75.64	12.55	16.77
+250 - 500 micron	12.47	88.11	11.19	15.98
+500 - 1000 micron	11.88	100.00	12.43	15.56

The sump was filled with 40 litres of water and approximately one kilogram of fine coal, to achieve a percent solids of 2.5%. The low percent solids was specifically chosen to minimise ash and coal entrainment and also to prevent any distortion of results through carrying capacity limitations. A diesel addition equivalent to 160g/t was added and conditioned for five minutes. An MIBC addition equivalent to 16ppm frother volume to fresh feed volume was added and conditioned for a further five minutes. No further reagents were added during the tests. As the Bowen Basin site, which supplied the feed sample, did not use wash water in its flotation operations no wash water was used in these trials.

Slurry was then pumped to the Jameson Cell at a rate of 14.1*l*pm. All tailing was continuously recycled back to the feed sump and pumped back to the Jameson Cell. The concentrate was collected at one, three, seven and fifteen minute intervals and then dried and analysed. In total 27 tests were conducted. The cumulative ash versus cumulative recovery curve for all tests is shown in Figure 2. This shows a sharp increase in combustibles recovery to 90%, with only a marginal increase in ash in concentrate. The maximum combustibles recovery achieved was 93% at a concentrate ash of 5.3%. The close grouping of results along the curve indicates that samples used for the varying tests was representative and exhibited similar flotation kinetic rates. All tailing and concentrate samples were sized at 63µm, 125µm, 250µm and 500µm. Ash and combustibles recoveries for the size fractions were then determined.

This method of test evaluation produces a series of curves of cumulative flotation time versus cumulative combustibles recovery. Although it provides a test of high reproducibility it does not directly equate to the performance of downcomers in production Jameson Cells. Production Jameson Cells operate with continuous new feed, where only a portion of tailing is recycled back to the feed sump. To determine how the continuous recycle test results equated to operation of production sized Jameson Cells a single pass test without recycle was conducted to a allow a recycle factor, f_r to be calculated, where

$$f_r = t_{st1} /(V_s/Q)$$

t_{st1} is the time required in the continuous recycle test for the cumulative combustibles recovery to equal the combustibles recovery achieved in the single stage flotation test, V_s is the sample volume treated in the single stage operation and Q is the volumetric flow through the orifice plate.

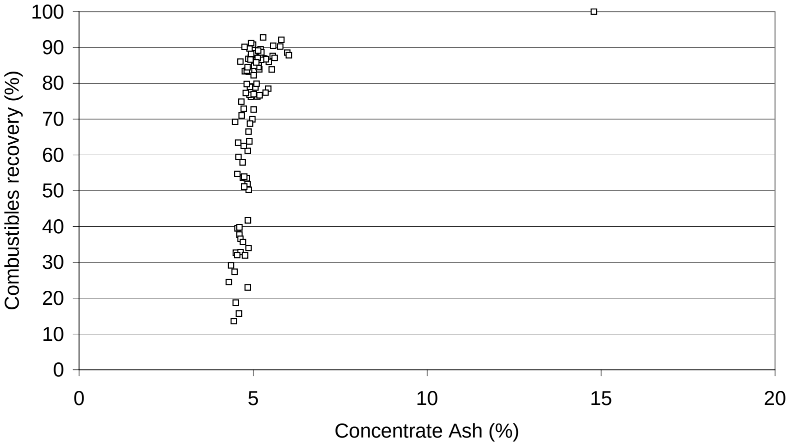


Figure 2.
Cumulative ash versus cumulative combustibles recovery.

Deleted: **Must give basis for ash results and make axis and legend**

For example single stage operation during this test program achieved a combustible recovery of 71%. With reference to Figure 3 this equates to 4.4 minutes of continuous recycle operation. With a volume of 40 litres and a flow of 11lpm, fr = 1.2. An fr < 1.0 would indicate material is short circuiting and potentially being preferentially floated while an $f_r = 1.0$ represents a uniform flow distribution and $f_r > 1.0$ represents a non uniform flow distribution. The f_r was calculated for different size fractions was 1.2 +/- 0.05, indicating various coal sizes were floated in a similar manner.

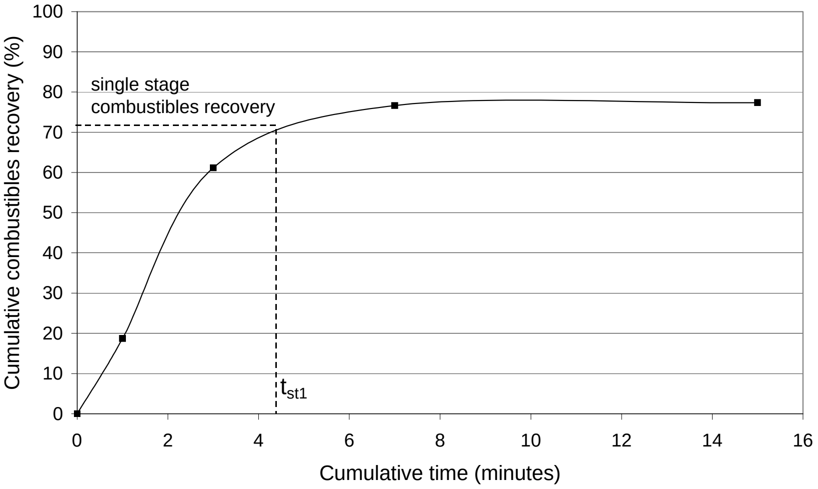


Figure 3.
Correlation between single stage and continuous tailing recycle.

Froth zone recovery

To determine the effect of the froth zone on combustibles recovery the depth of the froth zone was varied while other variables were held constant. The experimental conditions are shown in Table 2.

Table 2

Variables for froth zone recovery evaluation

Test no	Air-to-pulp ratio	Froth height (mm)	Froth residence time (min)
5	1.06	50	0.35
6	1.06	100	0.70
7	1.06	200	1.40
8	1.06	400	2.80

Pulp zone recovery

To determine the effect of the pulp zone recovery the air void fraction within the tank was varied. The air void fraction in the tank, ε was calculated by measuring the tank pulp volume without air addition, $V_{t(0)}$ and the tank aerated pulp volume immediately upon commencing air flow into the tank, $V_{t(1)}$, where:

$$\varepsilon = (V_{t(1)} - V_{t(0)})/ V_{t(0)}$$

The experimental conditions are shown in Table 3.

Table 3
Variables for pulp zone recovery evaluation

Test	Air-to-pulp ratio	Froth depth (mm)	Tank air void fraction (%)
16	0.90	100	2.59
15	0.90	100	5.17
14	0.90	100	7.76
13	0.90	100	10.34

Downcomer recovery

To determine the effect of the downcomer on combustibles recovery the air-to-pulp ratio and by association the downcomer vacuum were varied while other variables were held constant. The experimental conditions are shown in Table 4.

Table 4
Variables for downcomer recovery evaluation

Test no	Air-to-pulp ratio	Vacuum (kPa)	Froth Residence time (min)
4	0.22	18.5	0.70
3	0.45	15.5	0.70
2	0.90	10.5	0.70
24	1.06	3.5	0.70
23	1.06	0.75	0.70
26	1.06	0.5	0.70

The downcomer recovery was back calculated from the overall recovery, froth recovery and pulp zone recovery.

Results and discussion

Froth zone recovery

Figure 4 details the effect of varying the froth residence time on the recovery of the various size fractions. Linear lines of best fit have been inserted, which for size fractions above 63µm showed an r^2 correlation of 0.99. The slope of each line represents the rate of coal loss per unit time, with the Y axis intercept being the combined recovery of the downcomer and pulp zone. It can clearly be seen that loss in recovery in the froth zone for all size fractions is linearly proportional to the amount of time material spends within the froth. Typical Jameson Cells operating in coal flotation will operate with a froth residence time of approximately 1.5 minutes (Honaker et al, 1995). With reference to Figure 5, the rate of recovery loss versus average particle size it can be seen that particles less than 63µm in size are largely unaffected by froth residence time. Above 63µm the rate of coal loss increases dramatically, but then plateaus for particle sizes between 125µm and 1000µm.

The low recoveries at a residence time of zero for -63µm and 500µm to 1000µm particles are therefore an indication of lower pulp zone and downcomer zone recoveries for these size fractions.

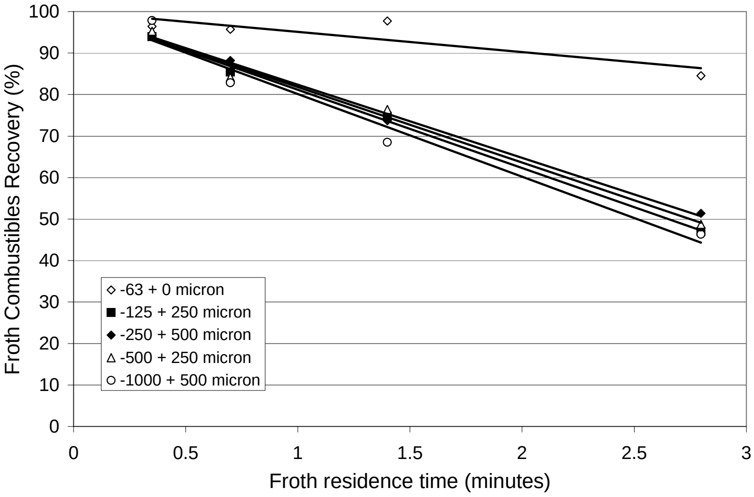


Figure 4
The effect of froth residence time on recovery in the froth zone

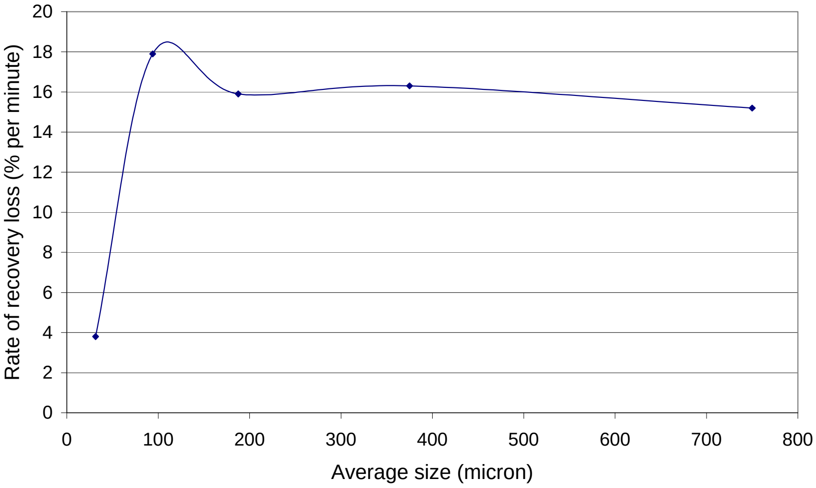


Figure 5
The effect of particle size on the recovery rate within the froth zone

Downcomer recovery

The effect of the downcomer air-to-pulp ratio on recovery can be described by with reference to the air-to-pulp ratio versus vacuum curve as shown in Figure 6. Below an air-to-pulp ratio of 0.2 the mixing zone within the downcomer is minor and the downcomer operates in what is essentially a bubbly flow regime. As the air-to-pulp is increased and vacuum decreases a distinct mixing zone is generated with high turbulence, followed by a pipe flow regime. A situation is eventually reached where attempts to increase the air-to-pulp ratio fail to entrain more air and only result in a decrease in vacuum. In this region the turbulent mixing zone dominates. For this paper evaluation of the downcomer will concentrate on the latter two areas.

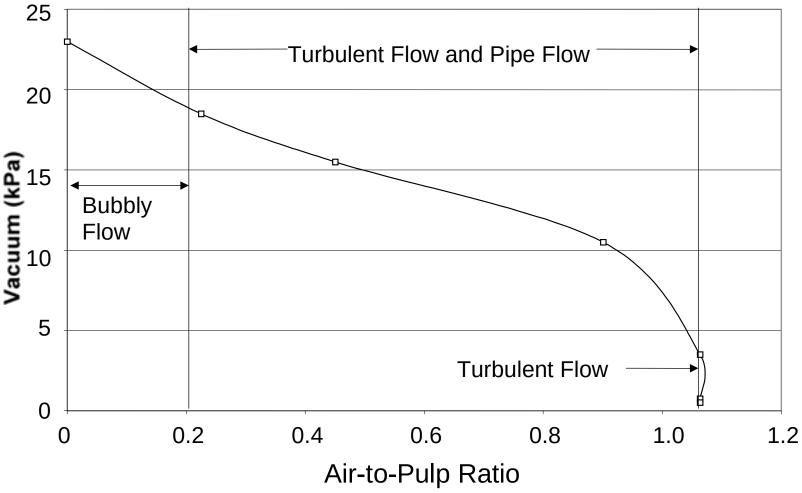


Figure 6
Flow regimes within the Jameson Cell downcomer

Figure 7 shows the change in downcomer recovery as the air-to-pulp ratio is increased from 0.22 to 1.06, but prior to the transition point where the vacuum falls suddenly. In this area the change in recovery per air-to-pulp ratio is approximately linear. Figure 8 details the average rate of recovery per change in air-to-pulp ratio over this range.

For the finer particle sizes, minus 125µm increasing the air-to-pulp ratio has a major affect on increasing recovery. Above this particle size the air-to-pulp ratio has a decreasing affect on recovery improvement, until the plus 500µm fraction where higher air-to-pulp ratios may in fact be causing combustible recoveries to fall marginally.

As one moves to the turbulent regime area of downcomer operation at the maximum air rate and low vacuums the rate of recovery is no longer linearly proportional to the air rate. To determine downcomer performance recovery is compared to the vacuum, Figure 9. For coal size fractions between -63µm to 500µm operation at low vacuums results in a loss of combustibles recovery. This may be due to the mixing zone no longer being contained within the downcomer and insufficient downcomer residence time for collection. The vacuum at which the drop in recovery occurs is dependent on the particle size, varying from 3.5kPa at –63µm to 0.75kPa at 500µm. It is interesting to note however that operation at a vacuum between 0.75kPa and 0.5kPa, or near the region of maximum turbulence, a substantial increase in combustibles recovery for plus 500µm coal particles occurs.

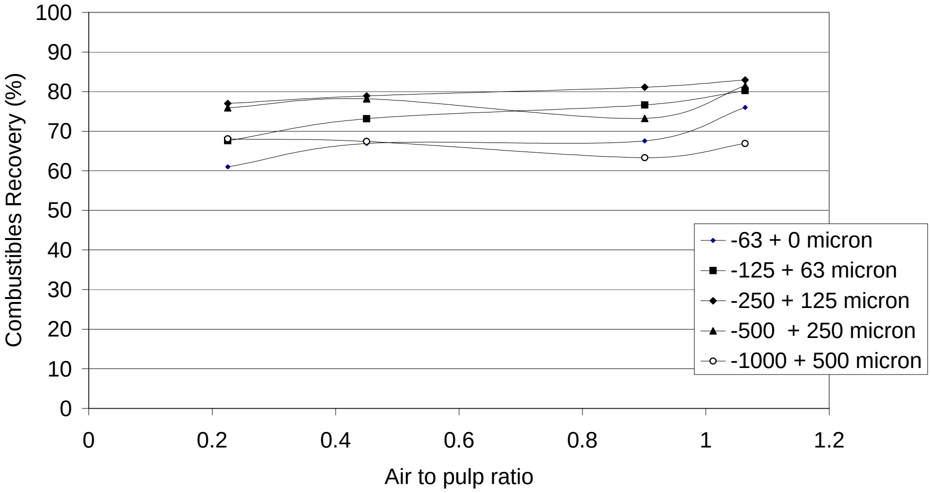


Figure 7

The effect of the downcomer air-to-pulp ratio on the downcomer combustibles recovery

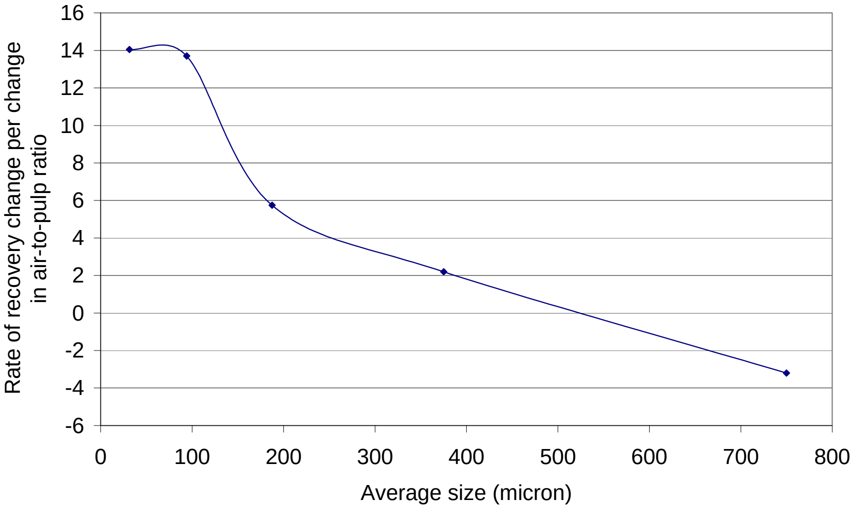


Figure 8

The effect of particle size on the rate of recovery within the downcomer

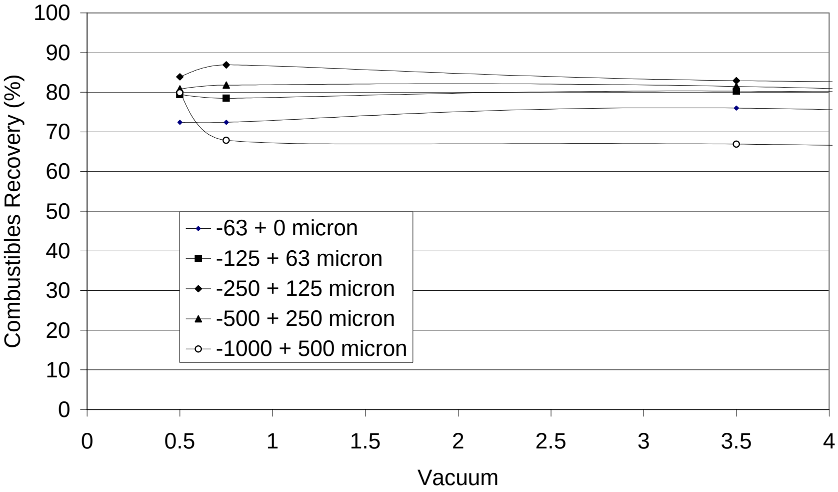


Figure 9
The effect of reducing vacuum on the downcomer combustibles recovery when operating at the maximum air-to-pulp ratio

Pulp zone recovery

It has generally been accepted that the pulp zone of a Jameson Cell does little more than maintain recovery generated within the downcomer. As evidenced in Figure 10 the pulp zone recovery is substantially less than the downcomer recovery and the downcomer must be considered the driving force for recovery within the Jameson Cell.

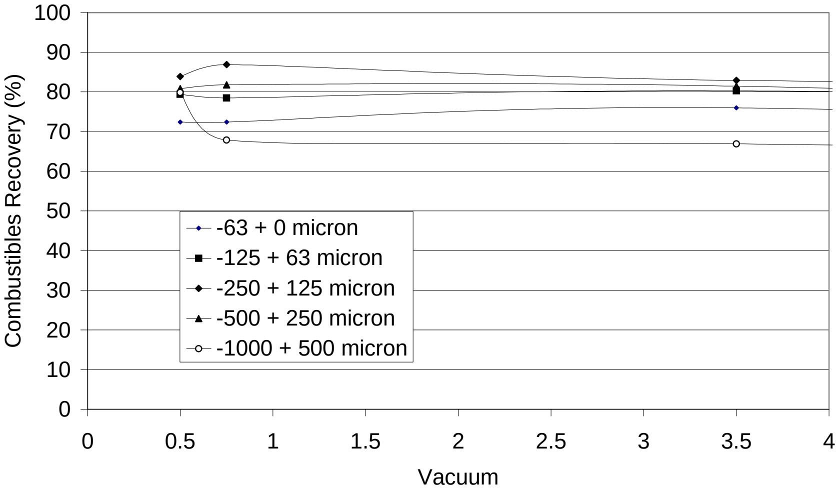


Figure 10
Pulp zone combustibles recovery

Although a full statistical analysis of pulp zone recovery is not possible due to the number of results available a brief review has been conducted. By reviewing data at an air-to-pulp ratio of 1.06, a vacuum of 3.5kPa, where the tank air void fraction was 10.3%, the following can be deduced:

1. For particle sizes above 125µm approximately 70% of coal lost in the froth zone is recovered in the pulp zone.
2. For particles below 125µm 90% of coal lost in the froth zone is recovered in the pulp zone.
3. For particle sizes between 63µm and 500µm approximately 17% of coal not recovered in the downcomer is recovered in the pulp zone.
4. For particles below 125µm approximately 12% of coal not recovered in the downcomer is recovered in the pulp zone.
5. There is no evidence of pulp zone coal recovery for particles greater than 500µm not recovered in the downcomer

Conclusions

1. The downcomer is the primary zone for combustibles recovery. Recovery of size fractions <500µm increase with increasing air-to-pulp ratios, while recovery of the >500µm fraction is either unaffected or decreases. At maximum air-to-pulp ratios the recovery of size fractions <500µm decreases with decreasing vacuum, while recovery of the >500µm fraction increases.
2. The pulp zone effectively recovers coal lost in the froth zone and to a lesser extent finer coal not initially recovered in the downcomer.
3. The froth zone recovery is linearly dependent of residence time within the froth.

Acknowledgments

The authors would like to thank Xstrata Technology and the University of Queensland for permission to publish this paper. One of the authors (GJH) is a recipient of an Australian Postgraduate Research Award (Industry) scholarship.

References

Australian Standards, 1994. AS4156.2.1 Coal Preparation: Part 2.1: Higher rank coal- Froth flotation- Basic test, Standards Association of Australia, Homebush, NSW, 1994.

Australian Standards, 1994. AS4156.2.2 Coal Preparation: Part 2.2: Higher rank coal- Froth flotation- Sequential procedure, Standards Association of Australia, Homebush, NSW, 1994.

Clarkson, C J, Manlapig, E V, Cheng, C Y, 1995. Pilot and laboratory scale trials on Jameson Cells

Harbort, G J, De Bono, S, Carr, D, Lawson, V, 2003. Jameson Cell fundamentals – a revised perspective. *Minerals Engineering*, Vol. 16, No. 11, Nov. 2003. Pp 1091-1101

Harbort, G J, Manlapig, E V, DeBono, S K, 2002. A discussion of particle collection within the Jameson Cell downcomer. Trans *IMM (Section C: Mineral Process. Extr. Metall*), 111 / *Proc. Australas. Inst. Min. Metall.*, 307, January/April, 2002, ppC1-C10

Evans, G M, Atkinson, B W, Jameson G J, 1995. The Jameson Cell. *Flotation Sci. Eng*, 1995, pp331-363

Finch, J.A., Dobby, G.S., 1990, Column Flotation, Pergamon Press, Toronto, Canada.

Honaker RQ, Mohanta NK, Ho K, 1995. Comparison of Flotation Cells. *12th International Coal Preparation Exhibition and Conference. Lexington, KY, USA May 2 – 4, 1995. pp 175 - 189*

Jameson, GJ, Belk, M, Johnson, NW, Espinosa-Gomez, R, Andreaditis, JP, 1988. Mineral flotation in a high intensity column. *Chemeca 88, 16th Australian Conference on Chemical Engineering*. Sydney, 1988. pp 507-510

Jameson, GJ, 1988. A new concept in flotation column design. Column '88 – *Proceedings of an International Symposium on Column Flotation*, SME, Phoenix Az, 1988. Sastry, KV, ed. pp 281-289

Manlapig, EV, Jackson, BR, Harbort GJ, Cheng CY, 1993. Jameson Cell coal flotation in 10th International Coal Preparation Exhibition and Conference, May 4-6 1993, Lexington, Kentucky. pp 203-219

Mohanty, M K, Honaker, R Q, 1999. Performance optimisation of Jameson flotation technology for fine coal cleaning. *Minerals Engineering*, Vol. 12, No. 4, 1999. pp 367 – 381

Vera, M, 1999. The Froth Zone. *The Optimisation of Mineral Processing by Modelling and Simulation 1996 to 1999 Volume 2- Flotation*. AMIRA Project P9L. Pp 117 - 123

Expansion of the Mount Isa Mines Copper Concentrator Phase One Cleaner Circuit Expansion

D Carr[1], G Harbort[2] and V Lawson[3]

ABSTRACT

Mount Isa Mines Limited operates one of the world's largest underground mines in North West Queensland. It's mining and smelting complex produces copper anode, crude lead bullion (containing silver), and zinc concentrate. Copper is recovered in the Copper Concentrator from chalcopyrite ore, converter smelter slag and rotary holding furnace (RHF) slag via grinding and flotation operations.

During 2002 a project to optimise cleaner circuit capacity was commenced. Its aims were to minimise copper losses and minimise slag/chalcopyrite cross contamination by reconfiguring existing flotation equipment and the addition of new flotation capacity. To minimise cross contamination a Jameson Cell was installed and dedicated to cleaning of slag concentrates, which had formerly been conducted by three flotation columns operating in series. The flotation columns were re-configured to operate in parallel and dedicated to upgrading chalcopyrite concentrate. In addition a Jameson Cell was installed to reject entrained copper from a prefloat concentrate consisting of talc and carbonaceous pyrite.

This paper discusses the logic behind the cleaner circuit expansion, recent test work relating to the Jameson Cell and the installation and commissioning phases. Difficulties relating to the commissioning are reviewed and the solutions to problems are highlighted. The benefits the reconfigured circuit has provided to Mount Isa Mines are also discussed.

JUSTIFICATION

The circuit capacity during commissioning of the Copper Concentrator in 1973 is now inadequate due to:

- increase in throughput rate and variability;
- increase in head grade and head grade variability resulting from Enterprise Mine ore;
- increase in talc in feed; and
- higher concentrate grade targets; meaning 15 per cent of installed roughing capacity is now lost to preflotation.

Chalcopyrite ore treatment justification

From 1975 to 1980 the Copper Concentrator treated 154 000 t Cu in feed per year. In 2001/02 it treated 230 000 t contained Cu in feed; a 50 per cent increase with 15 per cent less roughing capacity.

Ongoing metallurgical investigations and full plant demonstrations showed that a two per cent increase in copper recovery could be achieved if flotation capacity is increased. These demonstrations show that increased recovery is achieved when one line of grinding and the equivalent of double the flotation capacity is used. The concentrator flow sheet is shown in Figure 1 and illustrates the parallel grinding and primary flotation with a common cleaning circuit.

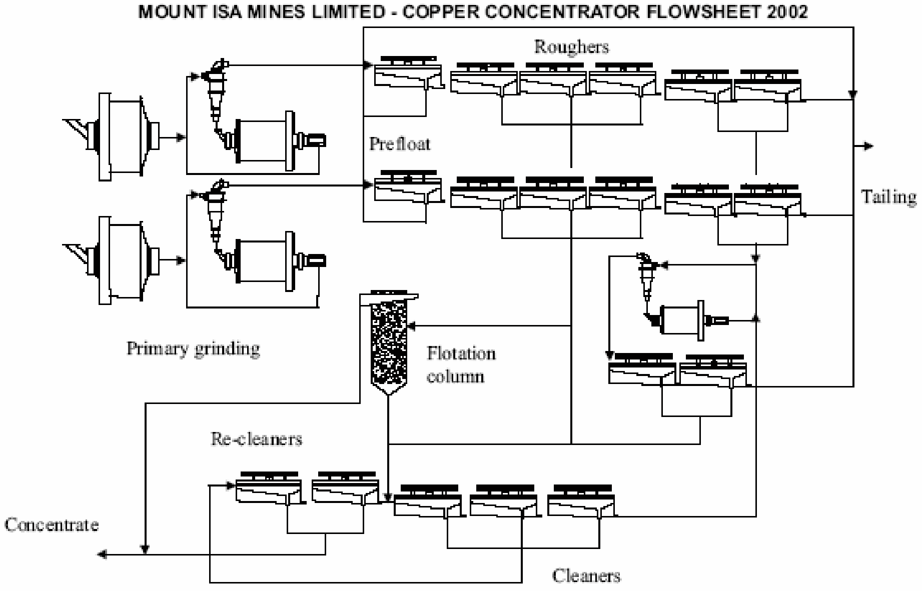


FIG 1 - Pre-expansion parallel milling flow sheet.

---

1. GRD Macraes. P O Box 84, Palmerston, Otago 9061 New Zealand.
2. Division of Mining and Mineral Process Engineering, University of Queensland, St. Lucia, Qld 4070.
3. MAusIMM, Senior Project Metallurgist, Mount Isa Mines Limited, 51 Darling Crescent, Mount Isa Qld 4825.
E-mail: vqlaws@isa.mim.com.au

Over time there has been an increase in the amount of naturally floatable species of carbonaceous pyrite and talc in ore. This would normally decrease concentrate grade for a given recovery; however at the same time the concentrator has been required to increase concentrate grade from 25 per cent to 27 per cent Cu. A preflotation circuit to remove naturally floating species has been necessary to achieve this. Fifteen per cent of rougher scavenger capacity has been reallocated to form the preflotation rougher circuit. There is no facility to clean preflotation rougher concentrate, causing higher than necessary copper losses by entrainment.

Slag treatment justification

Converter slag and RHF slag are both produced within the copper smelting complex of Mount Isa Mines and contain economically recoverable quantities of copper as chalcocite and metallic copper. When the Copper Concentrator capacity periodically exceeds that of the mine the opportunity is taken to treat converter and RHF slag.

The contamination of the chalcopyrite flotation with as little as 2 wt per cent slag results in total collapse of the flotation froth and loss of recovery. This collapse can be overcome with uneconomic additions of either sodium isobutyl xanthate (SIBX) or methyl isobutyl carbinol (MIBC). The mechanism by which slag removes the collector from the surface of the chalcopyrite particles is under investigation. This phenomenon is common to both converter and RHF slag. For this reason the flotation of ore and slag are completed independently of each other minimising cross contamination.

Performance at both Hilton Concentrator and in the Copper Concentrator indicated that froth washing techniques were the most effective forms of cleaning to achieve high-grade concentrates for smelting. Figure 2 shows the historical performance of converter slag in the Copper and Hilton Concentrators.

Plant test work was performed over the past five years to confirm the additional flotation requirements in the Copper Concentrator. Trials were conducted on a pilot scale and where possible on full plant scale. The level of both pilot plant and full plant investigations all but eliminated any significant technical risk. More recently, work was performed to confirm the proposed flotation flow sheet.

Proposed circuit expansion

Five areas were identified that would contribute to the increase in recovery. These areas were:

1. Jameson Cell for cleaning preflotation rougher concentrate;
2. Jameson Cell for slag cleaning;
3. changing column pipe work from series to parallel operation;
4. tank cells for scavenging; and
5. tank cells for cleaner scavenging.

This paper discusses the first three changes including the installation of two Jameson cells and changes to the copper columns. Benefits include the reduction of slag contamination in chalcopyrite flotation currently resulting in 0.3 per cent lower copper recovery per year and reduction in copper lost to preflotation (currently 0.5 - two per cent copper).

PROCESS DESCRIPTION

Mt Isa copper treatment

Chalcopyrite ore and converter slag are treated to produce copper concentrates at the Copper Concentrator of Mount Isa Mines Limited, Mount Isa, Qld. The concentrator was commissioned in 1973, replacing the original No. 1 Lead-Zinc Concentrator, for the processing of all the chalcopyrite ore coming from Mount Isa mine (Lumsdaine, O'Hare, 1980).

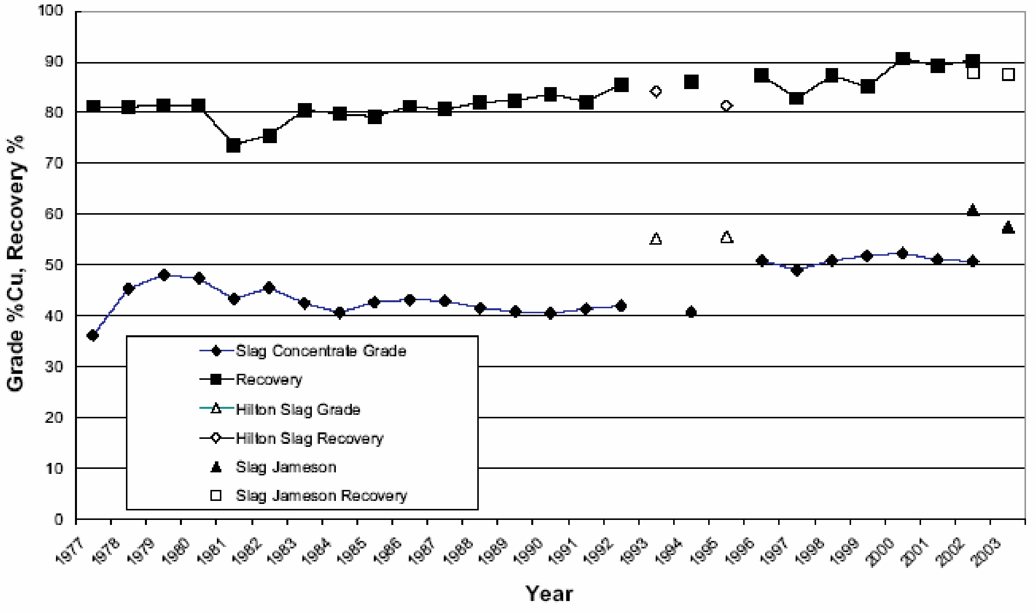


FIG 2 - Historical smelter slag metallurgical performance.

At present production ore is sourced from the southern 1100, 3000 and 3500 orebodies. Currently 50 per cent of mill feed is from the 3000 and 3500 orebodies. Chalcopyrite is the only significant copper mineral and occurs as a replacement deposit in a silica-dolomite host rock. Sulfide gangue consists of pyrite (FeS_2) and minor amounts of pyrrhotite and cobaltite. The ore averages eight per cent sulfur.

Future ore production will draw increasingly larger tonnages from the 3000 and 3500 orebodies. The host rock is very similar to that of the 1100 orebody but is slightly more siliceous in nature. Silica assays range from 60 to 70 per cent SiO_2 and the ore is both abrasive and hard with a current Bond Work Index of approximately 22kWh/t. Copper mineralisation is disseminated through the silica-dolomite. Mineralogical examination indicates liberation is about 75 per cent at a P_{80} of 150 µm. Flotation feed sizing is normally in the range 80 per cent passing 150 µm, depending on milling rate and ore type.

Smelter slag treatment

Treatment of converter slag in the 1970's and 1980's was on a batch basis during periods of ore shortage. The batch processing of Copper Smelter converter slag began in 1976 (Bartrum *et al*, 1978). Parallel processing of slag and ore commenced in 1992 and the flotation circuit was modified to simplify parallel processing in 1998 when it was determined that Copper Smelter Rotary Holding Furnace (RHF) slag also had economic retreatment value. Slag was treated through the concentrator using the existing flow sheet. In 1991 the comminution circuit consisting of secondary crushing, screening, open circuit rod milling and closed circuit ball milling was replaced with an AG/SAG/ball mill circuit. With the introduction of AG/SAG milling the treatment of slag could be performed in parallel with ore using one line of grinding, primary flotation and the flotation columns (Barnes *et al*, 1993). This flow sheet is shown in Figure 3. During 1993 and 1995 the nearby Hilton Concentrator was converted from treating lead zinc to converter slag to reduce the stockpile of slag that had developed due to high output from the mine and low concentrator throughput. For a period of this time Hilton Concentrator operated Jameson cells in the slag cleaner duty (Dawson, Harbort, 1996).

Preflotation

Periodically amounts of naturally floatable species of carbonaceous pyrite and talc occur in the Mt Isa copper ore. Early treatment in the Copper Concentrator consisted of flotation, followed by depression with ambiguous results (Hoffman, 1965; Lyon, Fewings, 1971). Investigation by Grano, Ralston and Smart, in 1990 identified the natural floatability of carbonaceous pyrite. An evaluation of using preflotation prior to roughing, rather than the flotation/depression route was conducted.

The preflotation circuit operated in the Copper Concentrator from 1992-1996. Originally the preflotation rougher concentrate was cleaned in the three flotation columns operating in series. Griffin (1993) reported that this circuit increased the concentrate grade from 25 per cent Cu to 27.2 per cent Cu for a 1.5 per cent loss in copper recovery. The loss consisted of 0.6 per cent Cu lost to the prefloat concentrate in the form of carbonaceous copper and 0.9 per cent loss due to the decrease in scavenger capacity.

In 1996 the duty of the columns was changed from preflotation cleaning to primary cleaners treating copper rougher concentrate and producing up to 50 per cent of the final concentrate. This circuit enabled consistently high concentrate grades to be achieved without the preflotation circuit. As the ore was sourced increasingly from the deeper copper orebodies 3000 and 3500 the level of talc increased making concentrate targets harder to achieve. The prefloat circuit was reinstated with the prefloat rougher concentrate being sent straight to final tail.

Work on a pilot scale using Jameson Cells for one stage preflotation and for preflotation cleaning had been performed over a number of years. A series of pilot tests were conducted in 2000 to confirm parameters for engineering design.

Jameson Cell operating principles

The principles of Jameson Cell operation have been discussed by numerous authors including Jameson (1998), Jameson *et al* (1998) and Evans *et al* (1995). Recent developments have been reviewed by Harbort *et al* (2003) and Carr *et al* (2003). Operation can be described with reference to Figure 4.

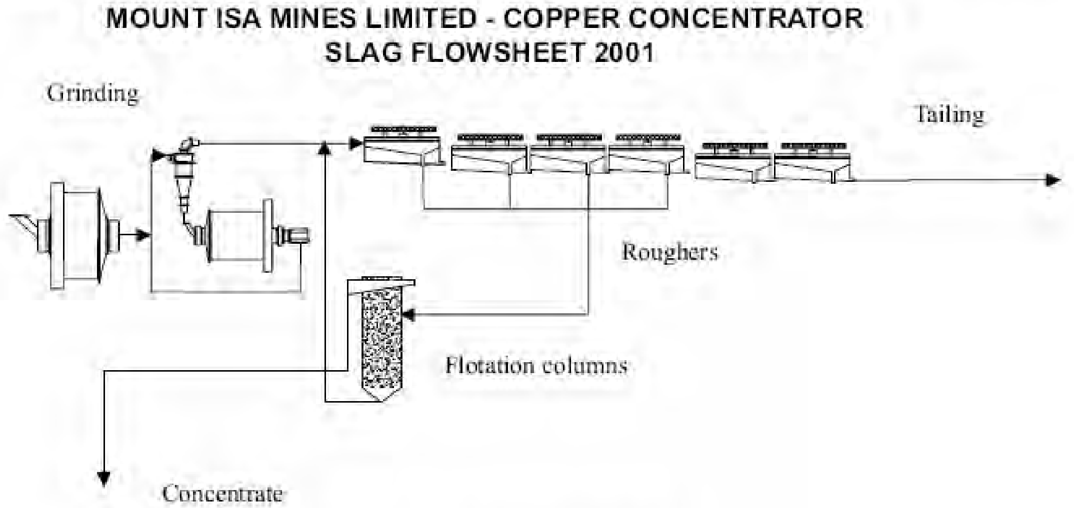


FIG 3 - Pre expansion slag circuit.

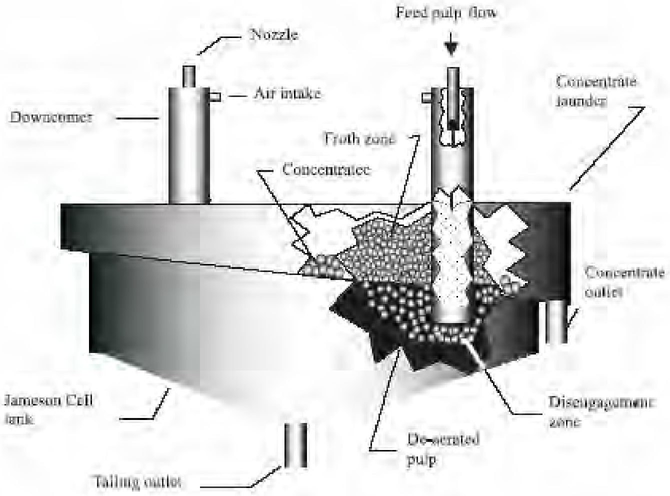


FIG 4 - Jameson Cell schematic.

The Jameson Cell can be divided into three main zones:

- The downcomer where primary contacting of bubbles and particles occurs. Feed pulp is pumped into the downcomer through an orifice plate, creating a high-pressure jet. The plunging jet of liquid shears and then entrains air, which has been naturally aspirated. Due to a high mixing velocity and a large interfacial area there is rapid contact and collection of particles.
- The tank pulp zone where secondary contacting of bubbles and particles occurs and bubbles disengage from the pulp. The aerated mixture exits the downcomer and enters the pulp zone of the flotation tank. The velocity of the mixture and large density differential between it and the remainder of pulp in the tank results in recirculating fluid patterns, keeping particles in suspension without the need for mechanical agitation.
- The tank froth zone is where materials entrained in the froth are removed by froth drainage and/or froth washing.

DESIGN

Jameson Cells

For cleaning of both converter slag and RHF slag a model E2532/6 Jameson Cell was installed. The model designation refers to a rectangular cell, 2.5 m in length × 3.2 m wide, equipped with six downcomers and an internal launder. A wash water system was included that consisted of a set of perforated trays suspended above the froth. At the design phase it was envisaged that operation on converter slag would require six downcomer operation with RHF slag only requiring three. As such the cell was designed with an internal baffle allowing a 50 per cent turndown. The design specification for both duties, plus commissioning information is given in Table 1 and 2. The new slag circuit is shown in Figure 5.

TABLE 1

Design and commissioning data for the Jameson Cell converter slag cleaner.

Converter slag	Design	Commissioning
New feed solids tph	148 - 188	200 - 252
New slurry flow m^3/hr	255 - 350	300 - 378
Downcomer feed m^3/hr	448	480 - 500
Downcomer feed pulp density kg/l	1.4	1.4 - 1.7
Tailing recycle (%)	40	0 - 30
Air flow m^3/hr	140	115
Jg (cm/sec)	0.43	0.45

TABLE 2

Design and commissioning data for the Jameson Cell RHF slag cleaner.

RHF slag	Design	Commissioning
New feed solids tph	51 - 95	124 - 202
New slurry flow m^3/hr	91 - 168	156 - 311
Downcomer feed m^3 per hr	224	400 - 500
Downcomer feed pulp density kg/l	1.4	1.35 - 1.65
Tailing recycle (%)	50	40 - 61
Air flow m^3/hr	45	110
Jg (cm/sec)	0.36	0.44

For cleaning of prefloat concentrate a model E2514/3 Jameson Cell was installed. The model designation refers to a rectangular cell, 2.5 m in length × 1.4 m wide and equipped with three downcomers, as shown in Figure 6. A submerged wash water system was included that consisted of a set of perforated pipes that were suspended in the froth zone. The cell was designed with adjustable concentrate lips that allowed the froth cross sectional area to be reduced by 50 per cent to achieve higher superficial air rise velocities. The design specification for prefloat cleaning, plus commissioning information is given in Table 3. The new chalcopyrite circuit is shown in Figure 7.

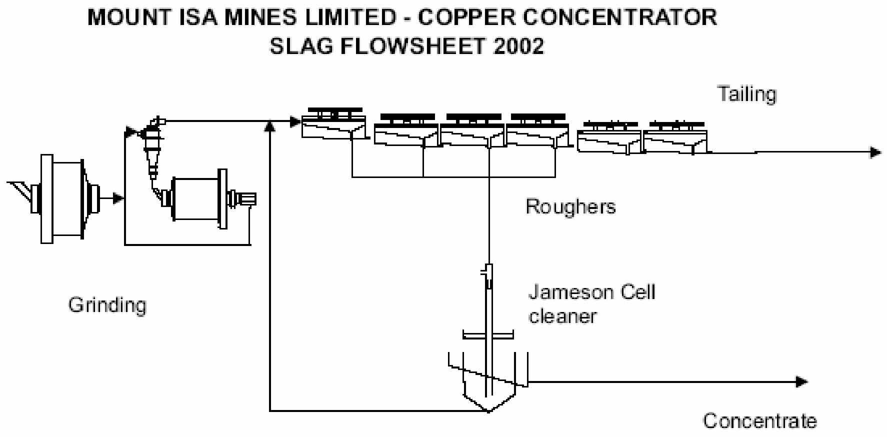


FIG 5 - Post expansion slag circuit.

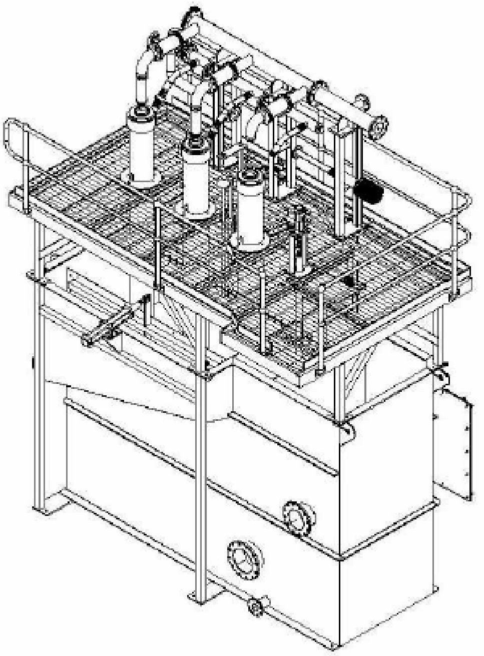


FIG 6 - Prefloat cleaner Jameson Cell.

Both Jameson Cells were fitted with an internal tailing recycle mechanism where a small amount of tailing was directed into the feed to provide constant pump feed and downcomer feed pressures.

A requirement of the installation was to minimise equipment and operator labour to operate the circuit as well as to complete construction and commissioning with a minimum of downtime. The location of the equipment required a design that would require the minimum involvement of the flotation operator during normal operation. For this reason the equipment was located approximately 12 m above ground level to allow gravity to be used to transport tailings to the head of the rougher circuit and for concentrate to travel to the final tailings and concentrate pump boxes. The feed pumps were located on a suspended concrete slab 6m above the ground at the same level as the existing flotation banks.

TABLE 3

Design and commissioning data for the prefloat cleaner Jameson Cell.

Prefloat cleaning	Design	Commissioning
New feed solids tph	54 - 80	5 - 20
New slurry flow m^3/hr	140 - 210	35 - 80
Downcomer feed m^3/hr	224	240 - 250
Downcomer feed pulp density kg/l	1.27	1.2
Tailing recycle (%)	20	64 - 93
Air flow m^3/hr	134	100 - 200
Jg (cm/sec)	0.6 - 1.0	0.9 - 2.3

Instrumentation and control

Both Jameson cells were fully instrumented with airflow control, level control, vacuum and slurry pressure measurement. A redundant 75 KW Variable Frequency drive was used on the slag Jameson cell to avoid delays in construction of a new Motor Control Cabinet and to provide the ability to vary the pump speed when changing from RHF to Converter slag. Two independent tail plugs were used on the slag Jameson cell to allow turndown for RHF operation without loss of level control. Pneumatic cylinders were used to control the angle of the adjustable launder on the prefloat cell as the simplest method of movement. The control room can move the launders over a range of approximately 600 mm, with automatic adjustment to the level calibration for the change in cell geometry undertaken in the plant DCS. Inline samplers of both concentrate streams were installed to allow online measurement of copper grade through the existing Courier 30XP on stream analyser and to provide shift samples for assay. The prefloat concentrate line passes through a U tube before discharging into the final tailings stream to allow flow measurement with a conventional magnetic flow meter.

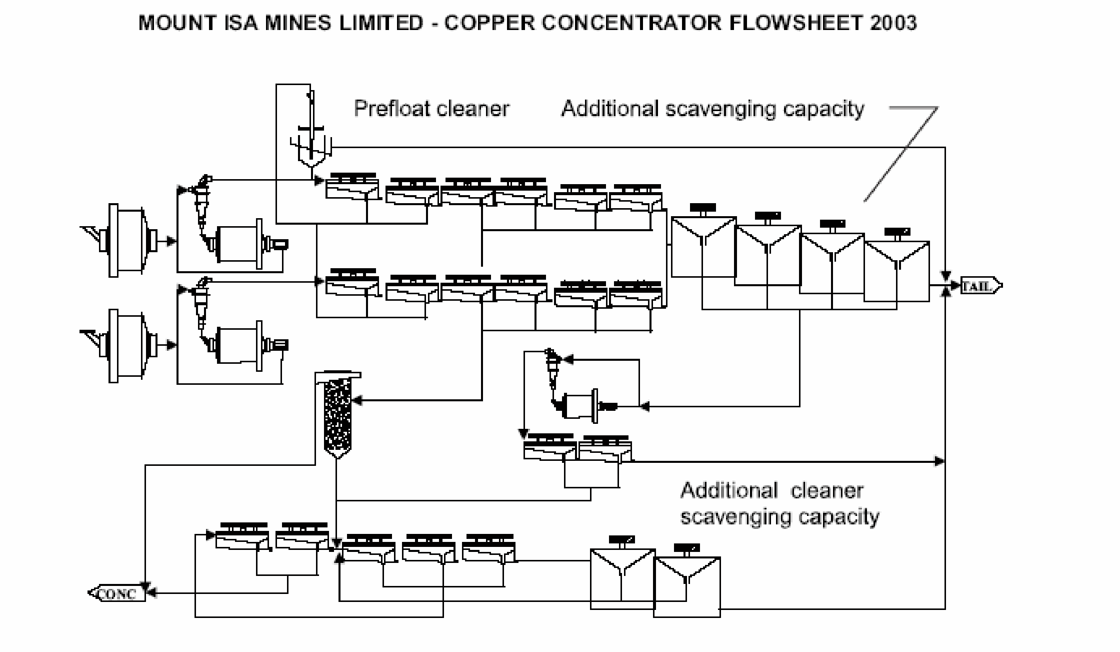


FIG 7 - Mount Isa Copper Concentrator post expansion.

A closed circuit Pan-Zoom-Tilt camera was installed above the cells allowing the control room operator to be able to see the froth surface, pulling rate of the cells and the level in the feed boxes. This has proved to be considerably useful in diagnosing problems with the cell without having to recall the flotation operator from other areas of the plant.

COMMISSIONING

Columns

The flotation column alterations consisted of re-piping feed, tailing and concentrate streams so that they could be operated in parallel, rather than in series. It did not require installation of additional equipment and as such the commissioning of the columns was relatively trouble free.

Slag cleaner

The maximum commissioning tonnage of new feed to the Jameson Cell on converter slag treatment exceeded the maximum design tonnage by some 34 per cent. On RHF slag treatment it exceeded the maximum design tonnage by some 113 per cent. This is due to a number of reasons including higher grinding tonnages and copper content in feed. The initial impact of the higher feed tonnages was high pulp densities in downcomer feed, which appeared to reduce Jameson Cell copper recovery and increase the re-circulating load in the circuit, causing further increases in feed. To minimise losses all six downcomers were operated during treatment of RHF slag. At times during converter slag treatment the tailing recycle mechanism was overloaded with feed bypassing the cell and going directly to tailing.

Air entrainment was found to be detrimentally affected by the high pulp densities. At pulp densities of 1.55 kg/l to 1.7 kg/l air entrainment was only half the design air requirement. Significant improvements in air entrainment were achieved at a pulp density of 1.45 kg/l with design air entrainment or better being achieved at the design pulp density of 1.4 kg/l. To ensure optimum feed densities during commissioning the Jameson Cell feed pump speed was increased by 20 per cent and dilution water was added to the feed box.

Figure 8 shows a comparison of the concentrate grade and recovery achieved following pulp density control with that achieved during the original test work in June 2000. During commissioning a maximum copper recovery of 63.5 per cent at a concentrate grade of 63.8 per cent Cu was achieved. This compares well with the original test series. It should also be noted that during commission numerous mill feed disruptions occurred with feed to the flotation circuit being disrupted for a significant portion of operating time. This resulted in attempts at complete optimisation of parameters and recovery being limited. It was expected that target copper recoveries of 70 per cent were achievable.

During the design phase discussion were held on whether the addition rate of SIBX to the roughers resulted in Jameson Cell instability and whether either reductions in rougher addition rate or a change to another collector would enhance Jameson Cell and overall slag treatment performance. There was some tenuous evidence from original test work that indicated that SIBX addition was detrimental and that increases in SIBX resulted in further deterioration of Jameson Cell performance. Operation during commissioning indicated that high SIBX addition would not affect Jameson Cell stability.

During original test work variation of MIBC had very little effect on vacuum generated and air entrained within the downcomer. During commissioning increased MIBC addition to the Jameson Cell feed was found to have a significant beneficial effect on downcomer vacuum and operating stability. The optimum MIBC addition was approximately 20 ppm MIBC in downcomer feed which is in the upper range of Jameson Cell installations.

Prefloat cleaning

The maximum tonnage of new feed to the Jameson Cell was substantially less than the maximum design tonnage at all times and specifically when only one milling line was treating chalcopyrite ore. This was due to lower than expected grades of prefloat material in run of mine ore. This resulted in levels of tailing recycle above design. Usually increased rates of tailing recycle are considered beneficial to operation as it increases the recovery of the valuable mineral without significant effect on concentrate diluents. This was not necessarily the case with the prefloat Jameson Cell. Although the true flotation talc recovery increased with increasing tailing recycle, true flotation pyrite recovery decreased and above a certain point copper recovery increased. A full understanding of the decline in FeS_2 recovery is not possible without detailed surface analysis of floated and non-floated particles. It is possible that only strongly carbonaceous pyrite is being recovered within the Jameson Cell. Alternatively multiple passes through the downcomer at high tailing recycle rates may have resulted in some passivation of the pyrite surface.

Immediately following start up a number of concerns were raised with the long-term efficiency of the submerged wash water system. It was noted that at a wash water to concentrate water ratio of 0.5, the low velocity water jets exiting the wash water pipes only penetrated the froth to a horizontal depth on 50 mm, compared to a required penetration of 250 mm. This generated sluggish froth zones between the wash water pipes, with low apparent recovery of prefloat concentrate. A number of comparisons with submerged versus 'top of froth' were conducted. These included the insertion of froth tracers in the froth zones and collection of timed lip samples. This allowed froth velocities, volumetric flow and tonnage produced to be evaluated between the two washing systems.

The submerged wash water piping resulted in a 20 per cent decrease in horizontal froth velocity, at a Jg of 1.25 cm/sec, with the two systems only showing equivalent horizontal froth velocities at a Jg of 2.0 cm/sec. The 'on top' wash water achieved a 50 per cent higher flow over the lip and a 50 per cent higher solids mass pull than the submerged wash water system. Although submerged wash water systems were shown to have major benefits in other applications it was considered that the area of wash water piping in the pre-float cell was too high a percentage of the froth area and the system was replaced with wash water trays.

The strength of the Jameson Cell air entrainment would generally indicate that sufficient frother is present to provide acceptable operation. There is however some indication that MIBC may be acting as a weak collector for naturally floating material as well as a frothing agent. A sample of prefloat rougher concentrate was collected and floated under nitrogen in a standard Mount Isa Mines prefloat bench flotation test, with no additional MIBC added. The test was then repeated with additional MIBC.

Plots of true flotation recovery (ie total recovery less quartz recovery) versus time were then generated to determine the effect of additional MIBC. The bench test indicated that additional MIBC addition to the Jameson prefloat cleaner would have only a minimal effect of talc recovery. Figure 9 shows the response of FeS_2. The bench test with no additional MIBC shows a talc recovery after eight minutes of seven per cent, or less than half that achieved on average by the Jameson Cell prefloat cleaner. With additional MIBC the eight-minute FeS_2 recovery increased dramatically to 47.5 per cent with the curve indicating further recovery gains were possible with time. There are a number of possible reasons for the improvement including smaller bubble size, a more stable froth or the MIBC acting as a weak collector, enhancing the flotation of partially carbonaceous pyrite.

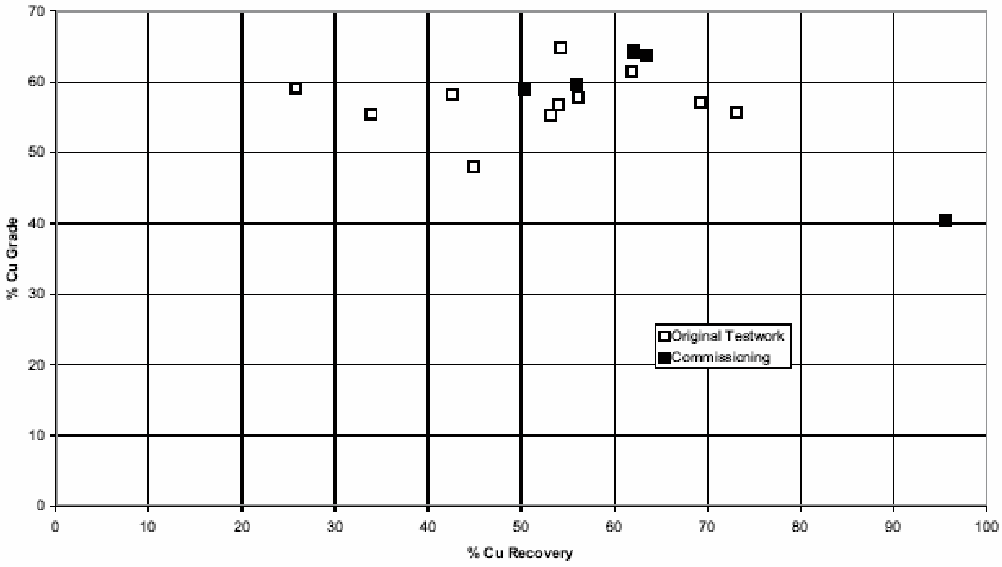


FIG 8 - Converter slag concentrate grade versus copper recovery.

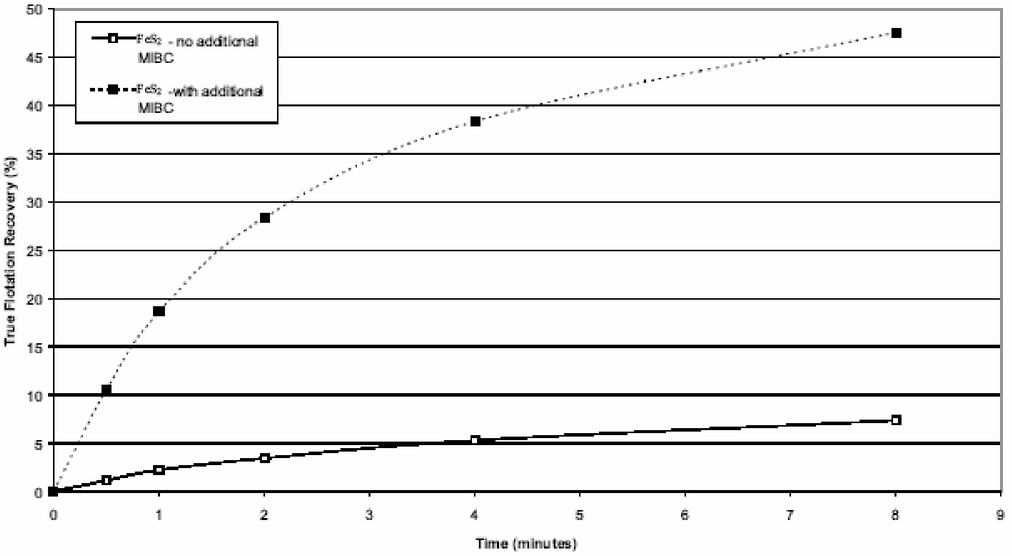


FIG 9 - Flotation of pyrite with and without additional MIBC.

The potential existed that additional MIBC addition to the Jameson prefloat cleaner could have a significant effect on improving FeS_2 recovery. Although additional MIBC is a promising route for increasing FeS_2 recovery there are some concerns. The current level of MIBC entering the Jameson Cell in rougher concentrate appears to be approaching saturation levels. Periodically it was observed that MIBC was causing bubble size within the downcomer to be so small that bubbles could not disengage from the pulp within the tank. This resulted in the tank 'foaming' with no effective froth/pulp interface, loss of level control and severe loss of recovery and concentrate quality. An initial trial of additional MIBC was aborted when this occurred.

Figure 10 shows total talc recovery versus copper rejection from concentrate for the commissioning period, the original test work and for bench scale tests taken during commissioning, respectively. Targeted performance was to achieve a copper rejection of 90 per cent, at a talc recovery of 70 per cent. The upper envelope for commissioning surveys indicates a minor shortfall from this point, with an interpolated copper rejection of 90 per cent at a talc recovery of approximately 67 per cent. At 90 per cent copper rejection talc recoveries of 50 per cent and 47 per cent were achieved in the bench scale tests and original test work respectively.

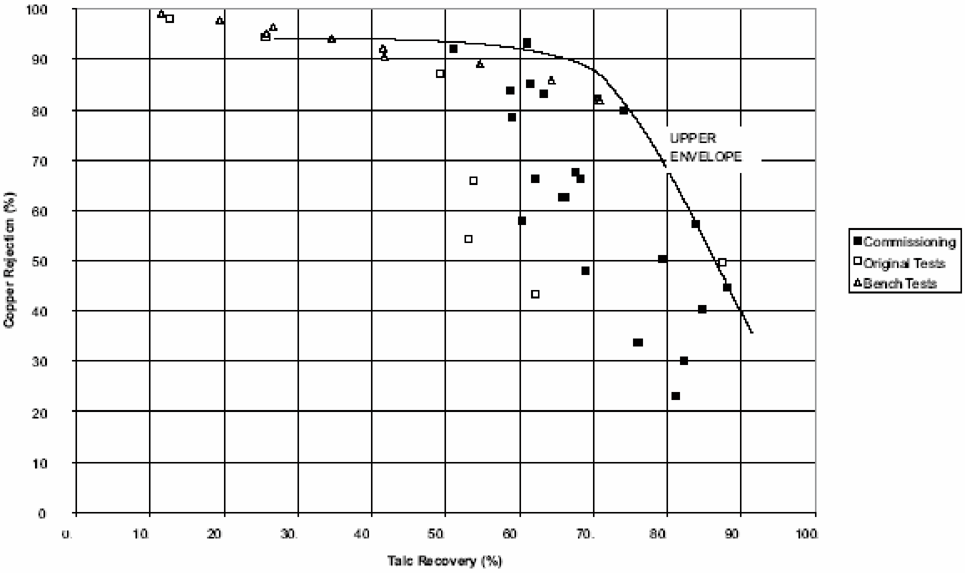


FIG 10 - Talc recovery versus copper rejection.

Figure 11 shows total FeS_2 recovery versus copper rejection from concentrate for the commissioning period, the original test work and for bench scale tests taken during commissioning, respectively. Targeted performance was to achieve a copper rejection of 90 per cent, at an FeS_2 recovery of 50 per cent. The commissioning surveys indicate a significant shortfall from this point, with a copper rejection of 90 per cent at an FeS_2 recovery of only 17 per cent. At 90 per cent copper rejection FeS_2 recoveries of 30 per cent and 28 per cent were achieved in the bench scale tests and original tests respectively. The commissioning results show a strong linear relationship between copper rejection and FeS_2 recovery indicating some degree of carbonaceous chalcopyrite being present and also FeS_2 recovery gains being due to entrainment rather than true flotation. There is some initial evidence that carbonaceous pyrite that did not float within the Jameson Cell has not been refloated in the prefloat roughers and does not report to final concentrate.

POST COMMISSIONING

To provide fewer fluctuations in feed tonnage to the slag cleaner blending of converter and RHF slag was commenced. This has resulted in a major reduction in the per cent solids entering the Jameson Cell and has produced very stable operation. A number of changes to the reagent addition were also trialed. All MIBC is added to the Jameson cell rather than the flotation feed to improve the air entrainment. Sufficient frother remains in the Jameson Cell tailing to meet rougher frother requirements. SIBX addition is used in the roughers to control rougher recovery. Additions greater than 200 g/t SIBX are used regardless of head grade whilst dilution water and wash water are used to control Jameson concentrate grade.

The prefloat cleaner was successfully refitted with wash water trays, which resulted in more effective rejection of entrained chalcopyrite and removal of talc.

The original butterfly valves supplied for commissioning have been replaced with V-notch knife gate valves, which have provided substantially better airflow control. The improved flow characteristics and pressure drop on the new valves has allowed the removal of substantial signal damping required for the air flow measurement and provided more consistent air flow control.

The expansion of the Copper Concentrator has been multifaceted and to quantify the benefit of one circuit change in isolation to the others is difficult. A number of qualitative benefits were rapidly evident. The dedication of the flotation columns for chalcopyrite cleaning only has resulted in a major decrease in slag contamination, allowing significantly more stable chalcopyrite flotation. The Jameson Cell slag cleaner operates with a tank residence time of only 3.5 minutes, compared to sixty minutes for the columns in series. The small volume of the Jameson Cell has allowed it to be rapidly stabilised during start-ups and following inadvertent shutdowns, releasing operators for other duties. Dumping of the cell, if required also results in much less spillage and cross contamination. The Copper Concentrator has been operating almost exclusively on the parallel milling circuit since December 2000 with the target of eliminating the slag stockpile. The prefloat Jameson Cell has reduced pyrite in concentrate and allowed the concentrator to handle increased talc head grades without increasing the copper loss, Figure 12. Run of mine talc head grades have been amongst the highest ever recorded, yet concentrate grades are still in excess of 25 per cent Cu.

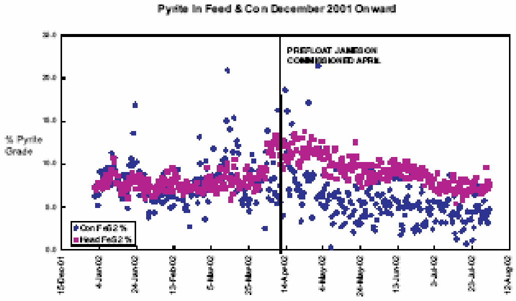


FIG 12 - Pre and post expansion pyrite in feed and copper concentrate.

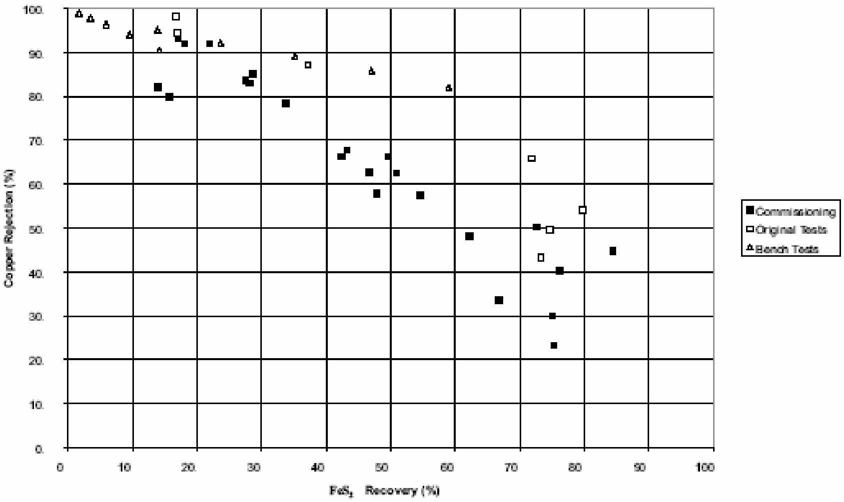


FIG 11 - Pyrite rejection versus copper recovery.

CONCLUSIONS

The slag Jameson Cell has allowed the Copper Concentrator to operate on the parallel milling circuit for extended slag campaigns due to the lower risk of cross contamination and the release of the columns back to the chalcopyrite circuit allowing higher ore treatment without loss in recovery. Also Jameson Cells have proved to be effective in their design role of copper recovery/rejection with a minimum operator involvement to control and very forgiving with variations in feed rate due to the nature of the internal tailing recycle. The prefloat Jameson Cell has reduced pyrite in concentrate and allowed the concentrator to handle increased talc head grades without increasing the copper loss.

ACKNOWLEDGEMENTS

The authors would like to thank Mt Isa Mines Limited, MIM Process Technologies and the University of Queensland for permission to publish this paper. One of the authors (GJH) is a recipient of an Australian Postgraduate Research Award (Industry) scholarship.

REFERENCES

Barnes, C D, Lumsdaine, I, O'Hare, S M, 1993. Copper converter slag treatment at Mount Isa Mines Limited, Mount Isa, Qld, *The AusIMM Proceedings*, 1993(1):31-37.

Bartrum, J, Schache, I S, Watsford, R M S, Slaughter, P J, 1978. Converter slag flotation practice at Mount Isa Mines Limited, in *Proceedings Mill Operators' Conference,* pp 113-120 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Carr, D, De Bono, S, Harbort, G J, Lawson V, 2003. Jameson Cell fundamentals: a revised perspective. Presented at *Flotation 03*, Helsinki, March 2003.

Dawson, W J, Harbort, G J , 1996. Jameson Cell recovery of copper from converter slag, in *Proceedings of Hidden Wealth – Unlocking values from low grade and refractory ores and wastes*, Minerals Engineering International.

Evans, G M, Atkinson, B W, Jameson G J, 1995. The Jameson Cell, *Flotation Sci Eng*, 1995:331-363.

Grano, S, Ralston, J, Smart, R St C, 1990. Influence of electrochemical environment on the flotation behavior of Mt Isa copper and lead-zinc ore, *International Journal of Mineral Processing*, 30 (1990), pp 67-97.

Griffin, L K, 1993. Results of the Second Extended Preflotation Plant Trial in the Copper Concentrator, Mount Isa Mines Technical Memorandum, MIM.

Harbort, G J, Manlapig, E V, DeBono, S K, 2003. A discussion of particle collection within the Jameson Cell downcomer, *IMM Trans / AusIMM Proceedings, Section C: Mineral Process Extr Metall, January/April,* pp C1-C10.

Hoffmann, C W, 1965. A review of developments in flotation treatment of Mount Isa chalcopyrite ores, in *Proceedings of 8th Commonwealth Mining and Metallurgy Congress Australasia*, 1965, Paper no 103, pp 1133-1149.

Jameson, G J, 1988. A new concept in flotation column design, *Column'88 – Proceedings of an International Symposium on Column Flotation* (Ed: K V Sastry) pp 281-289 (SME).

Jameson, G J, Belk, M, Johnson, N W, Espinosa-Gomez, R, Andreaditis, J P, 1988. Mineral flotation in a high intensity column, in *Proceedings Chemeca 88, 16th Australian Conference on Chemical Engineering*, Sydney, pp 507-510.

Lumsdaine, I, O'Hare, S M, 1980. Copper concentrator practice at Mount Isa Mines Limited, Mount Isa, Qld, in *Mining and Metallurgical Practices in Australasia* (Ed: J T Woodcock) pp 649-653 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Lyon, G C, Fewings, J H, 1971. Control of the flotation rate of the carbonaceous material in Mount Isa chalcopyrite ore, *The AusIMM Proceedings*, No 237, March 1971.

JAMESON CELL DEVELOPMENTS AT PHILEX MINING CORPORATION

by

Harbort, GJ,(1) Murphy, AS,(2) Budod A(3)

for publication in

The AusIMM Sixth Mill Operators Conference 1997

(1): Greg Harbort
ML Consultants
9 Marica Street
Bellbowrie, Qld 4070
Australia.

(2): Andrew Murphy
MIM Holdings Limited
GPO Box 1433
Brisbane, Qld, 4000
Australia

(3): Albert Budod
Philex Mining Corporation
PO Box 46
Baguio City, Benguet
Philippines

ABSTRACT

The Philex concentrator represents the first time Jameson Cells have been used in a cleaner scavenger application or to replace mechanical flotation cells in roughing and scavenging. It also represents the first concentrator to operate wholly with Jameson Cells. This necessitated a major change in Jameson Cell design as well as a change in operating philosophy within Philex.

This paper provides an update on cleaner circuit performance, together with refinements that were required for cleaner scavenger, roughing and scavenging operation. The extended commissioning period and operation to date are reviewed in respect to operating difficulties, changes in operational philosophy and metallurgical results. Comparisons of operating, costs are provided as are residence time and flotation area, together with scale up information leading to the final production cell design. Metallurgical results in relation to gold and copper recoveries are reviewed, with special attention given to direct comparisons between Jameson and mechanical flotation banks, size by size analysis and mineralogical performance.

INTRODUCTION

The Jameson Cell combines a novel method for air and slurry contact where a contained plunging jet naturally entrains air achieving high voidage and intimate particle bubble contact. The fundamentals have been described by numerous authors, including Jameson and Manlapig (1991) and are not discussed further here.

Philex Mining Corporation is one of the largest gold producers in the Philippines and a leading exporter of copper, gold and silver in the Far East today. The mine site is located at the southern tip of the Central Cordilleras of Luzon Island, 17 kilometres from Baguio City, Figure 1. Mining operations commenced in 1958 at a rate of 800 tpd. Plant expansions have increased tonnage to over 20,000 tpd with typical head grades of 0.3% copper and 0.5g/t gold, (Philex Mining Corporation, 1992).

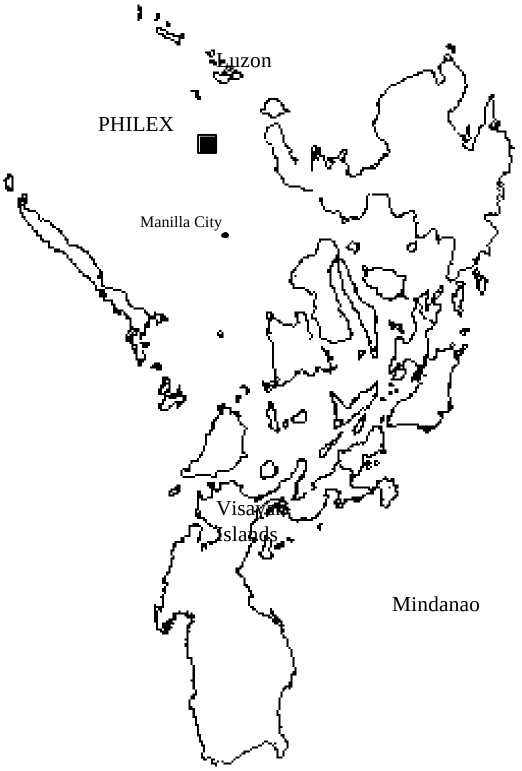


Figure 1. The Philippines

During 1992 Philex Mining Corporation conducted a major evaluation of their flotation circuit, Figure 2. This led to the conversion of their column/mechanical cell cleaner circuit to Jameson Flotation Cells in 1993. Following successful operation in cleaning further test work was conducted in 1993 in roughing and scavenging areas indicating additional metallurgical improvements could be made. In 1994 their mechanical cleaner scavenger circuit was replaced with Jameson Cells to provide floor area for a proposed rougher scavenger circuit. A phased introduction of Jameson Cell rougher and scavenger lines commenced in late 1994 and was completed in early 1996, producing the current Jameson Cell circuit, Figure 3.

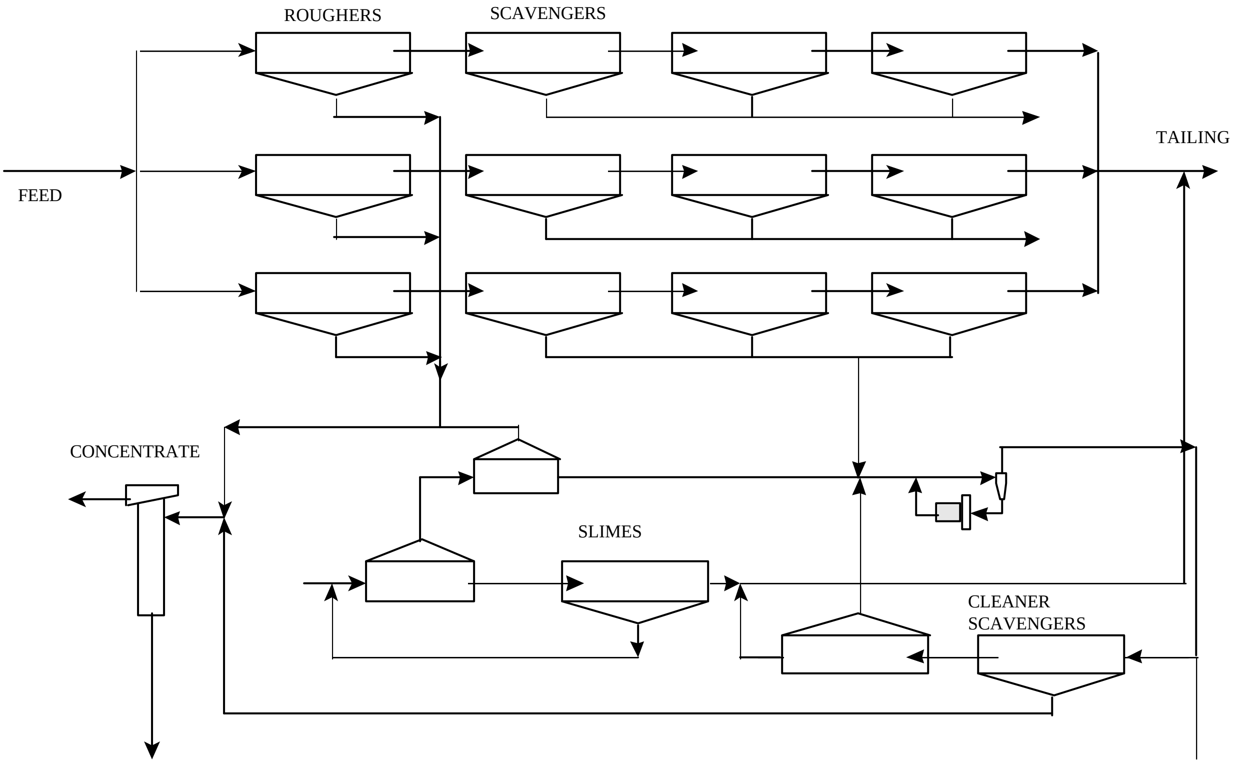


Figure 2. Philex Mining Corporation flotation circuit, 1992.

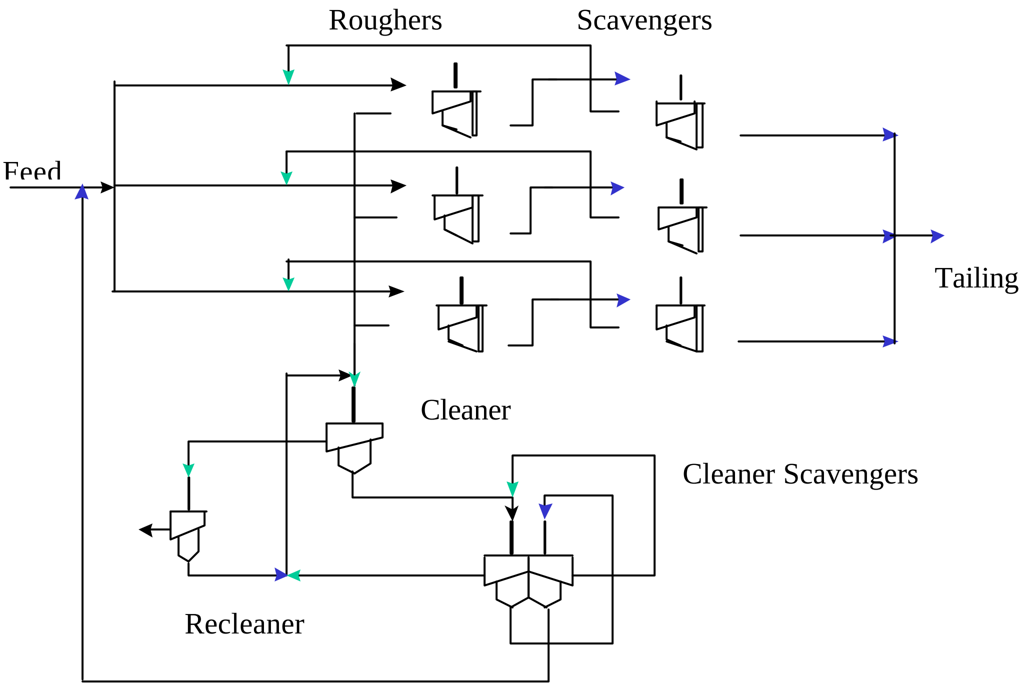


Figure 3. Philex Mining Corporation Jameson Cell flotation circuit, 1996.

CLEANER SCAVENGER INSTALLATION.

Motivation

Additional floor space created by the installation of the cleaner/recleaner cells had been used to increase cleaner scavenger residence time by sixty percent, through additional mechanical cells. As such, there had initially been no plan to replace the existing mechanical cleaner scavengers with Jameson Cells.

Following six months successful Jameson Cell cleaner/recleaner operation Philex's confidence in Jameson Cell technology was such to begin planning of a rougher/scavenger installation. This original proposal required a phased replacement of the mechanical banks, in their existing location, with Jameson Cells. A one third loss in plant capacity would have resulted at each equipment change over. In early 1994 a review of the rougher/scavenger proposal indicated that this entire circuit could be placed in the area where the mechanical cleaner scavengers were currently installed, without affecting production. This provided the motivation for a change of the cleaner scavenger circuit to Jameson Cells.

Design Considerations

Two major difficulties with this change over existed:

1. The only cleaner scavenger test work conducted (Harbort, 1992), was done with the aim of increasing circuit recovery with the addition of one Jameson Cell acting as a scalper. Scale-up calculations had shown that a four metre diameter cell with sixteen downcomers (4000/16) was required for this duty and initial cell design had been commenced. Typically it was to produce a concentrate of 15% Cu at a copper recovery of 60%. This compared with a new requirement to produce a recovery of 90% or better.
2. Available floor space was limited to an area 6.0m by 6.0m.

Alterations to initial design

The limited test work had indicated that recovery was very dependent on superficial air velocity, (Jg), with an increase in Jg from 0.4cm/sec to 0.8cm/sec giving a copper recovery increase from 60% to 70%. A further increase in Jg to 1.2cm/sec was expected to further improve recovery to 80%, still less than the new requirement, indicating that two stages of Jameson Cell treatment were required for the new application.

Until this installation Jameson Cell design had largely increased Jg by increasing the air to pulp ratio, (APR). Experience in other applications indicated that low APR values would not affect recovery provided the Jg was maintained. Later work at the University of Newcastle by Evans, Atkinson and Jameson (1995) subsequently confirmed theoretical aspects. Lower APR values would allow tank diameters to be reduced, and also allowed operation with lower frother addition. Reduction of APR from an initial value of 0.3 to 0.2, while maintaining Jg at 1.2cm/sec allowed the diameter required to be reduced from 4.0m to 2.75m.

Two 2.75m cells were still unable to fit into the allowable space. To overcome this problem the 4000/16 design was altered to produce a cell capable of two stages of flotation in the one shell. This was achieved by the use of additional baffles in the feed distributor, tank, internal and external launder - creating the 4000/16 2 stage Jameson Cell, Figure 4.

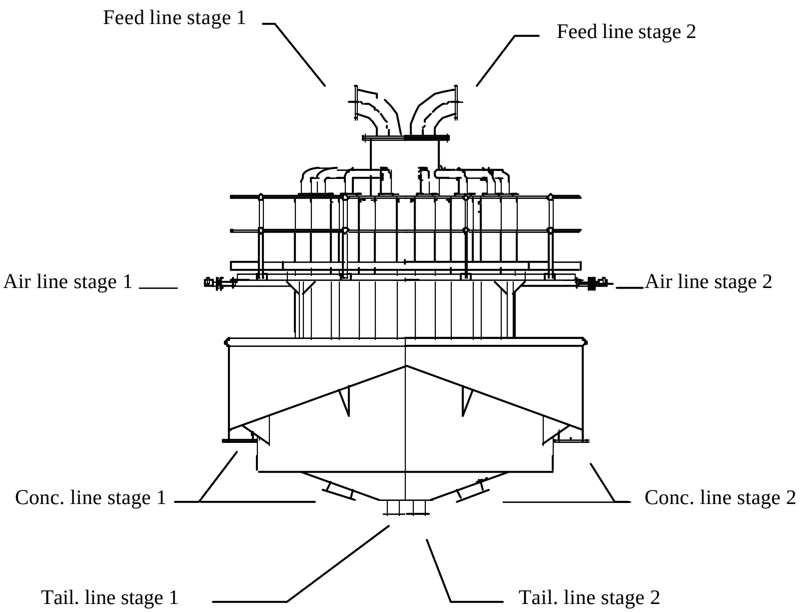


Figure 4. 4000/16 two stage Jameson Cell.

In March 1993 the Jameson Cell for the cleaner scavenger circuit was ordered. Fabrication of downcomer parts was done in Australia with steel work being built in the Philippines. The tank and platform were fabricated in two halves to allow movement into the concentrator. Equipment arrived on site in early August, 1994, with installation being completed three weeks later.

Commissioning

Commissioning commenced on August 30, and a number of problems, as detailed below were encountered.

- Height restrictions resulted in the cell being installed one metre lower than envisaged. This caused tailing pipe runs being close to the horizontal with subsequent sanding and restricted tailing flow, with pulp overflowing to concentrate, (Harbort, 1994).
- Bubble probes installed for tank level control were affected by larger than anticipated density fluctuations. They were replaced with float type level indicators.
- An instrumented control for feed pump box level control resulted in pump boxes and the Jameson Cell being hydraulically linked. Pump box level control was later achieved by placing tailing into a junction launder, one end of which flowed into the pump box, maintaining a constant level, the other end being directed to tailing.
- By mid September, after several modifications, recoveries across both stages was still only 50%, Figure 5. Investigation showed the problem was caused by the irregular trial use of a chemical defoamer. Used to suppress tenacious froth it resulted in severe loss of froth formation and extremely unstable Jameson Cell operation. Its use was discontinued and copper recovery increased to 75%.
- Several days later, corrosion in the recleaner feed box required running repairs to be conducted, limiting recoveries in the cleaner. After repairs were completed the cleaner scavenger copper recovery was increased to 85%.
- Reduced elevation had resulted in internal launders being too low to feed to respective areas of the plant and had been closed off. This caused dead zones to develop in the froth on the inner area of the cell. To overcome this piping was installed between the internal and external launder. This resulted in an eight percent increase in copper recovery to 93%, which was above the targeted value.

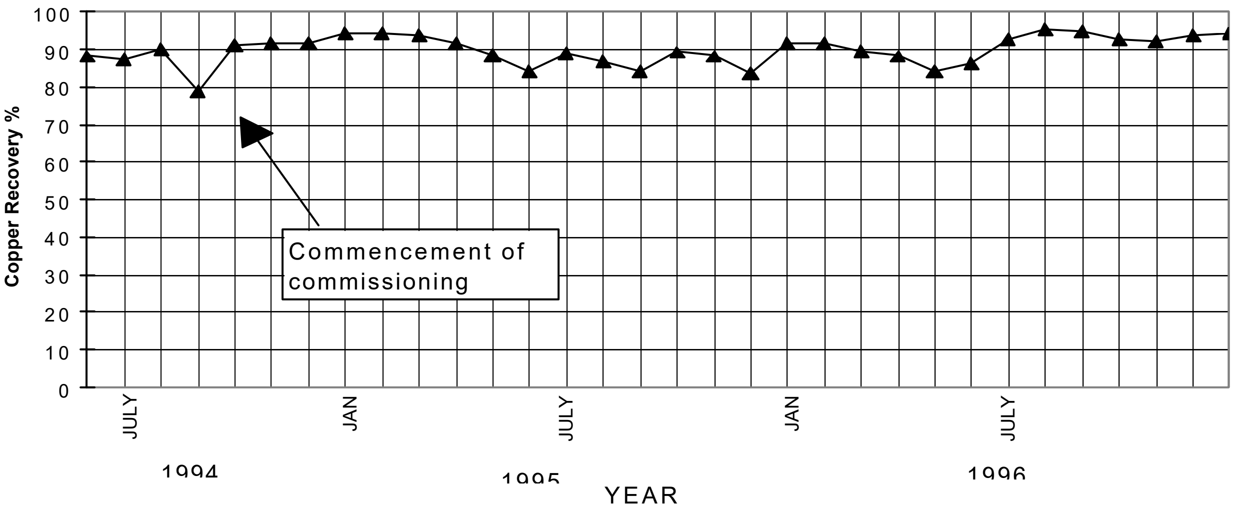


Figure 5. Cleaner scavenger copper recovery.

Size by size analysis

Figures 6 and 7 detail size by size recoveries of copper and gold across the 4000/16 Jameson Cell. The highest recoveries for both were achieved in the minus 74 micron fraction, being 93.4% for copper and 92.5% for gold. Recoveries decreased only marginally with increasing coarseness. At the coarsest size of plus 300 micron, copper recovery was 82.5% with gold recovery of 82.3%. An aberrant drop in recovery for both copper and gold was noted at the 149 micron size. The d_{80} of feed material was 75 micron.

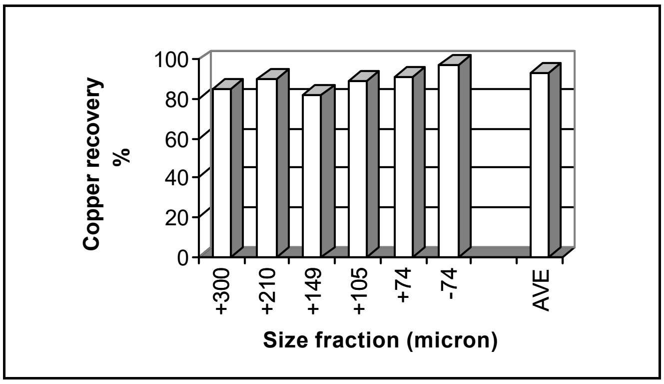


Figure 6. Cleaner scavenger copper size by size recovery.

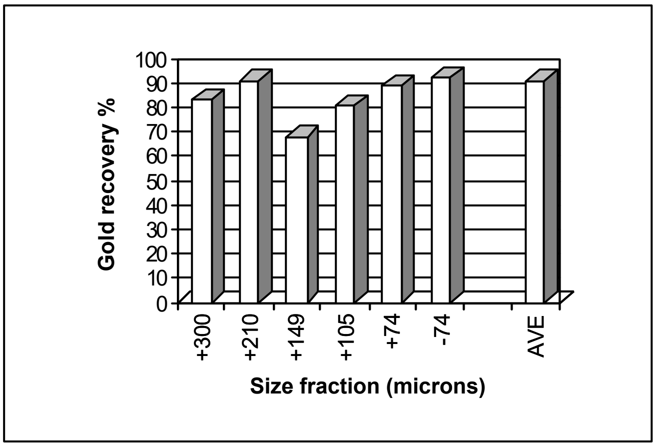


Figure 7. Cleaner scavenger gold size by size recovery.

Jameson Cell and mechanical cell comparisons

Table 1 details key comparisons between the Jameson and mechanical cleaner scavenger circuits.
As previously discussed the major justification for replacement of the mechanical cleaner scavenger was to create sufficient area for the replacement of the main roughers and scavengers. The Jameson Cell cleaner scavenger circuit achieved equivalent metallurgical performance, operating with eight percent of footprint area and 9.4% of the residence time of the mechanical circuit.

Power consumed by the Jameson Cell circuit at 197.8kW is half of the previous circuit. This is slightly above the anticipated power (Garcia, 1993), however the more stable operation produced has reduced the cleaner/recleaner power from its original consumption of 150kW, (Garcia, Sevilliano, Smith, 1995) to 112.7kW (Malicse, 1996), a fall of 37.3kW.

Orifice plate usage had initially been predicted at two sets of ceramic per year. The initial steel orifice plates, with an anticipated life of two months remained in service for nine months and the first set of ceramic orifice plates have been in use for 24 months.

Table 1. Comparison of Jameson and mechanical cell cleaner scavengers.

Comparison	Jameson Cells	Mechanical Cells
Power consumption - kW	198	393
Foot print area - sq metres	30	375
Residence time - minutes	2.9	30

CLEANER CIRCUIT UPDATE

The commissioning of the Jameson cleaner and recleaner cells have been discussed elsewhere (Harbort et al, 1993; Harbort, Sevilliano, Chui, 1994) and this paper presents an update on operation, since commissioning in 1993.

Froth formation

The first Jameson Cells installed at Philex were generally very well accepted by operators. One initial operational difficulty was the difference in froth formation between the Jameson Cells and the column/mechanical cells they replaced. The Jameson Cell cleaner and recleaner produce a fine froth, typically with bubble diameters no more than 5mm at the overflow lip. The replaced column by comparison had bubbles of 25mm diameter at the overflow lip. Operators often interpreted the small bubble size as meaning low grade concentrate, when this was not the case. Usual action was to first decrease the air rate to “improve” grade. As found by Atkinson and Espinosa-Gomez (1992) reduction of the air within a Jameson Cell causes an increase in vacuum within the downcomer and typically causes bubble size to get smaller. Visually, reducing air rates made the froth appear even “lower grade”. The next action was to increase wash water to “improve” grade. This produced a “wetter”, faster flowing froth, again interpreted as lowering grade, when in fact the opposite occurred.

The targeted concentrate grade of 25% copper, with equivalent pre-installation recoveries continued to be achieved until June 1995, 18 months after commissioning (Smith, 1996). In June 1995 both a drop in concentrate grade and cleaner circuit recovery occurred, Figure 8. Initial attempts to correct the fall concentrated on improvement in shift operation. This did arrest the fall, but a return to previous levels was not achieved. A detailed audit was then conducted around the Jameson Cells which highlighted a number of possible causes for the deterioration, (Harbort, 1995).

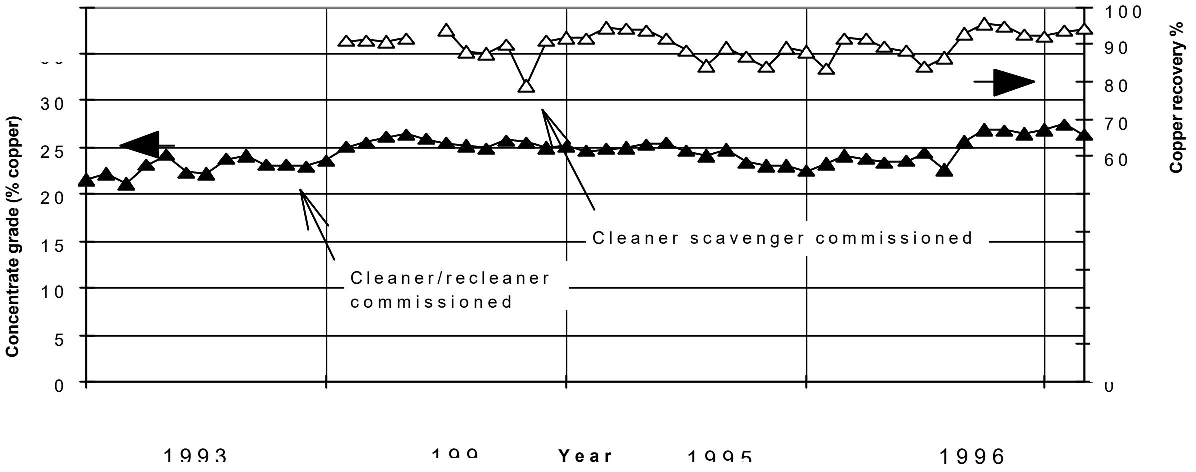


Figure 8. Cleaner circuit copper recovery and concentrate grade.

Feed flow rates

The recleaner cell had been designed to treat a flow of 51m^3 per hour. The actual operating flow had increased to 99.5 m^3 per hour through a combination of lower percent solids, lower cleaner concentrate grades and higher metal content. To cope with this increased flow the recleaner feed pump had been increased in speed and the orifice plate size increased. An investigation of the recleaner showed that a column of pulp was forming between each downcomer and the outer tank wall, giving inefficient froth/pulp separation. Jameson Cell modelling studies by the CSIRO, (Schwarz, 1992) had predicted this could happen, although this was the first time the phenomena has been observed in practice. To overcome the problem the downcomers were relocated inward toward the cell centre.

Carrying capacity

Froth carrying capacity limitations, causing a gradual copper build up in the cleaner circuit, was also considered a possible problem. Initial design indicated a recleaner carrying rate of 12.76g/min/cm^2. This value was routinely exceeded, (Manlapig, 1994), without a deterioration in recleaner performance. Indications from January 1995 were that even at 14.75 g/min/cm^2 carrying capacity was not limited. Although carrying limitations did not cause a high load system to develop, when one existed it did increase the time it took to remove copper from the circuit. Modifications to the cleaner cell were conducted to allow high grade concentrate to be removed from one downcomer area direct to final concentrate. This provided a rapid method for removing copper from the cleaner circuit.

Frother addition

Excessive frother was also diagnosed as a problem. Frother consumption had been gradually increasing over 1995, partly in an attempt to increase recovery and partly due to concerns on frother quality. In November 1995 frother addition to the rougher circuit was decreased by 25% and some improvement in cleaner circuit performance was observed. As the only frother addition was to the rougher circuit there were concerns that decreases in addition to improve concentrate grade would decrease overall recovery and no further reductions were conducted for several months. In July 1996 frother was again decreased by 25%. Within sixteen hours of the reduction cleaner circuit copper recovery improved from 85% to 95% and concentrate grade improved by approximately three percent copper.

A review of frother consumption over 21 months shows a direct relationship between increasing frother consumption and decreasing grade and recovery, Figure 9. Unlike mechanical cells where over addition of frother decreases concentrate grade without affecting recovery, in the Jameson Cell, under certain conditions, recovery can be affected. This was recognised earlier, (Harbort, 1993), but had typically only applied to small test cells. In the Philex cleaner circuit frother saturated feed had caused a “blurring” of the pulp/froth interface causing level indicators to drift. Since the reduction of frother the cleaner circuit has continued to produce record concentrate grades and recoveries.

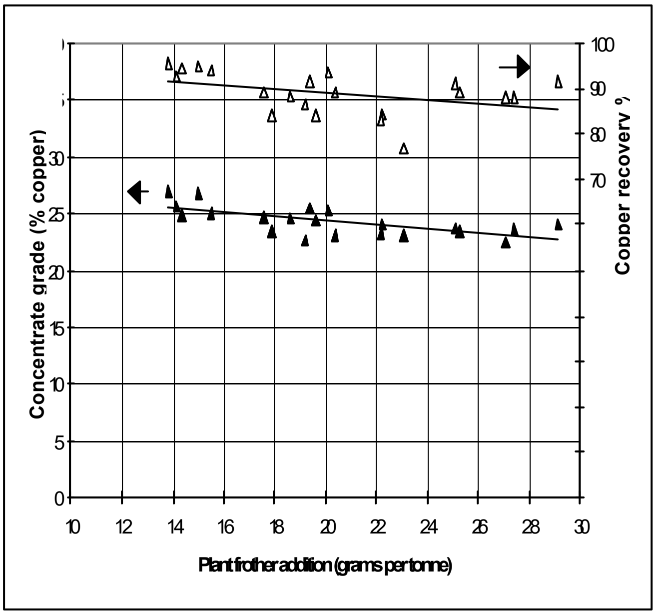


Figure 9. Effect of overall plant frother addition on cleaner circuit performance.

ROUGHER/SCAVENGER INSTALLATION

Design considerations

Where the cleaner scavenger cell required alteration to an existing design, the rougher scavenger Jameson Cells required a new design. Criteria for design included:

- ability to handle coarse feed
- sizing to incorporate space limitations
- efficient feed distribution to downcomers
- operator acceptance

The Philex concentrator operated with three parallel lines of rougher/scavengers and it was a site preference that the Jameson Cell installation would follow this layout. A number of installation options were reviewed.

1. *Six by five metre diameter Jameson Cells* (three for roughing/three for scavenging). This option failed the design criteria in terms of footprint area, lip length and froth travel distance, (Table 2).

2. *Twelve by 3.5 metre diameter cells*. There was operator resistance to this option in that circular cells were considered largely suitable for cleaning but it was felt that rougher/ scavengers should be rectangular.

3. *Twelve by R3330/8 cells*. Rectangular Jameson Cells had been operated at Newlands Coal Pty Ltd on coal fines, (Jameson, Goffinet, Hughes, 1991) and at the Philippine Smelting and Refining Corporation on copper slag scavenging, (Dawson, Harbort, 1996). However, these cells had operated at less than 20tph per cell, compared to the Philex requirement of 200tph per cell. Significant design work on rectangular cells for roughing duties had been done in 1993 Manlapig and this was further developed for the design of the R3330/8 model Jameson Cell for use at Philex.

Table 2. Jameson rougher/scavenger design criteria.

Model	Lip length/area ratio cm/cm^2	Maximum froth travel distance (mm)	Operator acceptance
5000/16	0.96	1000	medium
3500/8	1.46	625	medium
R3330/8	1.33	550	high
Target	>1.0	< 750	high

Design changes

Figure 10 shows a schematic of the R3330/8 Jameson Cell. Changes to the old design included:

- level control using a sand port and overflow weir
- use of a horizontal distributor, with reducing cross sections to maintain pressure
- vertical rather than horizontal feed inlet to downcomers
- use of ceramic at key locations within downcomers
- division of the downcomer into two contacting zones, separated by an aeration funnel that concentrates air around the plunging jet and reduces froth creep to minimise solids build up in air lines.

Twelve R3330/8 Jameson Cells were ordered by Philex in January 1995, with a staggered delivery and commissioning period.

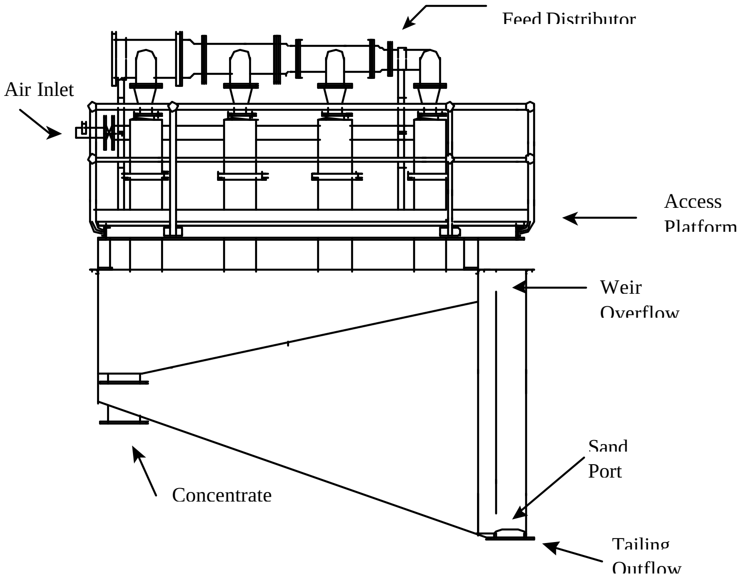


Figure 10. R3330 Jameson Cell.

Commissioning

Significant dates (Harbort, 1996), relating to commissioning were:

- December 7, 1994. Jameson bank C rougher commenced operation.
- February 14, 1995. Jameson Bank C scavengers commenced operation.
- June, 1995. Mechanical bank C de-commissioned.
- September 1, 1995. Jameson bank A commenced operation.
- October 10, 1995. Jameson bank B commenced operation.
- January, 1996. Mechanical banks A and B off line for refurbishment.
- September, 1996. Performance trial conducted showing the Jameson Cells were operating significantly better than predictions from test work.

A number of problems, as detailed below were encountered.

External Recycle Mechanism

Pump box level control to maintain steady feed pressures initially was via an instrumented control scheme. This was designed to keep feed pressures within +/- 10kPa of set points. Initial operation indicated that this was difficult. Jameson Cells operating in SX-EW duties had for several years made use of the Internal Recycle Mechanism (I.R.M.) to obtain stable pressures. (Dawson, Jackson, 1995). This concept was transferred to the rougher pump box at Philex to create the External Recycle Mechanism (E.R.M) as described by Harbort (1995) and Manlapig (1996). The E.R.M. is explained with reference to Figure 11. In simple terms it allows fluid in an intermediate box to flow into the main pump box, to maintain it at a constant level.

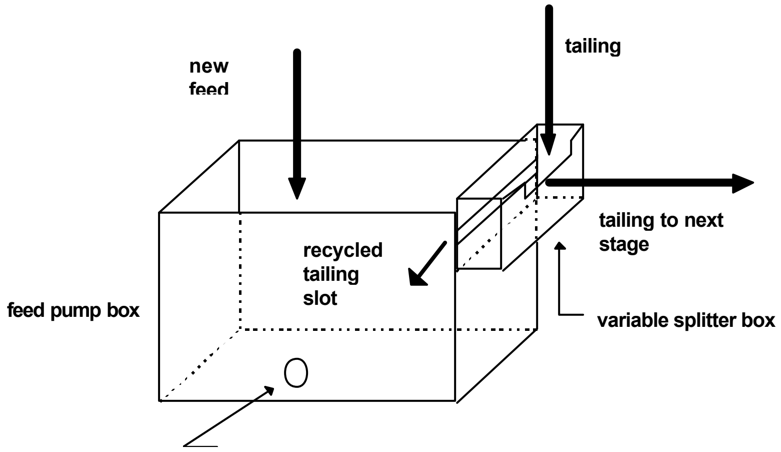


Figure 11. Schematic of External Recycle Mechanism (E.R.M.)

Operation with the E.R.M. commenced on December 16, 1994, with stable pump box level control being achieved with great success. Even wide fluctuations in flow were dampened allowing near constant feed pressures at all times. Figure 12 shows the beneficial effect of the E.R.M. in minimising flow fluctuations. At approximately 2am on September 12, 1996 operations were disrupted by cycloning problems in one of the ball mills. This resulted in varying flows to the flotation circuit, which continued until 11am when the problem was corrected. The short frequency of the flow disruptions resulted in mechanical cell level control being unable to stabilise, with consequential copper losses to tailing. By comparison, these fluctuations were adequately handled by the E.R.M., with no loss of recovery evident, even with manual adjustment on cell level control.

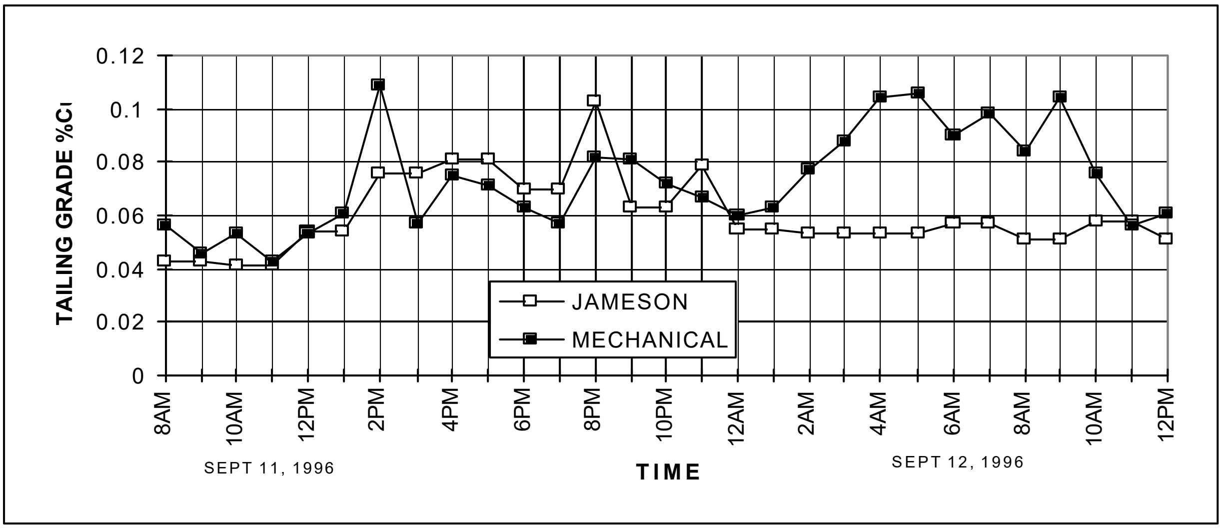


Figure 12. Minimisation of circuit disruptions with the E.R.M.

Froth formation

Once again the difference in froth formation between the Jameson and mechanical roughers was an area that caused some difficulty with operating personnel. Each bank of mechanical rougher/scavengers had a lip length of forty metres, with an average mass flow over the lip of approximately 0.29tph per metre of lip. By comparison, each bank of Jameson rougher/scavengers has a lip length of twenty six metres with an average mass flow over the lip of 0.65tph per metre of lip. Initially, experienced mechanical cell operators had a tendency to run Jameson Cells visually the same as the mechanical cells, with too low a mass pull, giving lower recoveries.

On going performance

Figure 13 shows the average operating performance in 1995/1996 compared with results from the mechanical banks. The difference between expected and actual tailing increases with increasing feed grade. A number of reasons appear to exist for this including improved copper oxide recovery, better scavenging ability and improved fine particle flotation.

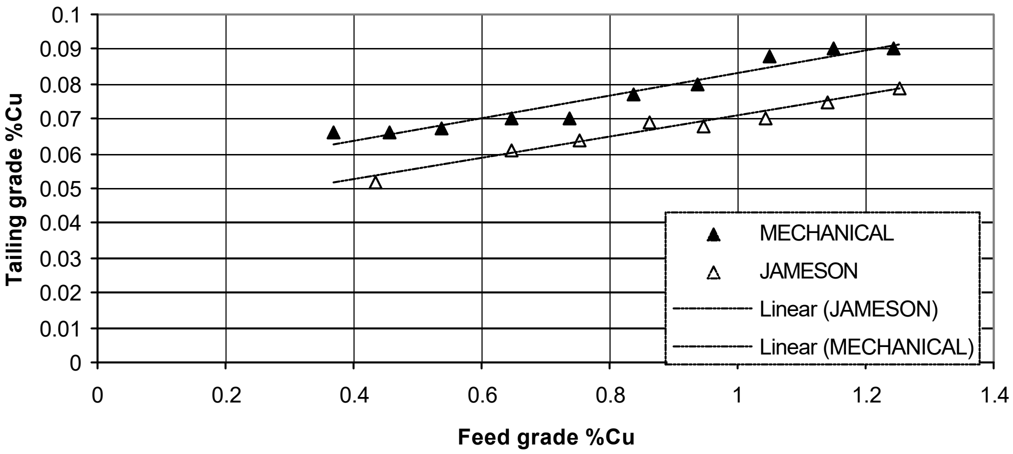


Figure 13. Comparison of Jameson and mechanical rougher scavengers.

Copper oxide content typically increases with increasing rougher feed grade. Immediately following commissioning it became apparent that the Jameson Cells were achieving higher recoveries of oxidised copper than the mechanical banks, Figure 14. During the initial tests feed material contained low copper oxide content and this advantage of the Jameson Cell was not apparent.

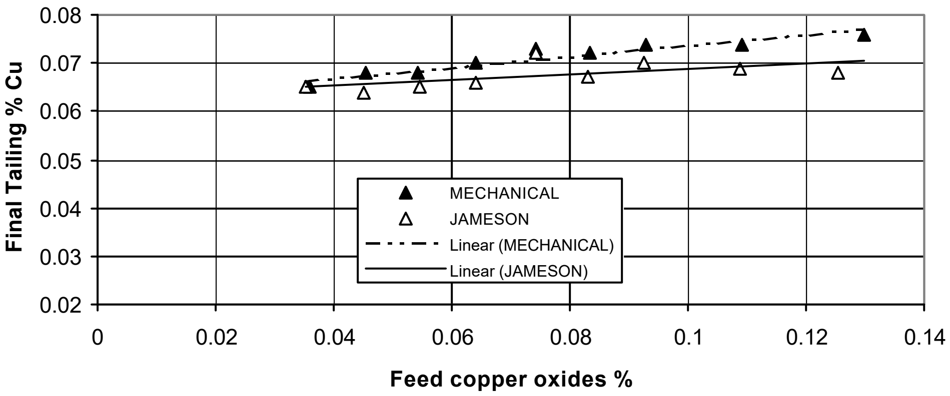


Figure 14. Effect of feed copper oxides on final flotation tailing.

Operation has shown that roughing performance was similar to test work and to that achieved by the mechanical cells. Scavenging in the Jameson Cells however shows a marked improvement over the mechanical cells. Figure 15 details the distribution of tailing results for September/October, 1996, with Jameson and mechanical banks having similar scavenger feed grades. Both operated for a similar period of time in the normal operating range, with the Jameson Cells performing better under minor upset conditions. Of major significance is that the Jameson Cells for twenty percent of the operating time produced a very low scavenger tailing, rarely achieved by the mechanical circuit at Philex. This is possibly due to the intense bubble/particle nature of contact which occurs in the downcomer. This is further highlighted by the fact that the Philex mechanical scavenging banks have twelve contacting stages compared to one for the Jameson scavengers.

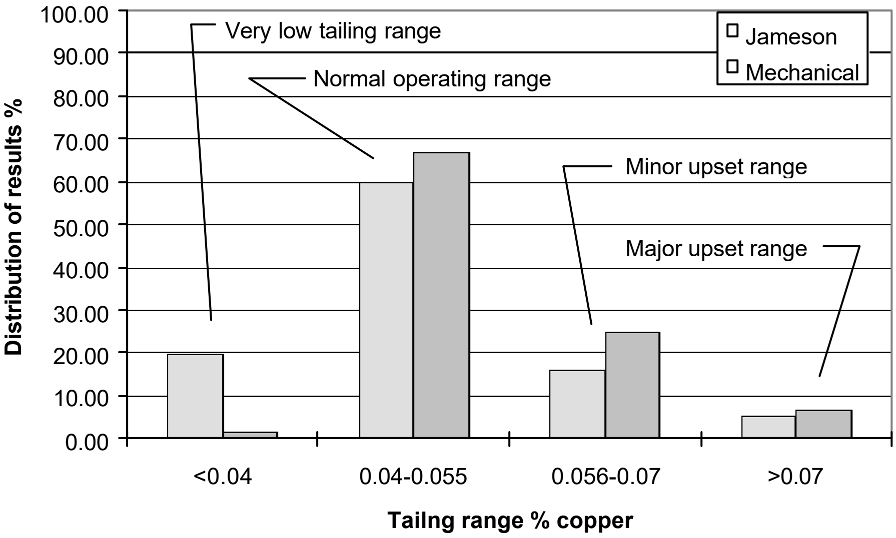


Figure 15. Distribution of tailing results for September/October, 1996.

Size by size analysis

Surveys have been conducted around the Jameson Cells and mechanical banks. These samples were sized by screening and cyclosizing and size fractions assayed to allow a size by size comparison of recoveries, Figure 16. The results show that the Jameson Cells and mechanical banks achieved similar recovery of copper in all size fraction greater than 14 micron. In the 10 micron to 14 micron size fraction the Jameson Cell achieved

2% additional copper recovery over the mechanical cells. For the minus 10 micron slimes fraction the Jameson Cells achieved 6.72% additional copper recovery. The samples were further analysed via QEM SEM, (CSIRO, 1996), showing losses of chalcopyrite to tailing in the +106 micron size range were as unliberated gangue binary composites. Liberated chalcopyrite up to 212 micron in size was observed in concentrate. By comparison, chalcopyrite losses in the 10 to 14 micron fraction were largely free particles, (60.2% of chalcopyrite lost), with 21.0% lost as composites with pyrite and only 16.8% being lost as gangue composites. It is likely that -10 micron material (unanalysed), would exhibit a similar form of loss. As such 20% of total copper losses were as free particles in the slimes fraction.

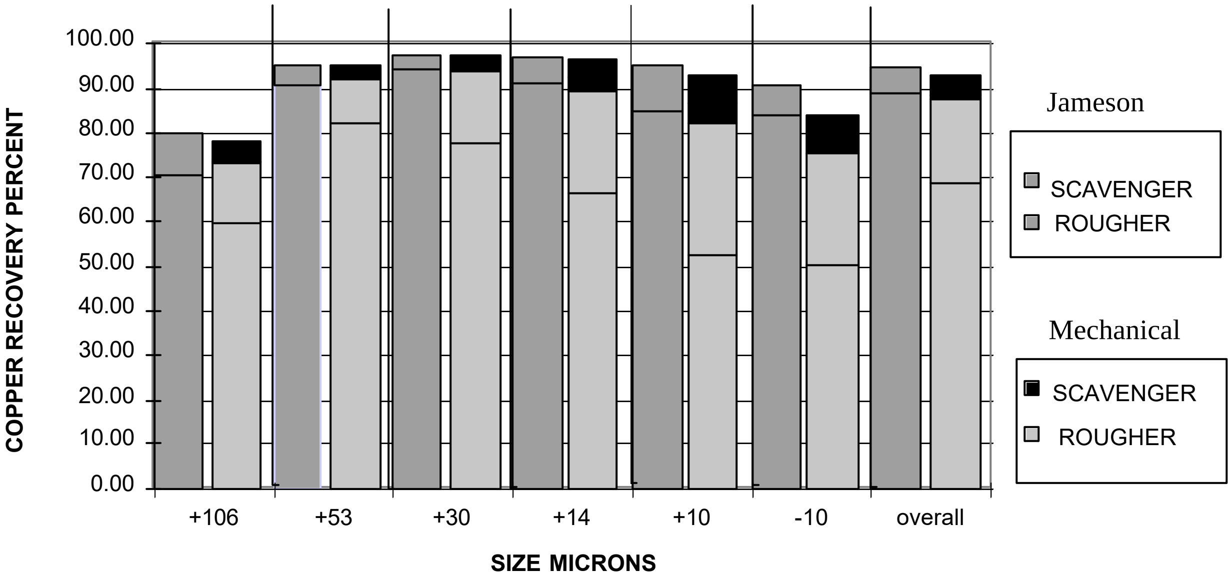


Figure 16. Copper recovery by size fraction.

Figure 17 details the effect of the Jameson Cell rougher installation on overall plant copper tailing. Early 1994 represents the plant's previous best performance in recent times where a tailing of 0.059% copper were achieved. The average tailing for the six months prior to Jameson C bank commissioning was 0.069% copper. Following completion of commissioning and during the early part of 1996 when mechanical banks were off line for refurbishment, the Jameson Cells, operating alone, achieved an average tailing of 0.064% copper. Peaks in tailing grade after this relate to periods where significant amounts of ROM spillage material was treated. Following normalisation of ore supply, tailing levels fell to 0.055% copper. With increasing confidence and expertise Philex have managed to reduce plant tailing further to an October 1996 average of 0.046% copper.

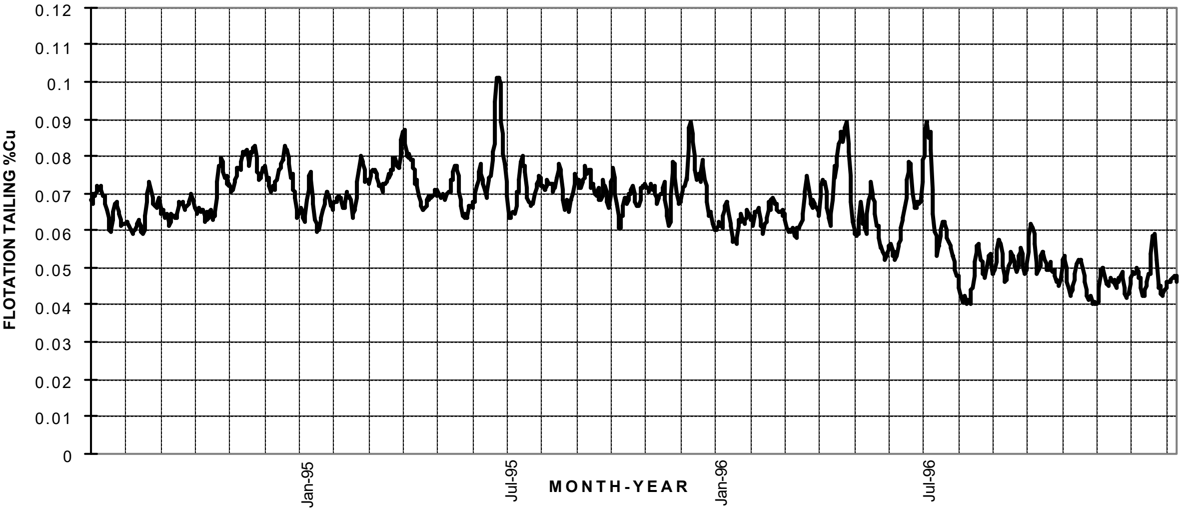


Figure 17. Plant copper final tailing.

In addition to improved copper recovery a significant increase in gold recovery, above that indicated by gold floating in sulphides, was observed. At typical plant recoveries the Jameson Cells improved gold recovery by approximately one percent. This appears in part due to elevated Jameson Cell recovery of free gold during plant upset conditions, Figure 18.

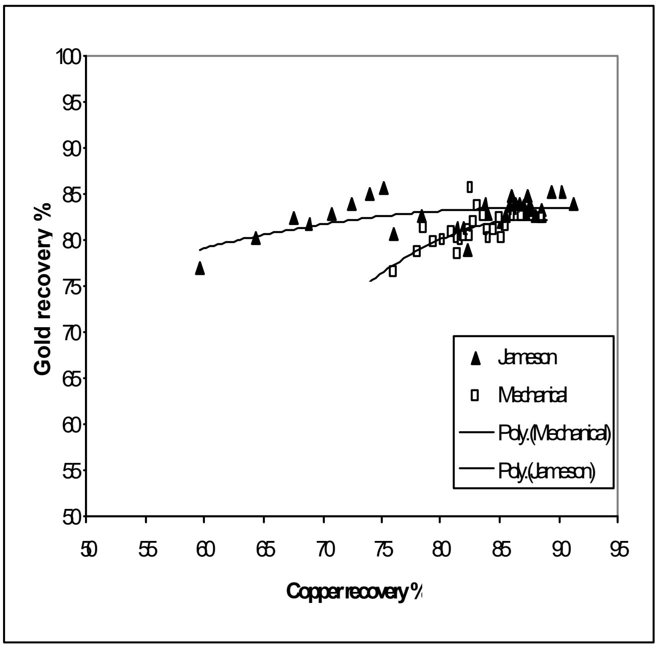


Figure 18. Copper recovery versus gold recovery.

Performance trial

The copper tailing grade achievable from test work was based on a sliding scale, largely governed by the copper feed grade to the roughers. In September 1996, a performance trial at rated capacity was conducted to confirm operating performance. For the trial, one Jameson bank was fed from milling line no 2, with milling line no 1 feeding to mechanical banks A and B, which acted as a control. Key parameters for the trial are shown in Table 3. (Harbort, Murphy, 1996).

During this trial the tailing achieved was 0.01% copper less than that anticipated.

Table 3. Performance trial key parameters

Parameter	**Line 1** **Control**	**Line 2** **Jameson A bank**
SIBX addition (g/t)	24.04	24.04
Frother addition (g/t)	18.35	15.2
Approx. pulp density (kg/*l*)	1.28	1.36
Grind (% passing 200#)	67.5	67.2

Jameson Cell and mechanical cell comparisons

Table 4 details other comparisons between the Jameson and mechanical rougher/scavenger circuits. The Jameson circuit occupies 60% less floor area and achieves equivalent results to the mechanical banks with 40% of their residence time. This is achieved with a power saving of 18%

Table 4. Comparison of Jameson and mechanical cell cleaner scavengers.

Parameter	Jameson Cells	Mechanical Cells
Power Consumption - kW	1396	1702.5
Foot print area - sq metres	360	900
Residence time - minutes	7.46	17.9

CONCLUSIONS

Philex Mining Corporation has successfully conducted a major installation of Jameson Cells into cleaner and rougher sections of their Banget concentrator. Indicated metallurgical improvements have been:

Cleaner/Recleaner	• 2.6% increase in copper concentrate grade • 3.5% increase in plant copper recovery • 2.6% increase in plant gold recovery
Cleaner Scavenger	• 1.5% increase in cleaner circuit copper recovery
Rougher/Scavenger	• 3.3% increase in copper recovery • 4.5% increase in gold recovery

REFERENCES

Atkinson, B W, Espinosa-Gomez, R, 1992. The Jameson Cell - Effects of downcomer length, feed pressure and air-pulp ratio. University of Newcastle report issued to Mount Isa Mines Limited.

CSIRO, 1996. Philex Concentrator - Jameson Cell rougher and scavenger circuit analysis. CSIRO Division of Minerals report issued to MIM Holdings Limited.

Dawson, W J, Harbort, G J, 1996. Jameson Cell recovery of copper from smelter slags, in Proceedings of the SAIMM Hidden Wealth conference, 2 - 4 October, 1996.

Dawson, W J, Jackson, B R, 1995. Evolution of Jameson Cells for solvent extraction applications, in Copper Hydrometallurgy Forum, Sept 18-19, 1995.(ALTA:Melbourne)

Evans, G M, Atkinson, BW, Jameson, G J, 1995. The Jameson Cell, *in Flotation Science and Engineering* (Ed:KA Matis), pp 331 - 363, Marcel Dekker, NY, 1995.

Garcia, T F, 1993. Economic benefits that can be derived from the operation of Jameson Cells at Banget Concentrator. Philex Mining Corporation internal report.

Garcia, T F, Sevilliano II, A C S, Smith, G A L, 1995. Economic benefit studies on the various Jameson Cell installations. Philex Mining Corporation internal report.

Harbort, G J, 1992. Pilot plant test work at Philex Mining Corporation. MTM report no JC45A/9209, issued to Philex Mining Corporation.

Harbort, G J, 1993. Companion guide to Jameson Cell operating software. MIM Holdings Limited general release.

Harbort, G J, 1994. Commissioning of middling circuit Jameson Cell at Philex Mining Corporation. MTM report no IS113\9411, issued to Philex Mining Corporation.

Harbort, G J, 1995a. External recycle mechanism for pump box level control. MIM Holdings Limited internal report.

Harbort, G J, 1995b. Flotation Circuit Review. ML Consultants report issued to Philex Mining Corporation.

Harbort G J, 1996. Commissioning of roughers and scavengers at Philex Mining Corporation. MTM report no IS133A/9601 issued to Philex Mining Corporation.

Harbort, G J, Jackson, B R, Jameson G J, Malicse, TB, Sevilliano II, A C S, 1993. Jameson Flotation Cell Operation at Philex Mining Corporation, paper presented at PSEM Symposium on Metallurgy, Manilla, November, 1993.

Harbort, GJ, Murphy, AS, 1996. Capacity and performance trial of Jameson Cell roughers and scavengers at Philex Mining Corporation. MIM Holdings Limited report no IS113A/9609 issued to Philex Mining Corporation.

Harbort G J, Sevilliano II, A C S, Chui, A V, 1994. Optimisation of the Philex Mining Corporation Concentrator with the Jameson Cell, in Proceedings of the AusIMM Fifth Mill Operators Conference, pp 37-42, (The Australasian Institute of Mining and Metallurgy: Melbourne)

Jameson, G J, , Goffinet, M, Hughes, D, 1991. Operating experiences with Jameson Cell at Newlands Coal Pty Ltd, in Proceedings of the Fifth Coal Preparation Conference, pp 146-159, (The Coal Preparation Society)

Jameson, G J, Manlapig, E V, 1991. Applications of the Jameson Flotation Cell, in Column ’91 - Proceedings of an international conference on column flotation (Eds Agar, Huls and Hyma).

Manlapig, E V, 1993. Jameson Cell - 2nd Generation flotation cell. MIM Holdings Limited internal report.

Manlapig, E V, 1994. Analysis of Jameson Cell performance - Philex cleaner/ recleaner. MIM Holdings Limited internal report.
Manlapig, E V, 1996. Scale model testing of pump box and launder (ERM Box). MIM Technologies internal report.

Schwarz, M P, 1992. Hydrodynamic simulation of the pulp zone of the Jameson Flotation Cell: Parametric studies. CSIRO report REP722 issued to MIM Holdings Limited.

Smith, G A L, 1995. Personal communication - November.

JAMESON CELL SCALE-UP EXPERIENCES AT BHP COAL'S GOONYELLA COAL PREPARATION PLANT

Michael F Carretta, (Process Engineer), James N Graham (Process Engineer)
BHP Coal Pty Ltd, Goonyella/Riverside Mine, Moranbah, Qld, 4744, Australia

Wendy J Dawson, (Metallurgist - Jameson Cell)
MIM Holdings Ltd, GPO Box 1433, Brisbane, Qld, 4001, Australia

Presented at: Coal Prep 97
April 29 - May 1
Lexington, Kentucky

ABSTRACT

In 1993, BHP Coal investigated several different flotation technologies for the replacement of the ageing mechanical (sands) flotation cells in it's Goonyella coal preparation plant. The study included pilot scale testing using two 0.5 metre (1.6') diameter Jameson Cells, having a throughput of 7.5 m^3/h (33 USGPM). The results showed that the cell could deliver superior metallurgical results compared to the mechanical flotation cells. In the following year, BHP Coal decided to replace the mechanical cells with Jameson Cells installed in a two stage (rougher - scavenger) configuration.

In 1995, BHP Coal initiated a project to confirm the ability of Jameson Cells to be scaled up from pilot to industrial scale. The aim of the study was to determine whether the full scale downcomer could deliver the results predicted by test work with a smaller sized downcomer. For the scale-up confirmation study, BHP Coal purchased two demonstration scale Jameson Cells, each having a diameter of 1.25 metres (4'), and a full-scale, 60 m^3/h (264 USGPM) capacity downcomer. The size of the demonstration cell's tank was calculated using parameters determined during the pilot scale test work. The results from the demonstration scale trials were found to be in line with the pilot scale results.

Having demonstrated the scale up from the pilot scale Jameson Cell to the demonstration scale cells with industrial scale downcomers, BHP Australia Coal proceeded with the installation of eight model 5000/16 cells (16.5' diameter cell with 16 downcomers) in two-stage, in series, configuration. These industrial scale Jameson Cells, each treating 960 m^3/h (4227 USGPM) used the same size downcomer as the demonstration cell. The installation of the Jameson Cells at Goonyella was part of a major fines circuit upgrade.

Commissioning of the refurbished fines circuit was completed in two weeks, with maximum throughputs achieved shortly thereafter. The full-scale Jameson Cell optimisation data are seen to closely match the pilot scale performance data. The results show that the pilot scale tests were able to accurately predict the performance of the full-scale Jameson Cells.

Several months of Jameson Cell operating data show that the full-scale installation has consistently been able to deliver the results predicted by the pilot scale testing. The Jameson Cell's performance is also demonstrated to be much less variable than that of the preceding mechanical flotation cells.

The ability of the Jameson Cell to deliver a consistently low ash product has allowed the recoveries from the coarse circuit to be increased, thereby paving the way for a 3.5% increase in plant yield, and the setting of BHP Coal production records.

INTRODUCTION

Goonyella / Riverside Mine is one of seven mines managed and operated by BHP Coal Pty Ltd (formerly BHP Australia Coal Pty Ltd) in central Queensland's Bowen Basin. While Goonyella (owned by Central Queensland Coal Associates) and neighbouring Riverside (owned by BHP Mitsui Coal) were amalgamated in 1989, the mines are on separate leases, and have their own coal preparation facilities. The two operations combined will produce approximately 10.7 million tonnes of coal this financial year, 7.2 million tonnes being from the Goonyella operation (Gosnell, 1996).

The Goonyella mine and coal preparation plant commenced operations 25 years ago, with an output of 4.4 million tonnes per year. In the original plant, the feed stock was split into two process streams:

- coarse coal (+0.5 mm wedge wire), comprising 80% of the plant feed, which was treated by six modules of heavy media cyclones, and

- fine coal (-0.5 mm wedge wire) which was treated by "sands" flotation in a common circuit consisting of four banks of 4 x 300 cubic foot Wemco mechanical cells. The ultrafines were discarded.

Subsequent modifications included the addition of a slimes flotation stage (1979), the incorporation of tailing scavenging spirals (1985) and a one third plant expansion in 1987 to increase the production rate to 6.7 million tonnes per year.

By August 1993, when Goonyella sold its one hundred millionth tonne of coal, the coal preparation plant could treat 1800 tonnes per hour, and consisted of eight heavy medium cyclone modules, split flotation consisting of four banks of 4 x 300 cubic foot Wemco cells for sands flotation and four banks of 4 x 300 cubic foot Wemco cells for slimes flotation, and scavenging spirals. Overall, the flotation circuit typically gave products with ash contents in the order of 9 to 10%, with combustibles recoveries ranging between 71 and 83%. However, the coarse flotation cells were severely corroded, and required replacement. New flotation technologies were investigated for this replacement due to the potential to increase the overall plant yield by achieving lower ash products from the flotation circuit.

In 1993, BHP Coal initiated an intensive on-site test work program to investigate the benefits to the Goonyella coal preparation plant of some novel, high intensity flotation technologies, namely:

- Jameson Cell, (MIM Holdings Ltd)
- Microcel, (Bulk Materials Coal Handling) and
- Centrifloat High Rate Cell (Century Oils).

The Jameson Cell, is a pneumatic flotation cell which has no moving parts and no requirement for compressed air or sparging mechanisms. A technical description of this flotation device is given elsewhere, (Evans *et al*, 1995).

TEST DETAILS

Pilot test work

Pilot scale models of the Jameson Cell, Microcel and Centrifloat High Rate cells were trialed in open circuit, with the following feed streams taken directly from the preparation plant:

- full flotation feed, 0.5 millimetre wedge wire x 0 coal (see Table 1),
- slimes flotation feed, 75 micron x 0 (minus 200 mesh) coal, and
- tailings thickener underflow.

Table 1 - Flotation feed characteristics

Size Fraction Mesh No. (nom. aperture size)		Mass (%)	Ash (%a.d.)
+16 mesh	(+1mm)	0.5	19.0
-16 +30 mesh	(-1 +0.5mm)	12.1	14.0
-30 +60 mesh	(-0.5 +0.25mm)	21.1	14.0
-60 +100 mesh	(-0.25 +0.15mm)	13.7	15.2
-100 +250 mesh	(-0.15 +0.063mm)	13.2	18.2
-250 mesh	(-0.063mm)	39.4	39.0

The flotation feed stream typically has an ash content in the range of 18 to 25%.

A mobile Jameson Cell pilot plant was used for the on-site test work at Goonyella. The pilot plant included two model JC500/1 Jameson Cells, which have a diameter of 500 mm (1.6'), fitted with one 100 mm (4") downcomer. The cells had an individual cell capacity of 7.5 m^3/h (33 USGPM), and could be operated as single stage, two stage, or single stage with partial tailing recycle. The various Jameson Cell configurations are illustrated in Figure 1.

Demonstration scale Jameson Cells

Jameson Cell flotation tests were performed using demonstration scale Jameson Cells to confirm that full scale 300 mm (12") diameter downcomers rated at 60 m^3/h (264 USGPM) could replicate the results of the pilot scale equipment. The demonstration scale Jameson Cell had a diameter of 1.25 m (4.1'), and had a section of the circumferential cell lip blanked off to ensure that the product mass rate per unit of lip length would be the same as for the full scale Jameson Cells which were proposed for installation at Goonyella. Two cells were skid mounted, with the intention of operating in a two-stage configuration.

Due to reagent conditioning constraints, the two-stage Jameson Cell configuration was difficult to operate. As a result, the demonstration scale trials were performed with one cell only operated in partial tailing recycle mode. The pilot scale test work showed that operating one cell in partial tailing recycle mode approximated the performance of two-stage operation.

The four-day demonstration cell test work campaign was undertaken at a time when the feed to the plant was described by site personnel as oxidised coal, and consequently, plant flotation performance was adversely affected. The feed coal was suspected to have been heat affected. Additionally, the flotation feed solids content was higher than usual (14 to 18%, compared to 10 to 12%).

Full scale Jameson Cells

Eight model 5000/16 Jameson Cells (16.5' diameter with 16 off 12" downcomers per cell) were installed in two-stage configuration, (four primary cells followed by four secondary cells) as shown in Figure 2. The specified flotation feed was 3800 m^3/h (16,700 USGPM) slurry containing 9.2% solids by weight.

Jameson Cell scale-up is performed using simple algorithms and the values of key parameters determined during test work. The model 5000/16 Jameson Cell was designed directly using optimised parameters from the pilot scale test work. The operating strategies and reagent usage were also determined during the pilot scale test work. Operating variables such as air flow rate, froth depth and wash water addition can be adjusted to achieve performance targets.

The Jameson Cell installation was part of a major fines circuit upgrade which also included design and construction of a coal thickener, building an annexe on the existing plant to house the Jameson Cells, upgrading the control system and relocating several items of equipment in the plant.

Figure 1 - Jameson Cell configurations for fine coal flotation

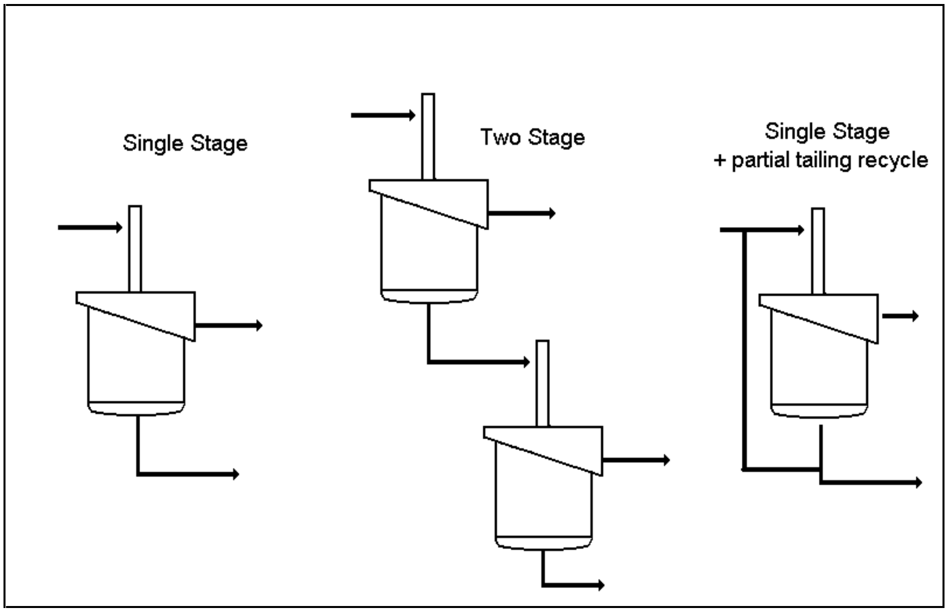
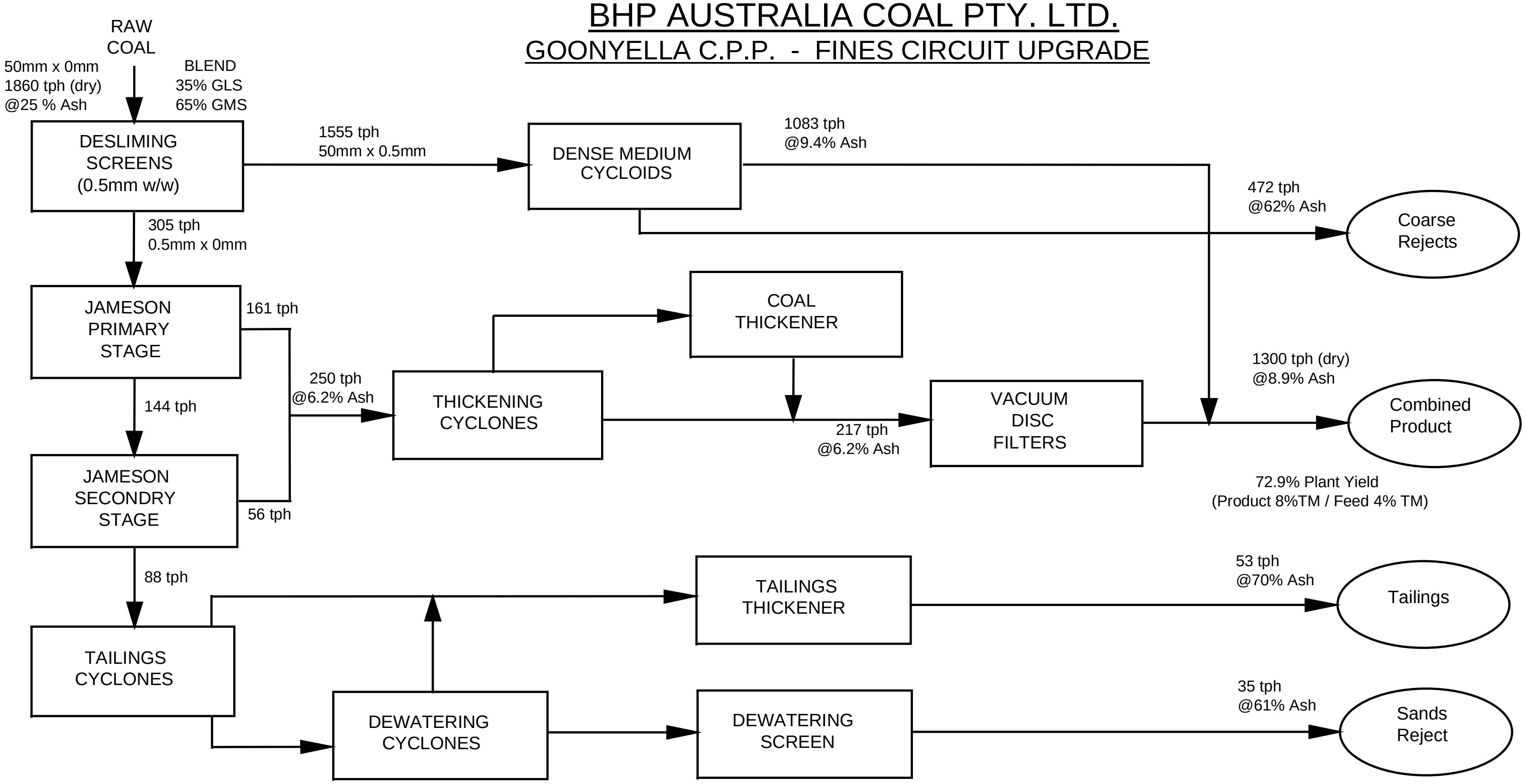


Figure 2

BHP AUSTRALIA COAL PTY. LTD.

GOONYELLA C.P.P. - FINES CIRCUIT UPGRADE



RESULTS AND DISCUSSION

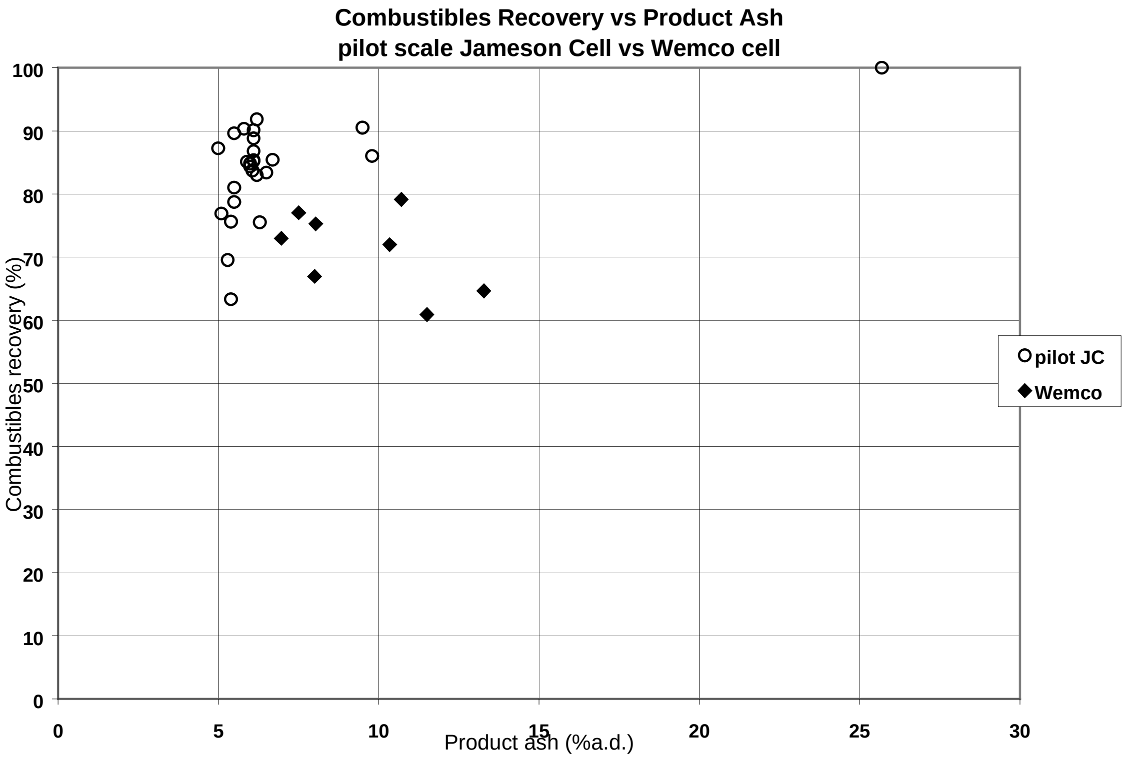
Pilot scale test work and evaluation

The pilot scale flotation test work showed that the Jameson Cell could achieve a significantly lower ash product than the existing Wemco flotation cells, while maintaining or increasing the combustibles recovery. Figure 3 illustrates the performance of the pilot scale Jameson Cell when treating the flotation feed stream, compared to the combined performance of the Wemco sands and slimes flotation cells. The Wemco data was obtained by an independent party performing flotation surveys.

During the pilot scale test work, the results from two stage Jameson Cell flotation were observed to be very similar to the results from tests performed with a single stage with partial tailing recycle. This is a typical result, and can lead to substantial equipment cost savings in full scale installations.

The test results from the three novel flotation cell technologies and operating data from the mechanical flotation cells were included in a comprehensive evaluation process, which examined capital and operating costs, forecast increases in revenue, operability and technical risk.

Figure 3



After rigorous analysis, BHP Coal decided to install Jameson Cells to treat the full flotation feed stream. The Jameson Cells were to be configured as two-stage units, with the tailing from the first stage being treated in the second stage. Although single stage Jameson Cells incorporating partial tailing recycle achieved results comparable to the two-stage cells, the latter configuration was selected, as it had been successfully installed and operated at BHP Coal's Blackwater Mine and MIM Holdings' Newlands operation (Manlapig *et al*, 1993). At that stage, single stage operation with partial tailing recycle had an element of risk, as the Goonyella pilot testing represented the first thorough investigation of this mode of operation, and there were no industrial scale Jameson Cells installed in this configuration. However, this Jameson Cell configuration has subsequently adopted by several coal preparation plants, (Dawson *et al*, 1996 and Le Page *et al*, 1995).

The Jameson Cell technology was selected over the competing technologies because of (Carretta, 1996):

- the lower capital cost for full scale installation,
- lower operating costs (power, reagents) and maintenance costs,
- the scale-up from the pilot rig had been proven at BHP Coal's Blackwater Mine,
- the Jameson Cell having achieved the same combustible recoveries and product ashes as the other technologies, and
- the Jameson Cell test rig was the easiest to operate and was insensitive to feed fluctuations of percent solids and feed rate.

The ability of the Jameson Cell to produce a low ash product meant that the recovery from the coarse circuit could be increased (at a slightly higher ash content), thereby increasing the plant's production rate while meeting the product ash specifications.

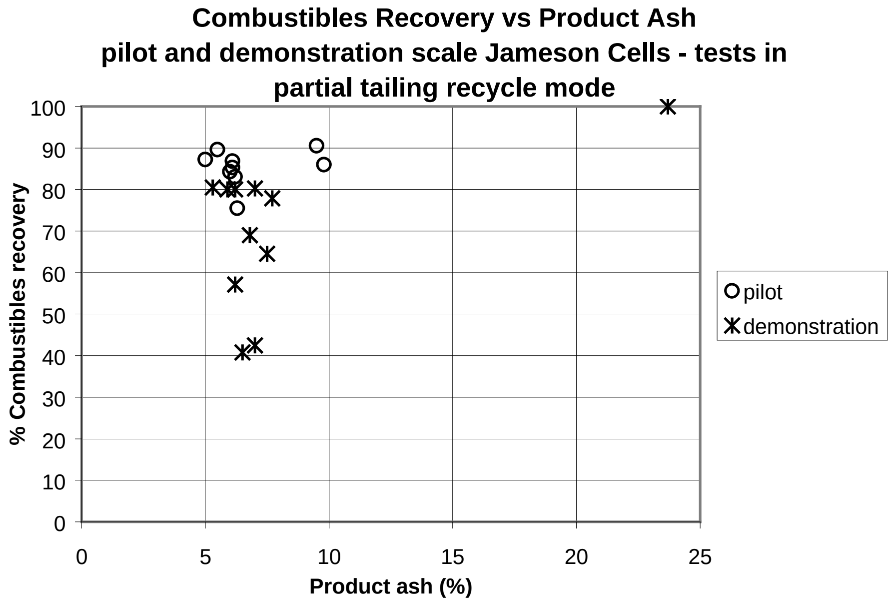
Demonstration scale test work

The combustibles recovery versus product ash results from the demonstration scale flotation tests were generally in line with those of the pilot scale tests, although at reduced combustibles recoveries, (Figure 4). The performance of the plant's Wemco cells when treating the poor quality feed used during the demonstration scale test work was also poorer than usual.

The demonstration cell test program was terminated prematurely when the feed type to the plant changed to a partially oxidised Riversids middle seam.

High feed solids contents are known to adversely affect the performance of the Jameson Cell, particularly when operating near carrying capacity limitations (Espinosa-Gomez *et al*, 1988). Therefore, the abnormally high feed solids content may have caused a limitation in the combustibles recovery. Given the less than ideal conditions under which the demonstration scale test work was performed, the results were viewed as encouraging.

Figure 4



Full scale

Commissioning of the upgraded fines circuit occurred in December 1995. The commissioning was completed in two weeks, with major difficulties being maintenance short-falls and operational problems with the one metre (3.3') diameter tailing classifying cyclones, the coal thickener, and the new reagent dosing system. The plant was able to operate at its maximum throughput of 2000 t/h (an increase over the pre-upgrade throughput of 1800 t/h) six weeks after re-commissioning the fines circuit.

Figure 5 illustrates the combustibles recovery versus product ash results of the full scale Jameson Cell optimisation trials, which were performed after commissioning, to the pilot scale test work results. The ability of the pilot scale test work to accurately predict the performance of a full scale Jameson Cell is demonstrated here.

The full-scale Jameson Cell's operating parameters such as reagent addition and froth depth were determined by pilot test work, and directly translated to full-scale operation.

The consistent performance of the full scale Jameson Cell installation is illustrated by Figure 6, in which one year's Wemco shift data is compared to approximately six months of Jameson Cell shift data. The Jameson Cell is demonstrated to achieve a consistently low ash product in spite of variations in the feed properties.

Figure 5

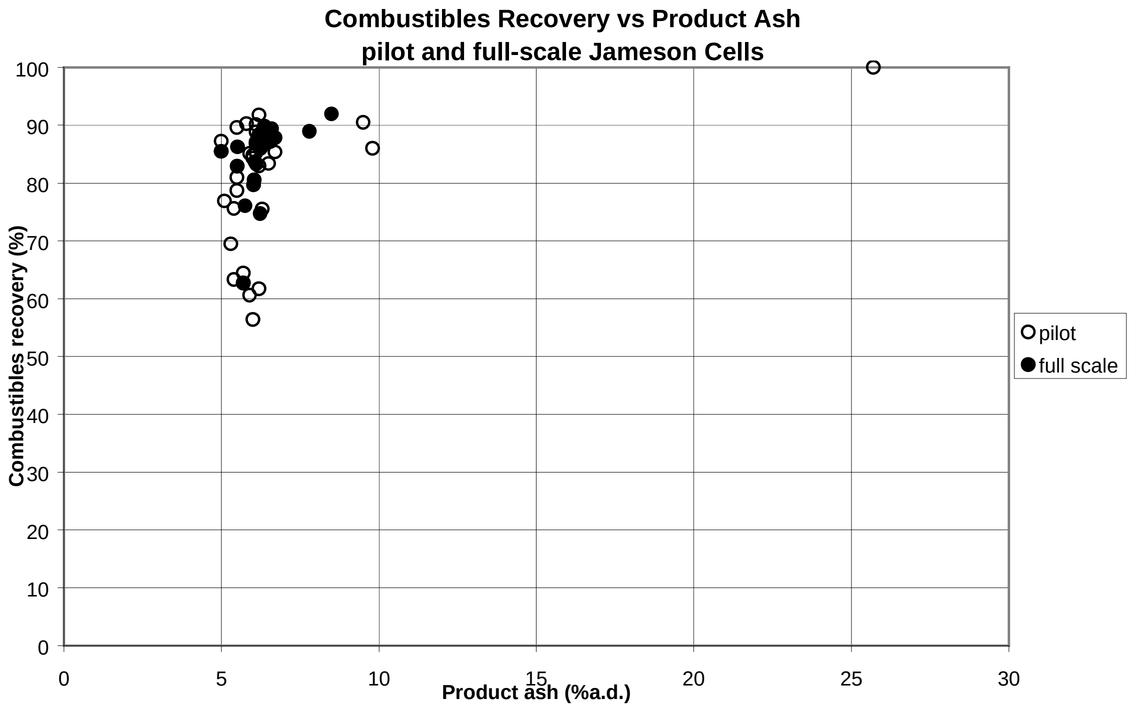
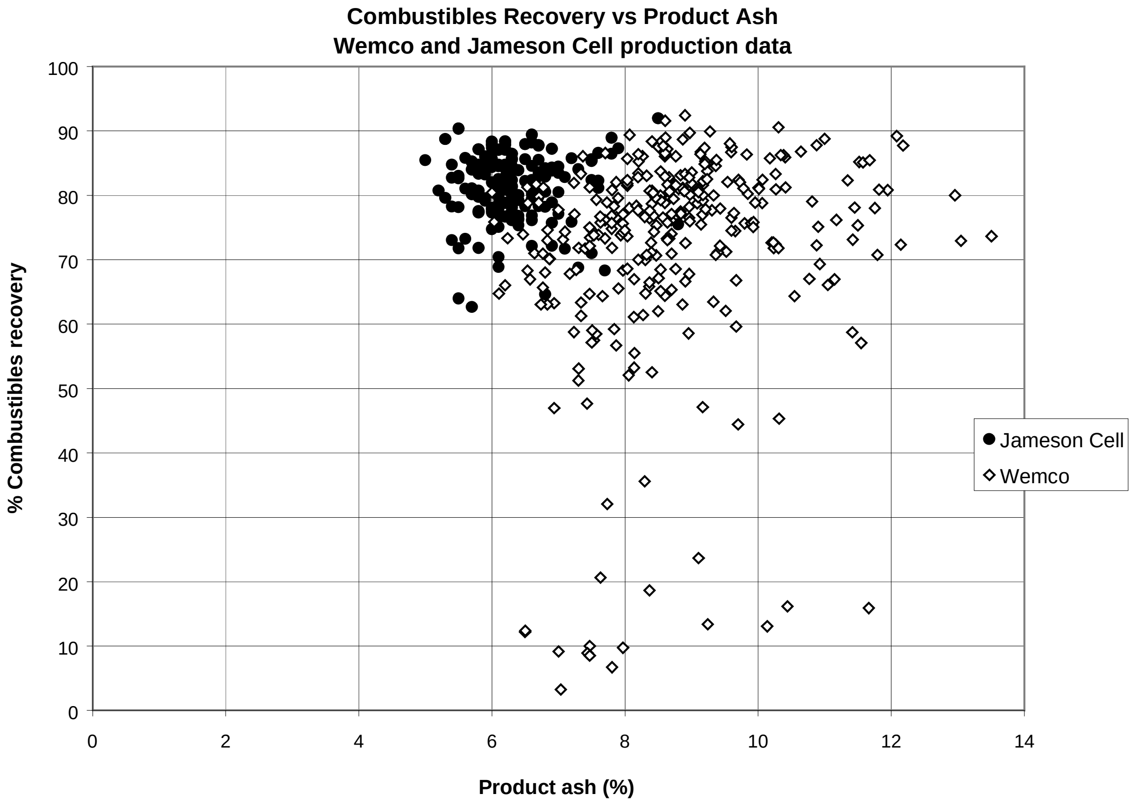


Figure 6



Figures 7 and 8 illustrate the distribution of the combustibles in the feed to the product and tailing when treated by the Jameson Cell and Wemco cells respectively. The Jameson Cell is demonstrated to maintain high combustibles recoveries across the range of particles sizes present in the flotation feed. The mechanical flotation cells reported poorer combustibles recoveries from the finest and coarsest particles, the former being particularly critical to the overall recovery.

Figure 7

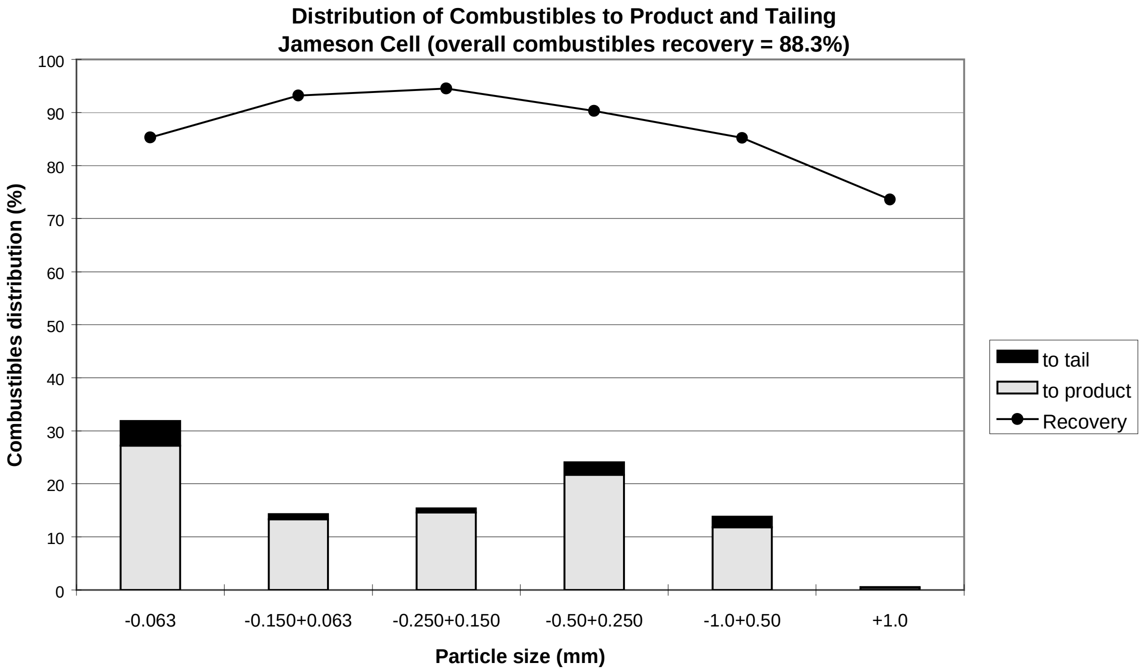
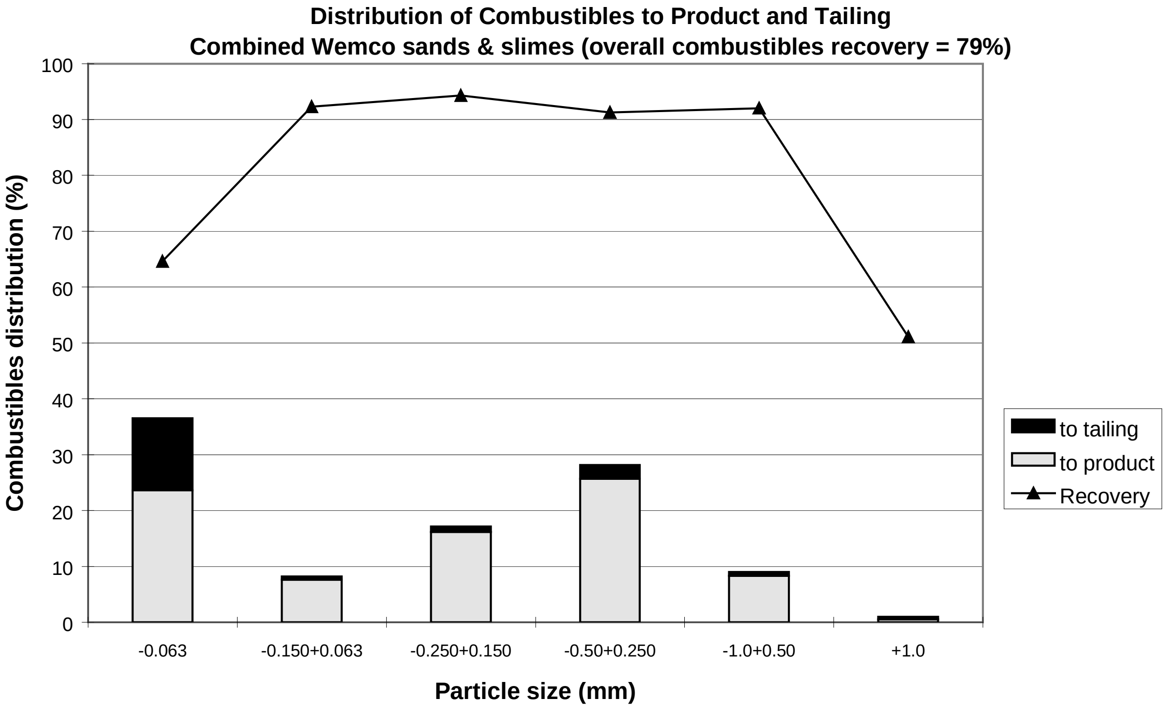


Figure 8



Performance data subsequent to commissioning the Jameson Cells has shown that the overall plant yield has been increased by 3.5%. The realised plant increase was significantly higher than the predicted yield increase of 2.1%. As a result, the payback for the fines upgrade project was reduced from the predicted 28 months to 17 months.

The success of Jameson Cell installation at the Goonyella coal preparation plant has contributed to the plant's record breaking performances. In March 1996, Goonyella broke the BHP Coal weekly production record by producing 198,645[1] tonnes of coal, overtaking the previous mark set by the nearby Peak Downs Mine. Goonyella has continued its record breaking performances, setting new weekly production highs in May 1996 (200,163 tonnes) and in January 1997 (210,476 tonnes).

CONCLUSIONS

1. Pilot scale Jameson Cell testing proved that this technology was superior to the existing mechanical flotation cells in producing a lower ash product and achieving higher combustibles recoveries.

2. During the demonstration scale Jameson Cell tests, the product ashes observed in the pilot scale test work were reproduced, although the combustibles recoveries fell short of the pilot scale results. The lower combustibles recoveries were most likely caused by the degraded nature of the coal feedstock.

3. The performance of the full scale Jameson Cell installation closely matches the performance of the pilot scale cells. Therefore, pilot scale Jameson Cell test work was able to accurately predict the performance of the full scale cells.

4. The Jameson Cell achieves higher combustibles recoveries from the finest (-0.063 mm) and coarsest (+ 1.0 mm) size fractions, and hence is able to give higher overall combustibles recoveries.

5. The ability of the Jameson Cell to consistently deliver a low ash product at high combustibles recoveries has contributed to an overall plant yield increase of 3.5%. This has lead to the setting of new production records.

ACKNOWLEDGEMENTS

The authors would like to thank BHP Coal Pty Ltd and MIM Holdings Ltd for permission to publish the data contained herein.

---

[1] All tonnages corrected to a moisture content of 8%, which is standard BHP Coal practice.

REFERENCES

Carretta, M. F., 1996. Goonyella/Riverside CPP1 fine circuit upgrade, *Technology Review Symposium - Coal Flotation, A Review of current Practice*, Brisbane.

Dawson, W. J., Yannoulis, G.F., Atkinson, B. W., and Jameson, G. J., 1996. Applications of the Jameson Cell in the Australian coal industry. *Column '96 - International Symposium on Column Flotation*, Quebec, August 26-28, pp. 233-246.

Espinosa-Gomez, R., Finch, L. A., Yianatos, J. B., and Dobby, G. S., (1988). Technical note: Flotation column carrying capacity: particle size and density effects, *Minerals Engineering*, Vol. 1, No. 1, pp. 77-79.

Evans, G. M., Atkinson, B. W., and Jameson, G. J., 1995. The Jameson Cell, *Flotation Science and Technology*, Matis, K. A., (ed.), Marcel Dekker Inc., New York, pp. 331-363.

Gosnell, P., 1996. Latest upgrade will see Goonyella/Riverside into 21st century, *Australian Mining*, Vol. 88, No. 11, December, pp.14-15.

Le Page, M., Goodman, P., and Halpin, B., 1995. Gordonstone coal preparation plant upgrade process design, in Smitham, J. (ed.), Proceedings of the Seventh Australian Coal Preparation Conference, Paper A4, pp. 46-70.

Manlapig, E. V., Jackson, B. R. Harbort, G. J. and Cheng, C. Y., 1993. Jameson Cell coal flotation, *Coal Prep '93: 10th International Coal Preparation Exhibition and Conference*, Lexington, pp. 204-219.

New Jameson Cell flotation of industrial minerals

by Steve Hall* and
Mark Harrison**

The efficiency of froth flotation is determined by a sequential series of probabilities (*1*). Paramount is the achievement of intimate particle-bubble contact ahead of value particle-bubble attachment, pulp-froth disengagement and finally, the transportation of concentrate froth to the product launders. Different flotation technologies utilise a variety of techniques (2,3,4) to maintain the above phenomena. An assessment of net operational savings in using any one technology is valid only for equivalent feed material. Economic assessment is then possible according to realised cost savings and any proven additional revenue from increased metallurgical performance.

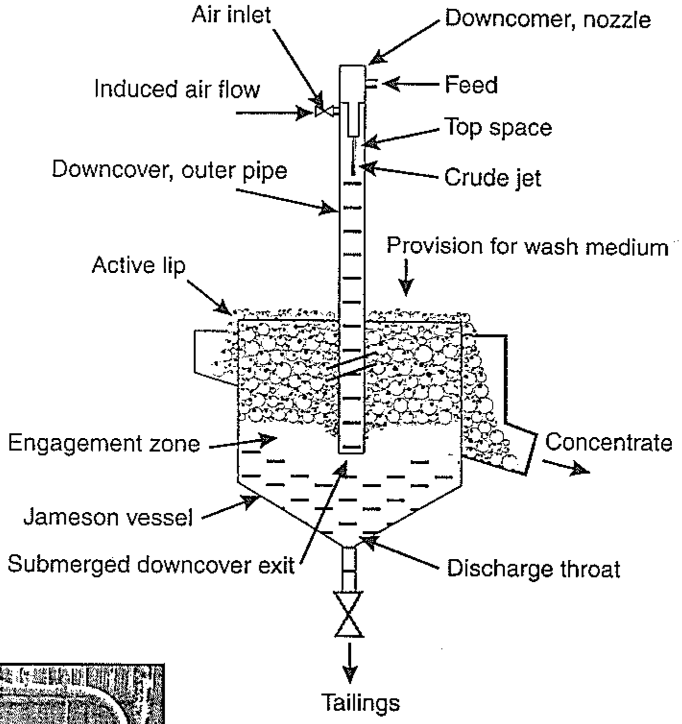
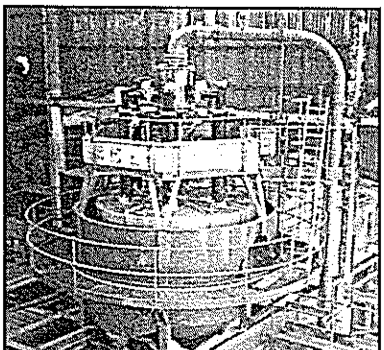


Figure 1. Section through a Jameson Cell indictating the relative location and the action of a solitary downcomer.

The Jameson Cell is an exceptional froth flotation device as it functions by hydrodynamical means. The associated technical and economic attributes of this non-mechanical technology impart revolutionary potential to commercial flotation. At present there are close to 130 commercial installations of the Jameson Cell at about 44 individual flotation sites throughout the world. On 6 September 1994, the first commercial application of the Jameson Cell to industrial mineral flotation was commissioned at Cleveland Potash Ltd in the UK.



Jameson Cell - technical operation

The heart of the Jameson Cell is the *downcomer*. The action of a solitary downcomer within a Jameson Cell is shown by *Figure 1*. A downcomer is a simple static device, comprising of a nozzle assembled to an outer pipe with a single air inlet. The downcomer is held vertically with its base positioned at a depth in an engineered Jameson Vessel.

During start up, feed enters the Jameson Vessel as a crude jet of slurry issuing from the nozzle. Quickly the base of the downcomer becomes submerged forming a liquid seal that ensures the downcomer fills up with pulp. A head of slurry is thereby elevated above the active lip inducing hydrostatic suction and thus creating a region of low vacuum in the top space of the outer pipe.

Initially the jet is submerged within the downcomer pulp. However, immediately upon opening the air inlet, air is drawn to the in situ vacuum from the local environment. Regulating the amount of induced air flow, stabilises the pulp at an intermediate height in the downcomer. An extent of the slurry jet plunging through the low vacuum region is thus exposed. High shear and mixing conditions are created by the impingement of the exposed jet within the pulp in the downcomer. Air is continuously entrained and dispersed as fine bubbles of diameter 300 to 600 μm (mean arithmetic) (*5*). A maximum of 50% to 60% void fraction (v/v) is attainable in the operating downcomer thereby thinning the film thickness between bubbles and increasing the interfacial area available to contact mineral particles.

Slurry is resident in the downcomer for between 8 to 10 seconds. Resultant fluid momentum ensures pipe flow of the mineral laden bubbles as they are carried downwards to the submerged downcomer exit.

In the vicinity of the exit, pulp and froth enter a disengagement zone. Froth rises forming a consistent froth bed that is washed as necessary to remove entrained gangue prior to concentrate froth overflow into product launders. Tailings flow through the discharge throat at the base of the Jameson Vessel. The resulting fluid flow characteristics satisfy all requirements for system agitation. In total, slurry is resident in a Jameson Cell for between 120 and 180 seconds.

Commercial Jameson Cells may use a number of downcomers within a froth-partitioned (FP) Jameson Vessel. Isolation of the rising mineral bearing froth produced by one or a group of

---

*University of Nottingham, UK.
**MIM Technology Marketing Europe, UK.

downcomers, maintains flotation performance at a fraction of the designed maximum feed volumetric flow. Aside from achieving an enhancement of metallurgical performance, the technical operation of the Jameson Cell alone contributes an innate number of unique money-saving attributes.

Jameson Cell: economic benefits

The eighteen economic attributes of the Jameson Cell are best described according to their intrinsic or extrinsic origin.

Table 1. Intrinsic attributes of the Jameson Cell

Economic attribute No.	*Definition*
1	Air aspiration
2	Bubble generation
3	Intimate particle-bubble contacting in short residence time
4	Tranquil, rising, particle loaded froth available for washing
5	System agitation
6	Rota on-line maintenance by rapid isolation of wear components
7	System turndown capability

Intrinsic attributes

The downcomer simultaneously provides seven intrinsic attributes as defined by *Table 1.*

Table 2. Conventional flotation components eliminated by Jameson Cell

System	*Description*
A	Air compressors, air blowers, and air delivery network
B	Spargers and/or sparging mechanisms
C	Agitators -: as rotor-stator assemblies, drive transmission and drive motors

Table 2 describes the mechanical systems eliminated by the intrinsic attributes of downcomers.

The removal of the mechanical entities from the flotation system leads to significant cost savings in absorbed energy, and in on-going maintenance.

Table 3. Extrinsic attributes of the Jameson Cell

Economic attribute No.	*Description*
8	Compact equipment size, ie. reduced footprint
9	Reduced numbers of cells per flotation stage
10	Simplified circuit layout
11	Simple process control and parameter monitoring
12	Effective response to changes of feed grade
13	Reduced operating manpower requirements
14	Greater operator understanding
15	Reduced total maintenance cost
16	Reduced total flotation floor area
17	Reduced spares inventory held in site stores
18	Rapid system stabilisation after start up

Extrinsic attributes.

The extrinsic attributes listed in *Table 3* occur as a consequence of a commercial Jameson Cell achieving flotation performance in a very short slurry residence time.

The purpose of this paper is to detail the new applications of the Jameson Cell to the treatment of industrial minerals. To date, commercial installations of the Jameson Cell exist in potash and graphite production. Other graphite, fluorspar and zircon duties are summarised by the conclusions of pilot study.

Commercial installations

(a) Site 1: Potash production

Cleveland Potash Ltd ("CPL"), unlocked the potential of the Jameson Cell in 1992. Initially, laboratory based potash recleaner tests were performed by the University of Nottingham as a Jameson Cell Testing Authority. A J100 pilot-scale Jameson Cell unit was used to test samples transported from CPL. The availability of a "hot floor" room enabled problems of potash recrystallisation from saturated brine slurries on temperature reduction to be avoided. Indicator tests showed that the downcomer could be applied to potash processing and that froth washing with saturated brine from a header tank was possible. These conclusions incited the development of a site based pilot plant programme.

At CPL, flotation is used to concentrate sylvite (KCl) from halite (NaCl) plus insolubles such as clays and sulphates. Stringent desliming of the all flotation feed is essential ahead of the addition of starch (clay depressant) and amine (KCl collector). Chloride mineral dissolution is minimised by the use of saturated brine as desliming and flotation media.

Two stage desliming to 100μm is achieved through primary and secondary cyclone banks. The secondary cyclone underflow (-1500+100μm) reports to the standard flotation feed pump box. Primary cyclone overflow is further deslimed to 30μm via tertiary cyclones.

Standard rougher tails are passed over DSM screens to recover middlings and any misplaced coarse potash. Rougher screen oversize (+450 μm) is distributed to three regrind ball mills. Rougher screen undersize reports to a tails centrifuge. Standard cleaner tailings are gravity fed to LH-cyclones and the cumulative bank underflow reports to the regrind mills. Standard recleaner tailings report to DSM screens. Recleaner screen oversize (+ 200μm) is also fed to the regrind.

Both LH-cyclone overflow and recleaner DSM underflow report to tertiary cyclones. The resulting total tertiary cyclone underflow (-100+30μm) is fed to the slimes flotation feed pump box. Tertiary cyclone overflow reports to a tails thickener for brine recovery from tails solids. Jameson Cell on-site pilot testing progressed by using key flotation streams of both flotation circuits (*6,7*). Results indicated that the greatest economic benefits could be gained by utilising the attributes of the downcomer as close as possible to the head of each flotation circuit.

Slimes flotation

On 6 September 1994, a single Jameson Cell 3250/6 (FP) was fully commissioned at CPL as a Total Slimes Flotation system. The single Jameson Cell replaced 16, 2.8 m^3 Denver No.30 DR flotation cells used as slimes roughers and cleaners, i.e., the entire original slimes flotation.

Client CPL

Jameson Circuit layout

Number of units	1
Number of stages	1

Feed detail

Origin	Tertiary cyclone underflow
Volumetric flow (m^3/h)	370
Solids (% (w/w)	18
Particle size specification	(-100+30) µm
Fraction of total plant product produced via Jameson, %	10

Commercial performance, mean production (30 day)

Feed, % KCl	24.5
Concentrate grade, % KCl	77.4
Tailings, % KCl	5.7
Recovery, %	82.8

A conservative economic assessment for the installation and operation of the Jameson Cell at CPL is summarised in *Table 4*.

Table 4. Conservative economic summary (*8*)

Item	*Value (£)*
Payback, days	213
Additional recovery revenue pa	201,939
Net operational saving pa	209,985

Table 5 highlights a 76.7% energy saving made by utilising the hydrodynamics of the downcomer instead of the conventional cell agitator motors of the original slimes circuit.

At CPL, the single Jameson Cell recovers fine potash from a gangue comprised of halite, clays and sulphates. The achievement of fine particle selectivity occurs at enhanced recovery leading to a conservative estimate of additional site revenue as £201,939 pa for a payback in 213 operating days.

Table 5. % Savings from the use of Jameson Cell vs original slimes circuit

Item	*% Saving*
Equipment footprint	80.9
Total float floor area	81.3
Absorbed power[{†}], pa	76.7
Maintenance[{Û}], pa	>79.0

[{†} Jameson Cell Additional Pumping Requirement vs. Original Slimes Flotation Circuit Cell Motors]

[{û} Jameson Cell Maximum Maintenance vs. Original Circuit-Previous Operating Year excl. manpower]

Standard flotation

The use of flotation at CPL reflects the achievement of good liberation of KCl from NaCl at coarser particle sizes. The use of two flotation circuits stipulates a requirement of the Standard Flotation circuit equipment to recover the +850 µm particles. *Table 6* compares the performance of the pilot Jameson Cell and the standard flotation sub-aeration cells when processing the standard flotation rougher concentrate.

Results of *Table 6* indicate that the Jameson Cell could increase the standard flotation circuit recovery from 75.5% to 81.7% and maintain product grade at 90.8% KCl, if pilot results were repeated on a commercial basis. The amount of high grade +850µm material recovered from a standard flotation rougher concentrate of grade 79.9% KCl is shown in *Table 7*.

With the Jameson Cell, the weight recovery from the high grade +850µm fraction has increased from 13.4% to 46.1% thereby enhancing the recovery of KCl from the standard flotation by 4.1 percentage points. By recovering the coarse material in the Jameson Cell, the regrinding of about one third of the +850µm fraction reporting to the rougher concentrate could be eliminated. In both slimes and standard flotation

Table 6. Single stage pilot Jameson Cell vs standard circuit cells [{*}]

Feed = standard flotation rougher concentrate, no froth washing

Feed		*Single Jameson Cell*				*Standard circuit cells*			
		Concentrate		*Recovery*		*Concentrate*		*Recovery*	
% KCl	*Std*	*% KCl*	*Std*	*%*	*Std*	*% KCl*	*Std*	*%*	*Std*
70.6	3.57	89.1	0.94	85.1	3.18				
77.6	2.09	90.8	1.42	81.7	2.17	91.0	2.09	75.5	1.76

[{*} Standard Circuit Cells = (28) off 2.8 m^3 Denver No.30 DR sub-aeration cells in cleaner/recleaner configuration]

Table 7. Single stage pilot Jameson Cell vs standard circuit cells [{*}]

Feed = standard flotation rougher concentrate, (grade = 79.9% KCl)

Technology	*No. of flotation stages*	*Wt. recovery of +850 µm size fraction*	*Conc. grade, % KCl*	*Recovery %*
Single Jameson	1	46.1	91.8	83.7
Standard circuit	2	13.4	91.7	79.6

[{*} Standard Circuit Cells = (28) off 2.8 m^3 Denver No.30 DR sub-aeration cells in cleaner/recleaner configuration]

Other studies

(a) Graphite: Nottingham University Testing Authority

Client *Confidential*

The University of Nottingham was approached to study possible improvements in graphite flotation performance, particularly concentrate grade, that might arise from using Jameson Cell Technology.

The client subsequently agreed confidentiality with MIM Technology Marketing Ltd and completed a feasibility test agreement prior to testwork. The graphite ore, of Russian origin, was supplied in lump form by the client and reduced in size to -106µm. Various flotation reagent chemistries were investigated and the J100 Jameson Cell could produce equivalent grades to those obtained by conventional laboratory housed Denver cell flotation (87% to 93% C). It would appear that, in this case, the concentrate grade was not limited by flotation performance, but by lack of liberation of gangue minerals from the graphite. Detailed mineralogical work is often necessary to support flotation testwork.

(b) Graphite: Namibian site pilot plant testing

Client *Confidential*

Pilot tests on location in early 1993 appraised the use of a Jameson Cell in primary and secondary cleaning duties at a site pilot plant.

The results of a 2^4 factorial design experiment identified the significant variables of (i) air flow and (ii) froth depth. The achievement of product grade was seen to be dependent on the selected rise velocity of the concentrate froth in the Jameson Cell. Equivalent or better grades were achieved with the Jameson Cell when compared to conventional sub-aeration cells or columns. The encouraging results obtained were to lead to a detailed study of the Jameson Cell as a final product cleaner. Unfortunately the pilot site closed due to the lack of a market for the plant product.

(c) Zircon: Australian site pilot plant testing

Client *Confidential*

The application of the Jameson Cell to zircon roughing duties was briefly explored at a mineral sands pilot plant. Zircon flotation was performed under conditions of pH control in the absence of frother. Collection with an amine occurred only after effective TiO_2 depression. A target performance for zircon roughing was set as a concentrate grade 22% ZrO_2 at 90% recovery with a maximum impurity of 2% TiO_2.

From single stage tests with a feed grade of 10% ZrO_2, the Jameson Cell produced concentrate grade of >21% ZrO_2 at a 65% unit recovery (<2% TiO_2) without any pH regulation of wash medium. It appeared therefore that zircon roughing could be completed in two stages of Jameson Cell although verification of the scavenger role remained outstanding. Further, a full characterisation of SiO_2 and Fe_2O_3 concentrate impurities as entrained or co-collected entities was not undertaken.

(d) Fluorspar: European on-site testing

Client *Confidential*

The application of the downcomer to fluorspar flotation yielded the results as shown in *Table 8*. The entire flotation test results were used to develop a flotation flowsheet and metallurgical account. The circuit contemplated that for PbS flotation, one stage of single unit Jameson Cell would replace the entire Pb flotation circuit, ie. 23 conventional cells as one roughing and three cleaning stages. In CaF_2 flotation, three stages of single unit Jameson Cell cleaning were deemed necessary to ensure the final product grade and thereby replace the current plant, ie. 49 conventional cells as six cleaning stages.

Any future application of the Jameson Cell at the site relies on stabilising the current commercial chemistry, and establishing a plant operating stratagem for the evolving plant feed mineralogy.

---

circuits at CPL, the high ionic strength of the saturated brine medium results in (*9*):

(i) High compression of electrical double layers

(ii) A four fold decrease in surface tension by frother addition when compared to the action of an equivalent concentration of identical frother added to an aqueous system

The tranquil flow of the rising potash bearing froth produced by the commercial Jameson Cell reflects a very quick change of surface tension within the downcomer as the potash is loaded to bubbles. Hence, coarse particle recovery is probably greatly assisted by the stable provision of a required and specific rate of interfacial surface area created rapidly within the Jameson Cell.

(b) Site 2: Graphite production

Graphite recovery from the primary cyclone overflow (PCO) was deemed critical to maintain the economic viability of the Southern Africa operation. Total flotation performance was adversely affected by the presence of fine slimes gangue in the PCO stream, and also by an inefficient primary cyclone performance. Combining the product of PCO flotation with the coarse graphite float product resulted in contamination of final flotation plant product grade to less than the required 98% C.

Attempts to improve the performance of the installed conventional sub-aeration cells failed to resolve the emergency. From remote site, approximately 30kg of sample of the PCO was flown to Johannesburg for limited testing in a laboratory housed J100 Jameson Cell.

Conclusions of Jameson Cell test work led to the rapid design, installation and commissioning of a commercial Jameson Cell to process the PCO. On 26 September 1994, a single 800/1 Jameson Cell was commissioned as a fines/slimes scavenger plus cleaner.

Client *Confidential*

Jameson Circuit layout

Number of units	1
Number of stages	1

Feed detail

Origin	PCO of bulk rougher concentrate

Volumetric flow (m^3/h)	11
Solids (% (w/w)	1.5 to 2.5
Particle size specification	d50 = 600μm, 10% passing 150μm

Commercial performance, single spot

Feed, % C	52
Concentrate grade, % C	90.2
Tailings, % C	25
Recovery, %	72

The incorporation of a single Jameson Cell in the graphite processing flowsheet has maintained the economic viability of the commercial operation. The Jameson Cell produces a sticky froth concentrate of fine/ultrafine graphite that is continuously fed to the tertiary cleaner feed. Plant recovery is thus critically enhanced at the required product grade of 98% C.

Table 8. Results of Jameson cell fluorspar flotation

Application	*No. of Jameson Cell stages*	*Feed*	*Concentrate*	
Lead		% Pb	% Pb	Recovery, %
PbS total system	1	6.5	69.8	93.4
Fluorspar:		% CaF_2	% CaF_2	Recovery, %
Cleaner 1	1	78.2	94.0	86.3
Cleaner 2	1	93.8	97.8	52.0

Summary

The hydrodynamic Jameson Cell uses static *downcomers* to eliminate the mechanical systems of conventional flotation equipment, and increase the reliability of the flotation circuit. The commercially operating Jameson Cell at Cleveland Potash achieves efficient fine particle selectivity from tertiary cyclone underflow. The single stage Jameson Cell replaces the entire original 370 m^3/h Slimes Flotation circuit; equipment footprint is thereby reduced by 80%. Net operational savings of £ 209,985 pa are verified by the additional site revenue gained from increased potash recovery and by the 80% cost savings in both the ongoing maintenance and the energy drawn.

Coarse potash +850 μm recovery from standard flotation rougher concentrate is achievable in a pilot Jameson Cell indicating a future possible potential of increasing the plant recovery at a premium product size. Associated improvements of concentrate debrining, centrifuging, and drying are envisaged whilst simultaneously reducing the mass load to regrinding. Fine graphite selectivity is commercially achieved from an inefficiently classified primary cyclone overflow thereby enhancing plant recovery of 98% C and ensuring an entire operation's economic viability.

Pilot tests have shown positive application of the downcomer to zircon roughing, galena flotation from fluorspar, and fluorspar cleaning. For galena and fluorspar duties, high grade concentrates are produced by the Jameson Cell in a greatly reduced number of stages when compared with the conventional equipment in current operating use. Further site work is essential to clarify the application of the Jameson Cell to zircon and fluorspar duties. The numerous intrinsic and extrinsic attributes of the Jameson Cell lend themselves to the consistent production of a rising, non-turbulent, mineral loaded froth available for continuous washing. This significant froth characteristic should greatly assist the profitable achievement of product grade for many other industrial mineral applications.

References

1. Atkinson B.W., Conway C.J. and Jameson G.J. High Efficiency Flotation of Coarse and Fine Coal, paper presented at: High Efficiency Coal Preparation Symposium, SME Annual Meeting, AIME 124th Annual Meeting, March 6-9, Denver, Colorado, USA (1995).
2. Couch G.R., Flotation, p38-45, section 5.1 in: Advanced Coal Cleaning Technology IEACR/44, pp96, IEA Coal Research, London, England (1991).
3. Hall S.T., Developing Flotation Technologies, p(1.1.1-1.1.5) in: Proceedings of Minprep '91 Symposium, Minerals Engineering the Challenges for the '90s, 9-11 April, Doncaster, UK, 09-11 April 1991, pp81, Mining Industry Promotions Ltd, Rickmansworth, England (1991).
4. Skillen A., Froth Flotation: New Technologies Bubbling Under, Industrial Minerals, No.305, February, p47-59 (1993).
5. Evans G.M., Atkinson B.W. and Jameson G.J. The Jameson Cell, p331-363 in Flotation Science and Engineering, pp576, Marcel Dekker, New York, USA (1995).
6. Burns M.J., Coates G. and Barnard L., Use of Jameson Cell Flotation Technology at Cleveland Potash Ltd., North Yorkshire, England, Trans IMM, Sect. C, Vol. 103, C162-167 (1994).
7. Burns M.J., Coates G., Barnard L., The Use of Jameson Cell Flotation Technology at Cleveland Potash, p290-300 in: Proceedings of International Fertilizer Association (IFA) Technical Conference, Amman, Jordan, 2-6 October 1994, pp492, IFA, Paris, France (1994).
8. Burns M.J, Pearson J. and Harrison M.E. Operating Experience and Savings with Jameson Cell Technology at Cleveland Potash, paper in press.
9. Nemedy L., Factors Influencing Froth Stability in Potash Processing, p605-609 in: Proceedings of First International Potash Technology Conference, Saskatoon, Saskatechwan, Canada, 3-5 October 1983, pp887, Pergammon Press, Oxford, England (1984).

The Jameson Cell

Geoffrey M Evans
Bruce W Atkinson
Graeme J Jameson

12

The Jameson Cell

Geoffrey M. Evans, Bruce W. Atkinson, and Graeme J. Jameson
University of Newcastle, Newcastle, New South Wales, Australia

I. INTRODUCTION

The Jameson cell is a type of flotation column in which the air and the pulp are brought together in a vertical tube. The air and pulp are dispersed into a dense foam of fine bubbles, which creates a favorable environment for particle–bubble collision and subsequent collection of hydrophobic particles (1). The bubbly mixture then discharges into the cell proper, which is essentially a disengagement chamber in which the bubbles carrying the floatable particles separate from the pulp. The disengagement zone behaves in a similar fashion to the froth zone in a conventional flotation column, and as in the column, it is possible to apply clean washwater to the froth to remove unwanted gangue particles.

The main advantages of the Jameson cell relate to the rapid collection of particles in the downcomer, leading to a compact space-efficient device, and to the ability of the cell to operate with self-aspiration, thus obviating the need for compressors or blowers.

Since its introduction at the commercial scale in 1989, the cell has been applied to a variety of ores, and full-sized cells are now operating on streams involving lead, zinc, copper, and nickel sulfides, as well as coal and industrial minerals. It has been applied as well to the removal of fine organic droplets from solvent extraction liquors.

In this chapter, a brief overview of antecedents to the Jameson cell is given. The general principles of operation are discussed, together with a description of the various phenomena that take place in the downcomer. The effect of the operational variables on cell performance follows, including some data from operating cells.

II. DEVELOPMENT

The cell was devised in the period 1985–1986, when the inventor was undertaking a collaborative research project at Mt. Isa Mines Limited, Mt. Isa, Queensland. This is a large mine, accustomed to using the best modern practices in the production of copper, lead, and zinc concentrates. A number of large flotation columns were being designed and introduced at the time in the copper and lead/zinc circuits. Although it was clear that the columns would offer significant advantages over the conventional mechanical cells in terms of the ability to wash the froth, it was also clear that the collection process was very slow, necessitating large residence times and consequently the large columns, which have become a familiar feature of this technology. Accordingly, alternative methods of bringing the particles and the air into contact were investigated. Following initial work in the laboratory at Newcastle University, prototypes were tested and further developed in the concentrator at Mt. Isa, resulting in a production-scale design. A number of Jameson cells were installed in the lead circuit and in the nearby Hilton concentrator then under construction. Further cells were installed in a coal operation at Newlands, Queensland, and on a copper stream at the Peko Warrego mine, Tennant Creek, Northern Territory, Australia.

The use of columns as gas–liquid contactors, as well as the use of plunging jets to entrain air, has been previously tried in a number of designs, dating from the time of the first introduction of flotation in mineral processing. Perhaps the first to appear in the patent literature is the column described by Norris (2) in 1907, in which air appears to be both dissolved and entrained into a stream of pulp, which is then introduced to the bottom of a flotation column complete with froth crowder. Various devices for entraining air into a pulp by means of a plunging jet are described by Taggart (3) as "cascade machines." None of those early machines was capable of control of the air and bubble sizes, and it is apparent that they were designed more by guesswork than from basic principles of fluid mechanics. A review of alternative flotation machines has recently been given by Jameson (4).

In the design of the Jameson cell, it has been possible to take advantage of the many advances that have been made in the dynamics of single bubbles and of two-phase bubbly flows. Thus the jet entry conditions are such as to make micro-bubbles of controlled size, so as to create a large surface area and hence to maximize the collection process in the downcomer. The downcomer is sized so that the bubbles created by the plunging jet are carried downward, but at the same time the downward pulp velocity is sufficiently low to allow bubbles to rise against the flow and create a mixture of high gas content. The downcomer diameter and the pulp flow must also be matched in such a way as to ensure that any large slugs that form are carried downward. In the froth zone, the rising velocity J_g of the gas must be such as to allow the froth to drain properly so that the gangue is not carried forward into the concentrate. At the same time, the J_g must be sufficiently high to minimize the drop-back of values and consequent loss of recovery. When washwater is used, the froth conditions must be such that again, the improvements in grade are not linked with reduction in recovery.

III. PRINCIPLES OF OPERATION

A. Introduction

In this section the different hydrodynamic regions occurring within the Jameson cell are discussed, with emphasis placed on the interactions between regions that determine the overall flotation performance. Before identifying the different hydrodynamic regions, however, it is useful to define the basic geometry of a Jameson cell and also to describe qualitatively what happens during startup and operation of a typical unit. As illustrated in Figure 1, the simplest geometry consists of a riser section, which is often referred to as the *cell*, and a single downcomer—although some of the larger units incorporate up to 30 separate downcomers in combination with a common riser section.

The downcomer consists of a vertical tube that is sealed at the top except for a regulated air inlet and a vertical nozzle through which the slurry feed is introduced. The base of the downcomer is located below the pulp level inside the riser. On startup, the air inlet at the top of the downcomer is closed, and the pulp feed is pumped through the nozzle. The air in the downcomer is entrained into the pulp, which forms the seal in the bottom of the cell; consequently, the pulp is drawn upward from the cell into the downcomer. The pulp level reaches the tip of the nozzle quite quickly, and as a consequence of the hydrostatic suction developed by the head of this pulp, the pressure in the head of the downcomer is less than atmospheric. When the inlet is opened, air is drawn into the headspace at the top of the downcomer where it is entrained into the downcomer contents by the plunging jet. The entrained air is broken up into fine bubbles that are quickly dispersed into the pulp and carried downward by the bulk fluid motion.

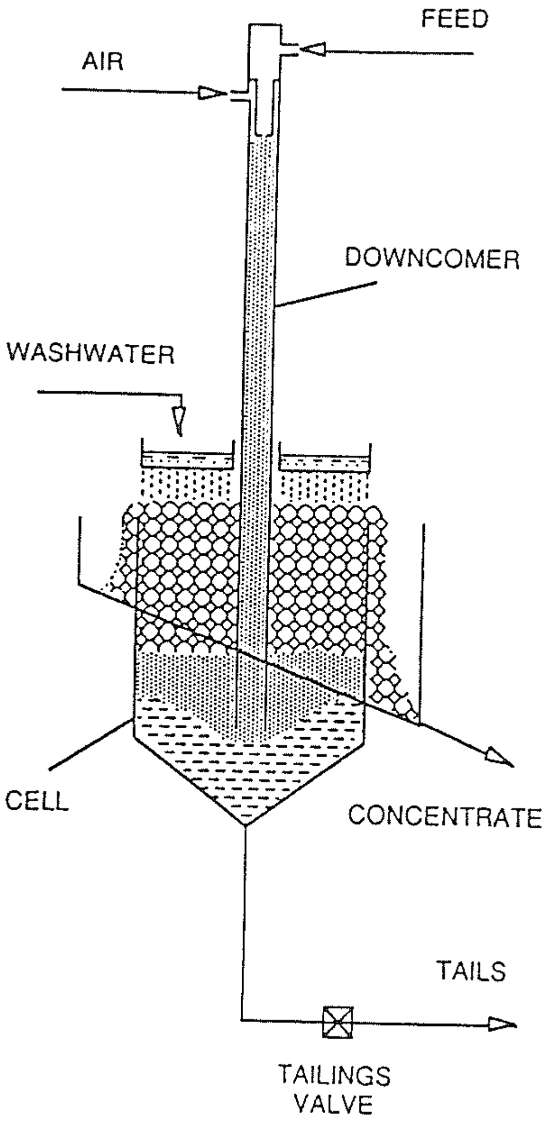


Figure 1 Layout of a single-downcomer Jameson cell.

The three-phase mixture passes from the base of the downcomer into the cell proper, which has a much greater cross-sectional area than the downcomer. Consequently, the downward superficial velocity of the mixture is reduced, allowing the particle-laden bubbles to disengage from the liquid, rise to the surface, and form a layer of froth. The froth then drains before overflowing into a collection launder, while the liquid phase and unrecovered particles leave through a valve at the base of the cell.

B. Hydrodynamic Regions

The different hydrodynamic regions that constitute the Jameson cell are shown in Figure 2. The free jet, submerged jet, mixing zone, and pipe flow zone occur within the downcomer, while the disengagement zone, which includes the froth layer, occupies the entire riser volume.

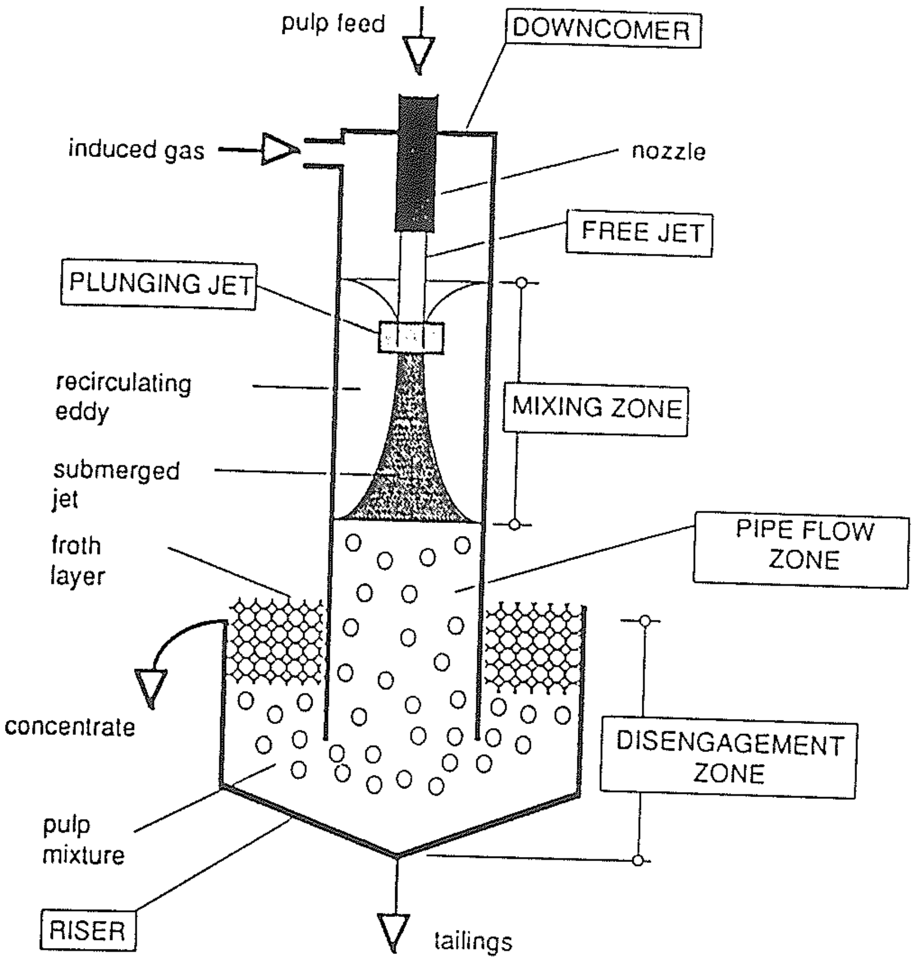


Figure 2 The various active zones in the Jameson cell.

1. *Free Jet*

The free jet is defined as the stream of liquid (pulp) feed between the tip of the nozzle and the horizontal free surface inside the downcomer. Once the jet leaves the nozzle, its diameter is determined by the relaxation of the velocity profile inside the jet and also by the interaction of the free surface of the jet with the surrounding atmosphere (5). The velocity relaxation inside the jet occurs as a result of the change from the velocity profile, when the pulp is inside the nozzle, to a plug flow velocity profile once the jet has left the confines of the nozzle. Further downstream the velocity profile changes from plug flow to parabolic flow due to the interaction of the outer boundary layer of the jet with the surrounding atmosphere. This interaction slows the velocity of the jet at its free surface resulting in an expansion in the free jet diameter.

Associated with the velocity relaxation within the jet once it leaves the nozzle is a lateral movement that creates undulations on the jet free surface (6). As shown in Figure 3, the undulations increase in magnitude with

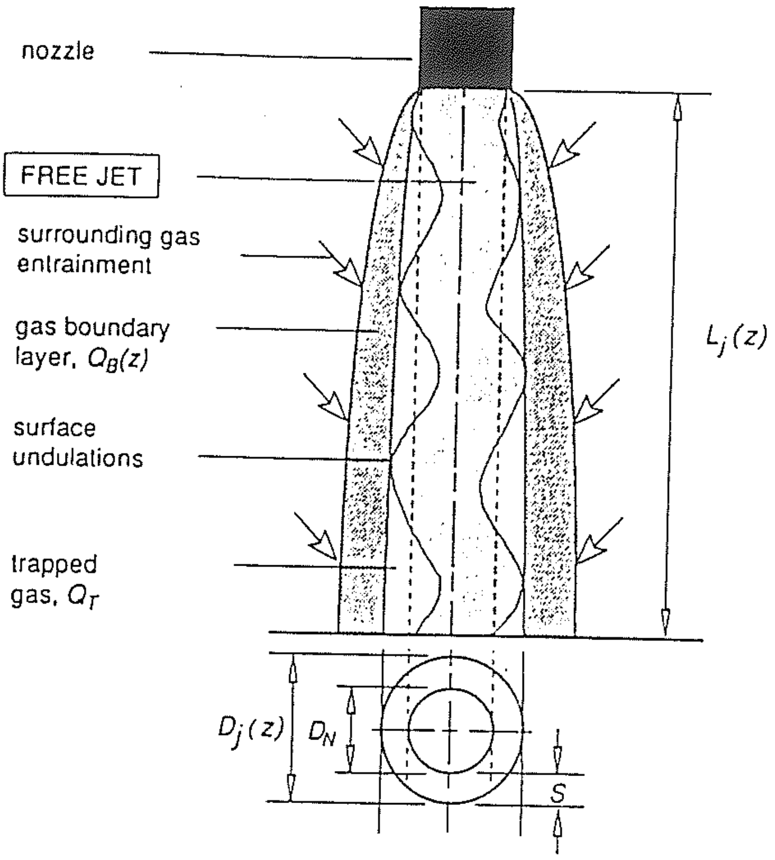


Figure 3 The mechanisms of air transport by a free jet.

increasing free jet length thereby increasing the effective diameter of the jet. The irregular nature of the jet surface created by the undulations is often called the *surface roughness*, and jets that exhibit this characteristic are called *rough jets*. Jets that display no surface roughness—and are generated using specially designed nozzles—are called *smooth jets*.

The surface roughness S is defined in Figure 3 as the difference between the nozzle diameter D_N and the effective diameter of the jet $D_j(z)$, which is a function of the length of the free jet $L_j(z)$. The surface roughness is strongly influenced by the upstream conditions of the jet delivery system. Disturbances such as swirl generated by a pump, vibration from a motor, or irregularities on the inside surface of the pipework are all amplified once the pulp passes through the nozzle, resulting in an increase in the surface roughness.

Also illustrated in Figure 3 is the transport of air by the free jet as it passes through the headspace at the top of the downcomer. First, a thin annular boundary layer of air $Q_B(z)$ is carried along adjacent to the jet free surface. Second, a quantity of air Q_T is trapped within the effective diameter of the jet. Therefore, the total amount of air transported by the free jet Q_F can be written as

$$Q_F = Q_B(z) + Q_T \tag{1}$$

Smooth jets that exhibit no surface roughness have no trapped air component. Hence, for two jets with the same velocity and pulp flow rate, a rough jet transports a greater volume of air than a smooth jet. Usually this results in a greater air/feed ratio Q_G/Q_L into the downcomer—where Q_G and Q_L are the air and liquid volumetric feed rates into the downcomer, respectively—which leads to higher air void fractions and interfacial areas. Therefore, it is desirable to operate the Jameson cell using rough jets, a situation that is not difficult to reach in practice since the jets normally have rough surfaces created by the disturbances generated within the delivery system.

2. *Plunging Jet*

The plunging jet is defined as the region where the free jet impacts with the horizontal free surface at the top of the downcomer, resulting in air entrainment. At the point of impact, a depression is formed in the horizontal free surface, which is often referred to by McCarthy (6) as the *induction trumpet*. A half-sectioned profile of the induction trumpet is illustrated in Figure 4, which shows the free surface at the top of the downcomer being drawn downward by the momentum of the free jet. The induction trumpet has a wide opening at the top, which tapers down to a thin annular film adjacent to the effective boundary of the free jet. The fluted entrance channels the moving air boundary layer into the induction trumpet, and a portion of it Q_f enters the thin film, while the remaining air inside the boundary layer Q_U travels radially out along the horizontal free surface and returns to the headspace (7).

The induction trumpet periodically collapses due to instabilities generated on its free surface, resulting in entrainment of the annular film. Combining this quantity with the air trapped within the effective jet diameter at the point of impact, the total entrainment rate of the plunging jet Q_E is

$$Q_E = Q_f + Q_T \tag{2}$$

It should be noted that the entrainment capacity of the plunging jet given by Equation (2) is not equivalent to the air feed rate into the downcomer. The reason is that air is being continuously circulated within the mixing zone, and a portion Q_H returns to the headspace, from which it is reentrained by the plunging jet. The air recirculated from the mixing zone effectively reduces the amount of new air Q_G that can enter the downcomer.

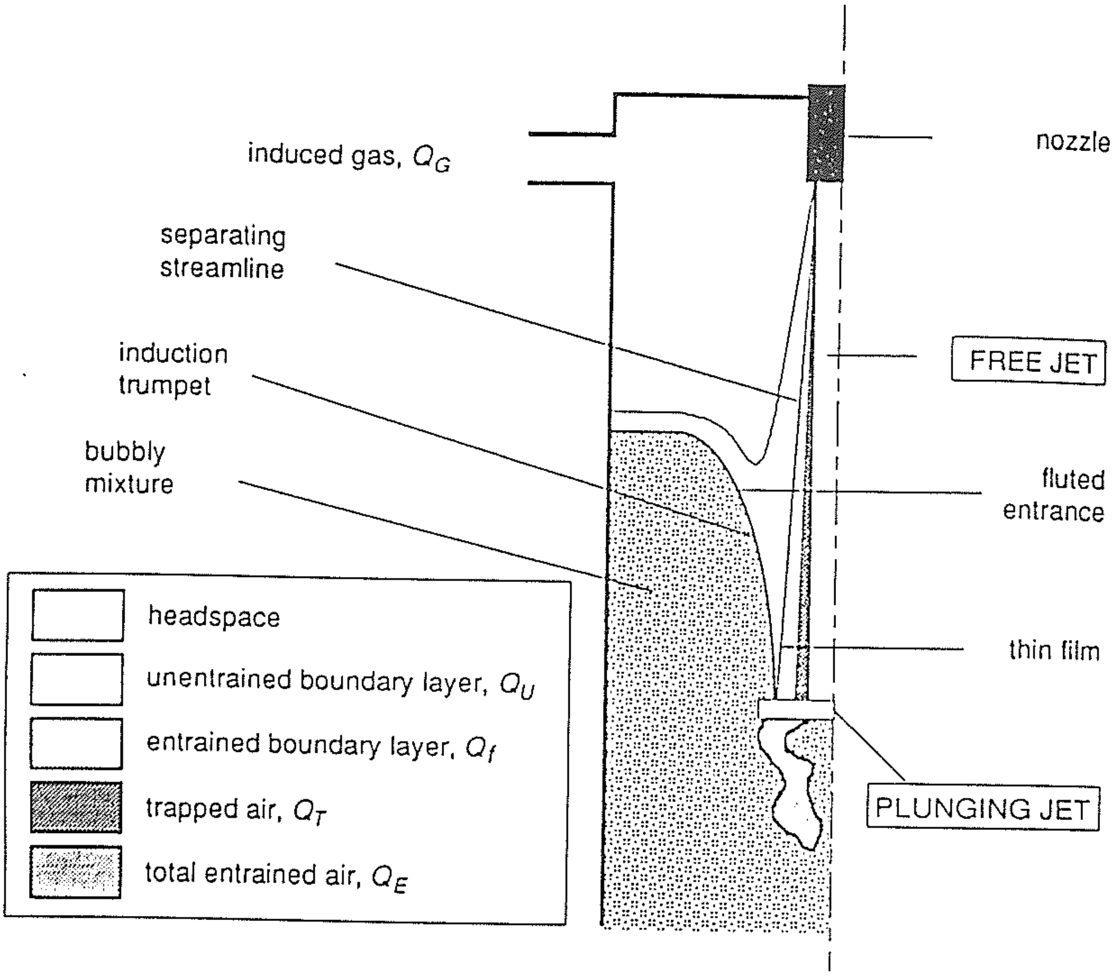


Figure 4 Components of air flow into plunging jet.

3. *Mixing Zone*

The mixing zone is defined as the volume occupied by (1) the fluid inside the submerged jet immediately below the plunge point that expands to occupy the cross-sectional area of the downcomer and (2) the body of fluid recirculating between the submerged jet boundary and the column wall—often referred to as the recirculating eddy. The high velocity gradients between the submerged jet and the recirculating eddy result in high energy dissipation rates within the mixing zone, which are responsible for the breakup of the air once it is entrained by the plunging jet. The entrained air is broken into fine bubbles before being transported downward into the pipe flow zone by the bulk fluid motion.

The size of the bubbles produced inside the mixing zone is related to the forces acting on the bubble surface. In low viscosity fluids, like those normally encountered in flotation circuits, the bubbles are deformed by forces arising from liquid velocity fluctuations acting over distances of the order of the bubble diameter. The restoring force resisting the deformation of the bubble is due to surface tension acting at the air–liquid interface. The ratio of these forces is known as the Weber number We and is defined as

$$We = \frac{\rho_L \overline{u^2} d}{\sigma} \tag{3}$$

where $\overline{u^2}$ is the average value of the squares of the velocity differences acting over a distance of the order of the bubble diameter d, and ρ_L and σ are the liquid (pulp) density and surface tension, respectively. If the bubbles are small compared to the turbulent macroscale but large compared to the microscale, the velocity fluctuations, following Hinze (8), can be related to the energy dissipation rate per unit volume $\dot{E}$ by the relationship

$$\overline{u^2} = C_1 \left(\frac{\dot{E} d}{\rho_L} \right)^{2/3} \tag{4}$$

where $C_1 \approx 2$ according to Batchelor (9).

For a given bubble size and at low Weber numbers, the deformation forces are dominated by the restoring forces, and the bubble is stable. However, as We increases, the shape of the bubble becomes increasingly distorted. Eventually a critical Weber number We_c is reached where the bubble becomes unstable and breaks up into smaller bubbles. The maximum bubble size d_m corresponding to the critical Weber number is

$$d_m = \left(\frac{We_c}{2} \right)^{3/5} \rho_L{}^{-1/5} \dot{E}^{-2/5} \tag{5}$$

Equation (5) has been used by Evans et al. (10) to predict the maximum stable bubble diameter inside the mixing zone, using estimates of the critical Weber number and the energy dissipation rate per unit volume.

The volume of the mixing zone is determined by the point of contact of the expanding jet and the wall of the enclosing downcomer. By considering the radial flow of momentum from the incoming jet to the recirculating eddies in the mixing zone, Evans (11) derived the following expression for the half angle β subtended by the expanding jet at the virtual origin:

$$\begin{aligned} \tan \beta &= \frac{D_d}{L_{MZ}} \\ &= \eta \left(\frac{P_1}{\rho_L U_j^2} \right) \left(\frac{\rho_{MZ}}{\rho_L} \right) \left\{ \frac{(0.37(R_c{}^2 - 0.5\,R_j{}^2))^{0.5} - 0.64 R_j}{R_j} \right\} \left\{ \frac{U_j - U_{e,\,max}}{U_j} \right\} \end{aligned} \tag{6}$$

where η is a jet energy transfer efficiency found by experiment to be 0.089, ρ_{MZ} is the density of the two-phase mixture in the mixing zone, R and R

are, respectively, the downcomer and jet radius, P_1 is the pressure in the headspace at the top of the downcomer, U_j is the jet velocity, and $U_{e,max}$ is the maximum return velocity in the eddy, which has been found by experiment to be 0.085 U_j.

The volume of the mixing zone is then assumed to be the volume enclosed by the cylinder formed by the downcomer of internal diameter D_d and the length L_{MZ} of the mixing zone.

Evans et al. (10) have compared the sizes of bubbles predicted by Equation (5), with measurements of the sizes of bubbles formed in an air water system, and found agreement within ±20%. Two downcomers of diameters 44 mm and 74 mm were used, and the jet velocity varied from 7.8 to 11.53 m/s. The bubble sizes were generally in the range 200–400 μm.

4. *Pipe Flow Zone*

The pipe flow zone is the region below the mixing zone inside the downcomer. The flow here resembles the downward flow in a vertical pipe. At low air flow rates, bubbly flow exists in which small discrete bubbles of different diameters move downward with the bulk liquid flow but not generally with the same velocity as the liquid phase. If the air flow rate is increased, the discrete bubbles coalesce and alternating air and liquid regions, or slugs, form inside the downcomer. The air slugs or Dumitrescu bubbles (12) have hemispherical caps that occupy nearly the entire cross-sectional area of the downcomer. The rise velocity U_r of these bubbles is a function of the downcomer diameter D_d and g, the acceleration due to gravity:

$$U_r = 0.496\left(\frac{gD_d}{2}\right)^{1/2} \tag{7}$$

At air flow rates when the Dumitrescu bubbles are initially formed, the drag and viscous forces are sufficiently great to give the bubbles a net downward velocity, and they are carried out through the base of the downcomer. However, if the air flow rate is further increased, the Dumitrescu bubbles become elongated, and the shearing forces present in the liquid make them unstable. This results in a chaotic mixture of air and liquid pockets known as churn-turbulent flow.

In terms of flow characteristics for the pipe flow zone, bubbly flow is the most desirable because it produces the greatest collection area for a given air/feed ratio, and it also provides for stable Jameson cell operation. With churn-turbulent and slug flow, the surface area per unit of gas volume is very low, and the collection efficiency is significantly reduced. Moreover, there is a much greater tendency for larger bubbles, with rise velocities greater than the net downward motion, to return to the top of the downcomer and be reentrained by the jet, reducing the amount of new air that can be introduced, and hence reduces the effective air/feed ratio.

5. *Disengagement Zone*

When the bubbly mixture reaches the bottom of the downcomer, it passes out into a vessel of larger cross-section, and the bubbles then disengage from the main pulp flow. For efficient action, the base of the downcomer is below the liquid level in the cell proper, and a hydraulic seal is maintained. Because of its low effective density, the bubbly flow initially hugs the outer wall of the downcomer, but as it rises it tends to entrain slower-moving fluid and then spreads laterally. The bubbles rise relative to the liquid and pass from the pulp layer into the base of the froth layer. The processes taking place in the froth are very important in the determination of the overall grade and recovery. If the gas superficial velocity J_g is too small or the froth depth too large, the bubbles coalesce and the froth degrades, leading to squeezing of the least hydrophobic particles and loss of recovery. If on the other hand the air rate is too high and the froth depth too shallow, the recovery may be high but at a low grade due to entrainment of gangue. Consequently, the cell must be operated to give optimum grade and recovery by manipulating the air rate and froth depth. If the air rate is too high altogether, the cell may "flood." This phenomenon occurs when the upward gas velocity exceeds the rate at which the liquid can drain back into the pulp layer and is evidenced in practice by the production of a very wet froth. In fact, when the cell floods, there is no distinction between the froth and pulp phases, and the interface that normally exists between them is lost. The phenomenon of flooding has been investigated in several recent papers. Pal and Masliyah (13) have given an empirical equation for the critical flooding velocity, and Langberg and Jameson (14) have given a more basic description, relating the flow in the froth and pulp phases to the flow of liquid in a packed bed, where in the present case the packing consists of the bubbles. Xu et al. (15) have given a comprehensive account of the various flow regimes and methods for calculating the limiting flow rates.

Washwater can be used with the Jameson cell as with other flotation columns, when high-grade products are required. In some cases it is possible to achieve the grades required by proper design of the cell, regarding the typical J_g allowed for, and by controlling the froth depth. However, the use of washwater adds another control variable of great power, whose proper use can lead virtually to total flushing out of the entrained gangue. The use of washwater does not unduly complicate the analysis of the flow presented in Refs. 13–15, and the same principles apply. The washwater flow is sometimes used to control liquid level as well as to wash the froth, but this practice seems to have nothing to recommend it and should be avoided as it adds further complexity, especially when the water flows needed for level control purposes are such as to be in conflict with optimum metallurgical conditions in the froth, for best grade and recovery.

The liquid layer in the cell forms a complex system of liquid and air recirculation patterns of varying turbulence and void fraction. The cell design is based on downcomer flows and downcomer placement to optimize this system to produce best grade and recovery. There is no limit on its volume, provided the net downward velocity of the pulp J_L is sufficiently low to avoid the entrainment of bubbles in the underflow. When the froth and disengagement zones have the same cross-sectional area, the two important velocities are the rate of rise of the bubbles in the pulp, and the rate of drainage of liquid in the froth. The former is usually greater than the latter, so that a column sized to give the correct J_g will also give the correct J_L, and bubble entrainment in the downward flow will not be a problem.

C. Interactions and Operating Stability

The interactions taking place between regions inside the Jameson cell and the various recycle paths taken by the air are illustrated by the flowchart shown in Figure 5. The performance of the cell is determined principally by the collection rate of the bubbles inside the downcomer and to a lesser extent by the separation of the product from the tailings stream in the riser. In general, for a given cell cross-sectional area, the collection rate is increased by increasing the air/feed ratio into the downcomer. However, an increase in this ratio can also lead to instabilities and in some cases to the complete collapse of the system. The operating stability of the Jameson cell as a function of air feed rate is shown in Figure 6.

At low rates of air induction, the level inside the downcomer is sustained just below the level of the nozzle, indicating that the plunging jet can effectively entrain all of the air in the headspace. The air is dispersed into very fine bubbles, which are carried downward by the net fluid movement, resulting in minimal recirculation of air back into the headspace from the mixing zone. In general, this is the ideal mode of operation as it provides (1) the most stable operating condition, (2) very small bubbles, and (3) the longest residence time inside the downcomer, thereby maximizing the bubble–particle contact time. The only disadvantage with this operating mode is the relatively smaller air/feed ratio inside the downcomer, which may lead to lower rates of production of interfacial area. (Recent unpublished work has shown, however, that in some practical cases a reduction in the ratio is compensated directly by the reduction in bubble size and hence an increase in the surface area of interface produced by a given volume of

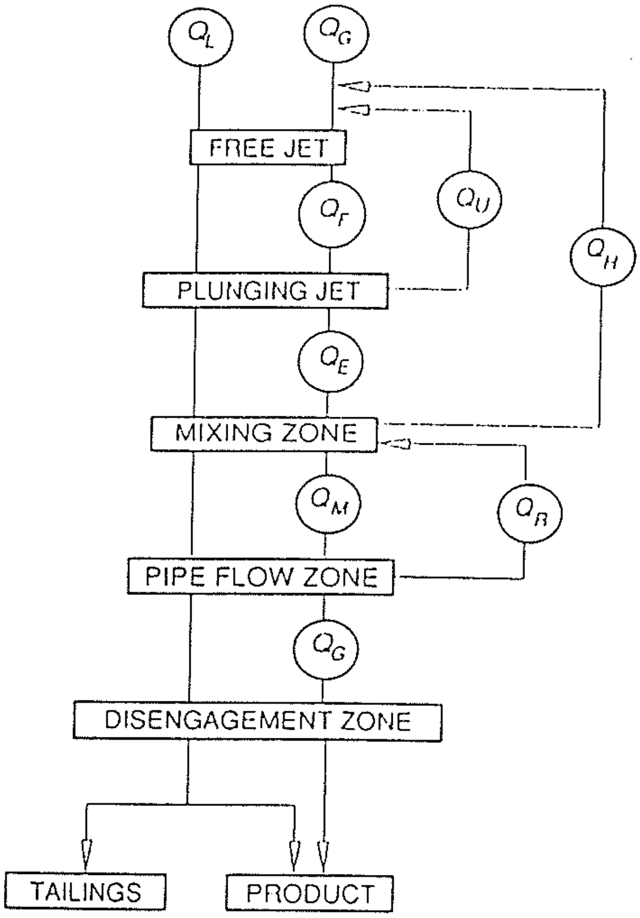


Figure 5 Interaction paths in the Jameson cell.

air. Thus a reduction of air rate, which gives much more stable downcomer operation, does not necessarily lead to a reduction in recovery, providing the J_g is maintained constant.)

If the air rate is increased, a point is reached where the jet can no longer entrain all of the air being introduced into the headspace, and the froth level in the downcomer starts to drop. Fortuitously, it appears that some type of compensatory phenomenon is at work. This phenomenon has the effect that the rate of entrainment of air by the plunging jet increases up to a point, as the length of free jet increases. The phenomenon is possibly linked to the increase in the effective jet diameter with increase in the free jet length. Whatever the reason, a new equilibrium height is reached inside the downcomer, marking the length where the jet can effectively entrain all of the entering air. In a sense the jet is self-regulating in that the length of the free jet will increase to accommodate the increase in the amount of air added to the downcomer (up to a point).

An increase in the free jet length also leads to (1) a reduction in the froth height inside the downcomer, resulting in a reduction in the residence time

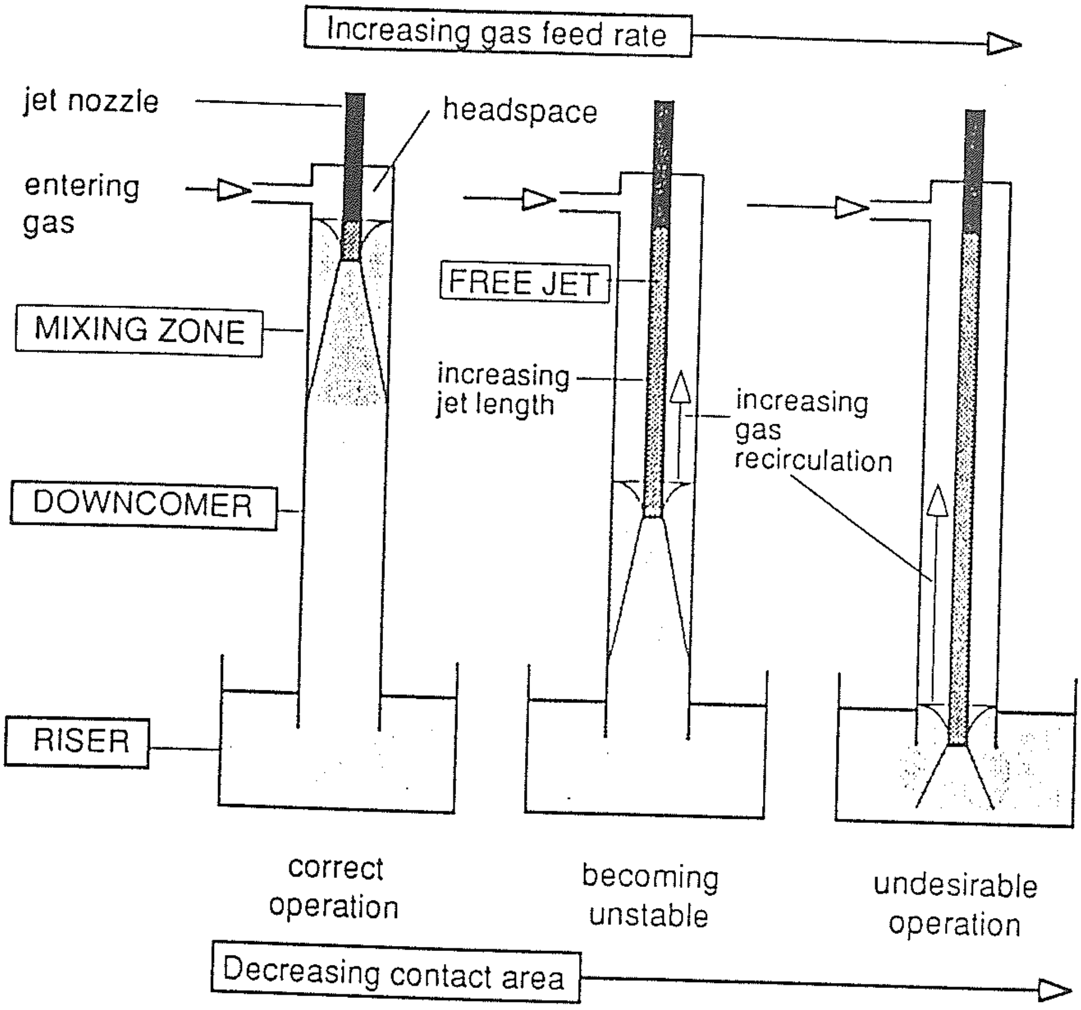


Figure 6 Variation of downcomer operation with airflow rate.

of the bubbles, and (2) an increase in mean bubble size generated within the mixing zone (11), resulting in a decrease in the amount of interfacial area generated per volume of air inside the downcomer.

More important than the potential loss in recovery rate is the possible effect of increasing the air feed rate on the overall stability of the system. At very high air rates, bubble coalescence results in the formation of large gas slugs in the pipe flow zone, which restrict the flow of air through the downcomer. This leads to an increase in the pressure in the headspace causing a drop in the froth level and, in the extreme case, the complete collapse of the system.

D. Conditions in the Downcomer

A number of investigators have studied the void fraction and gas holdup in the downcomer. Sanchez-Pino and Moys (16) measured the hydrostatic pressure between two points in a vertical downcomer and used a drift-flux method to analyze the data. The void fraction or gas holdup varied from 20% at an air/feed ratio of 0.1 to 55% at a ratio of 0.9. Marchese et al. (17,18) used a conductometric technique to measure the void fraction in the downcomer and compared their results with those from a total isolating method in which the contents of the downcomer are trapped between two valves in-line, which are closed simultaneously.

The values they obtained can be contrasted with those found in conventional columns that are typically of order 15%. The high void fractions explain the extremely rapid kinetics found in the downcomer; the values suggest that the bubbles are in fact approximating the close-packed spherical limit and that it would be unrealistic to expect void fractions in excess of 55 to 60%. Void fractions above 60% have been observed, but the most likely explanation is that the contents of the downcomer then consist of a dense foam of average void fraction 55 to 60%, embedded in which is a small number of relatively large slugs or Dumitrescu bubbles, which would contribute virtually nothing to the collection processes in the froth.

IV. JAMESON CELL OPERATING PARAMETERS

Some general operating characteristics of the Jameson cell are now discussed.

A. Froth Depth

The froth phase produced in a Jameson cell can be controlled in the same manner as in conventional columns. Shallow froth depths (less than 200 mm) are used where high recovery is necessary and grade is of secondary importance, while deeper froths (up to 1000 mm) are employed to obtain maximum concentrate grade. Shallow froths result in significant entrainment of very fine (less than 10 μm) gangue mineral components that accompany the pulp phase. Use of deeper froths results in significant drainage of hydrophilic gangue producing a higher grade concentrate and a higher percentage of solids in the concentrate. Under some circumstances, addition of counter-current froth washwater assists froth mobility, where otherwise excessive froth drainage would result in an immobile froth.

B. Superficial Gas Velocity

The superficial gas velocity J_g (cm/s) is the upward superficial velocity of air in a flotation cell. In the Jameson cell, the J_g is calculated by dividing the downcomer air rate (cm^3/s) by the cross-sectional area (cm^2) of the riser part of the cell. The cell is normally circular or rectangular in section, and the appropriate cross-sectional area is simply the area normal to the direction of the flow of the froth, excluding the area occupied by the downcomer(s). It is conveniently expressed in units of cm/s because values typically range from 0.5 to 4 cm/s in practice.

For a stream that is not carrying-capacity limited (see Sec. IV.H), the recovery and concentrate carrying rate (g/min/cm^2) tend to increase with increasing J_g, as in conventional columns (see Refs. 19 and 20). For a given stream and frother concentration, a maximum air rate (subsequently $J_{g,max}$) is reached, above which froth flooding occurs, resulting in the loss of froth-pulp interface, a very wet froth, and total loss of selectivity. In flooding, the entire cell fills with froth as the only stable phase, and there is no pulp phase.

Some data of interest are shown in Figure 7 for a lead/zinc circuit. Data were obtained using a Jameson cell in parallel with a conventional column. In each case, it is seen that there is a strong correlation between the production rate of solids and the superficial gas velocity J_g. For a given air rate, the Jameson cell gives greater production rates than the column, presumably a consequence of the smaller bubbles generated in this device.

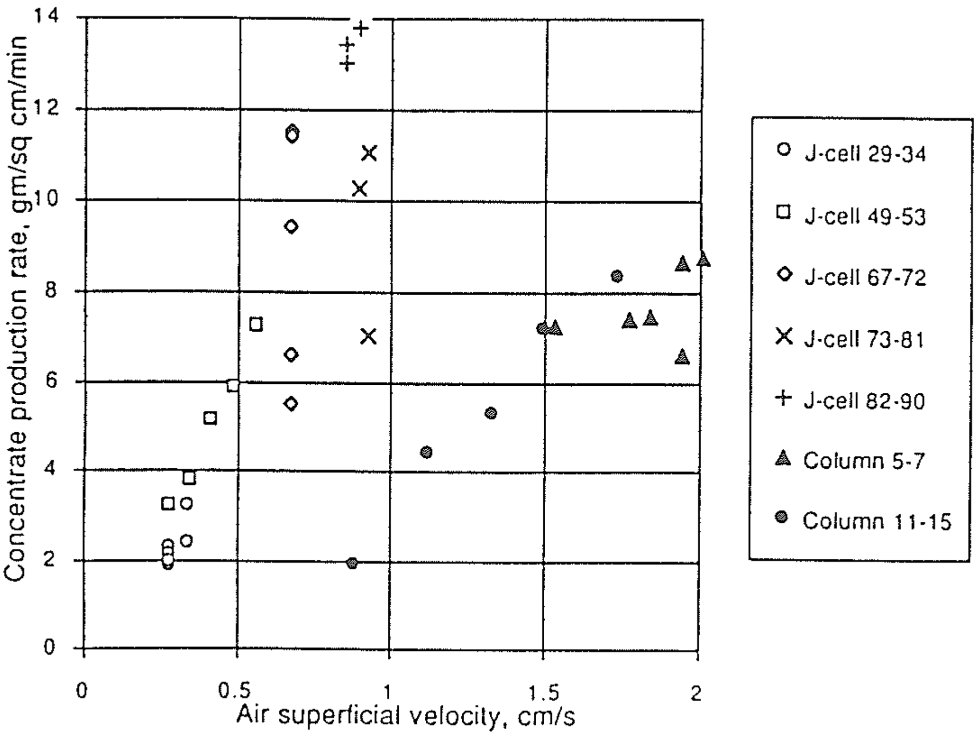


Figure 7 Variation of concentrate production rate with superficial air velocity J_g, for a lead/zinc flotation, with a conventional column and a Jameson cell in parallel. The spread in data is due to variations in froth depth. The numbers in the legend refer to various test runs.

The operating J_g used in the sizing of the Jameson cell depends strongly on the application, and on the residual reagent concentrations from any upstream processes. Generally speaking, low values (J_g = 0.4 to 0.8 cm/s) are employed in cleaning applications, and high J_g's (1.0 to 2.0 cm/s) are employed in roughing or scavenging applications. The reasons for these choices are as follows. In cleaning operations, a high proportion of the feed reports to the concentrate, and the froth loading tends to be high. Consequently, the bubbles are well coated with particles, which tend to stabilize the froth. The drainage rate of the interstitial liquid in the froth is retarded by the relatively high concentration of particles, so it is necessary to design for lower values of J_g to allow time for the gangue to drain from the froth to obtain the required high grade. In roughing applications, however, only a small fraction of the feed reports to the concentrate, and the froths formed tend to be less stable as a consequence. Also, gangue entrainment is not such a serious problem, because it can be coped with in the downstream cleaning circuit. In order to maintain a stable froth, it is therefore usual to design a Jameson cell for a roughing application with a higher J_g than in the cleaners.

In some circumstances, high residual concentration of reagents in the feed necessitates the use of low J_g's to avoid froth flooding. Although frother concentration is of primary importance to bubble size and hence the advent of froth flooding, circumstances have risen where collector (xanthate) and frother (methyl isobutyl carbinol, MIBC) interaction has been observed (21). In such a case, the frother dose should be decreased if the collector dose is increased, and vice versa. Too high a frother or collector concentration can lead to froth flooding, while too low a dose can lead to loss of froth stability.

Particle size can also have an influence on the maximum J_g, due to its effect on froth stability through bubble bridging. Small particles (less than 100 μm) are easily collected at low gas rates (J_g less than 1.0 cm/s), while recovery of coarser particles (greater than 100 μm) may be assisted by higher rates (J_g greater than 1.0 cm/s).

C. Bubble Size

Bubble sizing determinations for full-scale operating cells and test cells at the University of Newcastle show that the Jameson cell produces an arithmetic mean bubble diameter of the order of 300–600 μm, while the Sauter (volume-to-surface) mean diameter d_{vs} is of order 360–950 μm (21). These sizings compare very favorably with conventional columns where the Sauter mean bubble size is typically 2 to 3 mm.

Figure 8 shows bubble size versus air/feed ratio for a 300-mm diameter

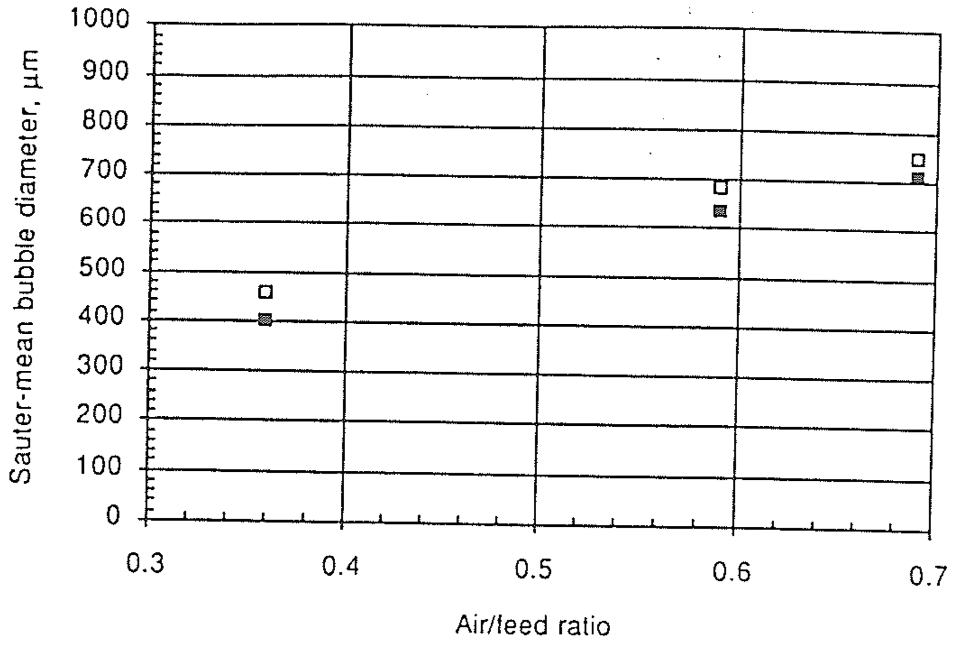


Figure 8 Measured values of the Sauter mean bubble diameter in exit stream from a Jameson downcomer as a function of the air/feed ratio, in flotation of a copper retreatment stream. Results for two separate determinations are shown.

Jameson pilot cell treating a fine copper retreatment stream ($d_{80} = 30\ \mu m$). These data confirm visual observations that bubble size decreases with decreasing air/feed ratio.

Chatiar (22) carried out experiments in the laboratory using air and water with MIBC as frother and found that as the concentration increased, the mean bubble size was reduced, up to a concentration of 30 ppm V/V, beyond which there was little change.

D. Air/Feed Ratio

Jameson cells generally operate with a volumetric air/feed ratio of 0.3–0.9. Experiences with large (2–3 m diameter) Jameson cells indicate that operation at a low air/feed ratio does not appear to detract from metallurgical performance *providing the superficial gas velocity J_g is maintained* above a certain minimum value. Operation at lower air/feed ratios has a stabilizing effect producing a more uniform and finer bubble size. A significant advantage of operation at lower air/feed ratios is that lower concentrations of frother are required.

Maintenance of metallurgical performance at low air/feed ratios can be explained by considering the flux of interfacial area S_b (interfacial area/s) per unit of column cross-sectional area (s^{-1}) as defined by Finch and Dobby (19):

$$S_b = 6\frac{J_g}{d} \tag{8}$$

where d is the bubble diameter. The superficial bubble surface rate (and hence the collection capability) can be maintained with reduced superficial gas rate providing the bubble size decreases accordingly. The data in Figure 9 were calculated using information from Figure 8 using Equation (8) and show the superficial bubble surface rate as a function of the air/feed ratio for a fixed cell cross-sectional area. It is seen that the superficial bubble surface rate remains almost constant at 55 s^{-1} approximately, despite two-fold changes in the air/feed ratio.

E. Washwater Ratio

The washwater ratio is defined as the ratio of the washwater addition rate to the flow rate of water in the concentrate:

$$W = \frac{Q_{WW}}{Q_{WC}} \tag{9}$$

where Q_{WW} is the washwater flow rate and Q_{WC} is the rate of flow of water in the concentrate. The washwater ratio is a relative measure of the amount of washwater applied.

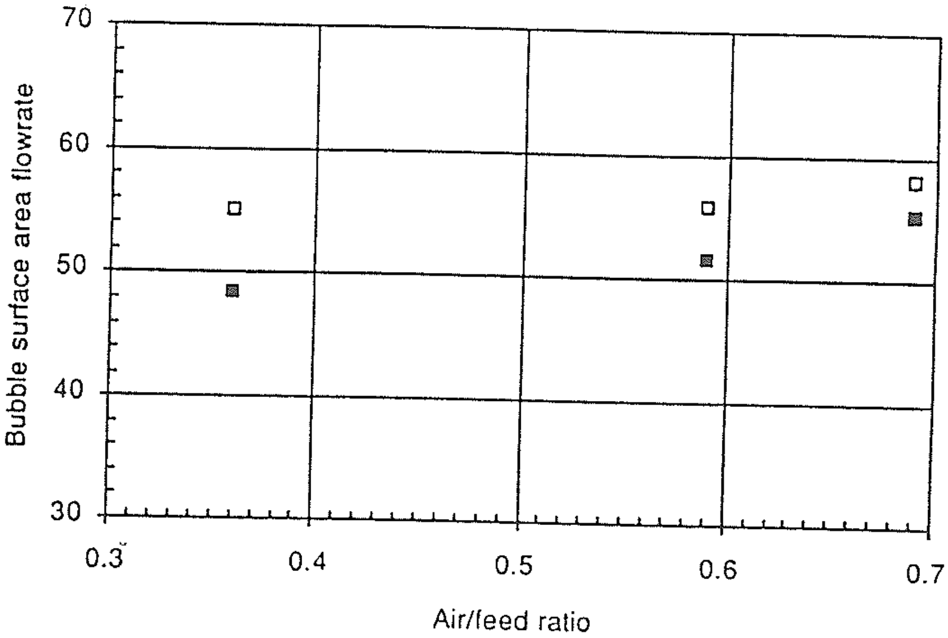


Figure 9 Flux of interfacial area for bubbles generated by downcomer and passing into the froth phase at the air/feed ratios shown in Figure 8. The interfacial flow rate S_b is calculated from Equation (8). The drop in airflow rate at low values of the ratio is mainly compensated by a corresponding decrease in the mean bubble size.

Another common method of describing the washwater addition is in terms of the bias, that is, the absolute excess of the washwater applied over the quantity of water being recovered in the concentrate, expressed as a superficial velocity J_b (cm/s):

$$J_b = \frac{Q_{WW} - Q_{WC}}{A_C} \tag{10}$$

where A_C is the cross-sectional area of the column. If no washwater is used, the washwater ratio is zero, and the bias is negative. When $J_b = 0$, $W = 1$; positive bias corresponds to washwater ratios greater than unity.

Although the bias does give an indication of the absolute amount of washwater being added, its use can be misleading because it does not take into account the wide variation in the absolute values of the rate of water entrainment in the concentrate. It is preferable to use the washwater ratio, which is a relative figure.

Consider, for example, a rougher application in which the mass flow of recovered solids is only a small fraction of the feed. In such a case, the flow of concentrate will be low and a correspondingly low volume of washwater will be needed to replace the water being carried out of the cell in the concentrate. Thus good froth washing may be achieved with a relatively low positive bias, say 0.005 cm/s. On the other hand, in a cleaning operation, where the mass flow of concentrate is high, a much higher superficial flow rate of washwater would be needed to replace the water in the concentrate, since there is much more water being removed in the froth.

Figure 10 shows a comparison of measured washwater ratios and bias rates for a range of operating conditions for a Jameson cell operating on a zinc rougher application. Although the bias covers a relatively narrow range, the washwater ratio has much greater variation. To an operator, the difference in bias between 0.02 and 0.04 cm/s may seem negligible, but it can be seen that in some circumstances, such changes in J_b may lead to a threefold increase in the washwater ratio – supplying in effect a 200% overload of washwater.

Although it might be thought that a cell should be operated with $W > 1$, there is evidence to show that the optimum operating value should be in the range 0.5 to 1. Figures 11 and 12, for example, show the effect of washwater ratio on recovery and grade, respectively, of a fine copper retreatment stream ($d_{80} = 30$ μm). As the washwater ratio is increased, the recovery decreases significantly up to $W \approx 1$ and then decreases more gradually for $W > 1$. The grade increases progressively up to a washwater ratio of 1.0, but no further grade improvement is observed at washwater ratios above this value. These data, together with other Jameson cell operating

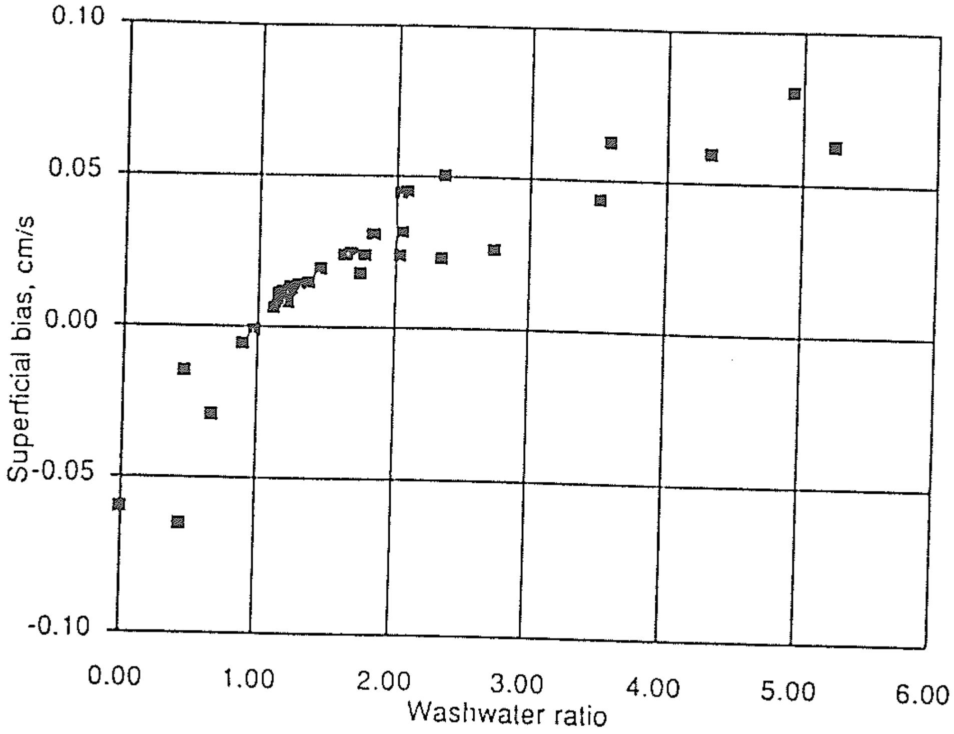


Figure 10 Plot of superficial washwater bias J_b against the washwater ratio calculated for the same zinc rougher flotation runs. Note the large variation in washwater ratio for relatively small changes in the absolute bias.

data, confirm that the best washwater operating point is at a washwater ratio of slightly less than unity, where recovery and grade are both high. Higher washwater addition simply results in decreased recovery without compensating improvements in grade. Precise on-line washwater bias measurement is highly desirable in order to maximize recovery at required grade.

F. Reagents

In streams tested to date, MIBC, long-chain alcohols, polyglycol propylenes, and polyglycol ethers have been used as the frothers in the feed to the Jameson cell, usually in the range 5 to 25 ppm. In cleaning applications, generally no frother addition is required due to the residual concentration from the roughing stage. In some cases, excessive frother from the upstream stage can lead to a reduction in maximum superficial gas rate J_g, which can be applied in the Jameson cell. Froth flooding is initiated at lower air rates due to the finer bubble size generated by an excess frother concentration (see Ref. 19, p. 22).

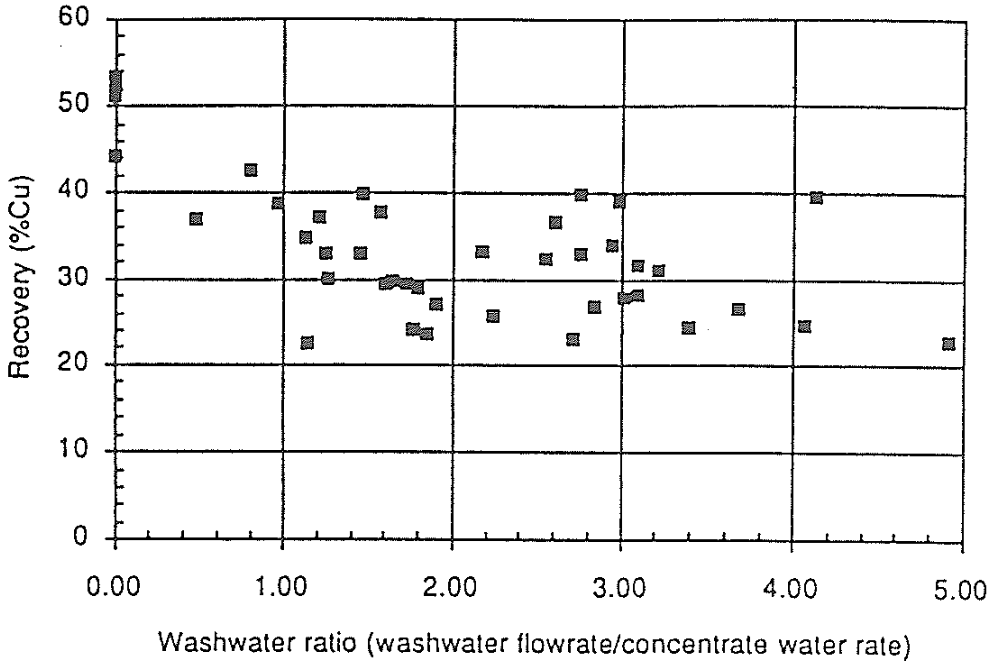


Figure 11 Effect of washwater ratio on recovery in the flotation of a copper retreatment stream. The apparent scatter here and in Figure 12 is due to differences in flotation conditions between individual runs, particularly in the froth depth and air rate. Note that the progressive decline in recovery as washwater is increased.

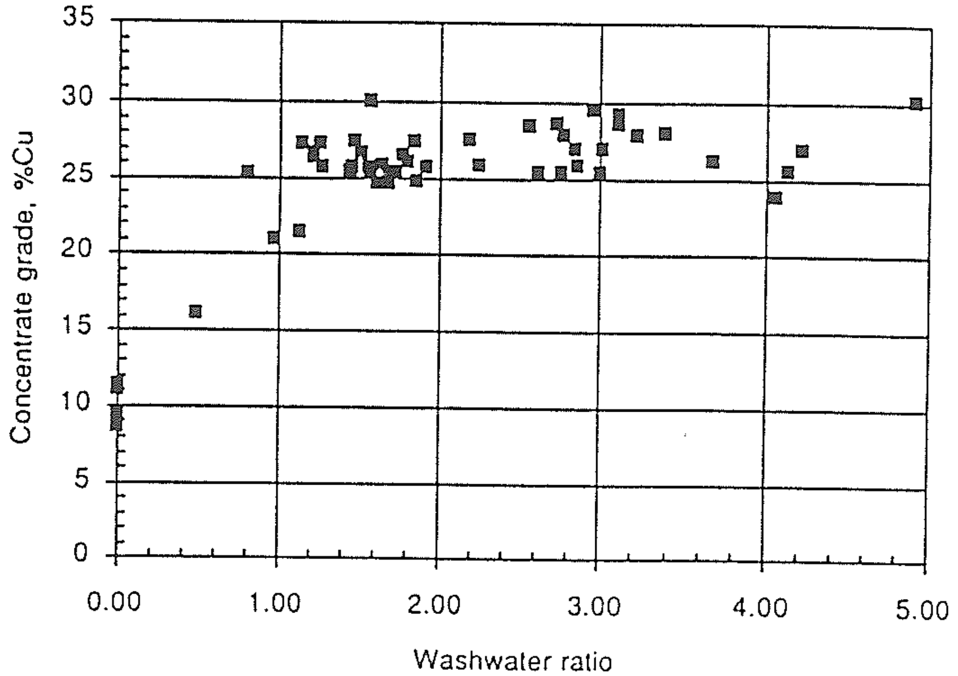


Figure 12 Grade vs. washwater ratio in flotation of a fine copper retreatment stream. The grade increases markedly until the washwater ratio reaches a value of 1 approximately and thereafter remains constant.

G. Downcomer Void Fraction

Experimental studies by Evans (11) and Marchese et al. (17,18) have shown that the void fraction (gas holdup) in a Jameson downcomer is generally in the range 50 to 60% by volume.

The downcomer void fraction can be measured directly, either from the isolating technique or from conductivity measurements. It can also be calculated using a downcomer momentum balance.

The gas void fraction ϵ in the dense foam inside the downcomer is defined as

$$\epsilon = \frac{V_G}{V_G + V_L} \tag{11}$$

where V_G and V_L are the volumes occupied by the gas and liquid (pulp) phases, respectively. In the isolating method, the contents of a section of the downcomer are trapped between two valves that are closed simultaneously, thus allowing direct measurement of V_G and V_L.

With the conductometric method, use is made of classical potential theory to relate the measured conductivity to the voidage. Thus Marchese et al. (17) applied the relationship of Maxwell (23) and found

$$\epsilon = 2\left(\frac{\kappa_L - \kappa_{LG}}{2\kappa_L + \kappa_{LG}}\right) \tag{12}$$

where κ_L and κ_{LG} are the measured conductivities of the liquid (pulp) and the liquid–gas mixture, respectively.

The void fraction can also be estimated from measurements of the pressure in the headspace in the downcomer. In the absence of dynamic effect, ϵ can be calculated from simple hydrostatic pressure considerations. Thus a pressure balance over the downcomer contents yields

$$P_2 - P_1 = \rho_m g h \tag{13}$$

where P_1 and P_2 are the pressures in the headspace and in the discharge end of the downcomer, respectively; h is the height of the dense foam in the downcomer; and ρ_m is the average density of the two-phase mixture inside the downcomer. When the gas density is negligible compared with the liquid density, it can be shown that ϵ is given by

$$\epsilon = 1 - \frac{\rho_m}{\rho_L} \tag{14}$$

Thus from Equation (14) we have

$$\epsilon = 1 - \frac{(P_2 - P_1)}{\rho_L g h} \tag{15}$$

Dynamic effects can be included by application of the momentum principle to the froth zone. The sum of the forces acting on the froth is equal to the gain in momentum flow rate between inlet and outlet, so

$$(P_1A_d - P_2A_d) + \rho_m ghA_d = \rho_m U_m(Q_L + Q_G) - \rho_L U_j Q_L - \rho_G U_G Q_G \quad (16)$$

where A_d is the area of cross-section of the downcomer and Q_G and Q_L are the gas and liquid flow rates, respectively. Ignoring the term in the gas density, we find:

$$\epsilon = 1 - \frac{(P_2 - P_1)}{\rho_L gh} + \frac{U_jA_j}{ghA_d}\left(U_j\left(1 - \frac{A_j}{A_d}\right) - \frac{Q_G}{A_d}\right) \quad (17)$$

where A_j is the cross-sectional area of the jet. It is seen by comparison with equation (15) that a dynamic correction term has been introduced into the expression for the void fraction. The correction is usually significant, contributing 25 to 50% of the calculated overall void fraction, depending on conditions.

H. Carrying Capacity

The carrying capacity *Ca* is the limiting or maximum concentrate production rate per unit of area of cell cross-section, usually expressed as concentrate solids rate (g/min) per unit of available cell cross-sectional area (cm^2). The carrying capacity is strongly influenced by the gas flow rate and the size of the bubbles in the froth discharging at the overflow lip, since these two factors influence the rate of surface area discharge into the launder. The mass of the hydrophobic particles that can be carried by the froth varies directly with the surface area. The other important factor is the size of the particles, because when a layer of particles is adhering to a gas–liquid interface, the mass of particles per unit of interfacial area varies directly as the mean particle size.

Carrying capacity limitations generally apply where high-grade streams are treated, as in cleaning applications, where the bubbles can easily become fully loaded with particles. The recovery is then determined by the carrying capacity rather than by the flotation kinetics.

The expected carrying capacity of conventional columns has been given by Espinosa-Gomez et al. (24) as

$$Ca = \alpha d_{80}\rho_p \quad (18)$$

where d_{80} is the size at which 80% by mass of the concentrate passes, expressed in μm, and ρ_p is the density of the particles (g/cm^3).

In conventional columns, the carrying capacity depends on column diameter. The parameter α has been found to be 0.068 in small columns (as reported by Espinosa-Gomez et al. (24) for a 50-mm column). For larger columns, up to 1.0 m diameter, Finch and Dobby (19) give $\alpha = 0.05$; and for columns greater than 2.0 m diameter, $\alpha = 0.035$ (25).

Figure 13 shows carrying capacities measured by Atkinson et al. (21) as a function of the washwater ratio in a Jameson cell treating coarse chalcopyrite (autogenous mill discharge, $d_{80} = 400\ \mu m$). It is seen that the maximum carrying rate is strongly influenced by washwater addition.

These data show that the carrying capacity for a Jameson cell must be defined at a particular washwater ratio. For maximum concentrate grade and recovery, it is generally necessary to aim for a washwater ratio of 1.0 or just slightly lower. Thus it would be appropriate to define the carrying capacity at a washwater ratio of 1.0. In all Jameson cell applications, concentrate carrying capacities have been shown to be at least as high as the values predicted for conventional columns using Equation (18).

V. APPLICATIONS

A. General

Table 1 shows a summary of sites employing Jameson cells in mineral or coal flotation applications. In most applications, the Jameson cell is capable of providing a final grade concentrate in a single stage, providing the

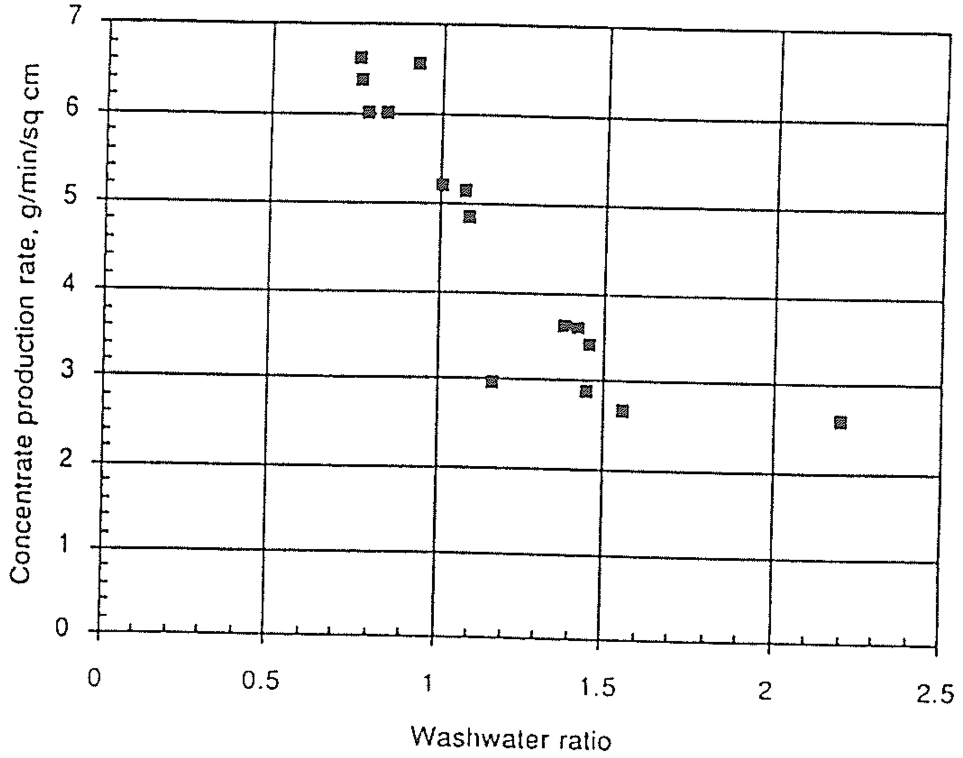


Figure 13 Concentrate production rate as function of washwater ratio for coarse chalcopyrite flotation, showing a strong reducing trend as washwater ratio is increased.

Table 1 Summary of Jameson Cell Sites (January 1993).

Company	Location	Project
Mount Isa Mines	Mount Isa, Australia	Lead/Zinc slimes
Mount Isa Mines	Mount Isa, Australia	SX-EW
Mount Isa Mines	Hilton, Australia	Low-grade middlings
Mount Isa Mines	Hilton, Australia	Lead cleaning
Peko Mines	Warrego, Australia	Copper cleaning
Amalg Syndicate	Moonta, Australia	Copper cleaning
Mamut Copper	Malaysia	Copper cleaning
Newlands Coal	Queensland, Australia	Fine coal
Amalg Syndicate	Spargoville, Australia	Nickel cleaning
Amalg Syndicate	Spargoville, Australia	Nickel roughing
Kidd Creek	Timmens, Canada	Copper/Zinc
Mount Isa Mines	Mount Isa, Australia	Fine zinc cleaning
Western Mining Corp	Olympic Dam, Australia	SX-EW
Phelps Dodge	Morenci, USA	SX-EW
Cons Murchison	South Africa	Antimony/gold
Matos Blancos	Chile	Copper cleaning
C.C.P.	Collinsville, Australia	Fine coal
Sth Atlantic Ventures	Oracle Ridge, USA	Copper roughing
White Mining	Nth Goonyella, Australia	Fine coal
BHP Australia Coal	Blackwater, Australia	Fine coal
Girilambone	NSW, Australia	SX-EW

values are sufficiently liberated and washwater is applied (where necessary). The Jameson cell is efficient for the recovery of the full range of particle sizes experienced in flotation feed streams. Examples of extremes include the recovery of coal slimes ($d_{80} = 25\ \mu m$) and very fine sphalerite/galena particles ($d_{80} = 10\ \mu m$) at the lower end of the particle size spectrum and, at the coarse end of the range, coal particles up to 1000 μm and chalcopyrite particles up to 400 μm.

Details of the metallurgical performance of operating Jameson cells have been given in a number of papers. Kennedy (26) described some of the early full-scale applications, in the lead/zinc concentrators at Mt. Isa, Queensland, and Hilton, Queensland; at the copper concentrator at the Peko Warrego mine, Tennant Creek, Northern Territory; and at the Newlands Coal operation in the Bowen Basin, Queensland, all in Australia.

The commissioning and operation of the Peko Warrego copper flotation cells has been described by Jameson et al. (27). The original concentrator produced a concentrate of 23% copper from a chalcopyrite ore containing pyrite, magnetite, hematite, and quartz, with traces of covellite and bismuthinite. The reason for the Jameson cell installation was to increase the concentrate grade from 23 to 28% or better, in order to reduce transport costs. Two 1.4-m circular cells were installed and commissioned in April 1990, taking a feed of up to 40 tph dry solids at 35% solids in the slurry. The required grade increase, obtained in the original pilot-plant tests, has been proven in plant operation.

A description of the Jameson cell installations at the Mt. Isa and Hilton mines has been given by Jameson and Manlapig (28) and Clayton et al. (29).

The flotation of fine coal slimes at the Newlands Coal operation has been described by Jameson et al. (30). The washery has an output of 1200 tph of coal, which is washed mainly in Batac jigs. The water stream from the jigs is allowed to settle and is returned for reuse. A side stream of 1250 m^3/h is taken off to prevent accumulation of slimes in the circuit, and previously this bleed stream had been sent to a tailings dam. However, the solids in the stream were in the range 3 to 8%, representing a considerable loss in ultrafine coal, around 200,000 tpa. Accordingly it was decided to recover the very fine coal, whose d_{50} is 22 μm, by flotation. Tests with mechanical cells showed that it would not be possible to recover the coal at the required ash level (12% maximum) except at unacceptably low yields, but with the Jameson cell it was found that the required ash level could readily be achieved, without the use of washwater. With a single stage of operation, the combustible recovery was 70%, and with two stages the overall recovery was 90% or better. A full-scale installation was commissioned in early 1990 and has operated ever since. Large installations are scheduled to start in 1993 at two other coal washeries in the Bowen Basin, at the BHP Australia Coal Blackwater mine and the North Goonyella joint venture.

B. Rougher Applications

The Jameson cell has been successfully employed in roughing applications and can produce a concentrate of final grade straight from run-of-mine ore. In such cases, it is possible to upgrade the capacity of a mill that has been operating with conventional flotation machines. The Jameson cell is then used as a scalper to remove a significant proportion of the feed, removing the load from the rest of the existing circuit, which can then be used to recover and clean the tails from the Jameson cell.

An interesting set of data has been obtained when a Jameson cell was set up to treat an autogenous mill discharge stream, on a chalcopyrite ore. Figure 14 shows the overall grade-recovery performance of the Jameson cell for single stage treatment of this stream. The average recovery was

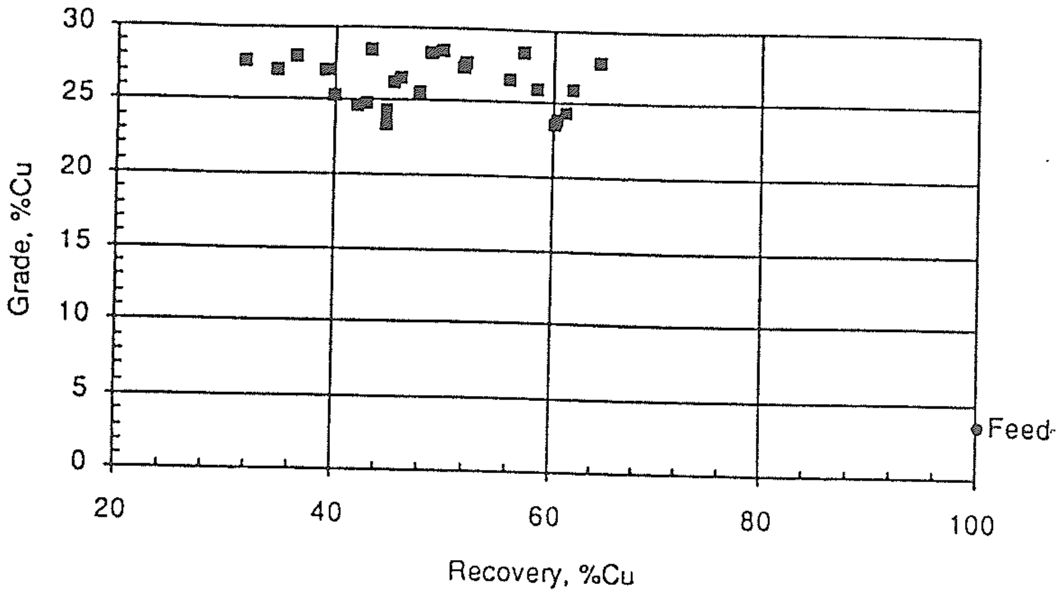


Figure 14 Grade-recovery data for single-stage Jameson cell flotation of discharge stream from autogenous mill discharge, chalcopyrite flotation circuit. Feed size d_{80} = 400 μm.

50%, at a concentrate grade of 27% Cu, from a feed grade of 3% Cu. The size analyses of the feed and the corresponding concentrate are shown in Figures 15 and 16, respectively. The d_{80} of the feed is of the order of 400 μm, while the d_{80} of the concentrate is 50 μm; most of the floatable particles are in the <20 μm fraction.

These data demonstrate the wide-ranging capability of the Jameson cell as an excellent flotation machine, capable of treating all particle sizes typically encountered in mineral processing.

VI. CONTROL REQUIREMENTS

The principal control requirements for a Jameson cell are

1. froth depth
2. washwater ratio

Froth depth control is essential for concentrate grade stability. It can be readily achieved by some type of direct or indirect level sensor (e.g., bubble pressure probe, diaphragm pressure sensor, ultrasonic sensor, float level indicator) or by a simple gravity overflow. In most applications the froth depth is between 300 and 800 mm, although for high-recovery operations where grade is not so important, froth depths as low as 50 mm are used. Where washwater is used, some measure of control is desirable in order to maximize recovery while maintaining required concentrate grade. There is

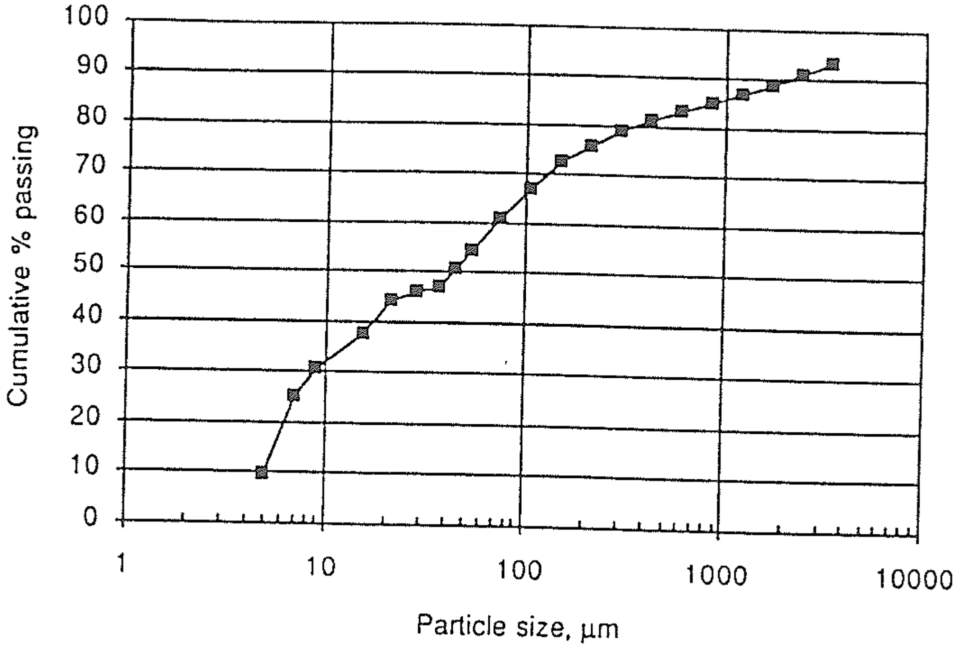


Figure 15 Cumulative size distribution of chalcopyrite flotation feed, autogenous mill discharge.

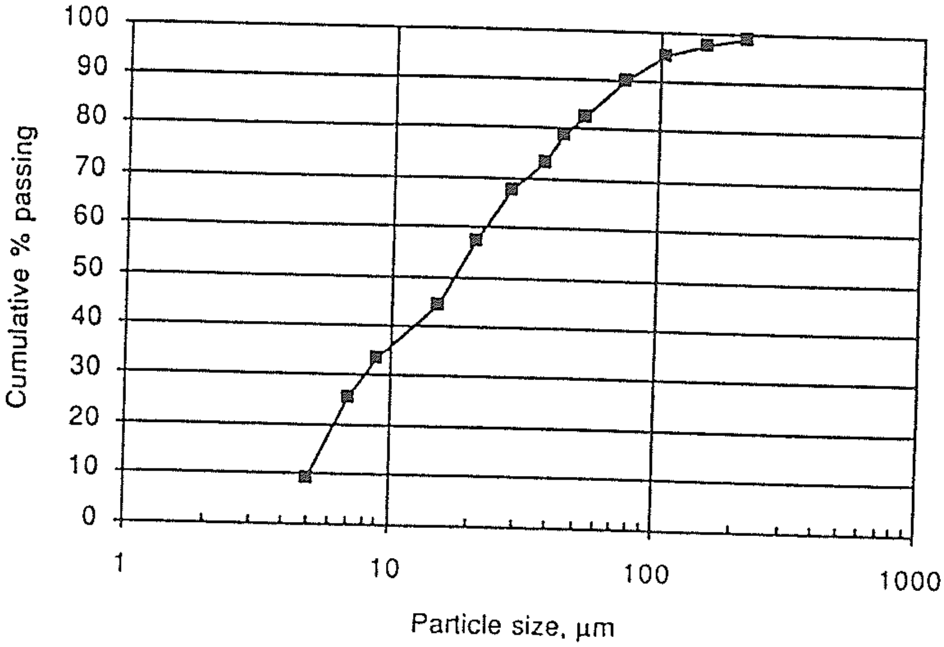


Figure 16 Cumulative size distribution, concentrate from flotation of autogenous mill discharge; $d_{80} \approx 50\ \mu m$.

some latitude in that a high grade is usually achieved for all washwater ratios greater than 1, but recovery may suffer if too much washwater is used.

Aeration of pulp occurs in the Jameson cell in the downcomer. In order to cope with feed changes, it is customary to design the cell for a maximum expected flow rate. Variations in feed can be dealt with by a number of strategies, including a variable-speed drive on the feed pump, controlling make-up water additions in the feed sump, and recycling of tails or an intermediate stream to maintain a constant flow. The cell design has been developed so that isolation of downcomers in conjunction with suitably positioned baffle plates allows fractions of the cell to be operated without affecting J_g or other design fundamentals. The solution chosen depends on the circumstances pertaining in the plant.

VII. AMENABILITY TESTING

Amenability testing is the term used to cover pilot plant testing for the purpose of assessing a technology for a given application. Table 2 provides a matrix covering the ranges of variables that should be tested in order to evaluate the capabilities of a Jameson cell fully.

In particular applications, some of the conditions may be omitted or given lower importance, e.g., mineral cleaning applications would always require washwater, whereas roughing applications or coal flotation may not. Each test should be repeated at least three times to ensure that experimental outliers are identified.

Table 2 Jameson Cell Amenability Testing Program for a Given Stream.

	Washwater ratio			Froth depth (mm)			Air/feed ratio		
Test	0	0.5	1.0	250	500	750	0.4	0.6	0.8
1	*			*				*	
2		*		*				*	
3			*	*				*	
4	*				*			*	
5		*			*			*	
6			*		*			*	
7	*					*		*	
8		*				*		*	
9			*			*		*	
10			*		*		*		
11			*		*				*

As described earlier, the washwater ratio is an extremely important variable, particularly in mineral slurry applications. For pilot-plant trials, washwater requirements can easily be estimated by collecting and weighing a timed concentrate sample. From knowledge of the approximate concentrate percent solids, the concentrate water rate can be found, which then allows setting of washwater addition to a nominated washwater ratio. In the absence of measured solids contents, it can be assumed as a good approximation that the flow rate of water in the concentrate (L/min) is the same numerically as the flow rate of froth concentrate (kg/min). The procedure should be repeated after initially setting up washwater addition, as the addition of washwater itself affects concentrate recovery rate.

ACKNOWLEDGMENTS

A worldwide exclusive license has been granted to MIM Holdings, Ltd., 400 Ann Street, Brisbane 4000, Australia, for the use and sale of the Jameson cell for metallurgical purposes. We are grateful to MIM Holdings, Ltd for their assistance in the preparation of this article. We also acknowledge the assistance of J. A. Finch, M. H. Moys, G. Harbort, and M. Marchese. One of us (BWA) acknowledges the support of an Australian Post-graduate Research Award, in conjunction with MIM Holdings, Ltd. and The University of Newcastle Research Associates Limited (TUNRA).

REFERENCES

1. G. J. Jameson, A new concept in flotation machine design, *Min. Met. Process.*, Feb., 44–47. (1988).
2. D. H. Norris, Apparatus for separating the metallic particles of ores from the rocky constituents thereof, US Patent 873586, 10 December (1907).
3. A. F. Taggart, *Handbook of Mineral Processing*, Wiley, New York, pp. 12-61. (1927).
4. G. J. Jameson, "Flotation Cell Development," Proc. Ann. Conf., Aus. Inst. Min. Metall., pp. 25–31 (1992).
5. M. J. McCarthy and N. A. Molloy, Review of stability of liquid jets and the influence of nozzle design, *Chem. Eng. J.*, *7*: 1–20. (1974).
6. M. J. McCarthy, "Entrainment by plunging jets," Ph.D. Thesis, University of Newcastle, Australia, 1972.
7. G. M. Evans and G. J. Jameson, "Prediction of the Gas Film Entrainment Rate for a Plunging Liquid Jet Reactor," AIChE Spring National Meeting, Houston, paper 34G (1991).
8. J. O. Hinze, Fundamentals of the hydrodynamic mechanism of splitting in dispersion processes, *AIChE J.*, 1: 289–295. (1955).
9. G. K. Batchelor, *Proc. Cambridge Philos. Soc.*, *41*: 359. (1951).

10. G. M. Evans, G. J. Jameson, and B. W. Atkinson, Prediction of the bubble size generated by a plunging liquid jet bubble column, *Chem. Eng. Sci.*, *47*: 3265–3272. (1992).
11. G. M. Evans, "A study of a plunging jet bubble column," Ph.D. Thesis, University of Newcastle, Australia, 1990.
12. D. T. Dumitrescu, Flow past an air-bubble in a vertical pipe, *Z. Angew. Math. Mech.*, *23*: 139–149. (1943).
13. R. Pal and J. H. Masliyah, Flow characteristics of a flotation column, *Can. Metall. Q.*, *29*(2): 97–103. (1990).
14. D. A. Langberg and G. J. Jameson, The coexistence of the froth and liquid phases in a flotation column, *Chem. Eng. Sci.*, *47*: 4345–4355. (1992).
15. M. Xu, J. A. Finch, and A. Uribe-Salas, Maximum gas and bubble surface rates in flotation columns, *Int. J. Miner. Process.*, *32*: 233–250. (1991).
16. S. E. Sanchez-Pino and M. H. Moys, "Characterisation of Co-Current Downwards Flotation Columns," Column '91, Proceedings of an International Conference on Column Flotation (G. E. Agar, B. J. Huls, and D. B. Hyma, Eds.), Sudbury, Ontario, July 1991, Vol. 1, 341–355 (1991).
17. M. M. Marchese, A. Uribe-Salas, and J. A. Finch, Measurement of gas holdup in a three-phase concurrent downflow column, *Chem. Eng. Sci.*, *47*: 3475–3482. (1992).
18. M. M. Marchese, A. Uribe-Salas, and J. A. Finch, "Hydrodynamics of a Downflow Column," to be presented at XVIII International Mineral Processing Congress, Australasian Institute of Mining and Metallurgy, Sydney, Australia, 23–28 May (1993).
19. J. A. Finch and G. S. Dobby, *Column Flotation*, Pergamon Press, Oxford. (1990).
20. J. A. Finch and G. S. Dobby, Column flotation: a selected review. Part I, *Int. J. Miner. Process.*, *33*: 343–354. (1991).
21. B. W. Atkinson, P. T. Griffin, G. J. Jameson, and R. Espinosa-Gomez, "Jameson Cell Test Work on Copper Streams in the Copper Concentrator of Mount Isa Mines Limited," to be presented at XVIII International Mineral Processing Congress, Australasian Institute of Mining and Metallurgy, Sydney, Australia, 23–28 May (1993).
22. R. Chatiar, "Report on Test Work on the Operating Parameters of the Jameson Cell," Internal Report, Department of Chemical Engineering, University of Newcastle, February (1992).
23. J. C. Maxwell, *A Treatise on Electricity and Magnetism*, 3rd ed., Vol. I, Part II, pp. 435–449, Oxford University Press, London (1892).
24. R. Espinosa-Gomez, J. A. Finch, J. B. Yianatos, and G. S. Dobby, Flotation column carrying capacity: particle size and density effects, *Miner. Eng.*, *1*(1): 77–79. (1988).
25. R. Espinosa-Gomez and N. W. Johnson, "Technical experiences with conventional columns at Mount Isa Mines Limited," Column '91, Proceedings of an International Conference on Column Flotation (G. E. Agar, B. J. Huls, and D. B. Hyma, Eds.), Sudbury, Ontario, July 1991, pp. 511–524 (1991).
26. A. Kennedy, The Jameson flotation cell, *Min. Mag.* October (1990).

27. G. J. Jameson, G. Harbort, and N. Riches, "Development and Application of the Jameson Cell," Proc. 4th Mill Operators Conference, Burnie, Tasmania, 10–14 March 1991; Melbourne: The Australasian Institute of Mining and Metallurgy (1991).
28. G. J. Jameson and E. V. Manlapig, "Applications of the Jameson Cell," Column '91, Proceedings of an International Conference on Column Flotation (G. E. Agar, B. J. Huls, and D. B. Hyma, Eds.), Sudbury, Ontario, July 1991, pp. 673–687 (1991).
29. R. Clayton, G. J. Jameson, and E. V. Manlapig, The development and application of the Jameson cell, *Miner. Eng.*, *4*(7–11): 925–933. (1991).
30. G. J. Jameson, M. Goffinet, and D. Hughes, "Operating Experiences with Jameson Cells at Newlands Coal Pty Ltd, Queensland," Proc. 5th Australian Coal Preparation Conference, Australian Coal Preparation Society, Newcastle, NSW, Australia, 12–17 May (1991).

BUBBLY FLOWS AND THE PLUNGING JET FLOTATION COLUMN

Graeme J. Jameson
Department of Chemical Engineering
University of Newcastle
Newcastle, NSW
Australia

ABSTRACT

A new type of gas-liquid contacting device has been invented and is now in commercial use worldwide for the separation of minerals and fine coal particles. The device is a radical departure from previous technology, and its development has drawn heavily on classical theory of bubble mechanics, and the fluid mechanics of two-phase flows.

The phenomena which underlie the new flotation cell are described, together with research results regarding the behaviour of confined plunging jets.

INTRODUCTION

Bubble columns are used in the chemical and mineral processing industries, as a means of contacting gas and liquid phases. In the minerals industry, the separation of valuable particles from the host rock or gangue is carried out by flotation, in which the finely ground mineral is suspended in water, and the suspension is treated with reagents which render the particles to be floated non-wetting or hydrophobic, leaving the gangue in a wettable state. The suspension is then brought into contact with air bubbles which adhere to the non-wetting particles and carry them to the surface where they form a froth which can be removed. The gangue leaves the contact vessel as a tailings.

For the past eighty years, flotation has been carried out in stirred vessels, with air bubbles being generated by a rotating impeller. In recent years there has been a considerable growth in other forms of contacting devices, such as the bubble column. A sketch of a typical bubble column is shown in Figure 1. The liquid feed containing the particles enters at the top of the column, and makes contact with a stream of bubbles which are generated in the base of the column. Contact takes place in the "collection" zone. The froth which forms is usually quite deep, of the order of a metre or more, and as shown in the sketch, the froth is often washed with clean water, to remove gangue material which is entrained with the bubbles as they pass from the liquid phase into the froth. Clean concentrate is removed from the top of the column and the gangue leaves in the tails stream.

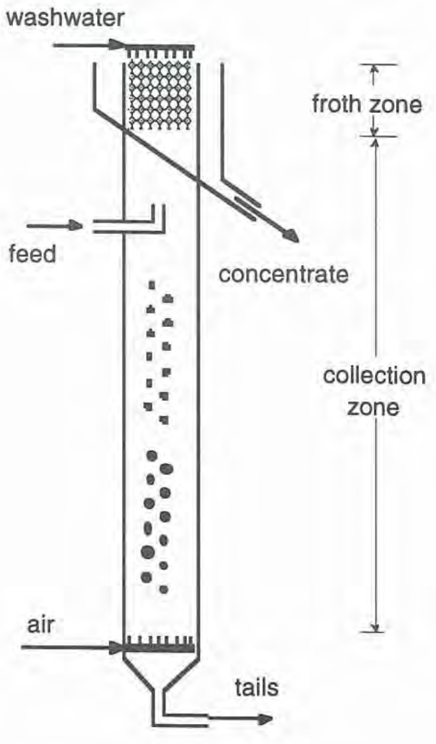


Figure 1. Diagrammatic sketch of a bubble flotation column. These columns are typically 10 to 15 metres in height and up to 4 metres in diameter.

Recently, another form of flotation column has been introduced, by the author, known as the Jameson Cell (Jameson, 1988). This cell works on a radically different principle to the bubble columns. It has now in use at over 100 sites around the world. In Australia, the value of coal recovered from tailing streams through its use is of order $75 million a year in exports, which would otherwise have gone to waste.

In this paper, some of the principles which led to the development of the Jameson Cell are discussed. Although a very simple device, it will be shown that its proper operation depends on a knowledge of classical theory for flow around bubbles, and of two-phase flows. Many of the phenomena on which it depends have been elucidated, but there are still some puzzling characteristics yet to be explained.

INITIAL CONSIDERATIONS

The Jameson Cell was the culmination of over 20 years of research into the physical mechanisms of flotation, and especially regarding the flotation of very fine particles. In the 1960's, it was recognised that the industry was moving towards finer and finer grinds, in order to liberate particles of values from the host rock, in so-called complex sulfide ore deposits. The flotation of fine particles, those below 20 μm in diameter, presented the conventional mechanical flotation cells with difficulties, in that the rate of flotation of such particles was known to be much slower than that of larger particles, so larger and larger cell volumes were required. Research by various authors, including Collins and Jameson (1976, 1977) showed that there was nothing mysterious about the behaviour of the fine particles, but that the diminished rate of flotation could be explained in hydrodynamic terms, including the effect of electrostatic forces arising from double-layer repulsion. Flotation was viewed simply as the result of a collision between a spherical bubble rising in a suspension of fine particles. These researches suggested that the flotation rate constant varied approximately as

$$k \approx d_p^2/d_b^3. \qquad (1)$$

where d_p is the particle diameter and d_b is the bubble diameter. Thus the rate of flotation could be improved by reducing the size of the bubbles, which were customarily made by the shearing action at the tip of a rotation impeller.

The invention of the Jameson Cell was rather serendipitous. The author had seen flotation columns being installed at Mount Isa in Queensland, and it seemed that there was a lot of "dead" space. The column volume needed for contacting was very large and residence times are typically of order 20 to 30 minutes. It was felt that if the contacting process could be made more efficient, the volume of the device could be considerably reduced, with consequent savings in construction and operating costs, and in convenience of operation. The collection zone, which occupied a total height of 12m approximately, seemed inordinately large, and it would be advantageous if it could be reduced in height or even done away with, so the author began playing with new ideas on paper and in the laboratory, aimed at an improved design.

Based on previous work, two guiding principles were followed:

(a) The bubbles in the system should be quite small, in the range 250 to 500 μm in diameter. This would give a high surface-to-volume ratio of the air dispersion, while retaining a sufficient buoyancy force so that the bubbles could rise in the disengagement zone, and not be carried out in the tailings.

(b) The void fraction should be as high as possible, so as to give a high surface-to-volume ratio for the contactor as a whole. It was realised that a countercurrent downwards design would best achieve this situation.

In order to control the bubble size, it was necessary to be able to predict the bubble diameter as a function of the flow environment, and the paper of Lewis and Davidson (1982) came to mind. They had investigated the splitting of a bubble in a shear flow, and had produced a simple criterion from which it would be possible to predict the maximum possible stable bubble size. In their analysis they assumed that a bubble of diameter d_b existed in a uniform shear field ω . Relative to the centre of the bubble, the velocity at the bubble extremities would be higher by an amount $\Delta U = \omega d_b/2$ approximately, leading to a pressure deficit of $(\rho_L \Delta U^2)/2$ by Bernoulli. The pressure reduction on diametrically opposite sides of the bubble would tend to split it apart, but would be opposed by the restraining force of surface tension, which is of order σ/d_b. Thus when $\rho_L \omega^2 d_b^2 >> \sigma/d_b$, the bubble will split, and for a given bubble diameter, there is a critical ratio of the splitting force to surface tension restoring force, which when exceeded, will cause the bubble to split. Lewis and Davidson termed this dimensionless ratio the critical Weber number

$$We_{crit} = \rho_L \omega^2 d_b^3 / \sigma \qquad (2)$$

and by a numerical analysis, found We_{crit} to be 4.7.

For various practical reason, the author decided to create a shear field by passing the liquid through an orifice. The shear rate is then of order

$$\omega \approx U_j/d_j \qquad (3)$$

U_j, d_j being the velocity and diameter of the liquid in the orifice, and since these two parameters can be varied in engineering terms, it should be possible to provide the shear rate appropriate to the formation of bubbles of the desired diameter.

A number of configurations were tried based on these considerations of bubble size, and the arrangement shown in Figure 2 was eventually arrived at. The design is of the utmost simplicity. The liquid issues as a jet , so the appropriate terms in equation (3) are thus the jet velocity and the jet diameter.

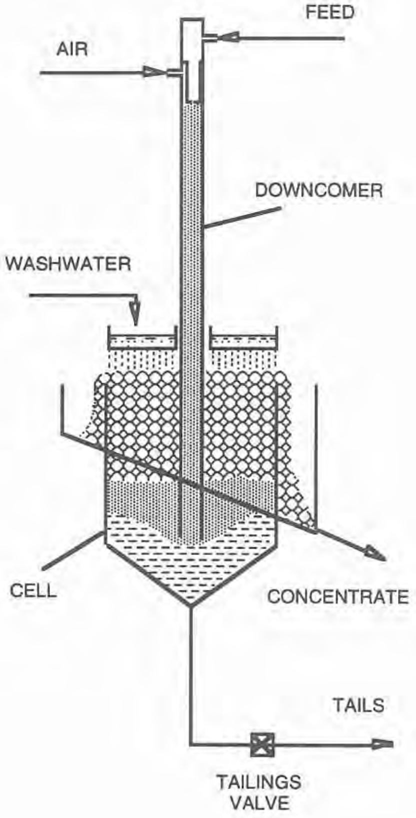


Figure 2. Diagrammatic sketch of a Jameson Cell. The downcomer is typically 3 m length and the cell itself is 1.5 m high. Largest cell diameter to date is 6 m. Downcomer diameter is up to 0.3 m.

In operation, the feed liquid is introduced in the form of a jet at the head of a "downcomer", which consists of a vertical tube which is sealed at the top except for a regulated air inlet and a vertical nozzle through which the liquid feed is introduced. The base of the downcomer is located below the pulp level inside the riser. On startup, the air inlet at the top of the downcomer is closed and the pulp feed is pumped through the nozzle. The air in the downcomer is entrained into the pulp which forms the seal in the bottom of the cell, and consequently, the pulp is drawn upwards from the cell into the downcomer. The pulp level reaches the tip of the nozzle quite quickly, and as a consequence of the hydrostatic suction developed by the head of this pulp, the pressure in the head of the downcomer is less than atmospheric. When the inlet is opened, air is drawn into the headspace at the top of the downcomer where it is entrained into the downcomer contents by the plunging jet.[1] The entrained air is broken up into fine bubbles which are quickly dispersed into the pulp and carried downward by the bulk fluid motion.

The three-phase mixture passes from the base of the downcomer into the cell proper, which has a much greater cross-sectional area than the downcomer. Consequently, the downward superficial velocity of the mixture is reduced, allowing the particle-laden bubbles to disengage from the liquid and rise to the surface and form a layer of froth. The froth then drains before overflowing into a collection launder, while the liquid phase and unrecovered particles leave through a valve at the base of the cell.

The fact that air can be drawn directly into the downcomer from the atmosphere is a strong practical advantage, because it obviates the need for a compressor. It should be emphasised that air induction is almost entirely due to hydrostatic effects, and dynamical effects such as are found in the conventional inductor or ejector are largely absent[2].

BUBBLE DYNAMICS IN THE DOWNCOMER

Once it had been demonstrated to work in the laboratory and in the field, it became necessary to provide design criteria and to carry out further studies of the behaviour of the plunging jet. Many of the design criteria were suggested by classical works in the field. For example, it is clear that the downwards velocity of the liquid should be such that it would exceed the rise velocity of the bubbles, including all bubbles which may result from coalescence. The rise velocity of air bubbles rising in water is shown in Figure 3, and it is very fortuitous that over a wide range of equivalent bubble diameters, the dependence is very weak. The reason for the flattening out of the rise velocity is that as the bubble increases in diameter, it changes shape. Small bubbles, less than about 0.3 mm in diameter, behave as solid spheres and their rise velocity can be calculated from Stokes law. As they approach 1 mm in diameter, bubbles start to flatten out like oblate spheroids, and Moore (1963) was able to relate the terminal velocity to the change in shape to the stresses acting over the surface of the flexible bubble. He found that in the vicinity of 2 mm in diameter, there was a maximum in the rise velocity.

Bubbles greater than about 4 mm equivalent diameter adopt a spherical-capped shape, and the rise velocity can be predicted by the equation of Davies and Taylor (1950):

$$U_t = 0.711(gD_e)^{0.5}, \tag{4}$$

where D_e is the equivalent bubble diameter. In the present case, any very large bubbles which form in the downcomer are broken and dispersed by the

---

[1] The operation described here is only possible if the rate of coalescence of bubbles in the downcomer is small. In mineral processing, bubble coalescence is inhibited by the presence of surface active materials, including frothers which are deliberately added to provide a stable foam, so that the foam layer containing the values can be removed.

[2] The novelty in this design owes more to Archimedes than to Daniel Bernoulli.

confined plunging jet, so a downwards velocity in excess of say 0.3 m/s is normally adequate.

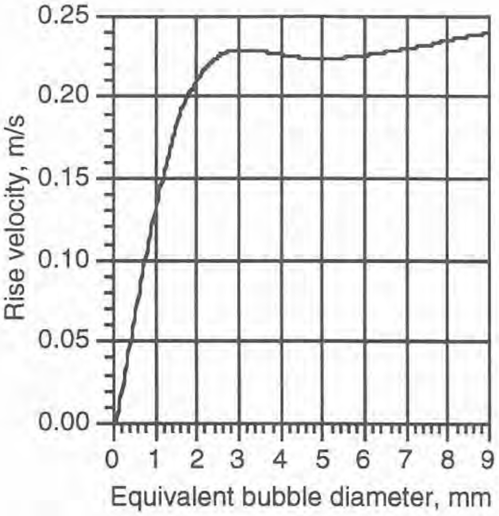


Figure 3. The terminal velocity of air bubbles rising in water contaminated with surface active agents, after Motarjemi and Jameson (1978).

Where the downcomer is very large, it is possible in principle for large air slugs to form, which could bridge the downcomer. Davies and Taylor (1950) showed that the rise velocity of such slugs was given by:

$$U_t = 0.46(ga)^{0.5}, \tag{5}$$

where a is the pipe radius. In practice, such slugs are not seen in the mixing zone in the downcomer, due to the disruptive action of the plunging jet. However, they do occur in the "pipe-flow" zone.

JET EXPANSION IN THE MIXING ZONE

The flow in the downcomer has been described by Evans (1990) and Evans et al. (1994). A number of flow regimes have been observed as depicted in Figure 4.

When the jet plunges into the confined bubbly mixture in the downcomer, air is entrained which is dispersed as fine bubbles in the shear layer adjacent to the jet. The jet itself loses momentum to the surrounding fluid, and expands with distance from the point of issue. At a certain point, the expanded jet touches the wall of the downcomer and from that point downwards, the gas-liquid mixture is essentially in a pipe-flow regime. Since the jet is confined, a recirculating eddy is established above the pipe-flow zone, and some bubbles escape from the eddy. The total air flow entrained into the jet is therefore the sum of air which is entering from the atmosphere and air which has escaped from the recirculating zone. The high velocity gradients between the submerged jet and the recirculating eddy result in high energy dissipation rates within the mixing zone, which are responsible for the breakup of the air once it is entrained by the plunging jet. The entrained air is broken into fine bubbles before being transported downward into the pipe-flow zone by the bulk fluid motion.

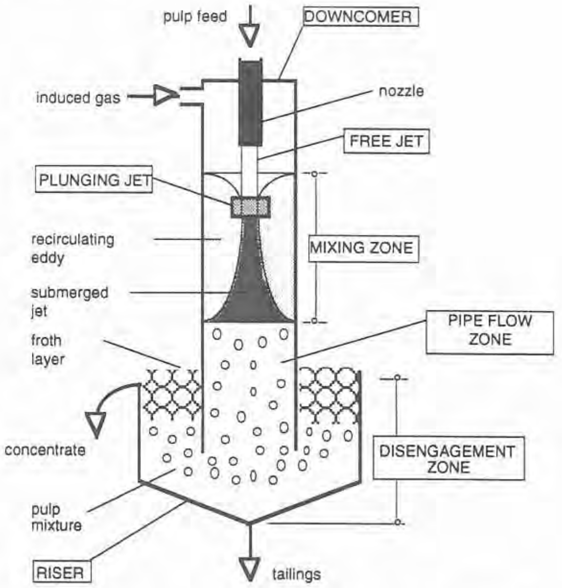


Figure 4. Flow regimes in the vicinity of a confined plunging jet.

The size of the bubbles produced inside the mixing zone – the volume confined between the surface of the froth and the point at which the jet touches the wall – is related to the forces acting on the bubble surface. In low viscosity fluids, like those normally encountered in flotation circuits, the bubbles are deformed by forces arising from liquid velocity fluctuations acting over distances of the order of the bubble diameter. The restoring force resisting the deformation of the bubble is due to surface tension acting at the air-liquid interface. The ratio of these forces can be written as a Weber number We, defined by

$$We = \rho_L \overline{u^2}\, d / \sigma \tag{6}$$

where $\overline{u^2}$ is the average value of the squares of the velocity differences acting over a distance of the order of the bubble diameter d ; and ρ_L and σ are the liquid (pulp) density and surface tension respectively[3]. If the bubbles are small compared to the turbulent macroscale but large compared to the microscale, the velocity fluctuations, following Hinze (1955), can be related to the energy dissipation rate per unit volume, $\dot{E}$, by the relationship:

$$\overline{u^2} = C_1 \left(\dot{E}\, d / \rho_L\right)^{2/3} \tag{7}$$

where $C_1 \approx 2$ according to Batchelor (1951).

---

[3] The derivation of equation (6) follows lines similar to those underlying equation (2), except that the pressure difference at the extremities of the bubble which lead to breakup is due to turbulent fluctuations rather than to a steady mean shearing flow.

For a given bubble size, and at low Weber numbers, the deformation forces are dominated by the restoring forces, and the bubble is stable. However, as *We* increases, the shape of the bubble becomes increasingly distorted. Eventually a critical Weber number, We_c, is reached where the bubble becomes unstable and breaks up into smaller bubbles. The maximum bubble size, d_m, corresponding to the critical Weber number is

$$d_m = (We_c/2)^{3/5}(\rho_L)^{-1/5}\dot{E}^{-2/5} \tag{8}$$

Equation (5) has been used by Evans *et al.* (1992) to predict the maximum stable bubble diameter inside the mixing zone, using estimates of the critical Weber number and the energy dissipation rate per unit volume.

The volume of the mixing zone is determined by the point of contact of the expanding jet and the wall of the enclosing downcomer. By considering the radial flow of momentum from the incoming jet to the recirculating eddies in the mixing zone, Evans (1990) derived the following expression for the half angle β subtended by the expanding jet at the virtual origin:

$$tan\ \beta = D_d/L_{MZ}$$

$$= \eta \left(\frac{P_1}{\rho_L U_j^2}\right)\left(\frac{\rho_{MZ}}{\rho_L}\right)\left\{\frac{U_j - U_{e,max}}{U_j}\right\} \text{x}$$

$$\left\{\frac{\left(0.37(R_c^2 - 0.5\ R_j^2)\right)^{0.5} - 0.64R_j}{R_j}\right\} \tag{9}$$

where η is a jet energy transfer efficiency, found by experiment to be 0.089; ρ_{MZ} is the density of the two-phase mixture in the mixing zone; R_c and R_j are respectively the downcomer and jet radius; P_1 is the pressure in the headspace at the top of the downcomer, U_j is the jet velocity and $U_{e,max}$ is the maximum return velocity in the eddy, which has been found by experiment to be $0.085U_j$. In operation, the jet half-angle is typically about 7°.

The volume of the mixing zone is then assumed to be the volume enclosed by the cylinder formed by the downcomer of internal diameter D_d and the length L_{MZ} of the mixing zone.

Evans *et al.* (1992) have compared the sizes of bubbles predicted by equation (5), with measurements of the sizes of bubbles formed in an air water system, and found agreement within ± 20%. Two downcomers were used, of diameters 44 mm and 74 mm, and the jet velocity varied from 7.8 to 11.53 m/s. The bubble sizes were generally in the range 200 to 400 μm.

TWO-PHASE FLOW IN THE PIPE ZONE

The void fraction in the downcomer, and especially in the pipe flow regime is regarded as important in providing an environment capable of providing the maximum probability of capture of particles by the bubbles. Measurements have shown that the void fraction can be as high as 0.5, which compares very favourably with values of 0.05 to 0.1 reported for mechanical flotation cells, and 0.1 to 0.15 for flotation columns. However, without further examination, it is not possible to say that the void fraction is entirely due to a dispersion of fine bubbles. For a given liquid flowrate, when the air to liquid ratio is small, less than 0.1, all the air introduced into the downcomer is dispersed into fine bubbles. As the ratio is increased however, it is possible that there will be a maximum in the air to liquid ratio at which it is possible for an assemblage of fine bubbles to be created which do not coalesce to any significant extent. It is quite possible for example, that as the air flow is increased, a point will be reached where the small bubbles are so tightly packed that there is no more room for them in the liquid, and any increase in the gas to liquid ratio will lead to the formation of a different flow regime, consisting of large bubbles or slugs moving together with a dispersion of fine bubbles. In the literature, two flow regimes have been identified which appear to coincide with the two cases discussed here: bubbly flow at low void fractions, and churn-turbulent flow at high void fractions.

Evans and Jameson (1994) have recently reported experiments in which the air void fraction, and the volumetric flowrates of the air and water were analysed using the drift flux model of Wallis (1962). The aim was to discover if there was any critical void fraction which would mark a transition from the bubbly flow to the churn-turbulent regime.

Wallis defines a drift velocity, v', for each phase which is the difference between the component linear velocity, v, and the total volumetric flux, J, for the two-phase mixture. For example, the gas drift velocity is equal to

$$v_g = j_g / \varepsilon, \tag{10}$$

where the total volumetric flux is the sum of the gas, j_g, and liquid, j_l, volumetric flux components, and is given by

$$J = j_g + j_\ell = (Q_g + Q_\ell)/A. \tag{11}$$

In eq. (11) Q_g and Q_ℓ are the gas and liquid volumetric flow rates respectively, and A is the cross-sectional flow area of the column.

Wallis also defines a drift flux, j', for the gas and liquid phases as being equal to the drift velocity multiplied by the fractional flow area occupied by the phase, i.e.

$$j'_g = \varepsilon v'_g \quad , \tag{12a}$$

$$j'_l = (1-\varepsilon)v'_l \quad , \tag{12b}$$

For a given bubble size, and at low Weber numbers, the deformation forces are dominated by the restoring forces, and the bubble is stable. However, as *We* increases, the shape of the bubble becomes increasingly distorted. Eventually a critical Weber number, We_c, is reached where the bubble becomes unstable and breaks up into smaller bubbles. The maximum bubble size, d_m, corresponding to the critical Weber number is

$$d_m = (We_c/2)^{3/5}(\rho_L)^{-1/5}\dot{E}^{-2/5} \tag{8}$$

Equation (8) has been used by Evans *et al.* (1992) to predict the maximum stable bubble diameter inside the mixing zone, using estimates of the critical Weber number and the energy dissipation rate per unit volume.

The volume of the mixing zone is determined by the point of contact of the expanding jet and the wall of the enclosing downcomer. By considering the radial flow of momentum from the incoming jet to the recirculating eddies in the mixing zone, Evans (1990) derived the following expression for the half angle β subtended by the expanding jet at the virtual origin:

$$\tan\beta = D_d/L_{MZ}$$

$$= \eta\left(\frac{P_1}{\rho_L U_j^2}\right)\left(\frac{\rho_{MZ}}{\rho_L}\right)\left\{\frac{U_j - U_{e,max}}{U_j}\right\} \times$$

$$\left\{\frac{\left(0.37(R_c^2 - 0.5\,R_j^2)\right)^{0.5} - 0.64R_j}{R_j}\right\} \tag{9}$$

where η is a jet energy transfer efficiency, found by experiment to be 0.089; ρ_{MZ} is the density of the two-phase mixture in the mixing zone; R_c and R_j are respectively the downcomer and jet radius; P_1 is the pressure in the headspace at the top of the downcomer, U_j is the jet velocity and $U_{e,max}$ is the maximum return velocity in the eddy, which has been found by experiment to be $0.085U_j$. In operation, the jet half-angle is typically about 7°.

The volume of the mixing zone is then assumed to be the volume enclosed by the cylinder formed by the downcomer of internal diameter D_d and the length L_{MZ} of the mixing zone.

Evans *et al.* (1992) have compared the sizes of bubbles predicted by equation (8), with measurements of the sizes of bubbles formed in an air water system, and found agreement within ± 20%. Two downcomers were used, of diameters 44 mm and 74 mm, and the jet velocity varied from 7.8 to 11.53 m/s. The bubble sizes were generally in the range 200 to 400 μm.

TWO-PHASE FLOW IN THE PIPE ZONE

The void fraction in the downcomer, and especially in the pipe flow regime is regarded as important in providing an environment capable of providing the maximum probability of capture of particles by the bubbles. Measurements have shown that the void fraction can be as high as 0.5, which compares very favourably with values of 0.05 to 0.1 reported for mechanical flotation cells, and 0.1 to 0.15 for flotation columns. However, without further examination, it is not possible to say that the void fraction is entirely due to a dispersion of fine bubbles. For a given liquid flowrate, when the air to liquid ratio is small, less than 0.1, all the air introduced into the downcomer is dispersed into fine bubbles. As the ratio is increased however, it is possible that there will be a maximum in the air to liquid ratio at which it is possible for an assemblage of fine bubbles to be created which do not coalesce to any significant extent. It is quite possible for example, that as the air flow is increased, a point will be reached where the small bubbles are so tightly packed that there is no more room for them in the liquid, and any increase in the gas to liquid ratio will lead to the formation of a different flow regime, consisting of large bubbles or slugs moving together with a dispersion of fine bubbles. In the literature, two flow regimes have been identified which appear to coincide with the two cases discussed here: bubbly flow at low void fractions, and churn-turbulent flow at high void fractions.

Evans and Jameson (1994) have recently reported experiments in which the air void fraction, and the volumetric flowrates of the air and water were analysed using the drift flux model of Wallis (1962). The aim was to discover if there was any critical void fraction which would mark a transition from the bubbly flow to the churn-turbulent regime.

Wallis defines a drift velocity, v', for each phase which is the difference between the component linear velocity, v, and the total volumetric flux, J, for the two-phase mixture. For example, the gas drift velocity is equal to

$$v_g = j_g / \varepsilon, \tag{10}$$

where the total volumetric flux is the sum of the gas, j_g, and liquid, j_l, volumetric flux components, and is given by

$$J = j_g + j_\ell = (Q_g + Q_\ell)/A. \tag{11}$$

In eq. (11) Q_g and Q_ℓ are the gas and liquid volumetric flow rates respectively, and A is the cross-sectional flow area of the column.

Wallis also defines a drift flux, j', for the gas and liquid phases as being equal to the drift velocity multiplied by the fractional flow area occupied by the phase, i.e.

$$j'_g = \varepsilon v'_g \quad , \tag{12a}$$

$$j'_l = (1-\varepsilon)v'_l \quad , \tag{12b}$$

where ε is the gas void fraction. By assuming that there is no net drift through the plane moving at the superficial velocity of the gas-liquid mixture then the system can be described by analysing the flow equations for either the gas or liquid phases.

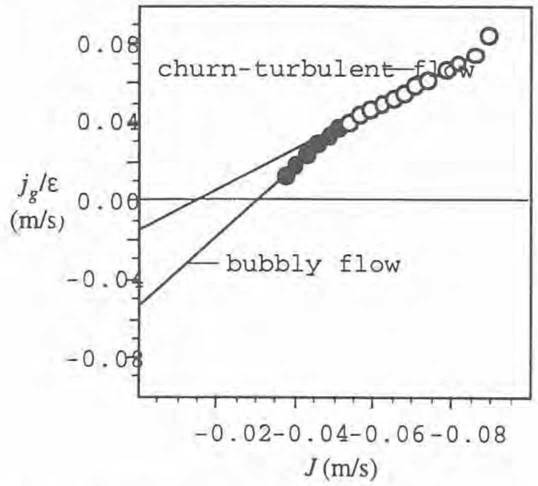


Figure 5. Drift flux plot for a particular experimental run.

Considering the gas phase only, equation (10) can be substituted into equation (12a), to give

$$j'_g = \varepsilon\left(v_g - J\right) \quad . \tag{13}$$

Expressing the total volumetric flux as the sum of the gas and liquid volumetric fluxes, equation (13) becomes

$$j'_g = \varepsilon\left(v_g - \left(j_g + j_l\right)\right) \quad , \tag{14}$$

and realising that

$$v_g = j_g/\varepsilon \quad , \tag{15}$$

the gas drift flux can be rewritten using only the volumetric flux terms and the gas void fraction, i.e.

$$j'_g = j_g(1-\varepsilon) - j_l\varepsilon \quad . \tag{16}$$

Equation (16) is a general expression based on the principle of continuity for the gas and liquid phases.

Wallis (1969) extended his drift-flux analysis and obtained an alternative expression for the gas drift flux, based on the terminal rise velocity of a single bubble in an infinite fluid, v_∞, and the gas void fraction, i.e.

$$j'_g = v_\infty g(\varepsilon)(1-\varepsilon)\varepsilon \quad , \tag{17}$$

where $g(\varepsilon)$ is a function of the gas void fraction. Ishii and Zuber (1979) developed expressions for $g(\varepsilon)$.

It will be appreciated that of the above variables, j_g, j_ℓ and ε are the only ones which can be measured directly, while the various drift velocities v_g, v_∞ etc are useful in providing insight into the phenomena. The value of the drift flux analysis can be seen from a number of representative plots of the data.

Figure 5 shows values of the drift velocity $v_g = j_g/\varepsilon$ at various values of the total superficial velocity J, following the analysis of Zuber and Findlay (1965). It is clear that in the experiment described, there is a transition in the slope of the data points at a particular value of J. The relevance of this point can be seen if the same data are plotted in the form $j'_g = j_g(1-\varepsilon) - j_l\varepsilon$ against the gas void fraction ε, as in Figure 6. Also shown on this plot is the volumetric drift flux curve predicted by the model of Ishii and Zuber (1979).

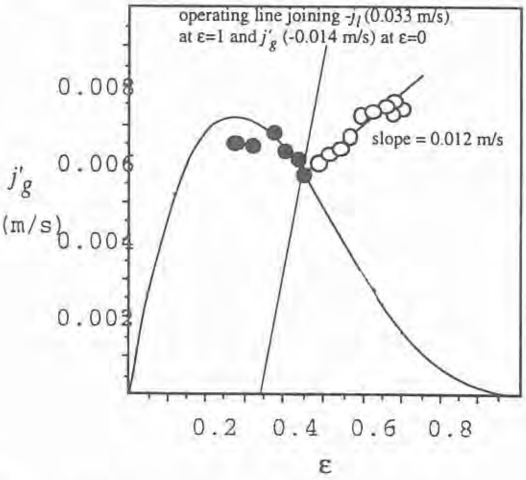


Figure 6. Values of the gas volumetric drift flux as a function of void fraction for a particular experiment. Conditions the same as in Figure 5.

It can be seen that for low void fractions, the experimental data follow the equation of Ishii and Zuber quite well, but at a value of the void fraction of 0.45 approx., there is a dramatic shift away from the model curve, indicating that there has been a change in the flow regime. The data suggest that above values of the void fraction of $\varepsilon = 0.45$, the bubbly flow is unable to accommodate further additions of air. Above this value, the bubbles are evidently so close-packed that any further air bubbles cannot be fitted in between the existing ones. Above $\varepsilon = 0.45$ for this particular flow, the two-phase mixture evidently behaves as if it were a dispersion of very large slugs or cavities moving in a dispersion of fine bubbles of void fraction 0.45.

STABILITY OF THE JET ENTRAINMENT SYSTEM

An effect which was not anticipated is the entrainment behaviour of the liquid jet as the air rate is increased. Referring to Figure 7, at low gas rates the downcomer is filled with a dense foam, but as the gas rate increases, it is seen that the height of the foam interface begins to drop, with corresponding increase in the headspace volume and the length of the exposed free jet.

It appears that at the higher gas rate, the system is able to come to a new equilibrium, in that for some reason, the rate of entrainment of gas into the dense foam is higher when the jet has been allowed to fall some distance from the point of formation, than when the foam interface is immediately adjacent to the jet exit point.

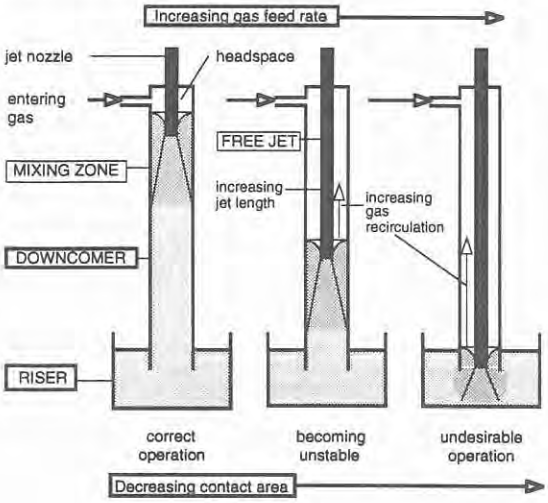


Figure 7. The effect of increased gas flow rate at constant liquid feed rate. At low rates the downcomer is filled with a dense two-phase foam. If the air flow is increased, the level of the top of the downcomer contents drops to a new equilibrium height at which the jet appears to be able to entrain the incremental air. When the air rate exceeds the maximum entrainment capacity of the jet, the foam bed collapses.

The explanation for this observation is not clear. The downcomer is typically 3 metres in length, and it can be shown that the increase in velocity due to free fall under gravity over the initial part of the downcomer is not very significant. It is possible that it is the relaxation of the velocity field inside the jet after departure from the nozzle which is responsible. Relative to the stationary downcomer, the surface velocity of the jet is initially zero, but momentum will transfer to the free jet from the interior, thereby tending to increase the surface velocity which must of course eventually become the same as the mean jet velocity. This phenomenon is still to be elucidated. If the air rate is increased further, a point is reached when the foam bed in the downcomer collapses totally, as shown in Figure 7.

It should be pointed out that the headspace in the downcomer is normally maintained at a pressure less than the ambient, so that air can be drawn into the downcomer from the atmosphere, obviating the need for a compressor or blower. The reason that the pressure is below atmospheric can be found by simple hydrostatic considerations, the pressure deficit due to changes in the momentum flux being of second order.

CONCLUSIONS

The Jameson Cell is an example of a practical fluid dynamical device which grew out of academic research relating to the behaviour of bubbles and bubbly flows. Cells based on this design are now in use at over 100 sites around the world. Many of the design considerations are based on classical results in bubble mechanics. The mechanisms of entrainment of air into the plunging jet are still not fully understood, and the unexpected stability of the system in response to changes in the gas flowrate is still be explained.

REFERENCES

Batchelor, G.K. 1951 *Proc. Cambridge Phil. Soc.*, *41*: 359.

Collins, G.L. and Jameson, G.J. 1976, "Experiments on the flotation of fine particles - the effect of particle size and charge" *Chem.Eng.Sci.*, 31, 985.

Collins, G.L. and Jameson, G.J., 1977, "Double-layer effects in the flotation of fine particles" *Chem.Eng.Sci.*, 32, 239.

Davies, R.M., and Taylor, G.I., 1950, "The mechanics of large bubbles rising through extended liquids and through liquids in tubes", *Proc. Roy. Soc.* A200: 375–390.

Evans, G.M. (1990), "A study of a plunging jet column", Ph.D. thesis, University of Newcastle, Australia.

Evans, G.M., Jameson G.J. and Atkinson, B.W., 1992. Prediction of the bubble size generated by a plunging liquid jet bubble column, *Chem. Eng. Sci.*, *47*: 3265-3272.

Evans, G.M. and Jameson, G.J., 1995, "Hydrodynamics of a plunging jet bubble column", (with G.M. Evans), *2nd Int. Conference on Gas-Liquid-Solid Reactor Engineering*, Cambridge, U.K., 27-29 March.

Hinze, J.O. 1955, "Fundamentals of the hydrodynamic mechanism of splitting in dispersion processes. *A.I.Ch.E. Jl. 1*: 289-295 (1955).

Ishii, M. and Zuber, N., 1979, "Drag coefficient and relative velocity in bubbly, droplet or particulate flows", *A.I.Ch.E. J.*, 25: 843-855.

Jameson, G.J., 1988, "A new concept in flotation machine design, *Min. and Met. Processing*, Feb., 44-47.

Lewis, D. A. and Davidson, J. F., 1982, "Bubble splitting in shear flow", *Trans IChemE.*, 60: 283–291.

Moore, D.W., 1963, "The boundary layer on a spherical gas bubble", *J. Fluid Mech.*, 16: 161-176.

Motarjemi, M. and Jameson, G.J., 1978, "Mass transfer from very small bubbles–the optimum bubble size for aeration", *Chem. Eng. Sci.* 33: 1415–1423.

Wallis, G.B., 1969, *One-dimensional two-phase flow*, McGraw-Hill, New York.

Use of Jameson cell flotation technology at Cleveland Potash, Ltd., North Yorkshire, England

M. J. Burns, G. Coates and L. Barnard

Cleveland Potash, Ltd. (CPL), is a mining company that extracts and refines sylvinite ore, primarily KCl and NaCl, from an evaporite deposit in North Yorkshire, England. The processing plant uses froth flotation to produce a KCl-rich product, which can be either a very fine concentrate (−100 μm) or a coarser concentrate (100–1100 μm). The Jameson cell is being tested in both circuits.

Although KCl is readily rendered hydrophobic by the amine collector, difficulty in recovering high-grade, coarse material is still experienced. Enhanced recovery of this slow-floating, +0.85-mm fraction would improve overall plant efficiency by assisting the debrining and drying of concentrates and by reducing pressure on secondary milling. Fig. 1 summarizes the primary aspects of the circuit.

Pilot trials

The pilot Jameson cell had an internal diameter of 520 mm and was approximately 5 m high. The cell was fed by a Warman slurry pump that generated a flow of 10 $m^3 h^{-1}$ at 2.5 bar, as measured at the feed inlet. The residence time in the cell was calculated to be 2 min. A schematic section of the cell is presented as Fig. 2.

Flotation of four different streams was undertaken with the Jameson cell; details of the streams are given in Table 1.

Flotation of recleaner concentrate

In tests on the flotation of recleaner concentrate methyl isobutyl carbinol (MIBC) frother was added directly into the pump suction line at the required addition of 5 ppm, which is critical in establishing and maintaining flotation. The slurry–froth interface was adjusted to a depth of 30 cm below the cell lip with the use of a bubble probe, relay and air-operated tails sleeve valve. At this stage no attempt was made to wash the froth with saturated brine.

The air flow to the cell was set at 50 l min^{-1}, equivalent to a superficial rise velocity factor of 0.41 cm s^{-1} and an air to pulp ratio of 1:3.33. Rise velocity is simply air flow divided by the cross-sectional area of the cell and is critical in determining scale-up parameters.

Flotation of cleaner concentrate

Cell conditions for the flotation of cleaner concentrate remained the same as those described above. Grade and recovery were measured over a wide range of feed grades, as was necessary to analyse the potential of the Jameson cell as a feasible production technology. The slurry density was maintained at 15–20 wt% solids to assist pumping and thereby stabilize feed inlet pressure and air flow into the cell at 2.5 bar and 50 l min^{-1}, respectively. Table 2 gives details of the results. The data in Table 3 illustrate how both the

Manuscript first received by the Institution of Mining and Metallurgy on 1 September, 1993; revised manuscript received on 21 December, 1993. Technical note published in *Trans. Instn Min. Metall. (Sect. C: Mineral Process. Extr. Metall.)*, 103, May August 1994.

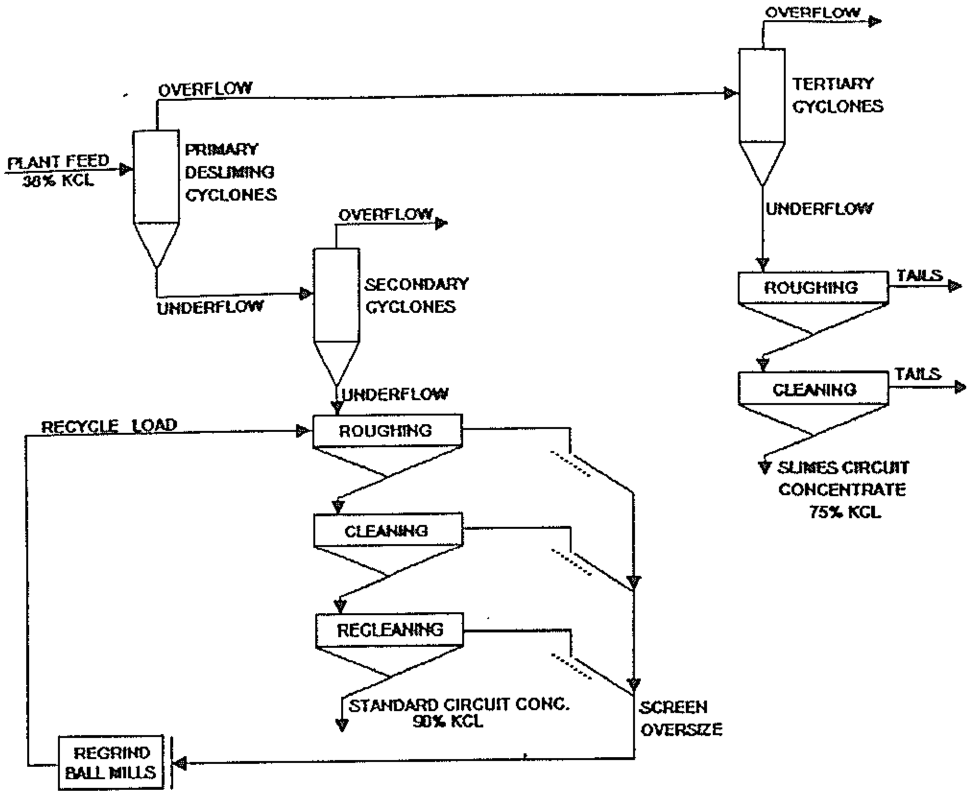


Fig. 1 Layout of standard and slimes flotation circuit at Cleveland Potash, Ltd.

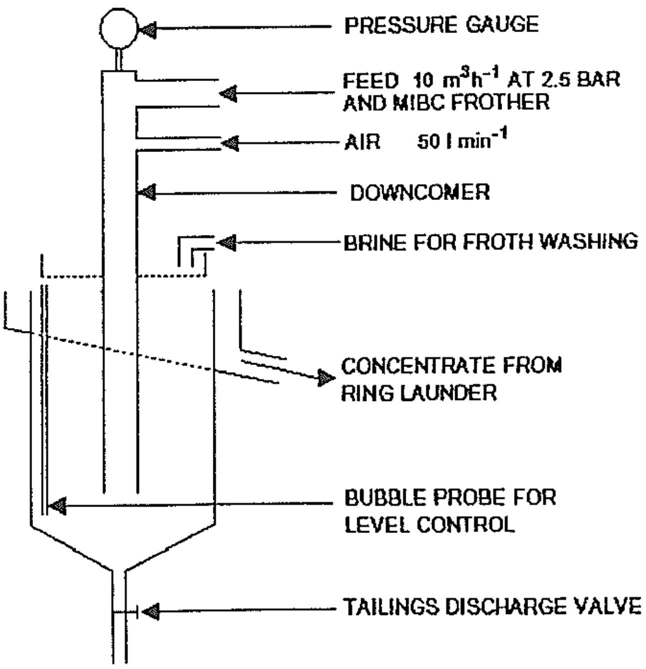


Fig. 2 Schematic diagram of Jameson cell

Table 1 Characteristics of streams tested

Size, mm	Concentrate Rougher	Cleaner	Recleaner	Slimes float feed
Amount retained, wt%				
+0.85	12.4	6.2	2.6	0.0
−0.85+0.60	14.3	15.4	14.6	0.3
−0.60+0.30	39.3	45.7	51.6	3.0
−0.30+0.10	24.6	24.8	26.7	34.6
(−0.10)	9.4	7.9	4.5	62.1
Solids, wt%	32	33	38	18
Sp. gr.	1.41	1.42	1.45	1.33
KCl, wt%	70–80	80–87	87–90	25–35

Jameson cell and present cleaners behave with extremely poor feed grades. Size analyses of the concentrates produced from the cells are documented in Table 4.

Flotation of rougher concentrate

In tests on the flotation of rougher concentrate the state of the cell was altered, the depth of the froth bed being increased to 40 cm. Air flow and MIBC additions remained the same. The sampling campaigns were constructed to enable comparison of recoveries obtained with the Jameson cell with those produced by the entire plant cleaner and recleaner circuit (Tables 5 and 6). At this stage froth washing was undertaken (runs 23, 24 and 27) in an attempt to remove entrained gangue.

In runs 11 to 21 tests were conducted at froth bed depths of 30 cm and no washing was carried out. The remaining operational parameters were unchanged.

During the flotation of rougher concentrate the recovery path of the coarse, +0.85-mm material was traced (Figs. 3 and 4).

Flotation of slimes float feed

In flotation of slimes float feed tests were carried out with the froth bed at 40 cm and no alteration of other variables. No dilution of the feed, however, was required as it was evident that feed inlet pressures and air influx were adequate. The results are given in Table 7.

Table 2 Performance of cells treating cleaner KCl concentrate

Run	Date	Jameson cell Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %	Plant recleaner cells Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %
1	29/04	75.4	34.7	91.2	87.0	75.4	44.3	89.0	82.1
2	04/05	80.0	22.5	88.1	96.5	80.0	48.1	91.5	84.1
3	05/05	65.5	20.1	89.2	89.5	65.5	34.0	88.1	78.3
4	06/05	75.8	44.3	90.6	81.3	75.8	48.9	89.5	78.2
5	06/05	70.0	32.6	88.6	84.5	70.0	49.2	89.2	66.3
6	06/05	73.7	38.8	88.5	84.3	73.7	48.5	90.6	73.6
7	06/05	68.3	31.9	88.0	83.6	68.3	37.5	90.3	77.1
8	13/05	72.0	18.8	87.4	94.1	72.0	27.0	88.0	90.2
Mean		72.6	30.5	89.0	87.6	72.6	42.2	89.5	78.7

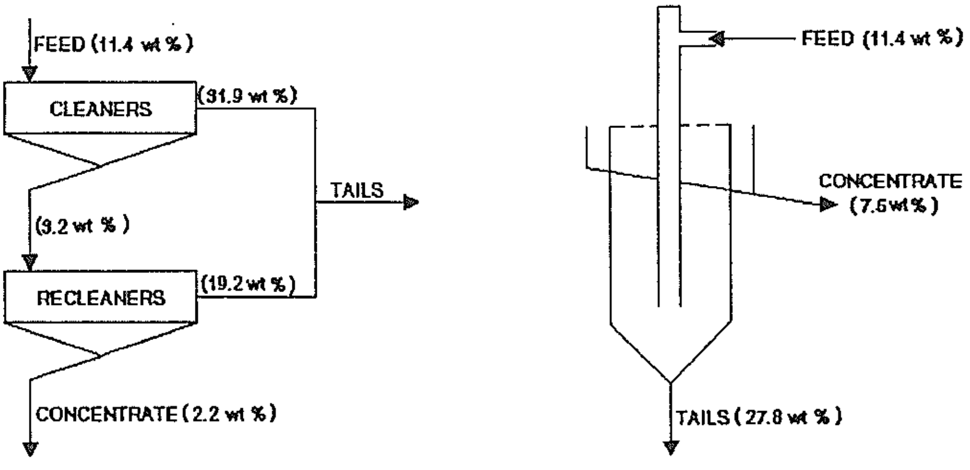
Table 3 Performance of cells treating low-grade cleaner concentrate

Run	Date	Jameson cell Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %	Plant recleaner cells Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %
9	06/05	41.3	16.5	89.7	73.6	41.3	28.1	88.6	46.8
10	06/05	43.2	20.5	88.5	68.4	43.2	26.9	88.5	54.2
Mean		42.3	18.5	89.1	71.0	42.3	27.5	88.6	50.5

Table 4 Size distribution of concentrates from Jameson cell and plant recleaner cells

Sieve size, mm	Retained, wt% 1	2	KCl, wt% 1	2	NaCl, wt% 1	2	Insolubles, wt% 1	2
+0.85	30.4	5.7	90.8	92.7	8.0	6.8	1.2	0.5
−0.85+0.60	31.3	21.3	85.1	89.0	10.5	8.0	4.4	3.0
−0.60+0.30	27.4	41.9	86.3	91.6	12.3	8.0	1.4	0.4
−0.30+0.10	9.7	26.3	91.1	92.1	8.0	7.4	0.9	0.5
(−0.10)	1.2	4.8	71.2	79.7	3.4	6.4	25.4	13.9

1, Jameson cell concentrate.
2, Plant recleaner concentrate.



(*a*) (*b*)

Fig. 3 Deportment of slow-floating +850-μm high-grade KCl particles to final concentrate in test run 24: (*a*) plant circuit; (*b*) Jameson cell. Data in figure refer to material retained. Total recoveries of +850-μm material to concentrate with plant circuit and Jameson cell were 14.75 and 53 wt%, respectively

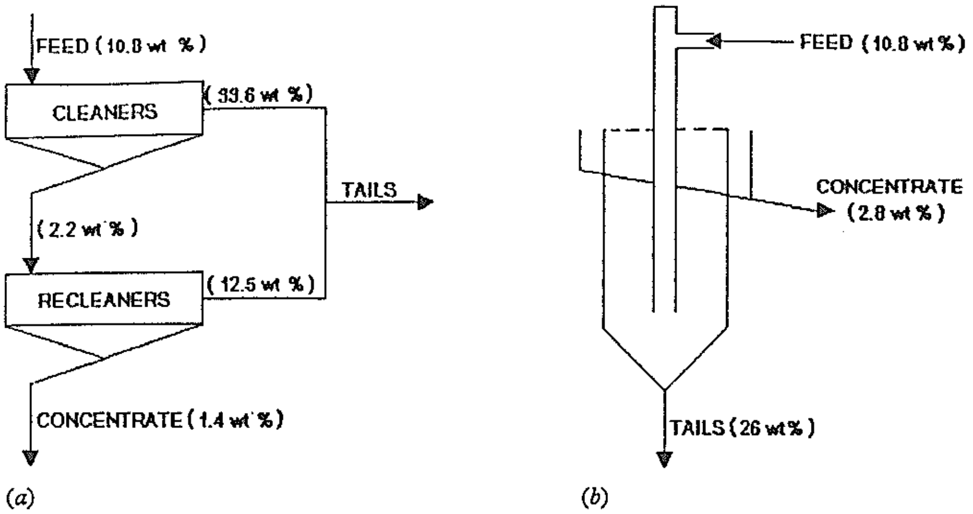


Fig. 4 Deportment of slow-floating +850-μm high-grade KCl particles to final concentrate in test run 25: (*a*) plant circuit; (*b*) Jameson cell. Data in figure refer to material retained. Total recoveries of +850-μm material to concentrate with plant circuit and Jameson cell were 8.74 and 17 wt%, respectively

Discussion and conclusions

Immediately after commissioning very little manipulation of metallurgical parameters—such as air flow, frother addition, depth of froth bed, levels of slurry in the downcomer and froth washing—was required to establish optimum performance from the cell. Neither was there any onerous requirement for testwork involving the analysis of small, incremental changes in cell variables. Most of the testwork was centred on the development of accurate sampling methods and continuous 24-h cell operation with minimum supervision.

Table 1 gives the characteristics of the streams that were tested; the solids content of the slurry proved to be significant throughout the pilot trials. For all streams except the slimes flotation feed it was necessary to dilute the feed to the Jameson cell with saturated brine at a ratio of 1:1. The total volumetric recovery from the CPL rougher flotation section is 500 $m^3 h^{-1}$. A full-scale production unit would, therefore, be required to accommodate an influx of 1000 $m^3 h^{-1}$. Preliminary scale-up calculations indicate that two units with diameters of 2.75 m would be needed to provide an effective replacement for 28 Denver no. 30DR 2.8-m^3 cleaner and recleaner cells.

The solids content of the rougher concentrate is, typically, 32 wt%, which does not produce an unduly viscous slurry. The concentrate, however, contains an appreciable amount of entrained air that does not easily escape from the pump box feeding the cell, resulting in low and surging feed inlet pressures of less than 2.5 bar. This poor operating condition gives rise to low degrees of aspiration, with subsequent loss of flotation performance. The prime role of the saturated brine diluent was, therefore, the removal of entrained air prior to pumping, its effect in reducing the apparent influence of viscosity not being overly significant. It remains to be seen if a method other than that of further dilution with brine can be found to reduce pressure loss at the feed inlet, thereby reducing the scale and cost of a larger industrial unit.

Flotation of recleaner concentrate

At present there is little scope for a fourth flotation step at CPL. Use of the Jameson unit did, however, result in an upgrade of the recleaner concentrate from 88.5 wt% KCl to 91.2 wt% KCl at a recovery of 94.7% KCl. This brief run yielded a sound base level and reference point for ensuing testwork.

Flotation of cleaner concentrate

Table 2 provides a comparison of the performances of the

Table 5 Performance of cells treating rougher concentrate

Run	Date	Jameson cell (with 30 cm froth-bed depth)				Plant recleaner cells			
		Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %	Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %
11	18/05	67.1	24.0	88.5	88.1	67.1	33.6	79.4	86.6
12	19/05	63.9	32.8	88.5	77.3	63.9	27.7	85.8	83.7
13	19/05	69.9	29.4	90.0	86.0	69.9	32.8	87.9	84.7
14	19/05	76.7	37.4	90.0	87.7	76.7	33.9	87.4	91.2
15	19/05	74.6	36.6	87.4	87.6	74.6	40.4	86.9	85.7
16	20/05	65.5	31.1	87.4	81.5	65.5	34.5	86.3	78.8
17	20/05	61.7	30.4	88.5	77.3	61.7	28.9	83.9	81.1
18	20/05	68.3	29.6	88.5	85.1	68.3	33.8	84.4	84.2
19	20/05	71.2	36.4	89.0	82.7	71.2	40.9	85.8	81.3
20	20/05	66.4	33.8	90.1	78.6	66.4	36.4	86.9	77.7
21	20/05	65.6	40.5	91.2	68.8	65.6	33.8	86.3	79.7
Mean		68.3	32.9	89.0	81.9	68.3	34.2	85.5	83.2

Table 6 Performance of cells treating rougher concentrate

Run	Date	Jameson cell (with 40 cm froth-bed depth) Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %	Plant recleaner and cleaner cells Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %
22	11/06	78.8	47.0	90.0	84.5	78.8	54.5	90.2	78.0
23*	15/06	80.0	52.9	90.2	82.2	80.0	63.0	92.7	67.0
24*	15/06	79.9	48.1	91.7	83.7	79.9	53.0	91.8	79.6
25	16/06	74.7	47.3	89.6	77.7	74.7	50.7	89.0	74.6
26	16/06	79.4	46.5	92.8	83.0	79.4	47.2	93.9	74.0
27*	24/06	77.5	52.6	91.1	76.0	77.5	53.5	91.1	75.0
Mean		78.4	49.1	90.9	81.2	78.4	53.6	91.4	74.7

*Froth bed was washed with saturated brine to minimize entrainment.

Table 7 Performance of cells treating slimes float feed

Run	Date	Jameson cell Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %	Plant recleaner cells Feed, wt% KCl	Tails, wt% KCl	Concentrate, wt% KCl	Recovery, %
28	07/07	27.8	4.9	77.4	87.9	27.8	8.5	78.4	77.9
29	08/07	18.4	6.0	75.9	73.2	18.4	6.5	77.9	70.6
30	09/07	23.3	4.6	71.8	85.8	23.3	6.6	82.3	77.9
31	12/07	21.8	4.9	70.0	83.4	21.8	4.3	78.5	84.9
Mean		22.8	5.1	73.8	82.6	22.8	6.5	79.3	77.8

Jameson cell and CPL recleaner cell with identical feeds. It is clear that the former achieved a superior grade and recovery of KCl at 89 wt% and 87.6%, respectively, the mean values of the plant recleaners being 89.5 wt% KCl at 78.7% recovery. The additional recovery is undoubtedly attributable to the reporting of high-grade, coarse particles to the concentrate on a scale that is not observed in plant recleaner operation (Table 4).

During the sampling campaign for which details are given in Table 4 the Jameson cell recovered 30.4 wt% +0.85-mm material to the concentrate compared to 5.7 wt% of the same fraction achieved by the plant recleaner cells. The benefits of such performance can be seen in terms of improved debrining, centrifuging and drying of concentrate, improved overall plant recovery and less recycle to secondary milling.

The results in Table 3 highlight the performance of the Jameson cell with very poor feed grades. It is apparent that it can operate at high efficiencies under extreme plant conditions. Indeed, recoveries are far higher with the Jameson cell, which would offer hope of a large reduction in recycle back to rougher flotation during times when the plant is required to process low head grades.

Flotation of rougher concentrate

Two distinct sampling campaigns were undertaken during the flotation of rougher concentrate, results for which are given in Tables 5 and 6. Table 5 compares the performance of the Jameson cell with that of the plant cleaner cell—again, with identical feed grades. The mean values of all campaigns show that the Jameson cell achieved a KCl grade of 89 wt% with 81.9% recovery. The plant cleaners produced a grade of 85.5 wt% KCl at 83.2% recovery. The poorer Jameson cell recovery simply reflects a high froth–slurry interface. A major advantage of the Jameson cell is the ease with which the slurry–froth interface can be altered in response to changing feed grades, which are seen to fluctuate quite markedly.

The results given in Table 6 are for more detailed sampling that was undertaken to allow direct comparison of the performance of the Jameson cell with that of the entire plant cleaner and recleaner circuit, as shown in Figs. 3 and 4. The mean values of the sampling campaigns given in Table 6 clearly show the difference in performance between the Jameson cell and the plant recleaners and cleaners. The Jameson cell produced acceptable final concentrate grades, similar to those achieved by the plant recleaners, with a recovery of 81.2% KCl—as opposed to a recovery of 74.7% for the whole of the circuit illustrated in Figs. 3 and 4. The extra recovery results from the fact that coarse material reports to concentrate and not to tails.

On three occasions the froth bed was washed with saturated brine (run numbers 23, 24 and 27), but no significant benefits were readily apparent. This is entirely to be expected owing to the nature of the froths at CPL, which are not prone to entrainment.

Flotation of slimes flotation feed

Two different grade/recovery relationships are evident from the data in Table 7, which gives results that were obtained in the treatment of slimes flotation feed. The recovery by the Jameson cell of 82.6% would result in the production of an additional 1 t h^{-1} potash, representing additional revenue of approximately £518 000 per year.

The argument that similar recoveries could be achieved with present plant cells by pulling the concentrate forward at a quickened pace is flawed owing to the practicalities of dealing with higher volumetric flows of pulp at low densities. The final slimes concentrate would be certain to suffer.

CPL has decided to purchase a production unit to treat slimes flotation feed. The unit will also be used to treat rougher flotation concentrate to assist in the decision to purchase further Jameson cells to replace the standard cleaner and recleaner circuit entirely. The cell would have a diameter of 3.25 m and would treat 325 $m^3 h^{-1}$ slurry. Effectively, it would replace two banks of roughers and one bank of cleaners, a total of 16 2.8-m^3 cells.

Envisaged savings

The power requirement of the motors of the present cells is an obvious area for savings; electricity costs would be reduced by at least £21 000 per year. The cost of maintaining 28 Denver cells in parts alone is close to £20 000 per year (not including the cost of manpower requirements). The fuel requirements for drying the final product would be reduced, possibly realizing savings of the order of £100 000. These estimates are based on the replacement of 28 2.8-m^3 cleaner and recleaner cells by two Jameson units. The payback period would be approximately one year. It should be noted that if the cells were replaced by traditional agitated units, there would be no savings and no payback period as there would be no metallurgical benefits.

It is readily apparent that Jameson cell flotation offers lower capital and running costs in addition to improved metallurgical performance.

Authors

M. J. Burns *Member* is a research and development engineer at Cleveland Potash. He previously worked in South Africa as a plant metallurgist for Johannesburg Consolidated Industries, gaining experience in precious- and base-metal processing. He is a graduate of the University of Leeds, England, holding degrees of B.Eng. in mineral processing and M.Sc. in engineering ceramics.

Address: Cleveland Potash, Ltd., Boulby Mine, Saltburn, Cleveland TS13 4UZ, England.

G. Coates has been a metallurgical technician at Cleveland Potash for six years. He holds H.N.D. qualifications in civil engineering and in business studies and finance.

L. Barnard worked as a research assistant at the National Physical Laboratory, Teddington, Middlesex, before joining Cleveland Potash recently as a metallurgical technician. She is a graduate of Imperial College, London, where she gained a B.Eng in material science and engineering.

FUNDAMENTALS OF JAMESON CELL OPERATION INCLUDING SIZE–YIELD RESPONSE

B W Atkinson, C J Conway and G J Jameson

Department of Chemical Engineering, University of Newcastle

ABSTRACT

The mechanisms of air-slurry contact in a Jameson cell are described to explain how the Jameson cell is capable of high recovery in a single pass. In particular, the fundamental differences in air–particle contact probability between Jameson cell, conventional column and mechanical cell flotation are highlighted.

Specific operational characteristics of the Jameson cell are reviewed including air rate and bubble size considerations, froth depth and wash water effects.

Initial applications of the Jameson cell in coal flotation have concentrated on recovery of fine coal (minus 0.1 mm) where conventional flotation technology is inefficient. Subsequent test work at a number of coal mine operations has shown the Jameson cell to be highly efficient at floating coarse coal up to and beyond 0.5 mm, and has resulted in installation of Jameson cells at major coal preparation plants to treat full fines feed. Results for various coals are reviewed, particularly in regard to recovery by size performance.

Wash water addition rate can have a large influence on yield, and at times is advantageous for concentrate ash reduction depending on the nature of non-coal material present. The concept of *wash water ratio*, as opposed to wash water bias, is explained.

Concentrate carrying capacity, which is a parameter widely used in the mineral column flotation industry, is a measure of the limiting concentrate removal rate based on bubble surface loading and froth transport phenomena and is not dictated by flotation kinetics. The relevance of the carrying capacity concept to coal flotation applications is discussed.

INTRODUCTION

The principles of operation of the Jameson cell have been described in detail in a recent review by Evans, Atkinson and Jameson (1993). Figure 1 shows a schematic representation of a single-downcomer cell. The downcomer acts as the air entrainment device and air-particle contactor, while the cell serves as a froth–pulp disengagement zone in the same manner as the upper zone of a conventional flotation column. In a Jameson cell, the air is self-induced and an external compressor or blower is not required.

Jameson cells are currently employed for fine coal treatment at four sites in Australia.

JAMESON CELL FUNDAMENTALS

Figure 2 shows a summary of the various active zones in the Jameson cell. The action of the free jet impinging on the surface of the bubbly mixture in the downcomer serves to entrain air into the downcomer by jet surface entrapment and film entrainment. The vigorous mixing and high shear conditions experienced in the mixing zone result in break-up of the air into very small bubbles of less than 1 mm diameter. The bubble buoyancy acts countercurrent to the overall fluid flow and results in a very high void fraction (gas hold-up) of 50 to 60% inside the downcomer. The high void fraction, in combination with the small bubble size, provides for very thin interfacial slurry film thickness between the bubbles, providing an ideal environment for particle-bubble contact.

Froth flotation efficiency is determined by a series of probabilities, namely the probability of particle-bubble contact, the probability of particle-bubble attachment, the probability of transport between pulp and froth, and the probability of froth collection into the product launder. Due to the extremely high interfacial area between pulp and air, the probability of particle-bubble contact is extremely high in a Jameson downcomer. The subsequent probabilities of attachment and detachment for the Jameson cell are similar to those for the conventional column and mechanical cells.

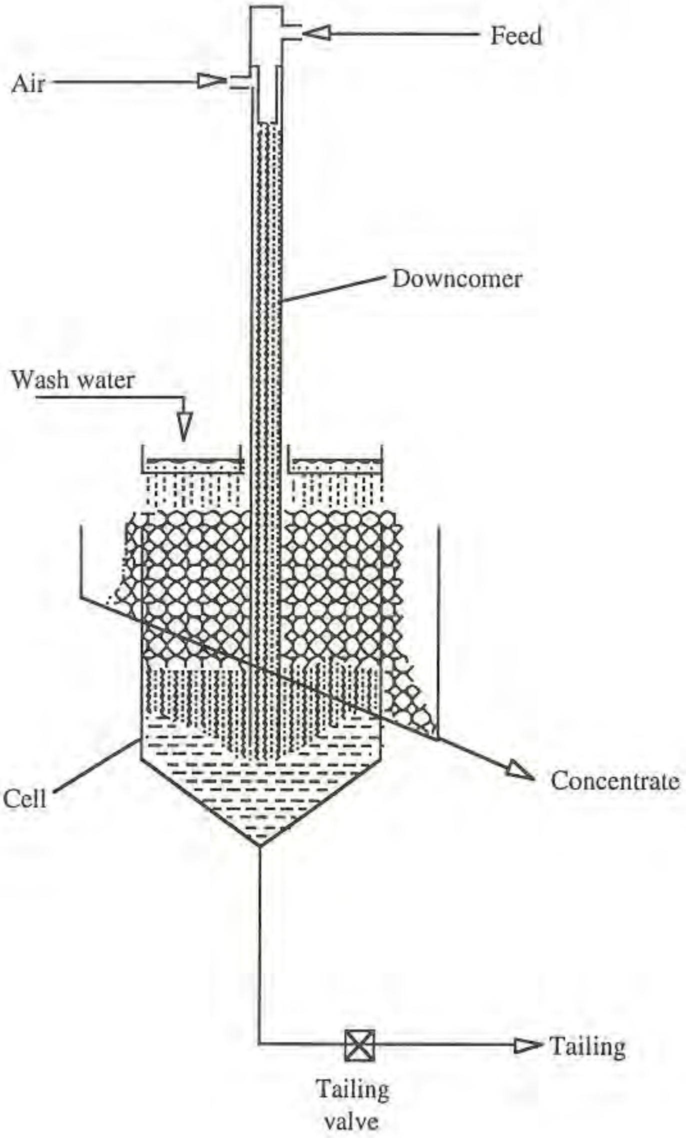


FIGURE 1
Schematic representation of a single-downcomer Jameson cell

In a conventional mechanically agitated cell, the void fraction is low (5-10%) and the bubble size is large (2-3 mm) and so as a result of the relatively low interfacial area, the probability of particle bubble-contact is low. Thus several cells in series are required to increase the particle residence time and improve the overall contact probability. In a conventional column, the void fraction is similarly low, and so the probability of particle-bubble contact has to be increased through increasing the particle residence time in the cell by providing a large column height. The Jameson downcomer provides an environment where particle–bubble contact probability is virtually 100% with a slurry residence time in the downcomer of 5–10 seconds.

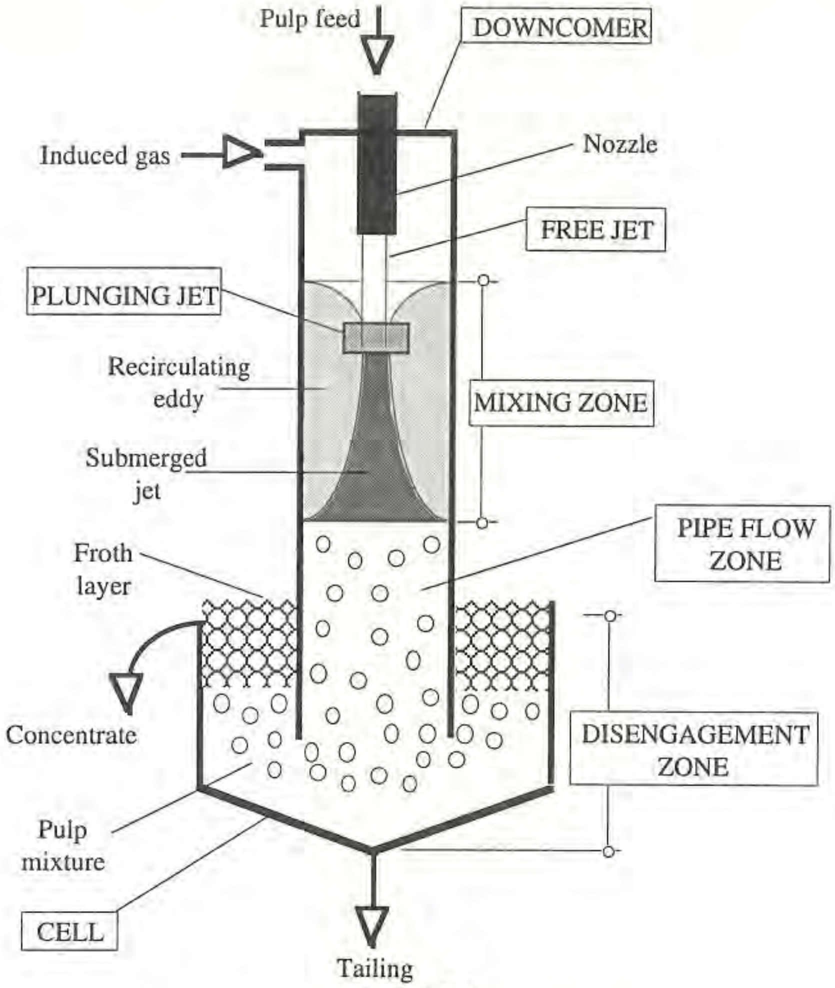


FIGURE 2
Various active zones in the Jameson cell (not to scale)

OPERATIONAL CHARACTERISTICS

Froth depth

The froth phase produced in a Jameson cell can be controlled in the same manner as in conventional columns. In principle, shallow froth depths (less than 100 mm) are used where high recovery is necessary and concentrate ash is of secondary importance, while deeper froths (up to 1000 mm) are employed to obtain minimum concentrate ash. Use of shallow froths can result in significant entrainment of very fine (less than 10 µm) gangue mineral (slime) components which accompany the pulp phase. Use of deeper froths results in significant drainage of hydrophilic gangue producing a higher grade concentrate, and a higher percentage of solids in the concentrate. Under some circumstances, addition of counter-current froth wash water assists froth mobility, where otherwise excessive froth drainage would result in an immobile froth.

In general for coal applications, Jameson cell performance is not very sensitive to froth depths above 200 mm. Figure 3 below shows yield and ash as a function of froth depth for a central Bowen Basin coal. This graph shows the expected increase in both yield and product ash at low froth depth. Other data (Cheng and Clarkson, 1992) has shown the opposite trend, but in both cases, froth depths above 200mm produce little change in performance.

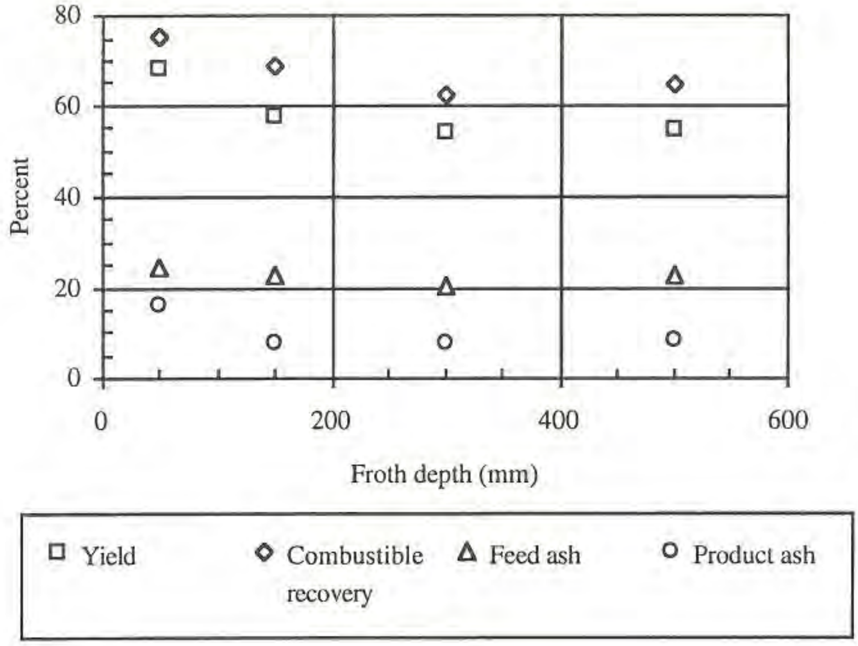


FIGURE 3
Influence of froth depth

Reagents

Figures 4 and 5 show the influence of collector and frother addition rates respectively, for the same coal as in Figure 3. Collector (diesel) addition rates above approximately 0.5 kg/tonne provide no yield benefit. For this coal, performance was relatively insensitive to frother (MIBC) additions above 30 g/m^3 of feed. These results represent the typical response to reagent changes for the Jameson cell.

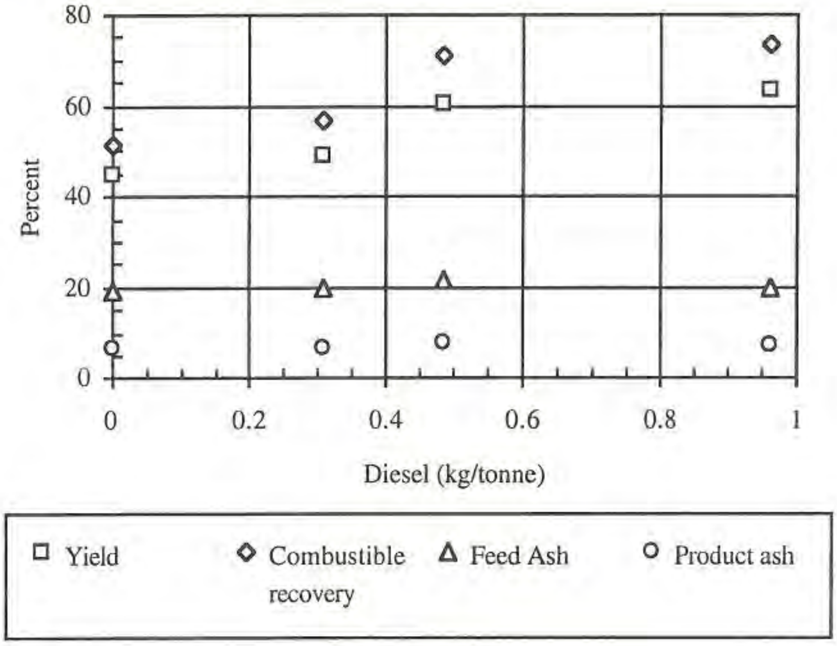


FIGURE 4
Influence of diesel (collector) addition

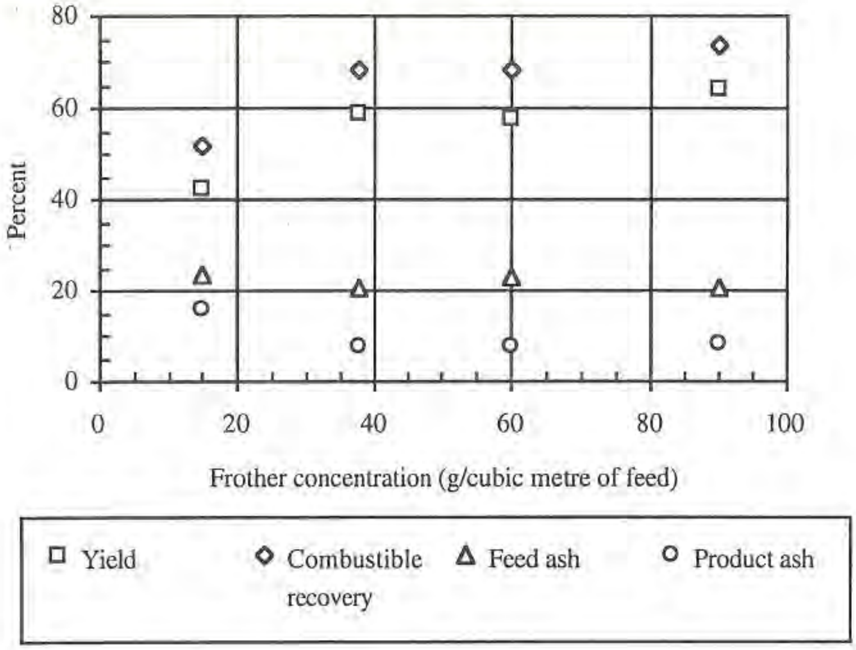


FIGURE 5
Influence of MIBC (frother) addition

Superficial gas velocity
The superficial gas velocity (J_g, cm/s) is the upward superficial velocity of air in a flotation cell. In the Jameson cell, the J_g is calculated by dividing the downcomer air rate (cm^3/s) by the cross-sectional area (cm^2) of the riser part of the cell. The cell is normally circular or rectangular in section, and the appropriate cross-sectional area is simply the area normal to the direction of the flow of the froth, excluding the area occupied by the downcomer(s). It is conveniently expressed in units of cm/s because values typically range from 0.8 to 1.4 cm/s in practice for coal applications.

For a stream which is not carrying-capacity limited (see below), the recovery and concentrate carrying rate (g/min/cm^2) tend to increase with increasing J_g, as in conventional columns (see Finch and Dobby (1990,1991)). For a given stream and frother concentration, a maximum air rate (subsequently $J_{g,max}$) is reached, above which froth flooding occurs, resulting in the loss of froth-pulp interface, a very wet froth, and total loss of selectivity. In flooding, the entire cell fills with froth as the only stable phase, and there is no pulp phase.

The operating J_g used in the sizing of the Jameson cell depends strongly on the application. Generally speaking, low values (J_g= 0.4 to 1.0 cm/s) are employed in mineral cleaning applications, and high J_g's(1.0 to 2.0 cm/s) are employed in mineral roughing or scavenging applications. As mentioned earlier, values of 0.8 to 1.4 cm/s are commonly employed for coal. When treating fine coal (< 100 μm) J_g's less than 1 cm/s are sufficient, while J_g's greater than 1 cm/s are required in coarse coal flotation (< 0.5 mm wedgewire). The reasons for these choices are as follows. In fine coal flotation and cleaning operations, a high proportion of the feed reports to the concentrate, and the froth loading tends to be high. Consequently, the bubbles are well-coated with particles, which tend to stabilise the froth. The drainage rate of the interstitial liquid in the froth is retarded by the relatively high concentration of particles, so it is necessary to design for lower values of J_g to allow time for the gangue to drain from the froth to obtain the required high grade (low ash). When treating coarse coal however, excessive froth drainage can result in an immobile froth unless higher J_g's are employed. In mineral roughing applications only a small fraction of the feed reports to the concentrate, and the froths formed tend to be less stable as a consequence. Also, gangue entrainment is not such a serious problem, because it can be coped with in the downstream cleaning circuit.

Bubble size
Bubble sizing determinations for full-scale operating cells, and test cells at the University of Newcastle, show that the Jameson cell produces an arithmetic mean bubble diameter of the order of 300 – 600 μm, while the Sauter (volume-to-surface) mean diameter, d_{vs}, is of order 360 – 950 μm, (Atkinson *et al.* ,1992). These sizings are appreciably finer than those obtained for mechanical cells and conventional columns where the Sauter mean bubble size is typically 2 to 3 mm. In the flotation process, for a given air flow rate, reduced bubble size results in increased bubble surface area and enhanced flotation kinetics.

Figure 6 shows bubble size versus air/feed ratio for a 300 mm diameter Jameson pilot cell treating a fine copper retreatment stream (d_{80} = 30 μm). These data confirm visual observations that bubble size decreases with decreasing air/feed ratio.

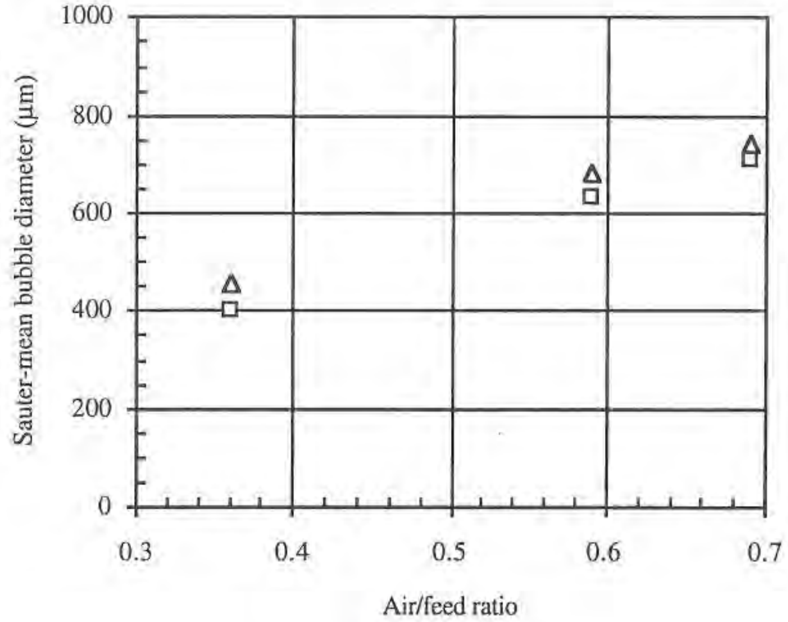


FIGURE 6
Bubble size versus air-feed ratio

Air/feed ratio
Jameson cells generally operate with a volumetric air/feed ratio of 0.3 – 0.9. Experiences with large (2 – 3 m diameter) Jameson cells indicate that operation at a low air/feed ratio does not appear to detract from metallurgical performance *providing the superficial gas velocity* J_g *is maintained* by decreasing the cell cross-sectional area. Operation at lower air/feed ratios has a stabilising effect producing a more uniform and finer bubble size. A significant advantage of operation at lower air/feed ratios is that lower concentrations of frother are required. For coal applications, yield has been reported to drop for air/feed ratios below 0.6, although this is most likely a reflection of the reduction in superficial gas velocity, rather than a result of the reduced air-feed ratio.

Maintenance of metallurgical performance at low air/feed ratios can be explained by considering the flux of interfacial area, (S_b, (interfacial area/s) per unit of column cross-sectional area; s^{-1}) as defined by Finch and Dobby (1990):

$$S_b = 6\,J_g/d$$

where *d* is the bubble diameter. The superficial bubble surface rate (and hence the collection capability) can be maintained with reduced superficial gas rate providing the bubble size decreases proportionately. The data in Figure 7 were calculated using information from Figure 6 using the above equation, and show the superficial bubble surface rate as a function of the air/feed ratio for a fixed cell cross-sectional area. It is seen that the superficial bubble surface rate remains almost constant at 55 s^{-1} approximately, despite two-fold changes in the air/feed ratio.

SIZE–YIELD RESPONSE

Cheng and Clarkson (1992) compared the performance of a two-stage(rougher-scavenger 7 m^3/hr, reagents to rougher feed only) Jameson cell versus 100 ft^3 mechanical Denver cells (5 in series) and 300 ft^3 cells (4 in series) at a southern Bowen Basin coal preparation plant. For comparison purposes, the mechanical cells should be thought of as being one rougher cell followed by 4 or 3 scavenger cells respectively. Figure 8 shows the combustibles recovery data for each case where the feed material was fine coal (raw "slimes flotation feed" approximately 26%ad ash) comprising 96% minus 500 μm and 55% minus 45 μm. The data shows very high recoveries for all size classifications when using the Jameson cell. This is a function of the highly efficient air-particle contact achieved in the downcomer.

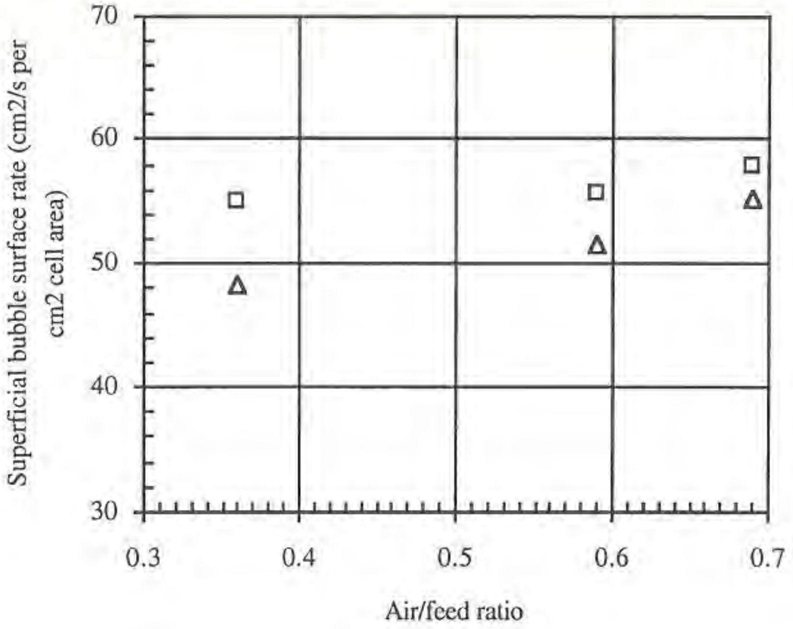


FIGURE 7
Superficial bubble surface rate versus air-feed ratio

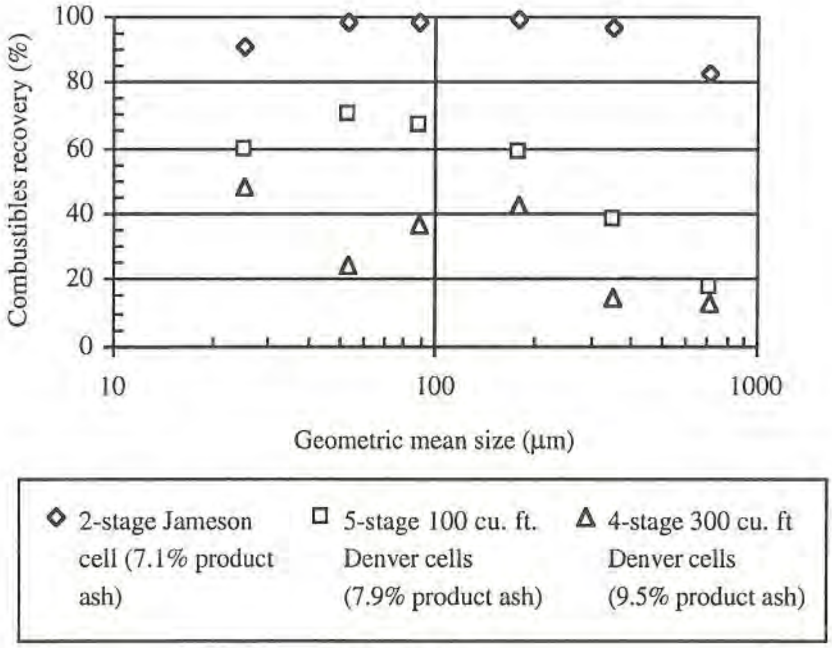


FIGURE 8
Combustibles recovery for Jameson cells versus conventional mechanical cells

Cheng and Clarkson (1992) also compared the results for the Jameson cell treating a coarse flotation feed. Figure 9 shows the results for the coarse feed of 69% minus 500 μm and 9% minus 45 μm. For other than the plus 500 μm material, the Jameson cell performed very well. In practice, this would allow spirals to be operated with a coarser feed resulting in significant improvement in spiral performance, with the Jameson cell recovering the fine material. The slightly higher recovery of minus 45 μm material for the spiral is a function of slimes entrainment to product as opposed to particle separation, and the ash of the minus 45 μm product fraction was 31% for the Jameson cell compared to 57% for the spiral.

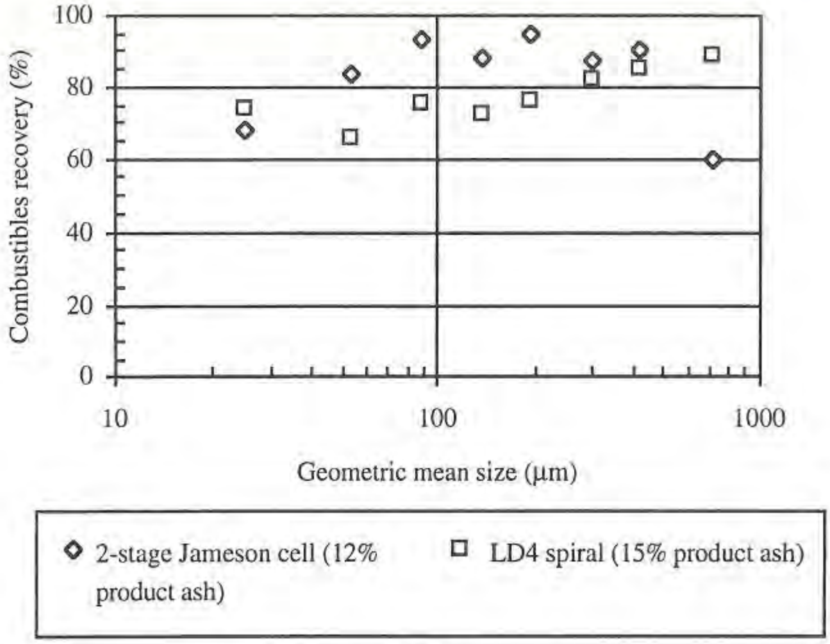


FIGURE 9
Combustibles recovery for Jameson cell versus LD4 spiral

Figure 10 shows a summary of all size-recovery data available for the Jameson cell for different coals at the time of preparation of this paper. The results for the Queensland coal examples represent results from raw plant feed and are excellent across all size fractions. The lower yield response for the other coals may be explained by considering the nature of the feed (generally high ash tailings material), and may be due to insufficient conditioning times and/or collector starvation due to the very fine nature of the feeds. It is also important to note that the results for New South Wales coals (c-e) are for single stage Jameson cell treatment only, whereas the two Queensland examples are for two-stage treatment.

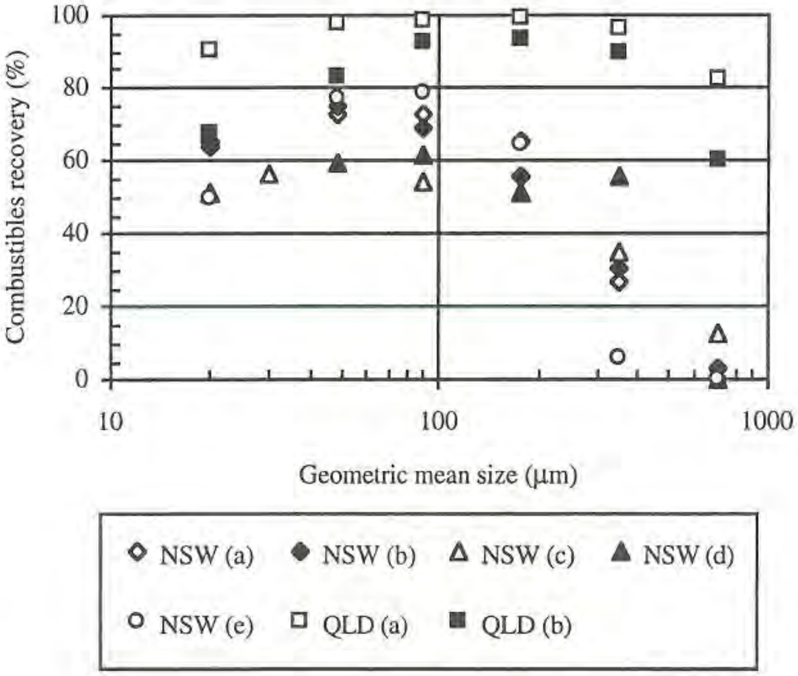


FIGURE 10
Combustibles recovery versus size – various coals

WASH WATER CONSIDERATIONS

In conventional column flotation literature, wash water addition rate is normally referred to in terms of the 'bias', which is the absolute excess of the wash water applied over the quantity of water reporting with the concentrate, expressed as a superficial velocity J_b (cm/s) where:

$$J_b \quad = \quad (Q_{WW} - Q_{WC})/A_C,$$

where Q_{WW} is the wash water flowrate, Q_{WC} is the rate of flow of water in the concentrate and A_C is the cross-sectional area of the column.

The wash water ratio is defined as the ratio of the wash water addition rate, to the flowrate of water in the concentrate:

$$W \quad = \quad Q_{WW}/Q_{WC}$$

The wash water ratio is a relative measure of the amount of wash water applied. If no wash water is used, the wash water ratio is zero and the bias is negative. When $J_b = 0$, $W = 1$; positive bias corresponds to wash water ratios greater than unity.

Although the bias does give an indication of the absolute amount of wash water being added, its use can be misleading because it does not take into account the wide variation in the absolute values of the rate of water entrainment in the concentrate. It is preferable to use the wash water ratio, which is a relative figure.

Figure 11 shows a comparison of measured wash water ratios and bias rates for a range of operating conditions for a Jameson cell operating on a rougher application. Although the bias covers a relatively narrow range, the wash water ratio has much greater variation. To an operator, the difference in bias between 0.02 and 0.04 cm/s may seem negligible, but it can be seen that in some circumstances, such changes in J_b may lead to a three-fold increase in the wash water ratio – supplying in effect a 200 percent overload of wash water.

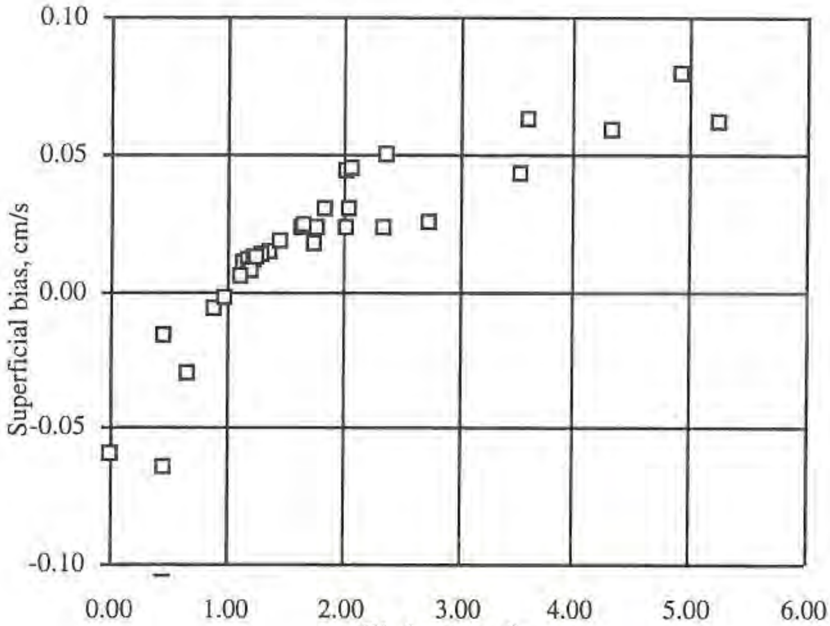


FIGURE 11
Superficial wash water bias J_b versus wash water ratio

CONCENTRATE CARRYING CAPACITY

The carrying capacity, *Ca*, is the limiting or maximum concentrate production rate per unit of area of cell cross-section, usually expressed as concentrate solids rate (g/min) per unit of available cell cross-sectional area (cm^2). The carrying capacity is strongly influenced by the gas flowrate and the size of the bubbles in the froth discharging at the overflow lip, since these two factors influence the rate of surface area discharge into the launder. The mass of the hydrophobic particles which can be carried by the froth varies directly with the surface area. The other important factor is the size of the particles, because when a layer of particles is adhering to a gas-liquid interface, the mass of particles per unit of interfacial area varies directly as the mean particle size.

Carrying capacity limitations may apply in coal flotation because the feed solids generally contain a high proportion of floatable material, where the bubbles can easily become fully loaded with particles. The yield is then determined by the carrying capacity, rather than by the flotation kinetics.

It is important to note that the concept of carrying capacity applies equally to conventional mechanical cells as well as to column flotation technologies.

The expected carrying capacity for conventional columns has been given by Espinosa-Gomez *et al.* (1988) as :

$$Ca = \alpha d_{80} \rho_p$$

where d_{80} is the size at which 80% by mass of the concentrate passes, expressed in µm, and ρ_p is the density of the particles (g/cm^3). In conventional columns, the carrying capacity depends on column diameter. The parameter α has been found to be 0.068 in small columns (as reported by Espinosa-Gomez *et al.* (1988) for a 50-mm column). For larger columns, up to 1.0 m diameter, Finch and Dobby (1990) give $\alpha = 0.05$; and for columns greater than 2.0 m diameter, $\alpha = 0.035$ (Espinosa-Gomez and Johnson (1991)).

The parameter, α, decreases with increasing cell diameter due to froth transport limitations, because the proportion of particles dropped back into the pulp increases as the distance over which the froth phase has to travel increases. In practice, the maximum predicted column carrying capacity values may be improved upon by increasing the effective lip-length (lip-length to cell area ratio) of the cells, hence reducing the transport constraints on the froth phase. Plant trials at Newlands on plant-scale Jameson cells has indicated that lip-length does not appear to be significant, as a reduction of one-third of the plant lip-length did not affect concentrate mass transfer rate.

In all Jameson cell applications, concentrate carrying capacities have been shown to be at least as high as the values predicted for conventional columns using the above equation. In direct Jameson cell and column coal comparisons, the Jameson cell has been shown to significantly exceed the column carrying capacity, based on actual results.

The potential occurrence of carrying capacity limitations in coal flotation can be demonstrated by investigating two different, but typical, feed size distributions which may be encountered (Table 1).

TABLE 1
Carrying capacity for typical coarse and fine flotation feeds

	Case 1 (Coarse)	Case 2 (Fine)
J_g cm/s	0.8	0.8
Air-feed ratio	0.5	0.5
Nominal feed size	–0.5 mm wedgewire	–100 µm
d_{80} µm (concentrate)	300	60
ρ_p g/cm^3 (concentrate)	1.30	1.30
Feed % solids	10	5
Calculated *Ca* g/min/cm^2	13.65	2.73

Figures 12-14 summarise the effects of each of the principle operational parameters for each case. For case 1 (coarse coal), carrying capacity limits yield only for an air-pulp ratio less than 0.4, a J_g greater than 1.1 cm/s, or for high feed % solids above 15%. For case 2 however, the concentrate production rate for the fines feed is carrying capacity limited for most typical operating scenarios.

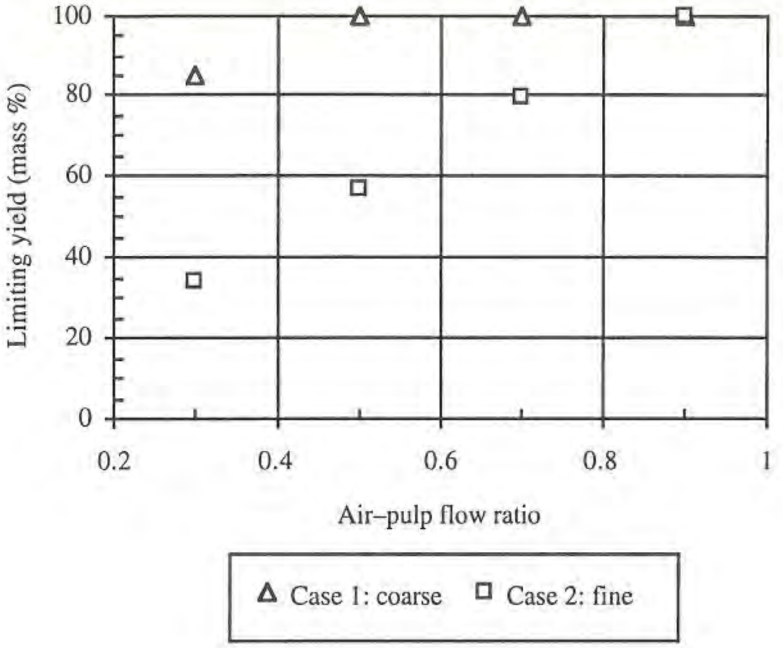


FIGURE 12
Limiting yield as a function of air-pulp flow ratio

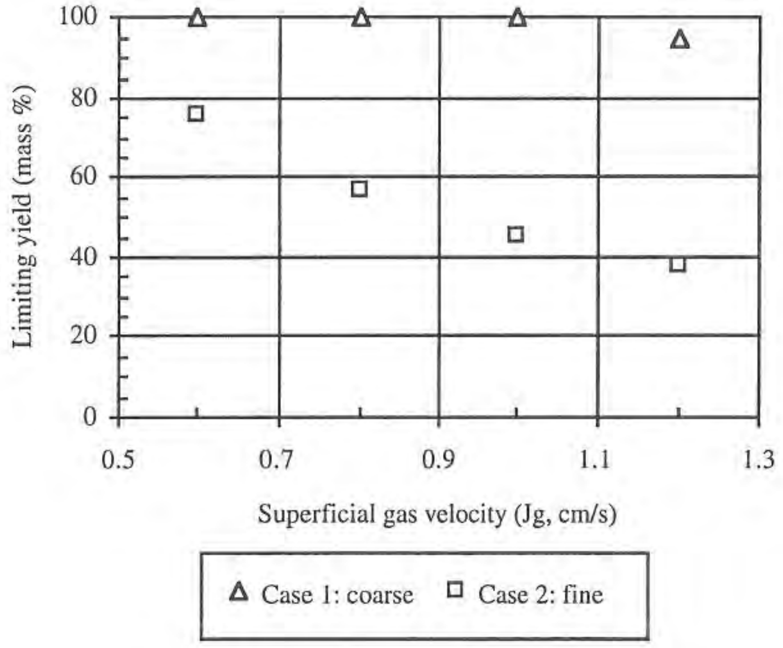


FIGURE 13
Limiting yield as a function of superficial gas velocity

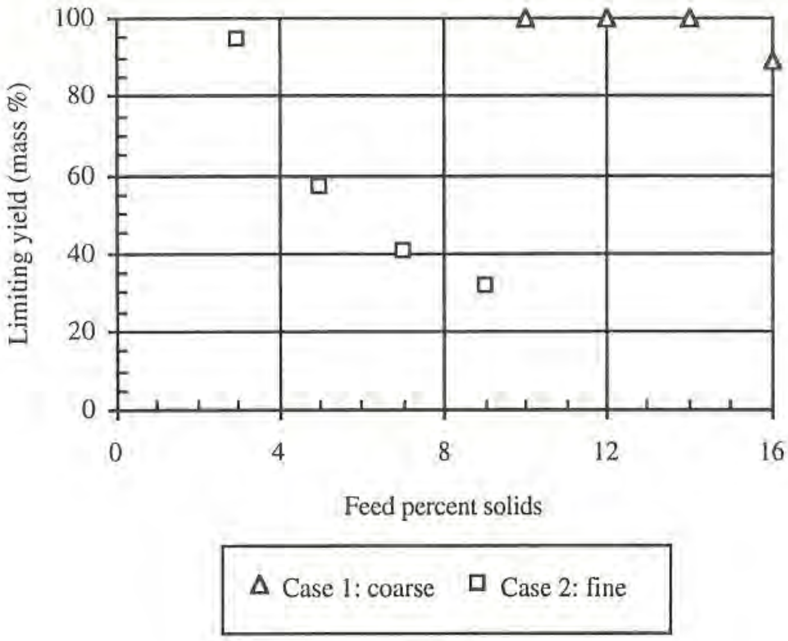


FIGURE 14
Limiting yield as a function of feed percent solids

SUMMARY

The probability of particle-air bubble contact in a Jameson cell has been shown to be extremely high because of the intensive mixing in the downcomer, and the high void fraction (50-60%) combined with small bubble size (< 1 mm).

The air–feed ratio in a Jameson cell may vary from 0.3 up to 0.9. Results indicate that operation at low air-feed ratio, while maintaining superficial gas velocity, enhances operation through reduced bubble size, improved stability, and the consequential reduction in frother addition. The net bubble surface rate, and hence metallurgical performance is maintained at low-air-feed ratio due to the associated reduction in bubble size.

In coal flotation, the Jameson cell appears to be relatively insensitive to froth depths above 200 mm.

The Jameson cell is capable of combustibles recoveries in excess of 80% for all coal particles less than 1 mm, when treating raw flotation feed in two stages. Lower recoveries have been observed for feeds other than the raw flotation feed, although this is most likely a function of a) tailings feed containing a high proportion of middlings material, and/or b) inadequate conditioning and/or collector starvation of coarse particles.

In some cases, depending on the nature of the feed material, wash water can be used to improve product ash. Where wash water is employed, the quantity added should be measured by the parameter *wash water ratio*, (*W*), with this value maintained at or below 1 to minimise loss of yield.

The concept of *carrying capacity* may apply to coal flotation systems, particularly when treating fines feed (minus 100 µm). In such an application, yield will be artificially limited by bubble loading constraints and froth transport phenomena, rather than by kinetic constraints.

ACKNOWLEDGMENTS

The authors wish to thank BHP Australia Coal Ltd (Blackwater mine), FAI Mining Ltd (Macquarie Coal Preparation Plant), Wambo Mining Corporation Pty Ltd, and Coal and Allied Ltd (Hunter Valley No. 1 Preparation Plant) for providing facilities and/or data to assist with the preparation of this paper. The authors also acknowledge MIM Holdings Limited–Marketing of Technology for permitting the publication of this paper. One of the authors (BWA) is a recipient of an Australian Postgraduate Research Award (Industry) scholarship which is also sponsored by MIM Holdings Limited and The University of Newcastle Research Associates Ltd (TUNRA).

NOMENCLATURE

A_C	free cross-sectional area of the cell	cm^2
Ca	concentrate carrying capacity	g/cm^2-min
d	bubble diameter	m
d_{80}	80% passing size	μm
d_{vs}	Sauter mean bubble diameter	m
J_b	wash water bias rate	cm/s
J_g	superficial gas velocity	cm/s
Q_{WC}	flowrate of water in the concentrate	m^3/s
Q_{WW}	wash water flowrate	m^3/s
S_b	superficial bubble surface rate	s^{-1}
W	wash water ratio	
	Greek Symbols	
α	empirical parameter	10^4 (min^{-1})
ρ_p	density of the particles	g/cm^3

REFERENCES

Atkinson, B.W., Griffin, P.T., Jameson, G.J. and Espinosa-Gomez, R., 1993, "Jameson cell test work on copper streams in the copper concentrator of Mount Isa Mines Limited", XVIII International Mineral Processing Congress, Australasian Institute of Mining and Metallurgy, Sydney, Australia, 23-28 May.

Cheng, C.Y. and Clarkson, C.J. 1992, "A report on the flotation of fine and coarse coals using a two-stage pilot Jameson cell", Internal report, Department of Mining and Metallurgical Engineering, University of Queensland.

Espinosa-Gomez, R. and Johnson, N.W. 1991, "Technical experiences with conventional columns at Mount Isa Mines Limited", in Column '91, Proceedings of an International Conference on Column Flotation (G.E. Agar, B.J. Huls, and D.B. Hyma, eds), Sudbury, Ontario, July 1991, pp 511-524.

Espinosa-Gomez, R., Finch, J.A., Yianatos, J.B. and Dobby, G.S. 1988, "Flotation column carrying capacity: particle size and density effects," Miner. Eng., Vol. 1, No 1, pp 77-79.

Evans, G.M., Atkinson, B.W. and Jameson, G.J. 1993, "The Jameson cell", Flotation Science and Technology, K.A. Matis, ed., Marcel Dekker, New York . (In print).

Finch, J.A. and Dobby, G.S. 1990, Column Flotation, Pergamon Press, Oxford.

Finch, J.A. and Dobby, G.S. 1991, "Column flotation: a selected review. Part I", Int. J. Miner. Process., Vol. 33, pp 343-354 .

JAMESON CELL TEST WORK ON COPPER STREAMS IN THE COPPER CONCENTRATOR OF MOUNT ISA MINES LIMITED

B.W. Atkinson[1]*, P.T. Griffin*[2]*, G.J. Jameson*[1]*, R. Espinosa-Gomez*[2]

1. Department of Chemical Engineering, University of Newcastle, N.S.W. 2308
Telephone (049) 21 6180 Facsimile (049) 601445

2. Mount Isa Mines Limited, Mount Isa, Qld. 4825

XVIII International Mineral Processing Congress, AusIMM, 23-28 May, 1993

JAMESON CELL TEST WORK ON COPPER STREAMS IN THE COPPER CONCENTRATOR OF MOUNT ISA MINES LIMITED

B.W. Atkinson[1], P.T. Griffin[2], G.J. Jameson[1], R. Espinosa-Gomez[2]

1. Department of Chemical Engineering, University of Newcastle, N.S.W. 2308
2. Mount Isa Mines Limited, Mount Isa, Qld. 4825

ABSTRACT

Two major issues are to be addressed in the Copper Concentrator of Mount Isa Mines Limited. The first of these is the need to increase concentrator flotation capacity to increase copper recovery, while the second is the need to reduce the iron sulphide dilution to meet future smelting requirements. Extra capacity in the cleaner and retreatment sections of the concentrator have been identified as key target areas, with additional copper recovery of at least 1% identified from plant trials.

Extensive pilot plant tests have been undertaken to evaluate the performance of the Jameson cell compared to conventional mechanical cell and column circuits. This work has shown that Jameson cells present an opportunity to achieve the desired grade and capacity specifications at lower capital cost than available alternatives.

This paper outlines metallurgical results of work undertaken in the Copper Concentrator, together with a summary of pertinent Jameson cell operating parameters investigated during the period.

INTRODUCTION

In the Jameson cell, feed slurry is introduced to the top of a vertical pipe (downcomer) through a constraining orifice to produce a jet of slurry in the downcomer. Air is self induced through an inlet in the downcomer resulting in a stable, high void fraction (approximately 60%) froth phase in the downcomer. At the exit of the downcomer, bubbles disengage from the tailing slurry and rise to overflow the cell lip into the froth launder. Froth depth and wash-water conditions may be varied as with a conventional column. Further descriptions of the Jameson cell are reported elsewhere - Kennedy (1990); Clayton, Jameson and Manlapig (1991).

The operations at Mount Isa and Hilton use Jameson cells in many applications, including generation of lead-zinc low-grade-middling (LGM) final concentrate and lead and zinc final concentrates. This paper outlines tests undertaken at the Copper Concentrator of Mount Isa Mines Limited (MIM), the impetus for evaluation, the results of testing, and the likely outcomes of the work at Mount Isa.

The MIM Copper Concentrator has been recently described (Espinosa-Gomez et al, 1991). Since that publication, the conventional crushing and grinding circuit has been replaced by two 9.75m diameter autogenous mills, followed by closed-circuit ball mill-classification to generate a flotation feed with d_{80} approximately 90 µm (Leung, Lumsdaine and Keran, 1992). Figure 1 shows a schematic diagram of the current comminution circuit.

The concentrator treats about 18 000 tonnes per day of 3% copper ore. Chalcopyrite is the only copper species and the main gangue minerals are iron sulphides (~ 13%) and non-sulphide gangue (~ 78%). Typical copper recoveries are 95% at a 25% copper concentrate grade. Figure 2 shows a schematic diagram of the concentrator circuit.

Pilot scale evaluation of Jameson cell technology has proceeded in the Copper Concentrator on the basis that it provides the cheapest option to increase capacity of the plant. The current bottle-neck in the plant is limited cleaning and retreatment capacity (Griffin, 1992). The Jameson cell offers the advantage of being able to produce final grade concentrate from virtually any stream, (where sufficient liberated chalcopyrite available and wash water applied), while being relatively low in capital cost, especially compared to conventional columns.

EQUIPMENT

Table 1 summarises the equipment used for evaluation.

Downcomer air rate was measured using a rotameter for the J100 rig. For the J500 rig, a differential pressure gauge mounted around a 12.0 mm orifice plate located inside a 25.4 mm induced air line was employed. J1000 air rate was measured using a vortex flowmeter. For each rig, wash water flowrates were calibrated by timing flows without the cell in operation. The J100 and J500 cells were fed from 200 litre and 2000 litre agitated conditioning tanks respectively. The J1000 cell was fed directly from a sump without conditioning. Reagents were applied to the conditioning tanks (feed sump for J1000) using Prominent dosing pumps. Where applicable, frother (MIBC) was applied undiluted; butyl xanthate (BX) was applied as a 9% w/w solution; and both cyanide (CN) and D101[3] were applied as 5% w/w solutions.

---

[3] D101 is a mixture of naphthalene sulphonate and dextrin.

TEST PROGRAM

The streams tested in the Copper Concentrator (Figures 1&2) were:

- autogenous mill discharge (using J1000 as "flash" flotation unit)
- flotation feed (using J500 for carbonaceous iron sulphide pre-flotation followed by J100 treatment of J500 tailing to produce final copper concentrate in a single stage)
- rougher concentrate (J500 - comparison with conventional columns)
- retreatment concentrate (J500 - comparison with conventional columns)
- retreatment feed (J500 and J100).

The J1000 and J500 units were used for batch-mode testing only, where feed, concentrate and tailing was sampled for each test. The J100 unit can be used in either batch-mode or tail-recycle-mode. In tail-recycle-mode, the tailing is continuously recycled to the feed conditioning tank, and the concentrate collected in timed intervals over a nominated period or until a barren froth is present. Batch-mode testing produces only a single point on a grade-recovery curve corresponding to single stage flotation, whereas the tail-recycle method leads to a full grade-recovery curve for a given feed.

RESULTS

AUTOGENOUS MILL DISCHARGE ("FLASH" FLOTATION)

The primary reasons for selecting this coarse stream (d_{95} of 4-5 mm and a d_{80} of 250 - 500 µm) for testing were:

a) high chalcopyrite liberation (e.g. chalcopyrite liberation as measured in product of similar size (d_{80} = 300 µm) is 82%[4] , Espinosa-Gomez et al, (1991a)). Thus by recovering this liberated chalcopyrite, the risk of unnecessary losses due to over-grinding is minimised.

b) high flotation rate due to non-reducing environment of autogenous mills, i.e. Eh = 240 mV SHE instead of a typical value of 50 mV at the head of the rougher bank (Trahar, 1984).

The results of single stage treatment of the autogenous mill discharge are summarised in Figure 3. From a 3.15% copper feed, an average final concentrate grade of 27% copper was

---

[4] Uncorrected liberation (2 dimensions) measured by QEM*SEM (Quantitative Evaluation of Materials by Scanning Electron Microscopy)

achieved with 50% copper recovery, (Atkinson, 1992). Recoveries up to 65% were observed at similar concentrate grades.

The installation of Jameson cells to treat the autogenous grinding mill discharge would result in a dramatic reduction in load on the conventional cleaning circuit. Assuming single stage Jameson cells were installed for this duty, the mass and copper loads in the cleaner circuit would be reduced by about 37% and 56% respectively, resulting in a 47% increase in flotation capacity. (The value of 47% is the average of 37% and 56% and it is a valid figure, considering that additional capacity will not only come from reducing the mass load, but also the copper load in a system which is already overloaded.)

Chalcopyrite recovery is maximum in the 8 - 100 μm range but decreases markedly outside this range, (see Figure 4). At equivalent recovery, the maximum size range recovered in the recleaner concentrate (see Figure 2) is much finer, 8 - 37 μm. Thus, flotation of this stream instead of the current flotation feed would avoid further regrinding of already liberated chalcopyrite. This is also reflected in the coarser sizing of the autogenous mill discharge concentrate versus recleaner concentrate, i.e. d_{80} of 50 μm versus 37 μm.

Entrainment rejection is efficiently controlled, as reflected by the low non-sulphide gangue recovery in the finest fraction (see Figure 4). Iron sulphide recovery is mainly due to true flotation of the naturally floatable carbonaceous pyrite (Grano et al, 1990).

Wash water ratios above 0.6 were required to obtain maximum concentrate grade, while wash water ratios above 1.0 significantly reduced recovery. *Wash water ratio* is defined as the ratio of the (wash water addition rate) to the (concentrate water removal rate). A wash water ratio below 1.0 indicates negative bias, while a value greater than 1.0 indicates positive bias.

Maximum concentrate grades were approximately 28% in this work. Concentrate was mainly diluted by fine naturally floating carbonaceous material. Small scale J100 tests indicate that concentrate grade can be increased to 31% when carbonaceous pyrite is removed in a pre-flotation stage.

FLOTATION FEED

Final concentrate is currently diluted by naturally floatable carbonaceous pyrite and non-sulphide gangue. This dilution accounts for about 18% of the final concentrate by mass. In the past (Grano et al, 1991) a pre-flotation stage has been investigated for the purpose of removing the naturally floatable material to improve the quality of final concentrate.

A Jameson cell pre-flotation stage was conducted on the flotation feed before collector addition and 2% of the flotation feed mass was removed as pre-flotation concentrate. This corresponds to removal of 50% of the total carbonaceous material. Copper losses in the Jameson cell pre-flotation concentrate were effectively 0.8%, which was similar to values obtained in the past (Grano et al, 1991).

A subsequent Jameson cell flotation stage on the pre-flotation tailing produced 35-45% recovery of copper at 32% Cu. As with the autogenous grinding mill discharge, the installation of single stage Jameson cells would reduce the mass and copper loads in the cleaner circuit, this time by about 23% and 33% , increasing cleaning capacity by about 28% (estimated as for the autogenous grinding mill discharge). Additional increase in cleaning capacity will also result from removing the carbonaceous pyrite. This is due to the lower mass load, and also due to the faster chalcopyrite flotation rate, as less carbonaceous pyrite depressant, D101, would be added (D101 does not only reduce the carbonaceous pyrite flotation rate but also the chalcopyrite).

ROUGHER CONCENTRATE

As expected, Jameson cell treatment of the rougher concentrate (~18% Cu) is carrying capacity limited. The Jameson cell concentrate produced from floating rougher concentrate has a d_{80} of the order 26 μm and a specific gravity of approximately 4.0. Using the principle of carrying capacity as defined by Espinosa-Gomez et al (1988), the expected carrying rate limitation for conventional columns is given by

$$C_a = \kappa * d_{80} * SG_{(conc)} \qquad \text{(at positive wash water bias)}$$

where d_{80} is expressed in μm, and $SG_{(conc)}$ in g/cm^3. The constant κ has been found to be 0.068 in small columns, particularly in respect to a 50mm column (Espinosa-Gomez et al, 1988); 0.05 for larger columns less than 1.0m diameter (Finch and Dobby, 1991); and 0.035 for columns greater than 2.0m diameter (Espinosa-Gomez and Johnson, 1991). For this case, the predicted column carrying capacity is 5.2 g/min/cm^2 for a 400mm diameter cell. Carrying capacity limitations generally apply where high grade streams are treated, as in cleaning applications, where the bubble (hence froth) loading constraint overrides any kinetic constraint.

Figure 5 shows measured carrying rates (Atkinson and Espinosa-Gomez, 1992) as a function of wash water ratio, and shows the carrying rate is strongly influenced by wash water addition. This effect may be less pronounced in columns where particles dropped back from the froth have a higher chance of recollection than in the Jameson cell.

This work shows that the concept of carrying capacity for a Jameson cell must be defined at a particular wash water ratio. For maximum concentrate grade and recovery, it is generally necessary to aim for zero to marginally positive bias wash water, i.e. a wash water ratio of 1.0 or just slightly higher. Thus it would be appropriate to define carrying capacity at a wash water ratio of 1.0. In this case, the resulting measured C_a is 5.3 g/min/cm^2, which compares favourably with the predicted column value of 5.2 g/min/cm^2. Measured carrying capacities in full-size cells at the Hilton Lead-Zinc Concentrator (J3000) and the Mount Isa Lead-Zinc Concentrator (J1900) have been shown to be at least as high as the predicted rates using a value of 0.035 for the coefficient, κ.

The average result from single stage tests treating the rougher concentrate was a chalcopyrite recovery of 20% at 31% Cu grade. Copper recoveries varied from 9% to 39%. The recovery will change with head grade and, perhaps more significantly, with feed rate, as the Jameson cell operation is fixed in this case by a maximum concentrate rate. Testing of the plant columns in the same application showed similar low recoveries (17 - 35%).

The Jameson cell matches the capability of conventional columns in terms of achievable concentrate grade, at a level significantly above that of the mechanical cleaning banks, which only achieve the order of 27% Cu. Figure 6 shows the grade-recovery curves for conventional columns and the Jameson cell (Riches, 1992).

RETREATMENT CIRCUIT

The retreatment feed consists of reground middling material from the cleaner tailing and scavenger concentrate. This material typically has a d_{80} of 30 μm and assays approximately 3% Cu. The current overall average performance of the retreatment circuit (including columns) is 75% copper recovery at 12% Cu.

Grade-recovery curves obtained for the Jameson cell are at least equal to that obtained for the 50mm pilot column. Figure 7 shows a comparison of performance between the Jameson cell, a 50 mm diameter column, and the existing mechanical cell retreatment bank. The grade-recovery curves for the Jameson cell were achieved using the J100 tail-recycle method treating the tail of a batch J400 test.

Recovery by size comparisons at the same overall chalcopyrite recovery showed the Jameson cell chalcopyrite recovery to be higher than the 50mm column by approximately 10% in the fine fractions (< 10 μm) and lower by approximately 10% in the coarse fractions (> 20 μm). Since this comparison was undertaken at the same overall recovery level, and under positive bias wash water conditions, it would appear that the Jameson cell is capable of higher fines recovery.

Figure 8 compares the performance of the plant columns and the Jameson cell on the retreatment concentrate. The Jameson cell produced superior concentrate grades compared to the plant columns, although the column results show that insufficient wash water was applied to column 1. With adequate wash water applied to the columns, equivalent grade-recovery performance would be expected for both the plant columns and the Jameson cell.

A recovery limitation of 40% per stage was observed for the Jameson cell. This limiting single stage recovery has also been observed on the plant columns. The recovery limitation of the Jameson cells floating the retreatment streams was investigated and found not to be due to carrying capacity constraints, as no limiting carrying rate was observed for either the retreatment feed nor concentrate (Atkinson, 1991). Figure 9 shows the concentrate copper carrying rate versus the feed copper rate and indicates a maximum single stage recovery of 40% recovery, regardless of feed copper grade in the range tested. It is thought that this recovery limitation is due to kinetic effects, as the retreatment material is known to be slow-floating, and in applications where slow floating material is present, low single stage recoveries are often observed.

Using the carrying capacity relationship described earlier, the predicted carrying capacity for the retreatment material would be 6.0 g/min/cm^2 (d_{80} = 30 μm, S.G. = 4.0). For a 25% Cu concentrate grade, this translates to a concentrate copper carrying rate of 1.50 g(Cu)/min/cm^2 (see Figure 9). The results shown in Figure 9 cover a range of wash water ratios greater than 1.0, (all positive bias), with the higher carrying rates corresponding to just-positive bias.

JAMESON CELL OPERATING PARAMETERS

Prior information has concentrated on actual survey results for various streams. The following sections discuss some general operating characteristics of the Jameson cell.

SUPERFICIAL GAS RATE

For a stream which is not carrying capacity limited, recovery and concentrate carrying rate tend to increase with increasing superficial gas rate (J_g) (see Finch and Dobby, 1990). For a given stream and frother concentration, a maximum air rate (subsequently J_g) is reached, above which froth flooding occurs resulting in the loss of froth-pulp interface, a very dilute froth, and total loss of selectivity.

AIR-PULP RATIO

Jameson cells generally operate with an air-pulp ratio of 0.3 - 0.9. Experiences with large (J3000) Jameson cells at MIM's Hilton Mine indicate that operating at low air-pulp ratio does not appear to detract from metallurgical performance, but instead can have a stabilising effect producing a more uniform and finer bubble size. This can be explained by considering the superficial bubble surface rate (S_b,(cm^2 bubble area/s)/(cm^2 column area)) as defined by Finch and Dobby (1990), given by:

$$S_b = 6 J_g / d_b$$

where d_b is the bubble diameter. Superficial bubble surface rate (hence collection capability) can be maintained with reduced superficial gas rate (lower air rate) providing the bubble size decreases accordingly. Qualitative observations support this explanation, and future quantitative research will investigate this characteristic.

BUBBLE SIZE

Bubble size experiments both in-plant and at the University of Newcastle show that the Jameson cell produces an arithmetic mean bubble diameter of the order 300 - 600 µm, while the Sauter (volume-to-surface) mean diameter is of the order 360 - 950 µm, (Atkinson and Espinosa-Gomez, 1992). Laboratory experiments on an air-water (with frother) system show bubble size is not affected by changing air-water flow ratio in the range 0.65 - 1.35, but decreases with increasing frother addition and also decreases with increasing volumetric flowrate for a given downcomer size (Chatiar, 1992). As discussed earlier, plant observations would tend to indicate that bubble size does decrease at low air-pulp ratios (~0.3), though this remains to be quantified.

Bubble sizings measured from the top of conventional columns at Mount Isa are of the order of 2500 - 3000 µm Sauter mean size (Ghag, 1989), considerably coarser than that achieved in the Jameson cell.

WASH WATER

Figures 10 and 11 show the effect of wash water ratio on recovery and grade respectively. In this case, the material being treated was the retreatment feed. Recovery decreases significantly up to a wash water ratio of 1.0 (zero bias), and then decreases more gradually with increasingly positive bias. Figure 5 shows the much more critical effect of wash water when treating the rougher concentrate.

Grade improves dramatically up to zero bias, but no further improvement is observed with increasingly positive bias. These data, together with other Jameson cell operating data, confirm that the desirable wash water operating scenario is around zero bias, where grade is maximised. Higher wash water addition simply serves to decrease recovery. Thus precise on-line wash water bias measurement is highly desirable in order to maximise recovery at required grade.

Wash water ratio is a much more direct measure of the water required as opposed to that described by the term wash water bias (J_b, cm/s). The purpose of wash water is to replace the pulp water in the froth with clean water to displace entrained gangue, thus the *wash water ratio has direct meaning in terms of the proportion of pulp-water-in-froth replaced.* Also, wash water ratio is a dimensionless term, independent of cell size, which is not the case for bias rate, J_b.

FROTHER REQUIREMENT

On all streams tested to date, the frother addition to feed required for the Jameson cell is in the range 5 - 25 ppm. In cleaning applications, generally no additional frother is required. In some cases, excessive frother from the upstream (e.g. roughing) stage can lead to a reduction in maximum superficial gas rate (J_g, cm/s) which can be applied in the Jameson cell. This is because froth flooding is initiated at lower air rates due to the finer bubble size generated by the excess frother concentration (see Finch and Dobby, 1990, page 22).

When treating the autogenous mill discharge, it was found that the level of frother required was extremely sensitive to xanthate addition, with frother requirement inversely proportional to xanthate addition.

SUMMARY

The quality of the final concentrate is likely to be increased in the near future (through reduced iron sulphide content). Extra capacity is also needed, particularly in the cleaner and retreatment areas. Results based on plant trials indicate a further potential increase in copper recovery of at least 1% is achievable, if required extra flotation capacity is added (Griffin, 1992).

Jameson cells present an opportunity to achieve the desired capacity and recovery increase at lower capital cost than alternative methods. As a flash flotation device treating the autogenous mill discharge, the Jameson cell is capable of producing a final copper concentrate of 27% Cu at an average recovery of 50% in a single stage. Alternatively, treatment of the flotation feed after secondary grinding results in a single stage recovery of 35-45% at a grade of

32% Cu, following a Jameson cell pre-flotation stage in which naturally floating carbonaceous pyrite is removed as pre-flotation concentrate.

It is likely that by the end of 1992, Jameson cells will be installed at the Copper Concentrator to pre-float iron sulphides and then produce a single stage final copper concentrate.

The cleaning ability of Jameson cells has been shown to be at least equivalent to that of conventional columns, while similar carrying capacity limitations apply to those observed for industrial conventional columns.

The Jameson cell can handle very coarse material without hindering cell performance, as demonstrated when treating the autogenous mill discharge, where particles up to 10 mm are common.

For most streams tested, no additional frother is required above that used for conventional cells. In order to maximise recovery, precise wash water control is necessary matched to the concentrate production rate. This observation would apply to any deep column froth device.

ACKNOWLEDGMENTS

The cooperation of Mount Isa Mines Limited in permitting the publication of this work is gratefully acknowledged. Bruce Atkinson receives a PhD scholarship which is sponsored jointly by the Australian Government (Australian Post-graduate Research Award), MIM Holdings Limited, and The University of Newcastle Research Associates Pty. Ltd. (TUNRA). Their support for this research is sincerely appreciated.

REFERENCES

Atkinson, B.W. (1991), Report on Jameson cell trials on retreatment streams in the No. 4 Copper Concentrator of Mount Isa Mines Limited, University of Newcastle Internal Report, submitted to Mount Isa Mines Limited, September, 1991.

Atkinson, B.W. (1992), Jameson cell treatment of autogenous mill discharge, University of Newcastle Internal Report, submitted to Mount Isa Mines Limited, March-April 1992.

Atkinson, B.W. and Espinosa-Gomez, R. (1992), Jameson cell treatment of flotation feed and rougher concentrate - Copper Concentrator, Mount Isa Mines Limited, University of Newcastle Internal Report, submitted to Mount Isa Mines Limited, May 1992.

Chatiar, R. (1992) Report on test work on the operating parameters of the Jameson cell, Internal Report, Dept. Chemical Engineering, University of Newcastle, February, 1992.

Clayton, R., Jameson, G.J. and Manlapig, E.V. (1991), The development and application of the Jameson cell, *Minerals Engineering*, v4 n7-11,, 925-933.

Espinosa-Gomez, R., Finch, J.A., Yianatos, J.B. and Dobby, G.S. (1988) Technical note: flotation column carrying capacity: particle size and density effects, *Minerals Engineering*, v1 n1 77-79.

Espinosa-Gomez, R., Griffin, P.T., Johnson, N.W. and Barnes, C.D. (1991), Development of the copper concentrator flowsheet at Mount Isa Mines Limited, *Copper 91 - COBRE 91*, Volume II, 53-67.

Espinosa-Gomez, R. and Johnson, N.W. (1991), Technical experiences with conventional columns at Mount Isa Mines Limited, *Column '91*, Agar, C.E., Huls, B.J. and Hyma, D.B. (Eds.), Volume 2, Sudbury, Ontario, 2-6 June, 511-524.

Finch, J.A. and Dobby, G.S. (1990), *Column Flotation*, Pergamon Press.

Finch, J.A. and Dobby, G.S. (1991), Column flotation: a selected review. Part I, *International Journal of Mineral Processing*, v33, 343-354.

Ghag, S. (1989), Direct measurement of bubble size in the LGM columns and its effect on metallurgical performance, Internal Memorandum, Mount Isa Mines Limited, January 30.

Grano, S.R., Griffin, L. K., Johnson, N.W., Smart, R.St. C. and Ralston, J. (1991), Treatment of naturally hydrophobic gangue minerals at the Copper Concentrator of Mount Isa Mines Limited. *Fourth Mill Operators' Conference*, AusIMM, Burnie, Tasmania, 10-14 March.

Griffin, L.K. (1992), Half tonnage trials to estimate extra flotation capacity required in the Copper Concentrator for 1% extra copper recovery (pre - autogenous mills). Technical report, Mount Isa Mines Limited.

Kennedy, A. (1990), The Jameson flotation cell, *Mining Magazine*, Oct 1990, 281-285.

Leung, K., Lumsdaine, I. and Keran, V.P. (1992), Decision of Mount Isa Mines Limited to change to autogenous grinding, *SME 121st Annual Meeting and Exhibit*, Feb 24-27, Phoenix, Arizona.

Riches, N.J. (1992) Comparative testwork using the Jameson cell and conventional columns using copper cleaner feed, Internal Memorandum, Mount Isa Mines Limited, January, 1992.

Trahar, W.J. (1984) Pulp potential in sulphide flotation, *The Wark Symposium on Principles of Mineral Flotation*, Jones, Woodcock (Eds), The Australasian Institute of Mining and Metallurgy, 132.

LIST OF FIGURES

Table 1 - Jameson cell configurations.

Cell type	J100	J500	J1000
Cell diameter (mm)	107	525	1000
Options[1]		(400,300)	
Downcomer diameter (mm)	28.5	107	250
Orifice diameter (mm)[2]	3.1	12.5	34.7
Slurry flow (litres/min)	4.8	125	780
Superficial gas velocity (Jg, cm/s)	0.4 - 2.0	0.6 - 3.0	1.0 - 1.4

---

[1] When optional cell inserts are used to reduce cell cross-sectional area, the abbreviated description of the cell becomes, e.g. J400 in the case of 400mm insert in J500. The abbreviation JXXX generally refers to a Jameson cell of XXX mm (approximately) diameter.

[2] Other orifice sizes may be employed.

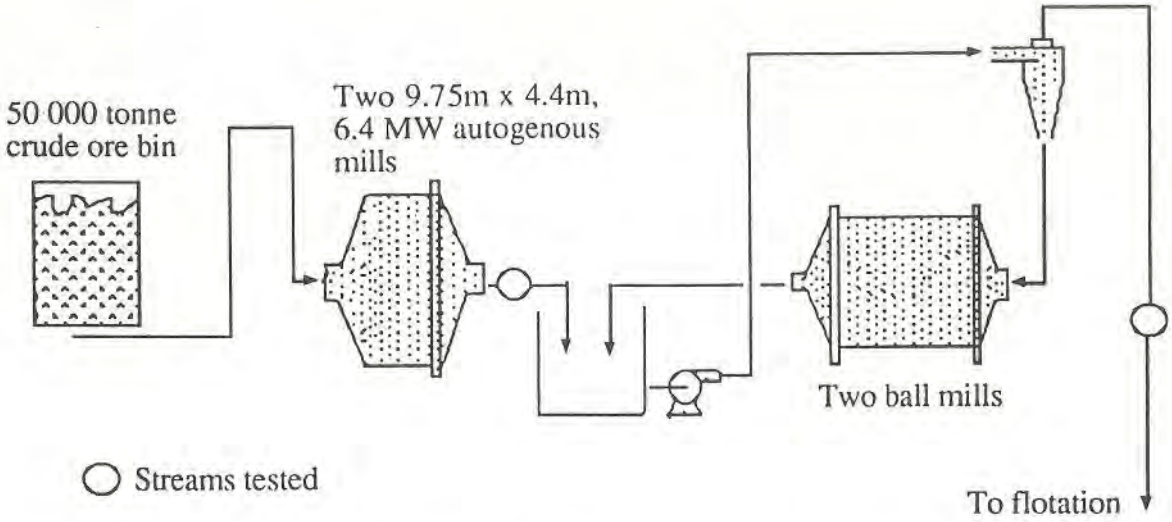


Figure 1 - Copper concentrator comminution circuit.

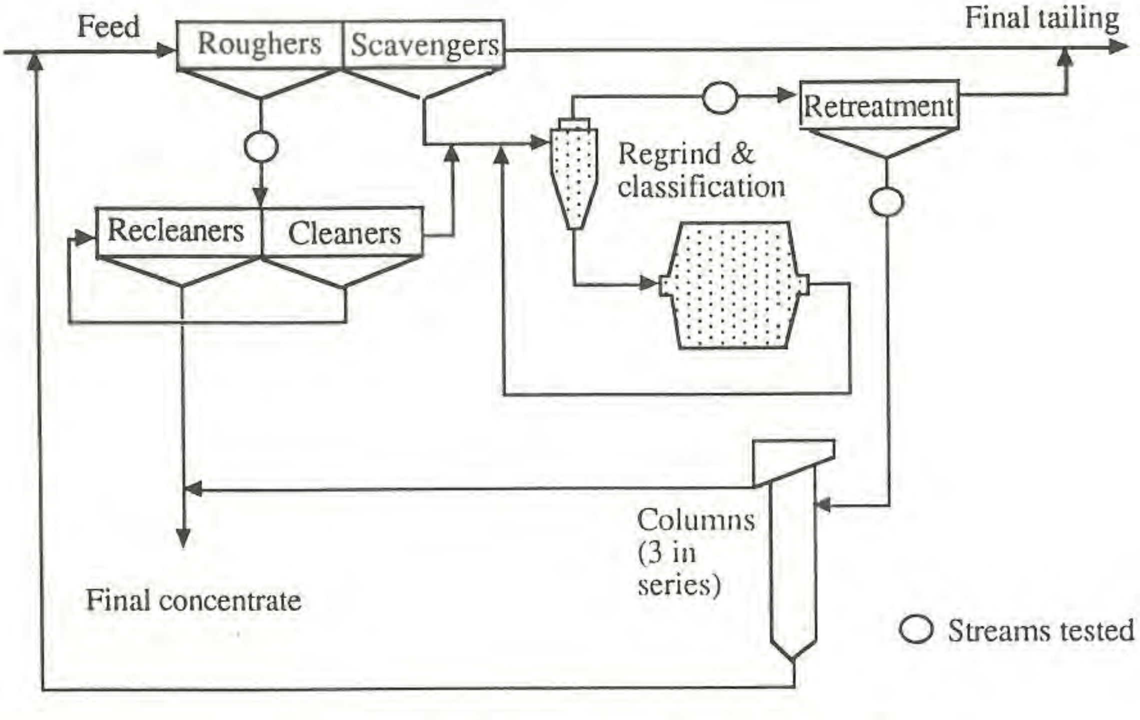


Figure 2 - Copper concentrator flotation circuit.

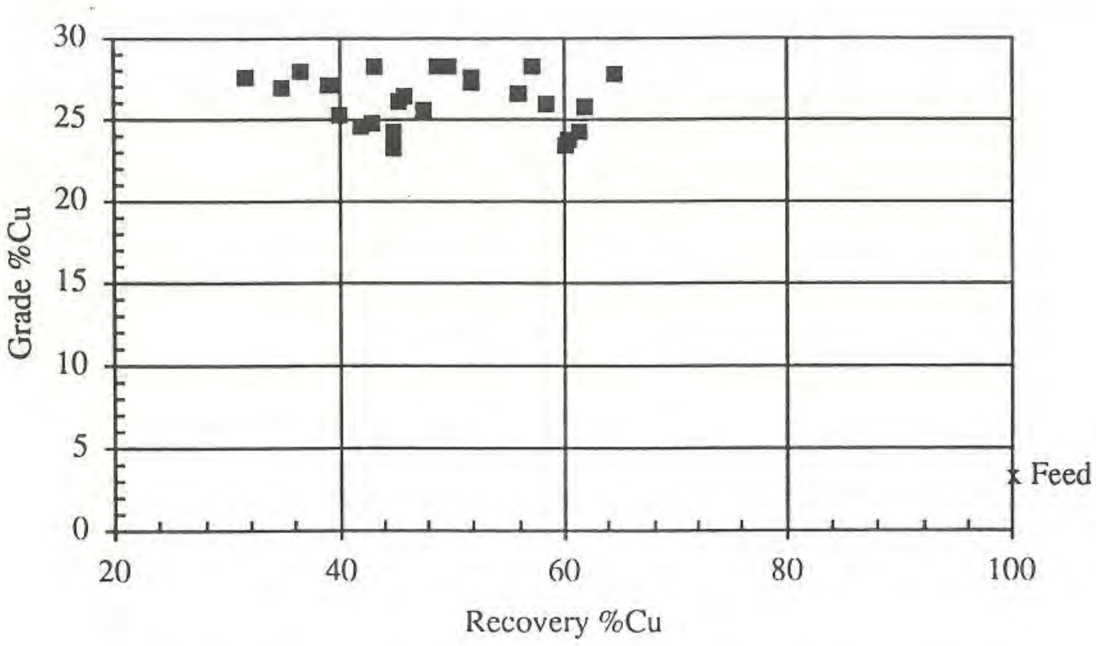


Figure 3 - Grade recovery results for single stage Jameson cell treating autogenous mill discharge.

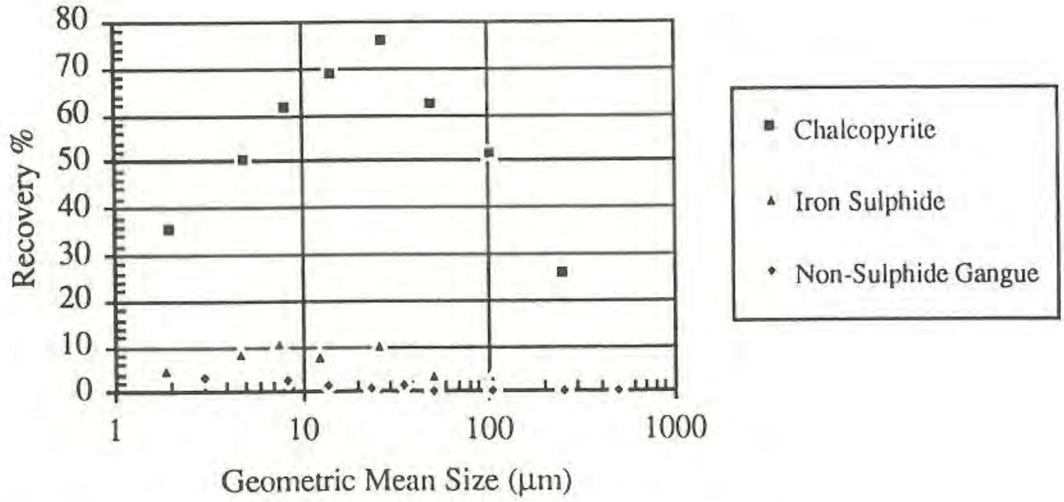


Figure 4 - Recovery by size - single stage Jameson cell treating autogenous mill discharge (concentrate grade = 27.5% Cu).

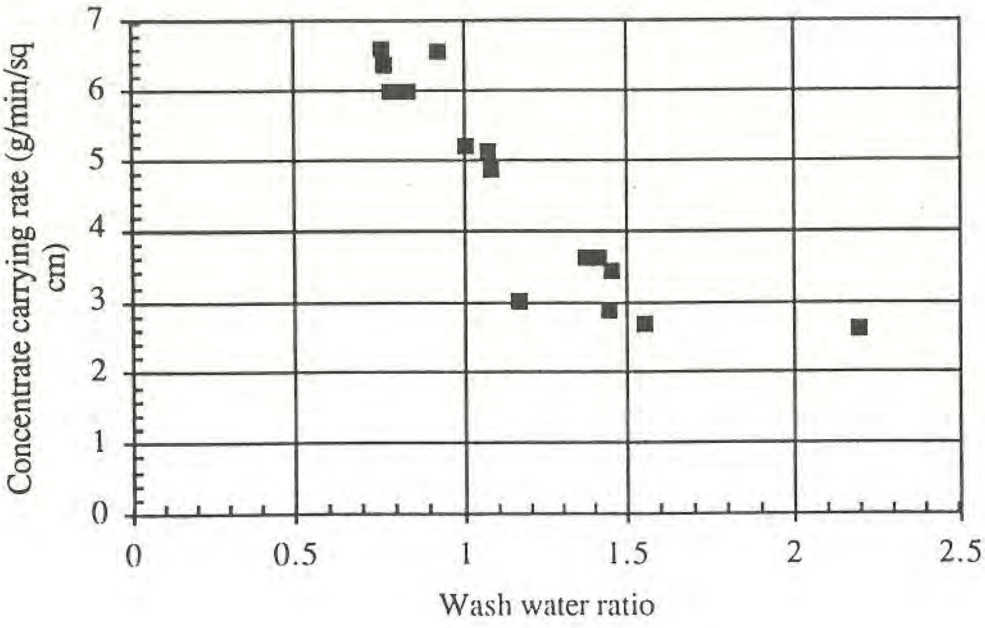


Figure 5 - Concentrate carrying rate versus wash water ratio.

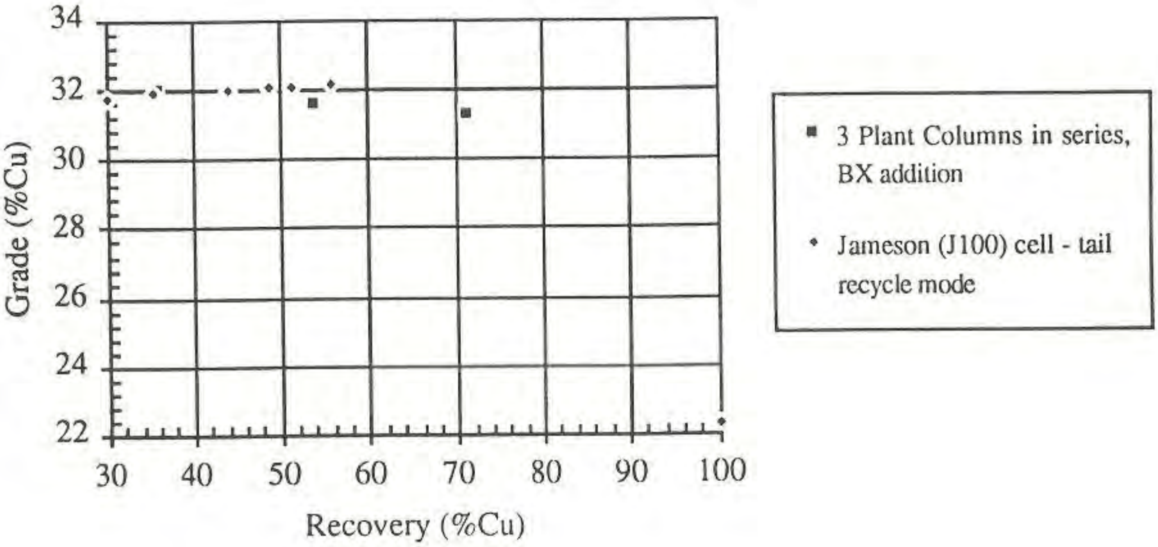


Figure 6 - Comparison of cleaning performance for Jameson cell versus plant columns treating rougher concentrate.

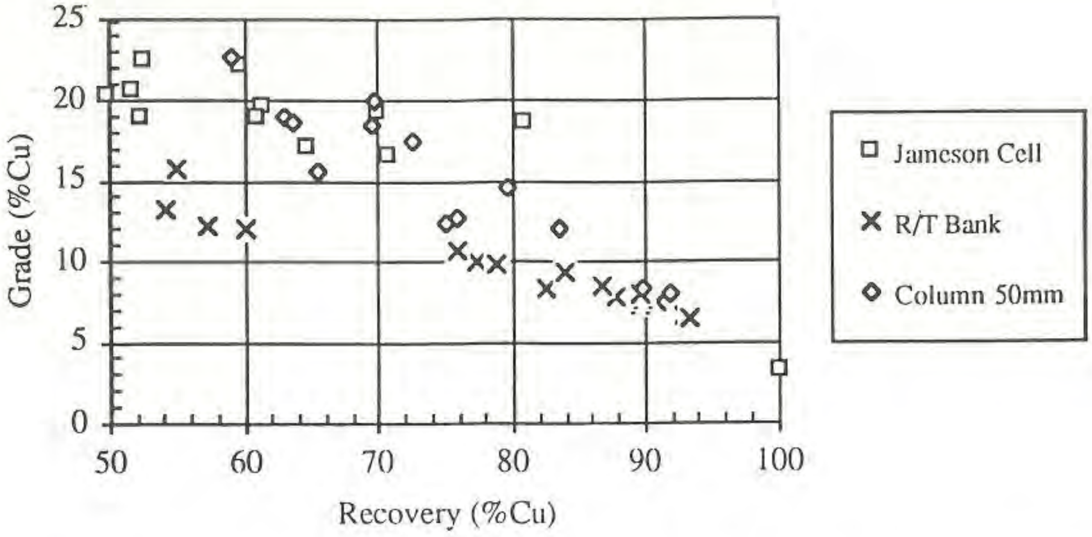


Figure 7 - Comparison of performance for Jameson cell versus 50mm pilot column and plant mechanical cells treating retreatment feed.

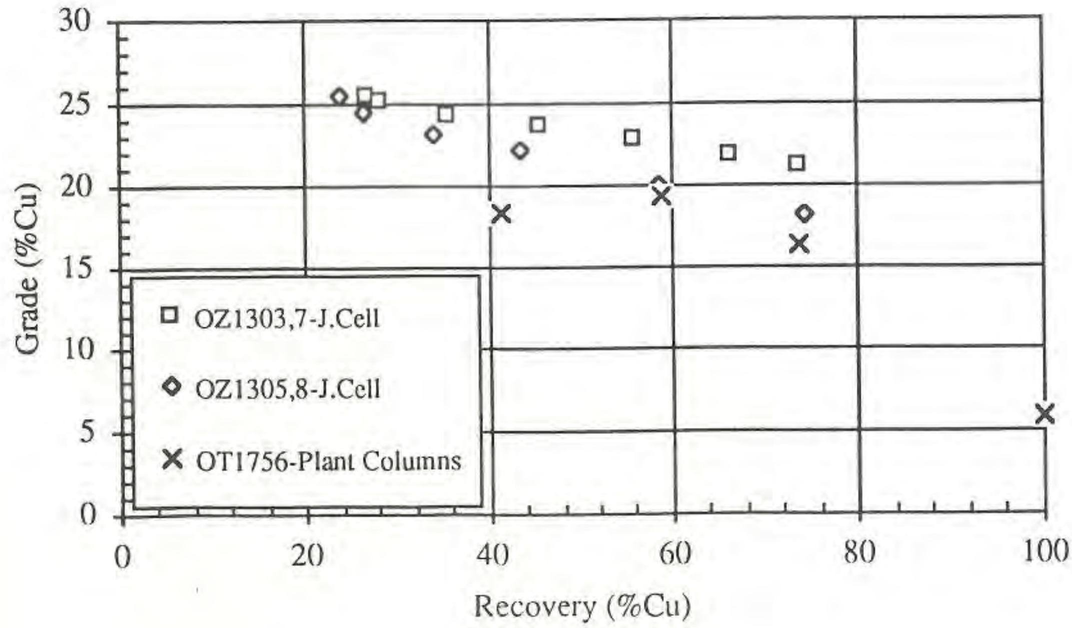


Figure 8 - Jameson cell versus plant columns treating retreatment concentrate. (Feed grade approximately equal for tests shown).

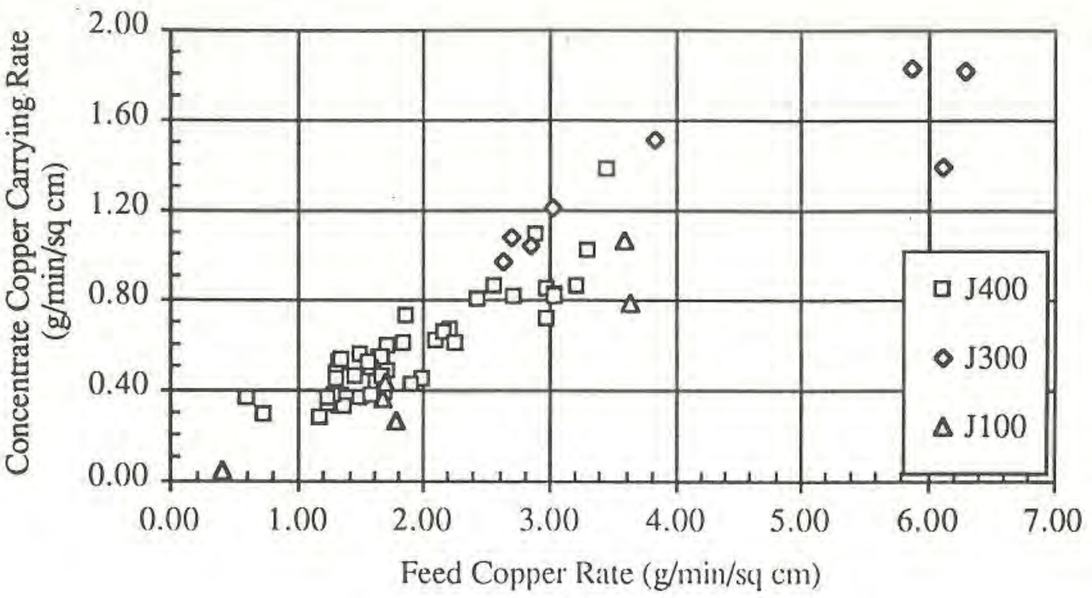


Figure 9 - Concentrate copper carrying rate versus feed copper rate for retreatment stream.

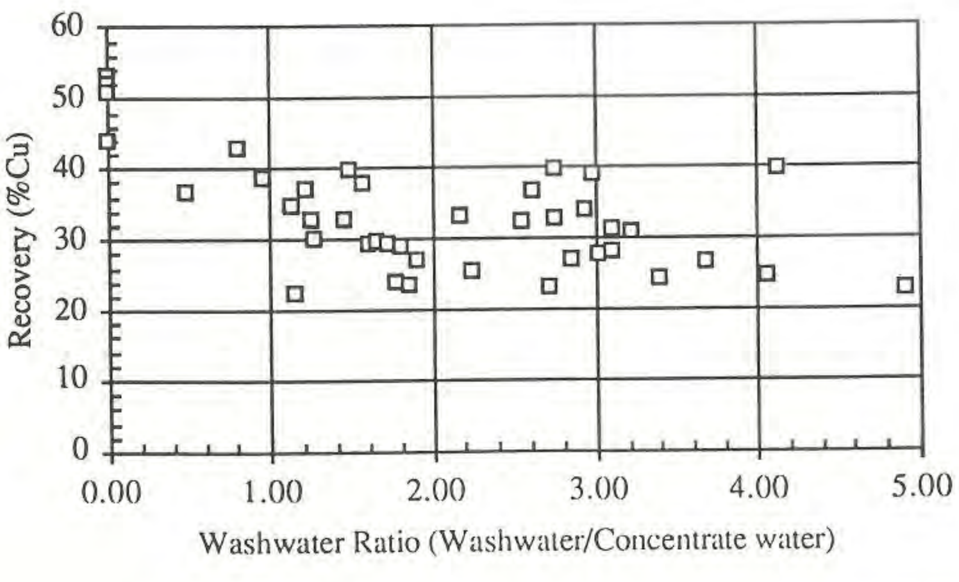


Figure 10 - Recovery versus wash water ratio (retreatment feed).

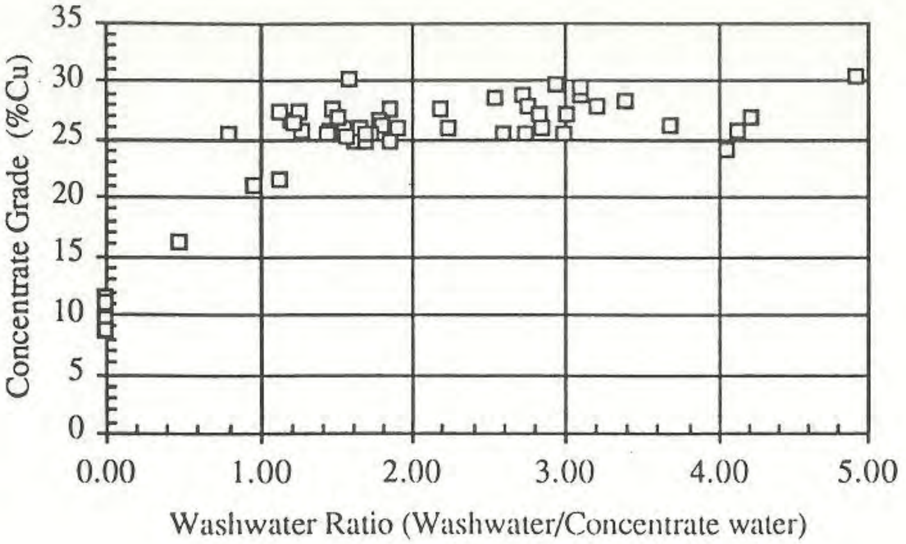


Figure 11 - Grade versus wash water ratio (retreatment feed).

Chemical Engineering Science, Vol. 47, No. 13/14, pp. 3265–3272, 1992.
Printed in Great Britain.

0009–2509/92 $5.00+0.00

PREDICTION OF THE BUBBLE SIZE GENERATED BY A PLUNGING LIQUID JET BUBBLE COLUMN

G.M. Evans, G.J. Jameson and B.W. Atkinson

Department of Chemical Engineering,
University of Newcastle, N.S.W., 2308, Australia

ABSTRACT

In this paper a model is presented to predict the maximum bubble size generated within the mixing zone at the top of a plunging liquid jet bubble column. The model uses a critical Weber number, where the energy dissipation rate per unit volume is derived from the theory of liquid-jet gas ejectors. The length of the mixing zone, and hence its volume, was determined experimentally from the vertical axial pressure profile along the wall of the column. The model was tested experimentally for a range of column and jet diameters, jet velocities, and liquid physical properties, and it was found that the measured maximum bubble diameter was in good agreement with the model predictions based on a critical Weber number of 1.2. It was also found that the bubble diameter distribution was fitted by a log-normal distribution, with a Sauter-mean-to-maximum-diameter ratio of 0.61 which is consistent with reported literature values.

KEYWORDS

Plunging liquid jet, Downflow bubble column, Bubble diameter, Weber number

INTRODUCTION

Figure 1 shows a schematic of a plunging liquid jet bubble column where both the gas and liquid are introduced at the top of the reactor. The liquid feed is in the form of a high velocity jet which passes through the headspace at the top of the column and entrains gas as it plunges into the liquid. A region of high recirculation and energy dissipation, known as the mixing zone, is generated at the top of the column where the entrained gas is broken into fine bubbles before being transported downward by the liquid flow, and discharged through the opening at the base of the column.

The size of bubbles produced within the mixing zone is determined by the forces acting on the bubble. In low viscosity liquids the bubbles are deformed by forces arising from liquid velocity fluctuations acting over distances of the order of the bubble diameter, d. The restoring force resisting the deformation of the bubble is due to surface tension acting at the gas-liquid interface. The ratio of these two forces is known as the Weber number, and can be written as $We = \rho\overline{u^2}d/\sigma$, where $\overline{u^2}$ is the average value of the squares of the velocity differences, and σ and ρ are surface tension and liquid density respectively.

The Weber number can be used to predict a maximum stable bubble diameter d_m, by assuming that a bubble will split once a critical Weber number, We_c, is reached, i.e.

$$We_c = \frac{\rho\overline{u^2}d_m}{\sigma} \tag{1}$$

where $\overline{u^2}$, in this case, is the average value of the squares of the velocity differences acting over a length scale equivalent to the maximum bubble diameter d_m.

In this paper the critical Weber number is related to the average energy dissipation rate per unit volume within the liquid to predict the maximum bubble diameter within the mixing zone at the top of a plunging liquid jet bubble column. The effects of jet and column diameter, jet velocity, and liquid physical properties on the resultant bubble size are measured and compared with the model predictions.

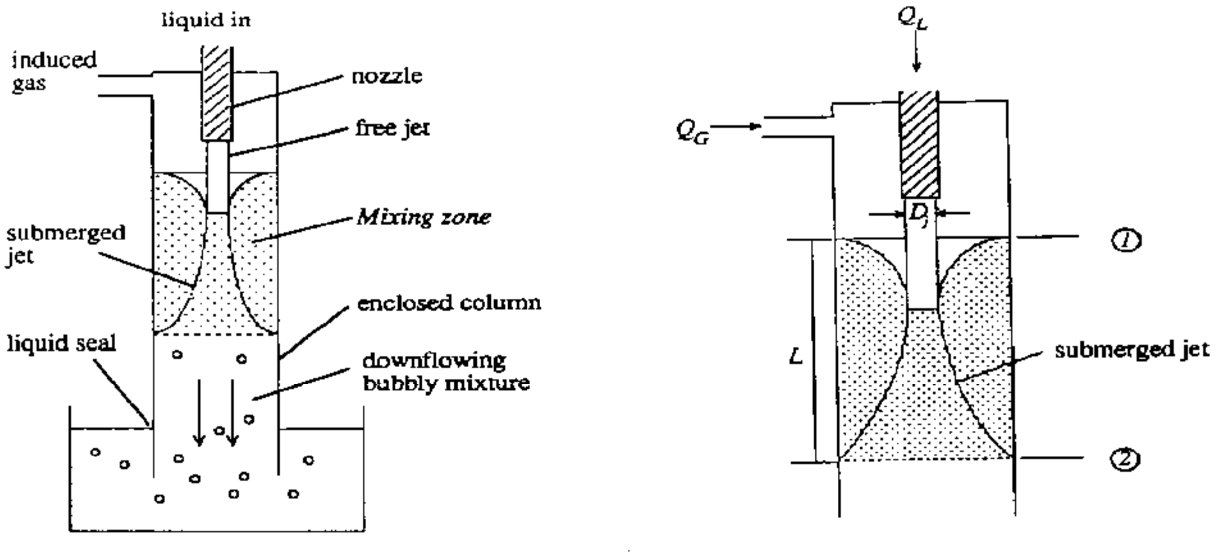


Fig. 1. Plunging liquid jet bubble column.

THEORETICAL

The maximum stable bubble diameter can be calculated from eq. (1) provided the liquid physical properties, critical Weber number, and average of the squares of the velocity fluctuations acting over length scales of the same order as the bubble diameter are known.

If the bubble is small compared to the turbulent macroscale but large compared to the microscale, the eddies responsible for breakup are isotropic and lie within the inertial subrange such that their kinetic energy is independent of viscosity and follows the Kolmogoroff energy distribution law, i.e.

$$\overline{u^2} = C_1\left(\frac{\varepsilon d}{\rho}\right)^{2/3} \quad , \tag{2}$$

where ε is the average energy dissipation rate per unit volume, and the constant $C_1 \approx 2.0$ according to Batchelor (1951). Substituting this expression into eq. (1), and rearranging, results in an expression for the maximum stable bubble diameter:

$$d_m = \left(\frac{We_c\sigma}{2}\right)^{3/5}(\rho)^{-1/5}(\varepsilon)^{-2/5} \quad . \tag{3}$$

Equation (3) has been derived assuming that the average energy dissipation rate per unit volume experienced by the bubble is uniform throughout the field. For flows where the energy dissipation rate per unit volume is not uniform throughout the field, such as in the mixing zone of a plunging liquid jet bubble column, the energy dissipation rate which determines the maximum bubble diameter for the entire flow field should be used.

The energy dissipation rate that determines the maximum bubble size in the mixing zone can be found by considering what happens to the entrained gas. Initially, all of the entrained gas enters the mixing zone at the plunge point of the liquid jet. The high energy dissipation in this region results in the generation of a number of very fine bubbles which are carried downward by the bulk liquid motion and leave the mixing zone. However, a number of larger bubbles are also generated, which recirculate within the mixing zone where they are eventually broken up and carried downward by the bulk liquid flow. Order of magnitude calculations based on the rise velocity of the bubble and the downward superficial liquid velocity, show that the residence time of the larger bubbles inside the mixing zone is of the order of one second, and this would allow a sufficient number of breakages to occur so that an equilibrium maximum stable bubble is reached. Furthermore, according to Hinze (1955), if the flow field is not too inhomogeneous, the powerful diffusive action of turbulence causes the average size of the largest bubble in the whole field to correspond to the average energy dissipation rate across this field.

The average energy dissipation rate per unit volume within the mixing zone of a plunging liquid jet bubble column can be calculated from an energy balance across the mixing zone volume. Following the analysis of Cunningham (1974) for a liquid jet gas pump, and realising that the mixing zone corresponds to the throat section of such a pump, the specific energy dissipation, e_s, within the mixing zone is given by

$$e_s = \frac{u_j^2}{2}\left[\begin{matrix} 1-2b-b^2(1+\gamma\lambda_1)(1+\lambda_2)^2+2b^2(1+\gamma\lambda_1)(1+\lambda_2) \\ -\dfrac{2\gamma\lambda_1 b^2}{1-b}+\gamma\lambda_1^3\dfrac{b^2}{(1-b)^2} \end{matrix}\right]-\frac{\lambda_1 P_1}{\rho}\ln\frac{P_2}{P_1} \quad , \tag{4}$$

where ρ is the liquid density; u_j is the jet velocity; γ is the gas/liquid density ratio, λ is the gas/liquid volumetric flow ratio; b is the jet/column area ratio; P is the absolute pressure; and the subscripts 1 and 2 refer to the inlet and outlet respectively of the mixing zone (see Fig. 1b).

Equation (4) can be simplified by assuming (i) that $\gamma = 0$, and (ii) that the change in pressure across the mixing zone is negligible, i.e. $P_1 = P_2$. This assumption is based on measurements by Evans (1990) who found that the work associated with compressing the gas within the mixing zone accounted for approximately one percent of the total work received by the gas. From the above-mentioned assumptions $\lambda_1 \approx \lambda_2$, which leads to

$$e_s \approx \frac{u_j^2}{2}\left[1-2b-b^2(1+\lambda_1)^2+2b^2(1+\lambda_1)\right] \quad . \tag{5}$$

The energy dissipation rate per unit volume is obtained by multiplying eq. (5) by the jet mass flow rate, $(\rho u_j \pi D_j^2/4)$, and dividing by the mixing zone volume, $(\pi D_c^2 L/4)$, where D_c is the column diameter and L is the length of the mixing zone (see Fig. 1b). This leads to

$$\varepsilon \approx \frac{\rho u_j^3}{2L}\left[b-2b^2-b^3(1+\lambda_1)^2+2b^3(1+\lambda_1)\right] \quad . \tag{6}$$

Equation (6) can be used in conjunction with eq. (3) to predict the maximum stable bubble diameter, providing the length of the mixing zone and the critical Weber number are known.

At present there is no theoretical model available that can predict the expansion of a submerged bubbly jet within a confined volume. Therefore, the length of the mixing zone is best determined experimentally.

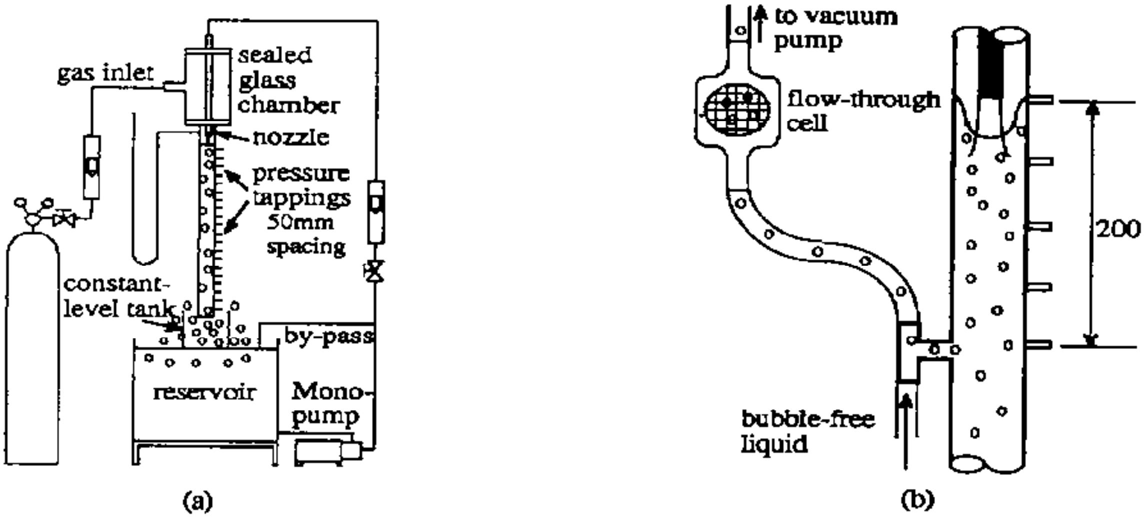
A critical Weber number can be chosen from a number of theoretical and numerical studies relating to the breakup of bubbles. Thus Hinze (1955) gives a value of 1.18 based on the breakup of a droplet or bubble in a viscous shear flow; Sevik and Park (1973) give a value of 1.24 based on the assumption that a bubble will oscillate violently and break up when the characteristic frequency of the turbulence is equal to one of the resonant frequencies of the bubble; Miksis *et al.* (1981) predicted a value of 3.23 for a bubble in a uniform flow; Lewis and Davidson (1982) found that a cylindrical bubble became unstable in an axisymmetric inviscid shear flow at a critical Weber number of 4.7; Ryskin and Leal (1984) predicted that the critical Weber number was a function of the Reynolds number and gave values in the range 0.95 – 2.76 for a bubble in a uni-axial extensional Newtonian flow.

Alternatively, bubble diameter measurements can be used to determine the appropriate critical Weber number for the mixing zone of a plunging liquid jet bubble column.

EXPERIMENTAL

Apparatus

A general layout of the experimental apparatus is given in Fig. 2(a). It consisted of a vertically positioned perspex column with its base extending 25mm below the surface of a constant level bath. Three columns were used in the study and their inside diameter measurements were 44, 74 and 95mm respectively. All three columns were identical in design, constructed from clear perspex, and consisted of a flanged section at the top which was bolted to the lower part of the chamber at the top of the bubble column. Below the flange was a straight length of clear perspex tube of length 1040mm. The top of the column was sealed from the atmosphere allowing air to pass into the column only through an inlet at the top. The air flow rate into the column was measured using a network of calibrated rotameters.



(a) (b)

Fig. 2. Experimental apparatus

The jet nozzle was fixed at the top of the column so that the liquid stream was directed along the vertical axis of the column and plunged into the constant level bath below. The overflow from the constant level bath was collected in a much larger vessel from which a Mono pump was used to pump liquid through a rotameter and back to the nozzle. The flow through the nozzle was coarsely adjusted by changing the pulley diameter on the drive shaft for the pump. Fine adjustments were made using the control valve on the by-pass line from the pump outlet and back into the reservoir.

Each nozzle was constructed of brass and consisted of a tapered machined entrance section which fitted neatly inside the nozzle delivery tube giving a smooth transition. The tapered entrance led to the throat of the nozzle which consisted of a straight length of precision-bore brass tubing. Pressure tappings were spaced 50mm apart along the axial length of the column. Another outlet situated 200mm below the top pressure tapping was used to draw the two-phase mixture through the optical flow-through cell as shown in Fig. 2(b). The optical cell was used to photograph the bubbles for determination of the bubble diameter distribution, and it consisted of two parallel glass windows spaced 2mm apart. A 2mm transparent grid was attached to the inside surface of the rear window, and this was used as a length calibration when measuring the bubble size.

Tap water was used as the standard liquid, and Teric 407, a nonionic ethylene oxide derivative produced by Imperial Chemical Industries, was used to control the surface tension of the water. Concentrations of 10, 150 and 1000ppm Teric 407 were used to vary surface tension over the range 65 – 48mN/m. Two aqueous sucrose solutions (16% and 28% by weight sucrose) were also used to vary the viscosity and density of the liquid; the surface tension of each of these solutions was adjusted to 62mN/m by the addition of Teric 407.

Procedure

The gas and liquid flow rates were set at the desired values, allowing the bubble column sufficient time to reach equilibrium. The column was assumed to be at equilibrium when the wall pressure profile became constant with respect to time. At this time the length, L, of the mixing zone was obtained from the wall pressure measurements (see Fig 3).

The two-phase mixture from the mixing zone was drawn through the optical cell with the aid of a jet-ejector pump located upstream of the optical flow-through cell (see Fig. 2b). The rate of bubbles passing between the parallel windows of the optical cell was adjusted in order to minimise the number of overlapping bubbles appearing in the photograph. This was achieved by diluting the two-phase mixture from the mixing zone with the clear liquid prior to it entering the optical cell. Preliminary experiments showed that the measured bubble size was not influenced by the rate at which the bubbly mixture was drawn through the optical cell, and no bubble coalescence was observed. The optical cell and tubing were thoroughly rinsed with 'Nonidet' detergent solution prior to each experiment to minimise bubble-surface interference.

The bubble photographs were taken using a Linhoff Polaroid camera fitted with a 156 mm telephoto lens and extension bellows. The shutter speed was set at 1/250 second with an aperture of f5.6. Backlighting to the flow-through cell was provided by an Olympus flash unit operating at a power ratio of 1/128. The bubble diameter was measured by projecting the negative of the bubble photograph onto a Summographics digitiser pad. The negatives also contained the original 2mm calibration scale, which was used to determine the true size of the bubbles. A minimum of 250 bubbles were measured to determine the diameter distribution for each experiment.

The length of the mixing zone and the resultant bubble diameter distributions were measured for a range of jet and column diameters, jet velocity, gas/liquid volumetric flow ratio, and liquid physical properties. A summary of the conditions for each experiment is given in Table 1.

Table 1: Summary of experimental conditions

Expt No.	D_c (m)	D_j (m)	u_j (m/s)	σ (N/m)	ρ (kg/m^3)	λ_1
1	0.044	0.0071	11.50	0.048[1]	997[2]	0.118
2	0.044	0.0071	11.53	0.054	998	0.116
3	0.044	0.0071	11.50	0.062	998	0.117
4	0.044	0.0048	11.50	0.047	997	0.129
5	0.044	0.0048	11.47	0.053	997	0.129
6	0.044	0.0048	11.50	0.065	997	0.127
7	0.044	0.0024	11.50	0.054	998	0.127
8	0.044	0.0048	11.50	0.064	1061	0.127
9	0.044	0.0048	11.50	0.065	1114	0.131
10	0.044	0.0048	11.47	0.065	999	0.125
11	0.074	0.0071	11.53	0.063	998	0.114
12	0.074	0.0048	15.00	0.065	999	0.126
13	0.074	0.0048	11.47	0.063	999	0.126
14	0.074	0.0048	11.47	0.063	999	0.296
15	0.074	0.0048	11.47	0.063	999	0.645
16	0.074	0.0048	7.80	0.064	999	0.125
17	0.074	0.0024	11.47	0.064	998	0.131
18	0.074	0.0071	11.50	0.061	998	0.114
19	0.074	0.0048	11.50	0.067	999	0.127
20	0.074	0.0024	11.50	0.062	998	0.130
21	0.095	0.0048	11.47	0.063	999	0.125
22	0.095	0.0048	11.50	0.062	998	0.124

[1] Measured using du Nuoy ring method.

[2] Values taken from Weast, R.C. and Astle, M.J. (1980). *CRC Handbook of Chemistry and Physics (60th ed.)*, CRC Press, Florida.

RESULTS AND DISCUSSION

Mixing Zone Length

The axial wall pressure profile was measured as a function of the distance z from the free surface inside the column. The general shape of the pressure profile was similar for each experiment, and a typical set of results is given in Fig. 3, where the absolute pressure P has been normalised with respect to the local atmospheric pressure P_0. In each case it can be seen that there is a distance beyond which the normalised pressure is a linear function of length. This trend indicates a region of relatively constant density and uniform flow, where the increase in pressure corresponds to an increase in the static pressure further down the column – at the base of the column the measured pressure is equal to the atmospheric pressure. In the region of uniform two-phase flow the wall frictional losses increase with increasing superficial liquid velocity as indicated by the increase in slope of the pressure gradient with increasing volumetric flow rate.

Each experimental curve has a section where the pressure profile is not a linear function of distance. This corresponds to a region of higher liquid velocity, due to recirculation of the liquid as a result of the rapid expansion of the confined jet. In the recirculation region the direction of the flow is reversed resulting in a negative dynamic pressure gradient which counterbalances the increase in the hydrostatic pressure, hence the sigmoidal shape of the curve. The mixing zone length was taken as the distance from the free surface inside the column where the pressure profile became linear. The results are listed in Table 2.

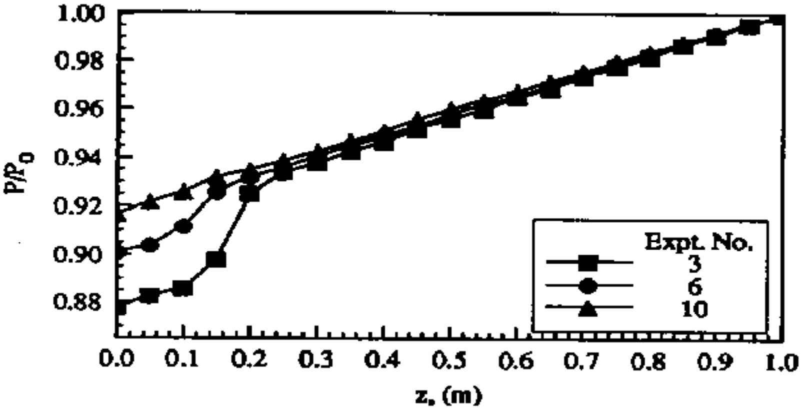


Fig. 3. Axial wall pressure profile

Bubble Diameter Distribution

The form of the bubble diameter distribution generated within the mixing zone was similar for each experiment. A typical cumulative frequency distribution plotted on a log-probability scale produced a linear relationship characteristic of a log-normal distribution. The log-normal distribution was skewed towards the smaller size ranges, possibly as a result of the preferential breakup of the larger bubbles over the more stable smaller bubbles. For many of the experiments the log-probability plots also showed two separate linear regions with different gradients, suggesting that the bubble size distribution was bi-modal. This could possibly be due to the different paths taken by the different size bubbles initially generated in the region close to the point of impact of the plunging liquid jet. The smaller bubbles spend only a short time in the mixing zone before being carried downward by the bulk liquid flow, whereas the larger bubbles are recirculated for a much longer time period before being broken up. The secondary breakup of the larger bubbles could possibly result in another distribution which was different to the initial distribution. If this is the case, then the bi-modal distribution represents (i) the initial distribution of fine bubbles generated by the plunging jet, and (ii) the distribution produced by the gradual breakup of the larger bubbles as they recirculate within the entire mixing zone volume.

Table 2: Summary of results

Expt No.	L (m)	ϵ (10^4kg/ms^3)	pred d_m (μm)	meas d_m (μm)	d_{vs} / d_m
1	0.175	21.40[1]	221[2]	222	0.58
2	0.170	22.20	233	280	0.56
3	0.175	21.40	257	387	0.63
4	0.150	11.50	279	248	0.62
5	0.148	11.60	299	222	0.61
6	0.148	11.70	337	406	0.67
7	0.148	11.70	337	380	0.59
8	0.197	9.31	358	318	0.59
9	0.221	8.67	372	294	0.62
10	0.072	6.06	392	412	0.57
11	0.246	5.63	443	441	0.60
12	0.249	5.53	455	584	0.60
13	0.170	3.61	529	543	0.60
14	0.215	2.86	581	523	0.65
15	0.261	2.35	628	698	0.67
16	0.074	2.62	608	718	0.60
17	0.113	1.37	780	594	0.63
18	0.249	5.52	438	519	0.61
19	0.166	3.73	527	614	0.62
20	0.115	1.34	778	716	0.68
21	0.157	2.38	625	531	0.63
22	0.153	2.46	611	611	0.59

[1] Calculated from eq. (6).

[2] Calculated from eq. (3).

The Sauter mean bubble diameter d_{vs} was calculated from a sample of N bubbles for each experiment, where d_{vs} is defined as

$$d_{vs} = \frac{\sum_{i=1}^{N} \left(n_i\, d_i^3\right)}{\sum_{i=1}^{N} \left(n_i\, d_i^2\right)} \quad . \tag{7}$$

The results are given in Table 2, and for comparison with other studies they have been normalised with respect to the measured maximum bubble diameter, d_m, which has been taken as the diameter that is greater than 99% of all the diameters in the cumulative frequency distribution. The average value for the bubble diameter ratio d_{vs}/d_m from all of the experiments, was found to be 0.61, with a standard deviation of 0.03. This result is in good agreement with published values, which are in the range 0.60 – 0.62 (Zhang *et al.*, 1985; Calabrese *et al.* 1986; Hesketh *et al.* 1987).

Critical Weber Number

The measured mixing zone length and maximum bubble diameter values listed in Table 2 were used, in conjunction with eqs. (3) and (6), to estimate the critical Weber number. It was found that a critical Weber number of 1.2 gave the best least-squares fit to the data. This value is in agreement with the critical Weber number calculated by Hinze (1955) for bubble breakup in viscous shear flow. Moreover, Sevik and Park (1973) provide some theoretical justification to this value based on the presumption that the bubbles within the mixing zone breakup due to oscillations of the bubble surface which occurs when the frequency of the turbulence corresponds to one of the resonant frequencies of the bubble. However, experimental verification of this breakup mechanism is required.

A comparison of the measured and predicted values using $We_c = 1.2$ is shown in Fig. 4. It can be seen that the predicted maximum bubble diameters were generally within 20% of the measured values.

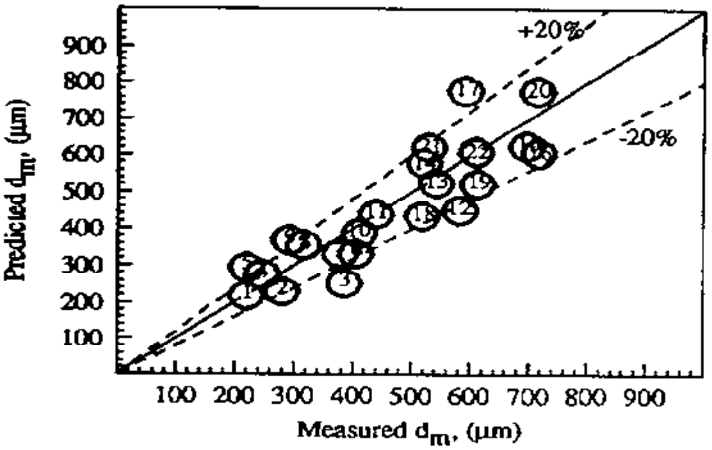


Fig. 4. Comparison between predicted and measured maximum bubble diameter

CONCLUSIONS

The maximum size of bubbles produced within the mixing zone of a plunging liquid jet has been measured and compared with predictions from a simple model. It was found that maximum stable bubble diameter may be predicted simply by determining the length of the mixing zone itself, and assuming the critical Weber number based on the average mechanical energy dissipation rate per unit volume, is equal to 1.2. The predicted maximum bubble diameter has been generally found to agree with measured values to within 20 percent over a wide range of column operating and liquid parameters. It was also found that the bubble size distribution was fitted by a log-probability relationship, and that the ratio of the Sauter mean diameter to the maximium measured bubble size was equal to 0.61, which is in agreement with the results from other experimental studies.

NOTATION

b	jet/column cross-sectional area ratio	
d	bubble diameter	m
d_m	maximum stable bubble diameter	m
d_{vs}	Sauter mean bubble diameter (defined by eq. (7))	m
D_c	column diameter	m
D_j	jet diameter	m
e_s	specific energy dissipation	m^2/s^2
L	length of mixing zone (see Fig. 1b)	m
Q_G	total volumetric gas flow rate rate	m^3/s
Q_L	liquid volumetric flow rate of jet	m^3/s
P	absolute pressure	kg/ms^2
P_0	atmospheric pressure	kg/ms^2
$\overline{u^2}$	average value of the squares of the velocity differences acting over a length scale d	m^2/s^2
u_j	jet velocity at nozzle	m/s
We	Weber number	
z	length down column measured from free surface	m
	Greek Symbols	
ρ	liquid density	kg/m^3
σ	surface tension	kg/s^2
ε	energy dissipation rate per unit volume	kg/ms^3
λ	gas/liquid volumetric flow ratio	
γ	gas/liquid density ratio	

REFERENCES

Batchelor, G. K. (1951). *Proc. Cambridge Phil. Soc.*, 41, 359.

Calabrese, R. V., T. P. K. Chang and P. T. Dang (1986). Drop breakup in turbulent stirred-tank contactors. Part I: Effect of dispersed-phase viscosity. *A.I.Ch.E. J.*, 32, 657-666.

Cunningham, R.G. (1974). Gas compression with the liquid jet pump. *J. Fluids Eng., Series I*, 3, 203-215.

Evans, G. M. (1990). A study of a plunging jet bubble column. Ph.D. Thesis, University of Newcastle, Australia.

Hesketh, R. P., T. W. Fraser Russell and A. W. Etchells (1987). Bubble size in horizontal pipelines. *A.I.Ch.E. J.*, 33, 663-667.

Hinze, J. O. (1955). Fundamentals of the hydrodynamic mechanism of splitting in dispersion processes. *A.I.Ch.E. J.*, 1, 289-295.

Lewis, D. A.and J. F. Davidson (1982). Bubble splitting in shear flow. *Trans. I. Chem. E.*, 60, 283-291.

Miksis, M., J. M. vanden-Broeck and J. B. Keller (1981). Axisymmetric bubble or drop in a uniform flow. *J. Fluid Mech.*, 108, 89-100.

Ryskin, G. and L. G. Leal (1984). Numerical solution of free-boundary problems in fluid mechanics. Part 3. Bubble deformation in an axisymmetric straining flow. *J. Fluid Mech.*, 148, 37-43.

Sevik, M. and S. H. Park (1973). The splitting of drops and bubbles by turbulent fluid flow. *J. Fluids Eng. Trans. ASME*, 95, 53-60

Zhang, S. H., S. C. Yu, Y. C. Zhou and Y. F. Su (1985). A model for liquid-liquid extraction column performance - the influence of drop size distribution on extraction efficiency. *Can. J. Chem. Eng.*, 63, 212-226.

ALBION PROCESS™

First Commercialisation of the Albion Process™ for Copper

Paul Voigt[1]*, David Littleford[2], Glenn Stieper[1], Mike Hourn[1]

1. *Glencore Technology, Australia*
2. *Sable Zinc Kabwe Limited, Zambia*

ABSTRACT

Sable Zinc Kabwe Limited ("**Sable**") is a base metal processing plant two km south of Kabwe Town in Central Province, Zambia. The region has a rich history in mining and minerals production, being a major lead and zinc producer from the 1900's up to the mid-1990's. In 2006, a copper plant was built at the Sable processing site to treat third party copper and cobalt ores and concentrates. The current flowsheet is based on whole ore leaching, solvent extraction and production of 8,000 tpa LME Grade A copper cathode. The plant is currently on care and maintenance and Glencore Plc ("**Glencore**") has taken the opportunity to convert the plant to an Albion Process™ plant to treat local concentrates unsuitable or uneconomic for smelting. The Albion Process™ plant comprises an M100 IsaMill™ operating in acidic conditions such that raffinate from the downstream process is recycled back to slurry the concentrate delivered to site thus maximising copper tenor in solution and reducing water consumption. The finely ground concentrate is then fed to existing stainless steel leach reactors converted to Albion Process™ duty by fitting HyperSparge™ supersonic oxygen injectors. The leach product is then directed to the existing solid / liquid separation equipment and onto copper and cobalt recovery circuits.

The plant was designed based on testwork of a nominal concentrate, however, a flexible approach to design allows the treatment of a wide range of concentrates. This positions the Albion Process™ plant at Sable a regional treatment facility for feed that is not economic to treat by conventional means.

INTRODUCTION

The Sable Zinc Kabwe Limited ("**Sable**") process plant is located 2km south of Kabwe town and 140km north of Lusaka in Central Province, Zambia. In 1902 rich deposits of zinc and lead were discovered with mines and processing plants operating in and around the town until mid-1994. In 2000, the Kabwe Mine tailings dams were sold to Sable , a newly established company, as part of a purchase deal for other plots and facilities. Up until 2004, Sable produced zinc cathodes in an SX/EW plant treating wash plant tailing material. In 2004, Sable was acquired by the South African mining company Metorex Limited ("**Metorex**") and in 2006 a copper/cobalt SX/EW plant was built to produce copper cathode from malachite ore supplied from the Democratic Republic of Congo. In 2011, Jinchuan Group International Resources Co. Ltd acquired Metorex and Sable was in turn sold to a subsidiary of Glencore Plc ("**Glencore**").

Glencore placed Sable on care and maintenance in November 2014 due to a decline in commodity prices coupled with the cash flow restrictions caused by the withholding of Value Added Tax refunds in relation to Rule 18(1)(b) by the Zambia Revenue Authority.

In 2015, Glencore evaluated the potential for the Albion Process™ to treat low grade concentrates that could not be treated by smelters due to a number of limitations including current blend fed to the copper smelter, low copper grade, low fuel content due to non-sulphide content, high cobalt content and high impurities such silica or alumina. Testwork was performed on a range of samples to make an assessment of process flexibility and, critically, copper and cobalt recovery.

Following this successful testwork, a decision was taken to convert the existing Sable copper/cobalt plant to an Albion Process™ plant capable of treating copper/cobalt concentrates containing refractory minerals such as chalcopyrite ($CuFeS_2$) and carrollite ($CuCo_2S_4$). The conversion will make use of existing infrastructure such as the stainless steel oxide leach tanks, the cobalt recovery circuit and the copper SX/EW plant. The plant nameplate is 16,000 tonnes per annum LME Grade A copper cathode. Cobalt will be recovered through the existing carbonate precipitation process.

The Albion Process™ is an atmospheric leaching process to oxidise refractory gold and base metals concentrate in a ferric/ferrous sulphate system for downstream recovery to final products. The technology is provided globally by Glencore Technology ("**GT**"). The process comprises two steps. The first is an ultra-fine grinding step performed in an IsaMill™ to grind the mineral concentrate down to a size where reaction products in the oxidation step to do not passivate the mineral surface and inhibit leaching. The second step is the oxidation of the ultra-fine ground concentrate through supersonic injection of oxygen with GT's HyperSparge™. The process runs at atmospheric pressure, autothermally and under mildly acidic conditions.

In the case of base metals such as copper, the solubilised metals are then recovered from solution by SX/EW. In the case of refractory gold the oxidised slurry is subjected to cyanidation or an equivalent process for gold and silver recovery.

The Albion Process™ has been commercialised in zinc and gold with five plants in operation and reported on extensively (Hourn & Turner, 2010; Hourn & Turner, 2012; Hourn et al., 2014; Voigt et al., 2015; Senshenko et al., 2016). The Albion Process™ installed at Sable will be the first commercialisation for leaching of copper and cobalt.

This paper describes the testwork development program and the process engineering performed to convert the Sable plant including the first application for an IsaMill™ operating in raffinate. The paper then describes the Sable Albion Process™ Plant and estimated project schedule.

DEVELOPMENT PROGRAM

In development programs for the Albion Process™, both technical and economic aspects of a potential project are evaluated to determine if it should be progressed to the next stage of development. Glencore Technology follows a development program in accordance with the Association for the Advancement of Cost Engineering International ("**AACEI**") Recommended Practice No. 18R-97 (Christensen and Dysert, 2005), which has been summarised in Table 1.

Table 1 AACEI Project Development Stages

Class of Study	Purpose	Accuracy LL (%)	Accuracy UL (%)	Project Completion (%)	Testwork
Class 5	Go / No-Go	-50 to -20	+30 to +100	0 - 2	Bench x 1
Class 4	Pre-Feasibility	-30 to -15	+20 to +50	1 - 15	Bench x 5
Class 3	Feasibility	-20 to -10	+10 to +30	10 - 40	Bench x 10
Class 2	Fixed Bid Prep	-15 to -5	+5 to +20	30 - 70	Pilot / Demo
Class 1	Execution	-10 to -3	+3 to +15	50 - 100	-

Each class of estimate represents a more detailed level of engineering study and a corresponding level of testwork to support the engineering study. In the case of the operation of pilot and demonstration plants, this is not required to perform detailed engineering and project execution since GT can obtain all design information from batch tests while still providing performance guarantees. Pilot and demonstration plant operation is normally completed to satisfy the risk management requirements of companies or financiers of projects.

At the Class 5 level of study, GT can make a reasonable evaluation of the suitability of the Albion Process™ for a project based on the approximate feed mineralogical and elemental composition, combined with our process experience without the need for testwork. Using a database of capital costs and operating information about the project, an estimate for capital and operating costs for a project can be quickly established. A subsequent Class 4 level study is then completed, supported by a preliminary testwork program, performed at a certified Albion Process™ testwork laboratory. The Class 4 level study provides a go/no-go evaluation of and the project and confirms key design information such as preliminary plant sizing, the extent of oxidation and metal recovery.

Since the Sable Albion Process™ Plant would treat a range of concentrates, four representative concentrates were obtained from the region and subjected to Albion Process™ testing at HRLTesting in Albion, Queensland.

Support Testwork

Four concentrate samples (A – D) were obtained from third party suppliers which reflected the typical quality available on the local market. The analysis of each concentrate is shown in Table 2.

Table 2 Analysis of Third Party Concentrates

Component	**A (%)**	**B (%)**	**C (%)**	**D (%)**
Cu	27.7	39.3	28.5	34.5
Fe	7.5	5.7	4.9	8.4
Co	0.04	0.03	3.2	3.7
S Total	5.6	8.5	12.4	18.7
S Sulphide	5.6	7.5	11.8	18.7
Carbonates	6.4	8.4	7.5	12.1
Malachite	23.7	31.1	12.1	10.2
Digenite	8.2	9.8	11.8	-
Chalcopyrite	9.3	15.2	16.2	14.2
Brochantite	7.9	15.1	8.8	-
Carrollite	-	-	8.5	9.8
Pyrite	-	-	-	4.5
Djurleite	-	-	-	13.1
Bornite	-	-	-	17.8
NS Gangue	50.9	28.8	42.6	30.4

The concentrates selected for testwork attracted significant smelter penalties due to either a relatively low fuel content or high insolubles content. Several of the concentrates also contained high cobalt grades which cannot be recovered in the smelting process. Low grade material has been previously tested and found to be well suited to treatment in Albion Process™ conditions (Voigt et al., 2016). The emphasis on this testwork was prove the performance and economics of these concentrates not suited to conventional flowsheets and determine the amenability of the sample to treatment in the Albion Process™.

A summary of the results for copper and cobalt extraction is shown in Table 3. Recoveries are expressed as recovery through the Albion Process™ leaching stage rather than overall plant recovery.

Table 3 Testwork Results

Component	**A**	**B**	**C**	**D**
Albion Cu Rec (%)*	98.9	99.2	99.5	99.3
Albion Co Rec (%)*	94.7	96.6	95.9	94.7
Sulphide Oxidation (%)	83	74	93	77
Leach Completed (h)	12	12	36	36
Test Residence time (h)	48	48	48	48

For all of the concentrates, the copper recoveries were at least equal to or better than would be achieved in a smelter even when considering the whole plant flowsheet. This is not unexpected since the Albion Process™ achieves over 99% copper recovery in samples where all the copper is present in primary sulphides rather than a mix of primary and secondary sulphides.

Importantly, the samples did not contain pyrite which is required to catalyse the decomposition of chalcopyrite in other atmospheric leaching processes.

Another important feature of the testwork was cobalt recovery. For concentrators where copper and cobalt cannot be economically separated or for concentrators that produce a cobalt by-product with high copper content, the Albion Process™ represents an option where both minerals can be recovered. This is expected to become increasingly important as cobalt demand continues to grow, placing upward pressure on prices.

The leach residence time was driven primarily by cobalt recovery, as the leaching kinetics for the carrolite minerals were slightly slower than for chalcopyrite. Copper recovery in samples C and D was maximised within 24 hours. Samples A and B contained less refractory materials and negligible cobalt.

Engineering Study

An Engineering Study was also prepared based on the testwork results. Unlike a greenfield engineering study estimate, the basis of the study was the conversion of existing facilities at the Sable site and to provide recommendations for new equipment that would need to be installed.

Important considerations from the testwork were incorporated in the design:

- Extent of sulphide oxidation – to size the oxygen plant and prepare the mass and energy balance
- Leach residence time – to allow sufficient leach capacity for different feed materials can be treated and ensure target copper and cobalt recoveries can be achieved
- Grind size and IsaMill™ size – the plant will be treat concentrates in the future which may have unknown grind characteristics and so the IsaMill™ needed to be flexible to respond to these variations
- Water balance – the IsaMill™ was configured to mill in raffinate to manage circuit wide water balance and copper tenor
- Acid soluble copper – it was found that some of the samples tested had up to 50% water soluble copper, and would need to be considered in the leaching profile and in the IsaMill materials of construction
- Variable feed – the plant will treat varying concentrates grades which could impact on plant performance if not scheduled and blended properly. Will need to blend the feeds to ensure that sufficient fuel is present in the concentrate
- Variable sulphide contents – the plant will treat concentrates containing varying levels of copper sulphides and oxides. Oxides do not require ultra-fine grinding, and so may at times bypass the IsaMill™. Oxides also contain minimal fuel value, which may impact the oxidative leach operating temperature.

A Block Flow Diagram for the Sable Albion Process™ plant and metal recovery circuits is shown in Figure 1.

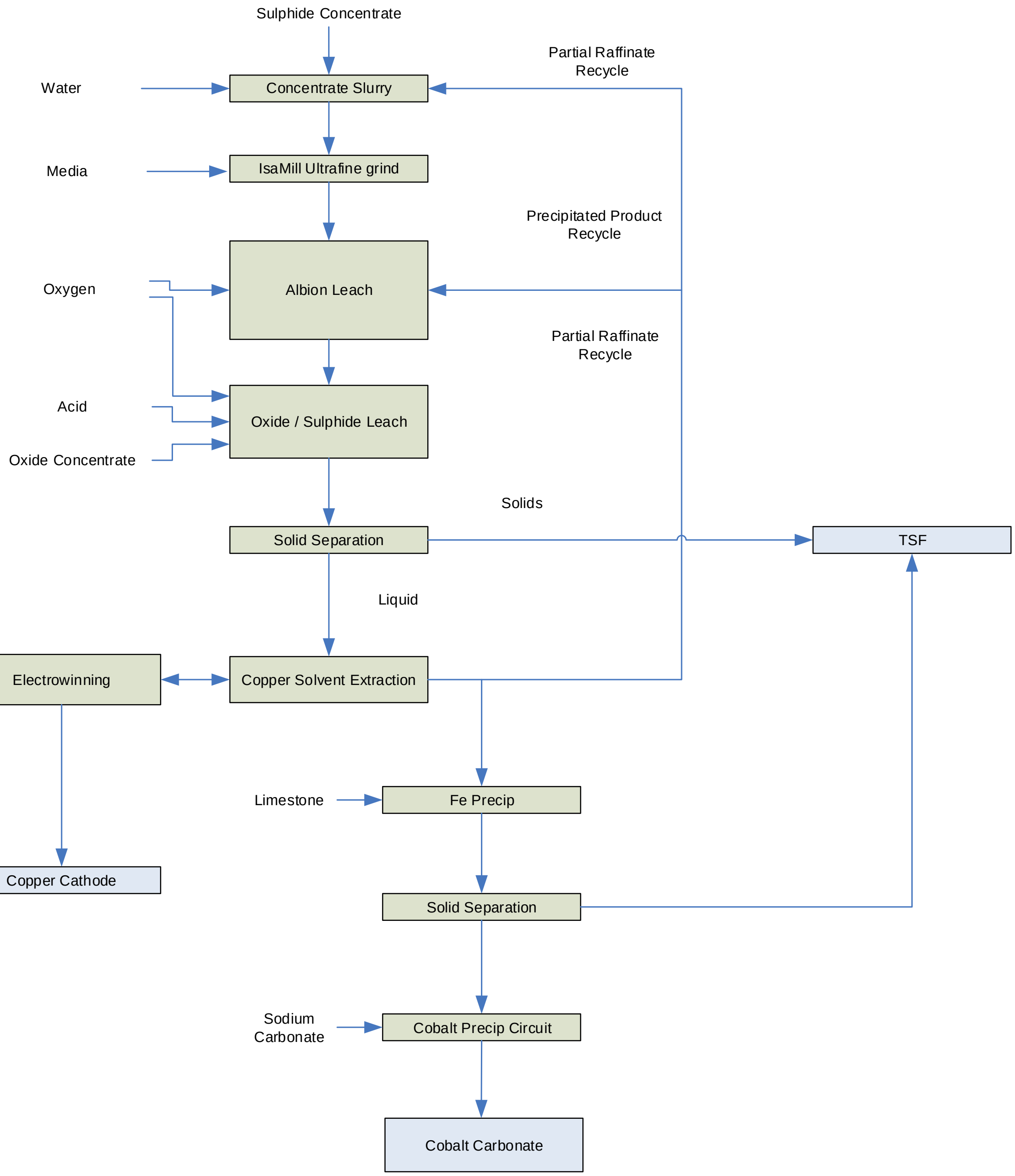


Figure 1 Sable Albion Process™ Plant Block Flow Diagram

Concentrate will be first slurried in raffinate and water in an agitated tank. The tank will be fed with via a hopper and conveyor on a batch wise basis depending on level. During the slurrying process some copper will leach out of the concentrate into solution.

The M100 IsaMill™ will be continuously fed and grinds the concentrate down to a p80 of 12 to 14µm, depending on the feed type. The IsaMill™ shell will be 316L stainless steel and all wetted parts materials selected to ensure performance in the acidic conditions.

The IsaMill™ will discharge directly into the first of five stainless steel oxidative leach reactors. The leach reactors will be covered with a roof and fitted with an exhaust stack. Dual bladed agitators shall be installed with appropriately sized agitators for gas dispersion and pumping duty. Oxygen shall be injected at the base of each leach reactor with the GT HyperSparge™ supersonic gas injector. Slurry will flow by gravity to downstream reactors with a launder system that will allow bypassing in the event of a maintenance event.

Depending on the leach profile, the relatively fast leaching oxides shall be added later down the leach train into leach tank 3 or 4 via a hopper and conveyor. Slurrying of the oxide material was avoided to manage the water balance. At this stage, supplementary acid may need to be added along with additional raffinate. Generally the acid demand of the system is met through in situ generation of acid from the chemical reactions and the acid credit from the raffinate.

After copper and cobalt is leached from the slurry into solution, a solid / liquid separation step produces solids that are disposed of in the Tailings Storage Facility. The copper rich solution is directed to the copper solvent extraction plant. The majority of the raffinate is recycled to the process. A bleed is taken from the raffinate to recover cobalt.

The bleed solution will first be neutralised with limestone to remove residual iron. The bleed solution, now containing mainly cobalt, will be directed to a carbonate based precipitation process for recovery of cobalt carbonate.

The rich electrolyte from solvent extraction is directed to the electrowinning plant for recovery of LME Grade A copper cathode.

From Figure 1, nearly all equipment required for the Albion Process™ is existing at the Sable facility. Three new stainless steel leach tanks, an M100 IsaMill™ and a small oxygen plant are the new equipment items required.

The main design criteria for the project are outlined in Table 4.

Table 4 Sable Albion Process™ Design Criteria

Criteria	Units	Nominal	Minimum	Design
Total Throughput (Sulph + Ox)	tpa	25,792	7,738	29,661
Total Throughput	tph	3.2	0.96	3.7
Copper Grade	%	25%	15%	35%
% of feed as sulphides	%	50%	50%	50%
Total Copper Units	tpa	6,448	1,161	10,381
Specific Energy for Fine Grinding	kWh/t	25	18	35
Oxidative Leach Residence Time - Sulphides	h	36	24	48
Terminal Cu Tenor	gpl	35	35	35
Raffinate Cu Tenor	gpl	4	4	4

A mass and energy balance was developed to assess a variety of concentrates based on the testwork results, vendor data and existing plant parameters at Sable. The mass and energy balance was used to develop a detailed Process Design Criteria set which was then used to build an equipment list and electrical load list.

The equipment list was then compared to existing equipment on site and a final equipment list and process flow drawings were developed.

A cost estimate was then prepared from the engineering study and input to an economic model showing the project significantly exceeded the investment criteria for a Glencore project.

SABLE ALBION PROCESS™ PLANT

The result of the Class 4 engineering study was then built upon to improve the cost estimation to a Class 3 level including improvement in the mass and energy balance to assess a wider range of feed materials and ensure equipment sizing was correct.

The Sable Albion Process™ Plant main equipment list and sizing is included in Table 5, highlighting where there is existing equipment for utilisation at the Sable facility.

Table 5 Main Equipment List

Equipment Name	Number	Size	Duty	Existing?
Concentrate Reslurry Tank	1	Vol - 10m^3	Slurry sulphide concentrate	Yes
IsaMill	1	M100	Fine grinding sulphide cons in acidic conditions	No
Oxidative Stainless Steel Leach Tanks	5	Vol - 100m^3	Leaching sulphide and oxide concentrate	Yes - 2 from 5
HyperSparge™ oxygen injectors	5	Nozzle - 4mm	Supersonic injection of oxygen	No
Oxygen Generator	1	10 tpd	Generate oxygen for leach process	No
Thickener	1	24m diameter	Thicken residue for filtration	Yes
Horizontal Belt Filter	2	Filt. Area - 44m^2	Solid/liquid separation	Yes
Tailings Storage Facility	1	-	Storage of solids	Yes
Copper SX/EW Plant	2	1500 tpm ea	Recovery of LME Grade A copper	Yes
Cobalt Precipitation Plant	1	50 tpm	Recovery of cobalt concentrate	Yes

Table 5 shows that the capital intensity of converting existing equipment to Albion Process™ duty is relatively low.

The existing plant had two of the five stainless steel oxidative leach tanks required. The existing stainless steel leach tanks were fabricated from duplex alloy 2304 which is suitable for the oxidative ferric leach duty and was the material used for the new tanks. For consistency the new leach tanks were made to the same dimensions and specifications as the existing ones.

In addition to the new tanks a small oxygen generator and an M100 IsaMill™ were also required. The M100 IsaMill™ was modified for the duty of operating in raffinate. This allows significant operational flexibility since the plant water balance is easier to manage, less total bleed is required and more acid from the raffinate can be utilised.

Oxygen mass transfer calculations suggest that the existing oxide leach agitators may not be optimum for efficient oxygen mass transfer. While the HyperSparge™ should contribute most of the power input required, contingency has been included to upgrade the agitators if this is required at a later time.

A high level schedule for the project including project development is shown in Figure 2.

	05/15	06/15		08/16	09/16	10/16	11/16	12/16	01/17	02/17	03/17	04/17	05/17	06/17	07/17
Initial testwork	🟩	🟩	//												
Process design			//	🟩	🟩										
Plant refurbishment							🟩	🟩	🟩	🟩					
IsaMill™ manufacture			//			🟩	🟩	🟩	🟩	🟩	🟩				
IsaMill™ trans to site			//									🟩	🟩		
Installation			//										🟩	🟩	
Commissioning			//												🟩

Figure 2 High Level Project Schedule

The schedule is driven by the time for manufacture of the IsaMill™ and transport to site. The other equipment lead times fall within this period. All refurbishment of existing equipment will be completed before the IsaMill™ is delivered to site.

In light of this, much of the equipment pre-commissioning will be completed before the IsaMill™ arrives at site. Based on the commissioning and ramp-up time of less than three months for a full Albion Process™ plant at GPM Gold in Armenia, the Sable Albion Process™ Plant should be commissioned within one month (Voigt et al., 2015).

CONCLUSIONS

The development Albion Process™ plant at Sable to treat locally available concentrates that are unsuitable for conventional treatment routes has been described. The economics of the project are very strong due to the utilisation and conversion of existing equipment. It is expected once the plant is established that it will provide an example of how existing equipment can be converted or with a small capital investment new equipment can be installed to treat refractory copper concentrates and integrated into an existing SX/EW flowsheet or plant site.

The Albion Process™ enters an exciting new period with the first commercial installation for leaching copper and cobalt concentrates becoming operational in early 2017. A further update will be provided upon commissioning.

REFERENCES

Christensen, P.C. and Dysert, L.R, 2005, Recommended Practice No. 18R-97 Cost Estimate Classification System – As Applied in Engineering, Procurement and Construction for the Process Industries, Association for the Advancement of Cost Engineering International (AACEI).

Hourn M. and Turner D.W., 2010, Albion Process for treatment of refractory ores, ALTA Conference Proceedings, Perth, Australia, June, 2010.

Hourn M. and Turner D.W., 2012, Commercialisation of the Albion Process, ALTA Conference Proceedings, Perth, Australia, June, 2012.

Hourn M., Voigt, P. and Turner D.W., 2014, Development of the Albion Process plant to treat refractory concentrates from the GPM Gold Project, Hydroprocess Conference, Vina del Mar, Chile, July, 2014.

Senshenko A.Y., Aksenov A.V., Vasiliev A.A. and Seredkin, Y.G., 2016, Technology for processing of refractory gold-containing concentrates based on ultrafine grinding and atmospheric oxidation, IMPC Conference Proceedings (yet to be released), Montreal, Canada, September, 2016.

Voigt, P., Hourn, M., Mallah, D. and Turner, W., 2015, Commissioning and ramp up of the Albion Process at the GPM Gold Project, World Gold Conference Proceedings, Johannesburg, South Africa, October 2015.

Voigt, P., Hourn, M. and Mallah, D., 2016, Treatment of low grade materials, MINEX Conference Proceedings, Moscow, Russian Federation, November 2016.

Ramp-up and long-term performance of the Albion Process™ plant at GeoProMining Gold Armenia

P Voigt[1], D Walker[2], O Kloiber-Deane[3] and A Tsvetkov[4]

1. Manager - Hydrometallurgy, Glencore Technology, Brisbane Queensland 4000. Email: paul.voigt@glencore.com.au
2. General Manager – Integrated Solutions, Core Resources, Brisbane Queensland 4010. Email: dwalker@coreresources.com.au
3. Principal Process Engineer, Core Resources, Brisbane Queensland 4010. Email: okloiber-deane@coreresources.com.au
4. CEO Deputy – Processing and Metallurgy, GeoProMining Gold LLC, Yerevan, Armenia. Email: atsvetkov@geopromining.com

ABSTRACT

The GPM Gold Operation has been employing an Albion Process™ plant at its Sotk (Zod) gold mine in Armenia since 2014. Long term data from the operation is now being published for the first time. The data presented includes concentrate throughput, as feed to the Albion Process™ plant (tonnes per day), sulphur grade in concentrate (% S), total sulphide sulphur oxidation in the Albion Process™ plant (% SOx) and gold recovery from Albion Process™ plant residues (%). The data provides a basis for evaluation of the effectiveness of the commissioning, ramp up and operation of the Albion Process™ plant. Notwithstanding the impact of a slower concentrator commissioning than expected, the plant demonstrates Series 1 behaviour on McNulty ramp up curves, indicating successful commissioning and fast ramp up. Interrogation of the data and operational experience indicates that this successful operation is a result of a process that is flexible, robust and stable, one that was designed well based on solid testwork results and pilot plant trials during the study phase undertaken by Glencore Technology and Core Resources, and where the local operators were provided with appropriate training in operation of the plant.

BACKGROUND

The GPM Gold Project

GeoProMining Gold LLC (GPM) owns and operates the Sotk (Zod) gold mine and Ararat processing facility in Armenia (“GPM Gold”). The open cut mine is located near the border with Azerbaijan, and gold bearing ore is transported via a state-owned rail link to the process plant at Ararat, near the Turkish border.

Reserves at the Sotk mine (as at an August 2011 estimate) were 14.2 Mt at 4.3 g/t Au. Indicated resources were estimated to contain 28 Mt of ore at 4.2 g/t Au, and inferred resources were estimated to contain 16 Mt at 4.2 g/t Au. The gold-bearing mineralogy is dominated by arsenopyrite and pyrite. Gold is preferentially associated with arsenopyrite and to a lesser extent pyrite. Gold occurs as free milling and finely dispersed through arsenical sulphides and tellurides. The geology of the deposit has been discussed separately (Voigt, Hourn, Mallah and Turner 2014).

The project has been in operation for several decades, originally mining weathered oxide ores overlaying sulphides. GPM acquired the project in 2007 and began initially treating gabbro and low sulphide ores. Testwork performed on samples obtained from exploration drilling into the underlying sulphide zones showed gold recoveries of around 20-30 per cent when treated through the conventional carbon in leach (CIL) process. Further testwork was performed to evaluate oxidation of the sulphide material prior to cyanidation. Oxidation of arsenopyrite and pyrite before cyanidation of the resulting residue was successful, with gold recoveries lifting from 20-30 per cent to over 90 per cent.

GPM evaluated four oxidation technologies, namely roasting, pressure oxidation (POX), biological oxidation (BiOX) and the Albion Process™, an atmospheric oxidative leaching process. An environmental-technical-economic evaluation was performed, with the Albion Process™ prevailing as the most suitable technology. The main driver was a lower capital cost and reduced implementation time.

Development testwork for the project began in 2009. Four sulphide ore samples (600 kg each) from across the ore body were tested, culminating in a continuous pilot plant oxidative leaching program. The final flowsheet consisted of milling and sulphide flotation to generate a sulphide concentrate, which was then treated through an Albion Process™ oxidative leach plant. Residue from the Albion Process™ plant is combined with flotation tails and fed to the existing CIL circuit.

Nameplate design for the plant is 100,000 tpa concentrate, generating on average ~100,000 ozpa gold as doré. The Albion Process™ plant was delivered to GPM as a lump sum technology package linked to performance guarantees by Glencore Technology. Mechanical design was completed in December 2012. Construction of the operation was completed in April 2014, and commissioning completed in July 2014.

The Albion Process™

The Albion Process™ is an established atmospheric oxidation technology for treatment of refractory sulphide concentrates. It is currently installed in six operations globally, with two Albion Process™ plants applied to the leaching of gold hosted with refractory sulphides. The chemistry of the Albion Process™ has been widely reported (Voigt, Mallah and Hourn, 2017).

In applications for gold, the Albion Process™ consists of fine grinding of the sulphide concentrate, followed by oxidation in tanks operating at atmospheric pressure with gaseous oxygen injected at supersonic velocities to maximise oxygen mass transfer and facilitate the oxidation reactions. Oxidised residue from the Albion Process™ is then treated with a standard cyanide leach circuit for gold recovery.

The oxidation reaction is carried out under a mildly acidic pH (5.5), and is maintained by the addition of limestone. The reactors are auto-thermal and typically operate in the range 85-98°C. Oxygen utilisations of 80-90 per cent are typical. The Albion Process™ for gold does not solubilise any base metals and does not produce any elemental sulphur or jarosites, meaning discharge slurry does not require washing and cyanide consumption remains low in the leaching circuit. Residues typically have good filtration characteristics, and materials of construction requirements are typically lower than for acidic POX or BiOX circuits.

THE GPM ALBION PROCESS™ PLANT FLOWSHEET DEVELOPMENT AND DESIGN

The overall GPM plant flowsheet is shown in Figure 1.

Before the implementation of Albion Process™ the GPM plant comprised the existing milling, CIL and gold recovery plants to treat ore directly from the Sotk mine. A flotation plant was existing at the site but was no longer used. The implementation of the Albion Process™ required the refurbishment of the flotation plant and installation of the new Albion Process™ plant equipment.

The flowsheet components for an Albion Process™ plant treating refractory concentrate for gold recovery comprises more or less the same unit operations between different projects; fine grinding followed by oxidative leaching and in some installations thickening. The GPM plant comprises an M3,000 IsaMill™ fine grinding plant and nine (9) 270 m^3 Albion Process™ leach reactors followed by a 10m high rate thickener where leach residue is directed to the CIL at a target 42 per cent solids. The process plant has been described in detail elsewhere (Voigt, Hourn, Mallah and Turner, 2015).

Key design criteria for the GPM project are provided in Table 1.

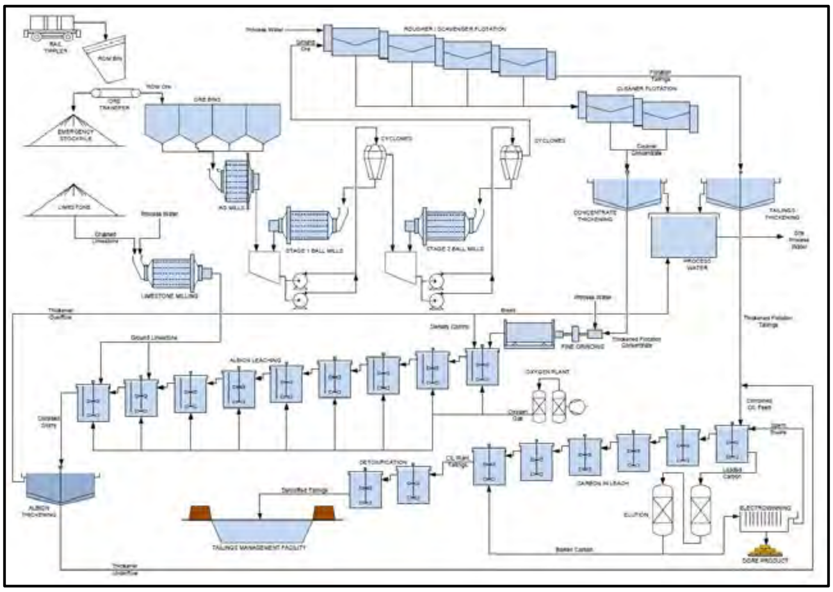


Figure 1: GPM flowsheet.

GPM OPERATIONAL PROCESS DATA

This paper presents process data from 2014 to 2017, supplied by GPM, with a view to evaluating the ramp up and the performance of the Albion Process™. This set reflects the largest series of operational data from an Albion Process™ plant made publicly available to date. It demonstrates the performance of the Albion Process™ under a number of conditions, some of which are normally adverse for hydrometallurgical sulphide processing circuits.

Key process data has been collected over the duration of plant operation and analysed. Key process data includes:

1. Concentrate throughput, as feed to the Albion Process™ plant (tonnes per day);
2. Sulphur grade in concentrate (% S^{2-});
3. Total Sulphide sulphur oxidation in the Albion Process™ plant (%); and
4. Gold recovery from Albion Process™ plant residues (%).

Oxygen utilisation data is not collected by site and has been reported elsewhere based on plant surveys (Voigt, Mallah and Hourn, 2017). Since the power supply contract is arranged on a take-or-pay basis, excess oxygen that is generated is directed to the CIL.

Monthly data has been compiled in Figures 3 to 6. Charts of plant data are shown based on monthly averages, with bars showing one standard deviation calculated from the individual daily data for each month. The data presented is based on monthly averages of daily data, with no data conditioning – i.e. downtime for plant maintenance etc. has not been removed from the data prior to reporting average monthly figures.

Interrogation of the detailed process data reported along with operational experience indicate that, in addition to achieving the main design parameters of throughput and recovery, the project has enjoyed success because of several key factors:

1. Fast ramp up – the Albion Process™ was commissioned and ramped up within three months of construction;
2. Flexible, robust and stable – the Albion Process™ has been able to effectively treat a feed concentrate with highly variable throughput and quality;
3. Designed right, the first time – the project was designed and executed based on solid testwork and engineering design so that the commercial plant performed better than laboratory tests (Voigt and Walker, 2018); and
4. Technology transfer – the technical know-how about the Albion Process™ was effectively transferred to the client team so they knew how to operate and maintain the plant.

These factors are discussed in greater detail throughout this paper.

Parameter	Units	Value
Albion Nominal Feedrate	tph	12.1
	tpd	290
Albion Design Feedrate	tph	13.5
	tpd	334
Flotation Concentrate S^{2-} Concentration	%	17.6
Degree of S^{2-} Oxidation	%	76
Design Oxygen Utilization	%	80
Design Gold Flotation Recovery	%	88
Design CIL Gold Recovery (Albion residue)	%	92
Design CIL Gold Recovery (flotation tails)	%	40
Design Overall Gold Recovery	%	85.8
Annual gold production	oz	100,000
Annual ore throughput	tpa	1,000,000

Table 1: Key process design criteria.

Fast ramp up

The GPM Albion Process™ Plant commissioning was completed in July 2014. This was approximately four months ahead of the completion of the concentrator commissioning, with the resulting reduction in concentrate supply affecting the Albion Process™ plant ramp up rate. Furthermore one of the two oxygen plant air blowers failed soon after commissioning, however the Albion Process™ plant was not oxygen limited at this time and was able to deal with the concentrate production rate.

The GPM plant gold production data was plotted on a McNulty curve to assess ramp-up performance as shown in Figure 2.

Figure 2 shows that the GPM plant exhibited ramp up behaviour between Series 1 and Series 2 during the first two years with Series 1 performance exceeded after 2 years. Series 1 is indicative of a successful implementation of project, typically representing a higher degree of “care exercised during the project development phase.” (McNulty, 2014). The basis for analysis is the gold production which was taken as the principal measurement of the success of the overall project.

The early stages of the project demonstrated Series 2 behaviour, in part constrained by the slow concentrator commissioning. However, the ramp up characteristics of the project accelerated rapidly, and the plant exceeded Series 1 behaviour within the first 3 years of operation.

This was a major achievement for the technology. McNulty and others have observed that "projects exhibiting Series 1 behaviour generally relied on mature technology" (McNulty 2014) and that the presence of a licensed technology with "few or no predecessors" (McNulty 2014), was one of several risk factors that might contribute to a plant lowering ramp up performance from Series 2 through to 4. The experience at the GPM plant is contrary to this observation, where ramp up constraints in the concentrator were more significant than in the Albion Process™ circuit.

Concentrator ramp up

During 2015, concentrator throughput and recovery were poor due to ramp up problems in the concentrator and treatment of transitional ores with poor flotation response. The second half of 2015 onwards saw primary sulphide ore presented to the flotation plant with a corresponding increase in recovery, grade and throughput. Concentrator performance achieved a significant step-change in performance around the middle of 2016 following a de-bottlenecking project.

Figure 3 shows that concentrator performance has improved significantly since commissioning in terms of absolute grade, recovery and throughput as well as stability. The flotation gold recovery and concentrate production remains below nominal and design levels (88 per cent, 282 tpd and 324 tpd respectively) however the gold grade and sulphur grades are higher than design resulting in design gold units processed through the plant. Gold units that tend not to float are typically non-sulphide and more amenable to direct cyanidation, and these flotation tails are passed direct to the CIL circuit.

Albion Process™ramp up

During the corresponding period since commissioning, the performance of the Albion Process™ has been above design and stable, oxidising any feed that is directed to the process despite significant variation of feed quality and quantity, as discussed in the following section. The Albion Process™ plant is limited by concentrate feed and tends to operate at design rates over an hourly basis until feed is depleted. This is to maintain efficient and steady operation of the IsaMill™.

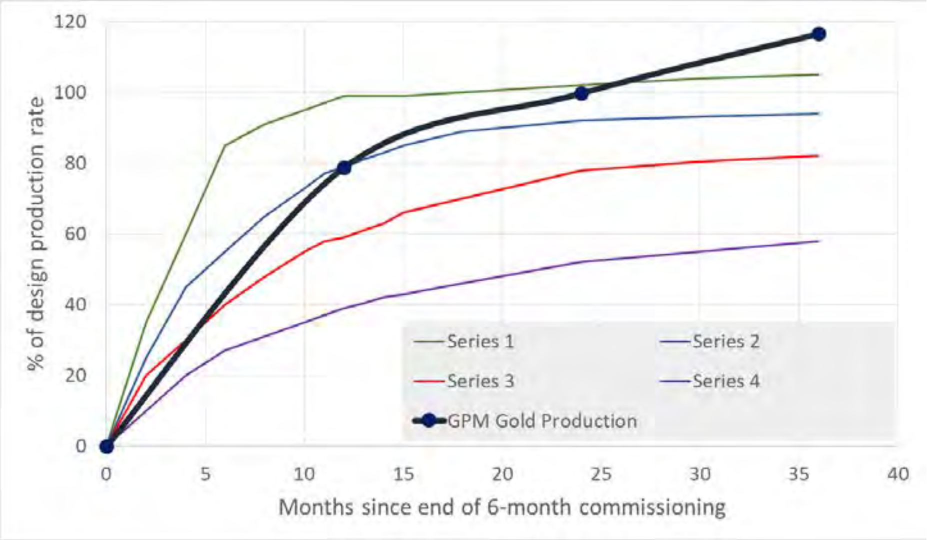


Figure 2 – McNulty curve for GPM process plant.

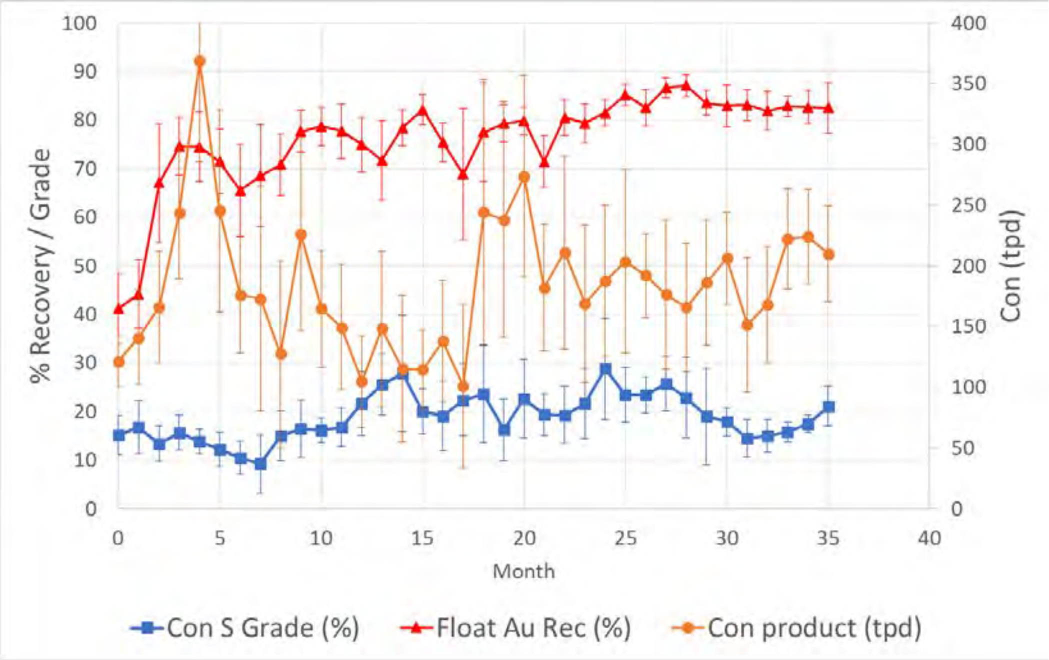


Figure 3: Concentrator performance.

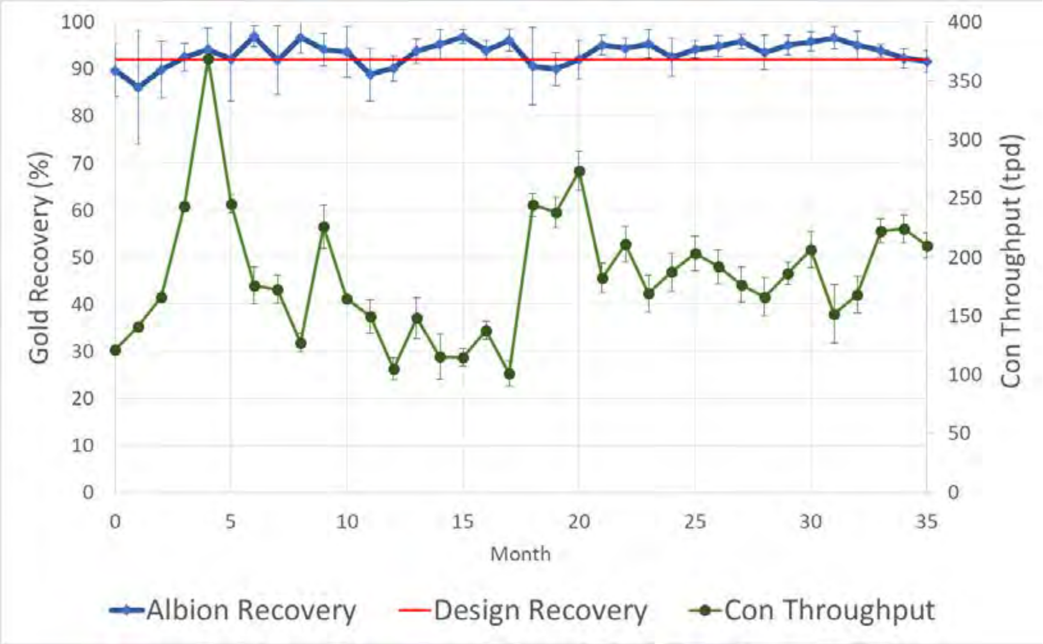


Figure 4 - Gold recovery versus throughput.

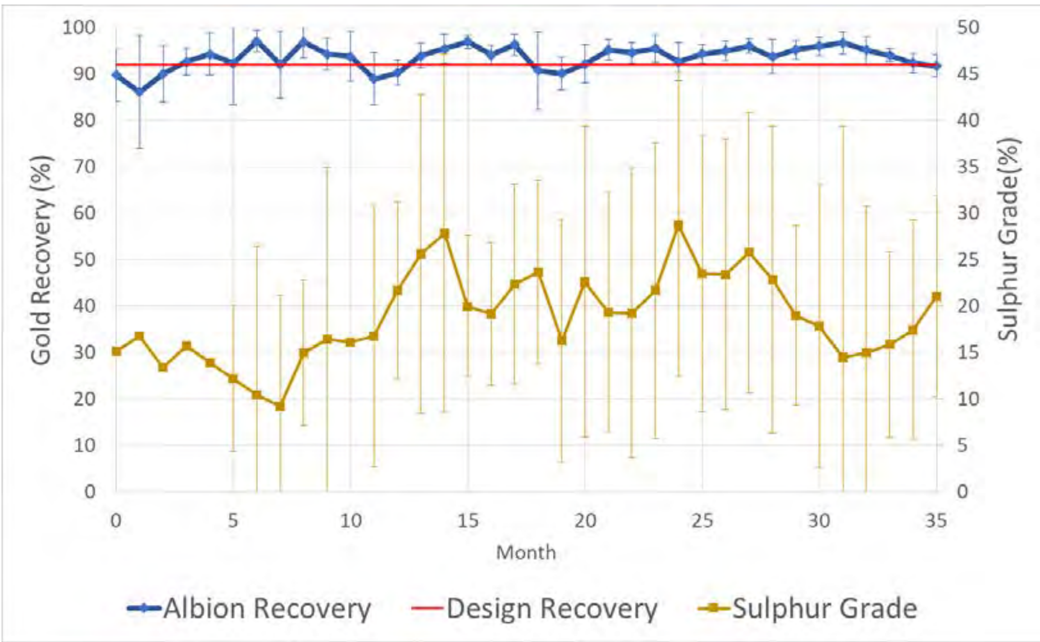


Figure 5 - Gold recovery versus concentrate sulfur grade.

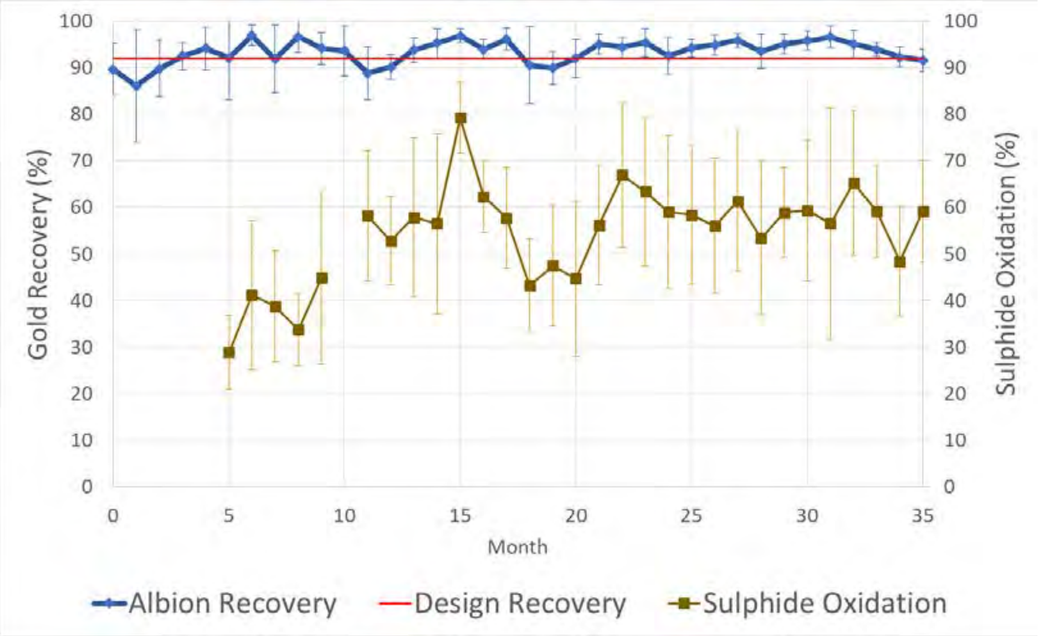


Figure 6 – Gold recovery versus degree of sulfide oxidation.

FLEXIBLE, ROBUST AND STABLE

A key feature of the data is the stable performance of the Albion Process™ when treating a range of throughput rates and a range of quality of feed in terms of sulphur grade and gold grade. Gold recoveries have remained consistent since commissioning and have predominantly been above design levels.

Consistent gold recoveries above design have been maintained in spite of unexpected variability in the throughput rate to the Albion Process™ plant, and significant variability in the sulphur grade being fed to the plant. The figures below represent graphically the consistency of the Albion Process™ plant gold recoveries against the variability of these parameters:

Throughput variability has reduced over time as the operation matures. Nevertheless gold recoveries were consistent early in the life of the operation, a significant advantage for the operator of the plant as it works from that early base towards better process control.

The ability of the Albion Process™ to process variable sulphur grades at GPM is due to the auto-thermal nature of the Albion reactors. Once a minimum sulphide-sulphur throughput is met, the reactors operate in the range 85-98°C. If the plant encounters lower sulphide levels it can usually operate for several hours due to the large thermal inertia in the process and has no material impact.

By that time, the issues causing the low sulphur grade in the concentrator have usually been resolved. If the low sulphur condition persists and temperature drops in the first few reactors, the first reactor can be partially scuttled with slurry draining to the bunded area below the thickener and recycled to the process via spillage pumps. This allows the reactors to be re-filled with higher grade material with the temperature increasing again for regular operation. The reactors at GPM are open to atmosphere, so when the sulphide feed rate increases the additional exothermic heat released by oxidation of these units immediately raises the vapour pressure of the slurry, which in turn drives a greater evaporation rate that removes the additional heat. The reactors at GPM operate near the boiling point of water.

This contrasts with Pressure Oxidation circuits which are typically sensitive to variations in sulphur grade due to the requirement to manage heat of reaction in an autoclave. During an environmental-technical-economic evaluation phase of the project, a more stable throughput and sulphur grade was assumed as a basis for comparison of the POX and Albion Process™ options. The Albion Process™ was selected in part for its ability to handle any unexpected variability in feed arising from the ore body or from operation of the concentrator; selection of an autoclave circuit may have resulted in significant reduction in throughput in order to manage the ore variability seen as the project has operated.

Additionally, during project development, different ore bodies required different levels of sulphide sulphur oxidation. Selection of VPSA type oxygen plant allows turn-down of oxygen generation to match the required oxidation level and subsequent oxygen requirements for the system. Due to the lower oxidation level requirement than design, the leach reactions are normally completed in Albion Leach Reactor 6, and the full 9 reactors are not required.

Oxidation capacity

It is also noted that to achieve and exceed the design recovery, the extent of sulphide sulphur oxidation required has varied and typically is around 55 per cent, well below the design value of 76 per cent. The Albion Process™ plant generally allows more control of oxidation levels than in an autoclave circuit. There is a reduction in oxygen operating costs associated with lower oxygen requirements.

The principal duty of the Albion Process™ plant is the oxidation of sulphur, to release gold finely disseminated in the sulphide matrix for recovery in the downstream cyanide leach circuit. The oxidation capacity of the Albion Process™ plant is constrained either by the residence time in the circuit, or by the available oxygen, and was a major consideration in the design and engineering of the plant during the design phase. The nominal and design oxidising capacities are 1.56 and 1.80 tonnes sulphide sulphur per hour respectively assuming an 80 per cent oxygen utilisation.

The design for the GPM Albion Process™ plant is to achieve 76 per cent sulphide sulphur oxidation, with a nominal and design sulphur oxidation of 37.6 and 43.2 tonnes oxidised per day. The design oxygen consumption was 336 kg / t concentrate assuming 80 per cent utilisation. Actual consumptions are around 215 kg / t concentrate based on plant experience due to higher than design oxygen utilisation and lower oxidation rates (Voigt, Hourn, Mallah and Turner, 2015). The reported plant data indicates the oxidation extent has been controlled and adjusted in order to maintain gold recoveries, demonstrating the flexibility of the process.

Periods of oxidation that exceeded the design maximum were demonstrated in 2016 and 2017. The spare oxidation capacity evident in the Albion Process™ plant is demonstration of the contingency applied in design, and is to the superior performance of the HyperSparge™ supersonic gas injection lance. Supersonic gas injection maximises the shear imparted to the slurry resulting in the resistance to gas to liquid oxygen mass transfer being reduced and in turn maximising oxygen mass transfer rate. The actual plant oxygen utilisation was found to be around 90 per cent or higher based on plant surveys and has been reported by Voigt, Mallah and Hourn (2017).

DESIGNED RIGHT – THE FIRST TIME

A pillar of fast ramp up and achievement of design throughput and recovery is a successful design process from the laboratory to the full-scale production plant. The GPM plant was designed and supplied by Glencore Technology based on results from batch and continuous testwork campaigns conducted as part of the BFS managed by Core Resources during 2011 and 2012. The purpose of the testwork was to obtain the following key design parameters which culminated in the mass and energy balance and process design criteria:

- optimise the grind size to minimise the IsaMill™ energy input,

- IsaMill™ Signature Plot to determine the grind energy required to achieve a certain particle size in order to size the IsaMill™ (The design specific energy to achieve 80 per cent passing 10µm was 59.0 kWh/t and recent plant data shows 59.7 kWh/t although not always required to run at 10µm),
- sulfide oxidation against gold recovery to optimise gold recovery and minimise oxidation requirements,
- measurement of slurry oxygen up-take rate under various conditions for designing the oxygen mass transfer system,
- rheological measurements for sizing of the pumps,
- determine the sensitivity of sulphide leach kinetics to key parameters such as operating temperature and slurry density,
- dettling data for sizing of the thickener, and
- test the process performance against samples that represent potential variance in the feed material during the life of the plant.

Testwork was structured around defining metallurgical performance of ore types and composites based on the life of mine schedule. During the testwork it was found that different ore types gave a different flotation response and required different levels of oxidation to achieve greater than 90 per cent gold recovery in cyanidation.

The design approach was to select a conservative sulphide sulphur oxidation target of 76 per cent to ensure that different combinations of ore types could be treated and still achieve the design target of greater than 90 per cent gold recovery in the CIL. The oxidation target largely drove the tank volume required and the oxygen plant size. Since oxygen is recognised as a large portion of the operating costs, two Vacuum Pressure Swing Adsorption (VPSA) type oxygen plants were selected for oxygen generating duty to allow oxygen output to be turned down when not required for the process and save on operating costs (as illustrated in Figure 6). VPSA technology was selected over PSA due to the lower power cost per unit of oxygen produced. The oxygen purity generated is 93 v/v %.

The important part of oxidation design is oxygen mass transfer to the slurry system. The scale up of oxygen mass transfer process is critical to the process success and is well understood and proven through the use of the HyperSparge™ supersonic gas injection technology (Voigt, 2017). A conservative oxygen utilisation was assumed at 80 per cent although plant measurements have shown this to be closer to 90 per cent (Voigt, 2015).

It can be observed from Figures 2 to 6 that there are no large drops in production rate over a monthly period. This stability of the production is a result of a well-designed plant. Specific factors which have contributed to high-availability of the plant include:

1. No major works or process modifications were required following commissioning, with most activity focussed on bringing the grinding and flotation concentrator up to full production;

2. The concentrator has two lines of operation and one line can be kept going while a mill re-line occurs; The Albion Process™ has been designed to avoid major downtime, through:
    - the surge capacity around the IsaMill,™
    - the ability to remove, inspect and insert HyperSparge™ without stopping the process or emptying a leaching tank,
    - the capability to by-pass the leaching section, and
    - the process comprises well known and understood equipment (tanks, pumps, thickeners, agitators).

As a result of this design, the process operates with very high availability (more than 95 per cent) and stable throughputs.

TECHNOLOGY TRANSFER

The Albion Process™ is an established technology but had not been previously deployed into Armenia or the wider CIS. This meant that training of operations and maintenance personnel was required to ensure safe and timely ramp up of the process. Training was performed in a class room environment during the final stages of construction and then on the plant during pre-commissioning checks and commissioning. Training was performed by Glencore Technology experts as well as representatives from sub-suppliers.

The training performed prior to commissioning meant that Glencore Technology (GT) was positioned to take the role as commissioning manager with minimal site presence (three personnel) while GPM personnel operated the plant. This ensured that a high level of ownership of the plant process management was taken on by the project owner at the very beginning of the project.

Following commissioning and ramp up, bi-annual visits from Core Resources and Glencore Technology personnel ensured that regular communication on technical issues occurred.

As a testament to the technology transfer process and ownership of the process, GPM have arrived at a number of process modifications and improvements:

1. By-passing the thickener

   The Albion Process™ leach residue is directed to a thickener to increase the solids content from 30-42 per cent suitable for feeding to the CIL process. Since the Albion Process™ residue is mixed with flotation tailings and sent to CIL, the Albion Process™ thickener has been by-passed and CIL feed density specification met by slightly increasing the flotation tailings thickener underflow density.

2. Reduced reagents and Closed overflow lines on tanks

   The Albion Process™ leach reactor design comprises inter-connecting launders for slurry transfer and each tank is fitted with an overflow arrangement that works with a weir to keep the head space sealed so only gases exit the ventilation stack. In the event of high carbonates in the feed or excessive flotation reagent accumulation in the process water, the tanks can overflow due to excessive foam production which is designed to be transported between tanks through the launders but often finds the path of least resistance through the reactor overflow pipes. Although this material is recovered back to the process with the bund and spillage system, it sometimes results in the oxygen flowrate to be reduced in the offending reactor and then increased in another causing gas surging and further foaming issues. The recycle of the spillage material also causes process surges.

   Site personnel undertook an improvement project to gradually reduce flotation chemicals (mainly frother) and to close up the overflow weirs. Froth generation has been significantly reduced through reduced reagent additions in the concentrator. The overflow weirs were also welded shut which directs all foam through the interconnecting tank launders. These two measures have eliminated froth overflow.

3. Control system

   The Albion Process™ was supplied with a distributed control system (DCS) with the plant operated from a centralised control room. The existing milling and CIL plant was controlled with minimal instrumentation and field control panels so transition to a centralised control system was a new concept. During the implementation of the Albion Process™ the milling and flotation plants were integrated into the DCS for fully centralised control from ore feeding to Albion Process™ discharge.

4. Tellurides

   The presence of gold hosted in tellurides was identified during the initial testwork phases and based on drilling would be encountered after a few years of mining primary sulphides. As such, GT designed the Albion Process™ plant to be able to raise the pH to around 9.0 in the last leach reactor allowing oxidation to occur at elevated pH with the aim of oxidising and liberating gold hosted in tellurides. This had the added benefit of conditioning before the CIL which is operated around 10.5. Since the Albion Process™ runs around pH 5.5 this is a simple process transition to make. The lime ring main from the CIL was extended to Albion Process™ leach reactor 9 allowing elevated pH operation and recovery of gold and silver from tellurides. Tellurides have been observed in the feed concentrate ranging from 80 ppm to 1200 ppm. Initially the plant suffered notable reductions in gold recoveries (around 2 to 3 per cent) and was mitigated through the implementation of the elevated pH in the last leach reactor.

COLLABORATION AGREEMENT

During 2017, GPM and Glencore Technology signed a collaboration agreement for the marketing of the Albion Process™ in Russia. This is a significant milestone for the two companies because it allows new companies installing an Albion Process™ plant to leverage off the experience of the collaboration for the implementation of the Albion Process™ plant at GPM. The details of the agreement allow new project personnel to visit the GPM plant to learn more about the Albion Process™ and allows access to the GPM site for training of personnel in operations and maintenance.

ACKNOWLEDGEMENTS

The authors gratefully acknowledge the provision of the data by the owners of the GPM operation in Armenia.

REFERENCES

Hourn, M, and Turner, D, 2012. Commercialisation of the Albion Process, in *Proceedings ALTA Gold Conference 2012, pp 231-248* (ALTA Metallurgical Services: Melbourne).

McNulty, T, 2014. Plant Ramp-Up Profiles: An update with emphasis on process development, in *COM 2014 – Conference of Metallurgists Proceedings* (Canadian Institute of Mining, Metallurgy and Petroleum: Westmount)

Voigt, P, Hourn, M, Mallah, D and Turner, D, 2014. Development of the Albion Process plant to treat refractory concentrates from the GPM Gold Project, in *Proceedings – Hydroprocess Conference,* 2014, Chapter 5 (Gecamin: Santiago).

Voigt, P, Hourn, M, Mallah D and Turner D, 2015. Commissioning and Ramp-Up of the Albion Process at the GPM Gold Project, in *Proceedings MetPlant 2015*, pp 207-219 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Voigt, P, Mallah, D and Hourn, M, 2017. Oxygen Mass Transfer in the Albion Process™: From the Laboratory To The Plant, in *Proceedings ALTA 2017*, pp 135-144 (ALTA Metallurgical Services, Melbourne, Australia).

Voigt, P and Walker, D, 2017. The Albion Process™ at the GPM Gold Project – The Success of a Technology, in *Proceedings CIM Convention 2018*, (Canadian Institute of Mining, Vancouver, Canada).

MAKING THE RIGHT SELECTION: A COMPARATIVE ANALYSIS FOR THE TREATMENT OF REFRACTORY GOLD CONCENTRATES

ABSTRACT

Building on work completed in earlier papers, the current paper presents a techno-economic evaluation of two recently demonstrated technologies for treatment of arsenic containing refractory gold concentrates. Using defined project parameters and inputs, the authors compare capital and operating cost estimates for pressure oxidation (POx), and the Albion Process^{TM}. The paper incorporates data now publicly available from the Albion Process^{TM} plant, which has operated at the GeoProMining Gold (GPM Gold) project in Armenia since 2014, as well as recent POx circuits.

Key Words: Albion Process, Autoclave, Pressure Oxidation

INTRODUCTION

The genesis for this paper began in 2005 when a similar refractory gold extraction comparative study was undertaken by authors mainly from Jacobs' predecessor Aker Solutions [3]. The previous study compared existing established refractory gold extraction processes with the newly emerging Albion Process^{TM}. With a continued strong demand for options to treat refractory gold deposits, the Albion Process^{TM} plant which has been in operation for more than three years at GPM Gold in Armenia now provides operational and capital cost information.

We have chosen to constrain this paper to only comparing pressure oxidation (POx) with Albion Process^{TM}, with both processes fed by sulphide flotation concentrates. To make the comparison more direct we have also assumed that the upstream materials handling, crushing, grinding and flotation plant sections are identical for both options, as are the downstream gold extraction (Carbon in Leach [CIL] circuit and gold room) and tailings handling systems.

We acknowledge there will be some inherent differences due to the extraction processes used and their management needs in regards to chemistry and plant water balances.

It is acknowledged that there are other relevant processes which can be used for treating refractory gold in both whole of ore and concentrates feed applications; these include roasting (using suppliers such as Outotec [Lurgi] and Technip [Dorr Oliver] and biological oxidation such as Outotec BIOX). These require close liaison with specialty vendors and extensive test work with those particular vendors. There is also the option of producing a concentrate for sale to a larger smelter complex.

Including all these options in this review would require a high degree of open exchange with a number of competing parties and therefore this review has been constrained to comparing the POx process with the Albion Process^{TM}. It is considered that the Albion technology is sufficiently proven and mature to be viable for practical consideration for refractory gold applications. The GPM plant has now been in operation for more than three years and there are six Albion plants in operation globally in a variety of locations and treating a variety of commodities.

For the ease and accuracy of comparison we have based comparative costs on a project in North America with a similar gold output to the GPM Gold project.

A conservative design availability of 85% is used in this paper for the POx circuit. In comparison the Albion Process^{TM} availability is considerably higher at 90%. Recognising the difference in plant availability for the purposes of this comparison is fundamental for the sizing of major equipment in each circuit.

It is strongly acknowledged that, for any refractory gold project, there is a need to have sufficient and early high quality testing and studies completed in areas such as mineralogy, geometallurgy, floatability, gravity recovery, grindability, rheology and cyanide extraction. These are typical requirements before the critical work of developing engineering options for the process route commences. A formal and staged engineering study approach using experienced teams and established study benchmarking is vital to working towards selection of a preferred process option(s) for the specific project.

ASSUMPTIONS

This is a comparative study between two refractory gold oxidation methods. As such the philosophy adopted has been to limit the comparison to the specific plant areas that are significantly different. Therefore, this gives an input stream of flotation concentrate and an output stream of oxidised slurry being fed to the CIL and subsequent downstream recovery and tailings.

The concentrate tonnage and composition used in this paper is the same as the actual GPM Gold material and POx performance has been implied for that feed material. The actual GPM Gold plant also directs the flotation plant scavenger tailings to the CIL plant and this approach has been adopted for both the POx and Albion Process^{TM} circuits studied in this comparison. Based on the actual GPM Gold plant operating experience and current practice, any future plant similar to the GPM Gold plant would be modified to eliminate the discharge thickener and change the Albion Process^{TM} leach reactors from nine 240 m^3 tanks to six 340 m^3 tanks. These flowsheet changes have been incorporated into this comparison.

Other assumptions that have been made include the following:

- A North American generic site with a 2018 cost structure is the comparison. The specific site is flat with stable foundations and good ground conditions at an elevation of 1200 metres above sea level.
- Any reagents required for the oxidation process are included in the costings developed for this paper. This includes sourcing, mixing, storage and distribution as needed for lime, limestone, oxygen etc.

MINERALOGY

For the ore body studied in this comparison, pyrite, arsenopyrite and pyrrhotite are the major sulphide minerals present which report to the sulphide concentrate. Gold occurs as native free gold, finely dispersed gold in arsenical sulphide, gold tellurides and secondary native gold remaining after oxidation of sulphides and tellurides. Silver is present in its native form in quartz, chalcopyrite and pyrite as well as silver tellurides. The main gangue minerals are quartz, talc and chlorite, with minor magnesite, dolomite and calcite.

The sulphide concentrate is produced from a typical milling and bulk flotation circuit. The elemental and mineralogical composition of the sulphide flotation concentrate used to develop the comparison between the POx circuit and the Albion Process^{TM} circuit is summarised below:

Table 1 - Flotation Concentrate Elemental Assay

Chemical Element		Units	Value
Arsenic	As	% w/w	2.66
Iron	Fe	% w/w	18.71
Sulphur	S	% w/w	17.26
Silicon	Si	% w/w	17.89
Oxygen	O	% w/w	30.76
Magnesium	Mg	% w/w	4.80
Calcium	Ca	% w/w	1.79
Other	-	% w/w	6.13
Gold	Au	g/t	47.28
Silver	Ag	g/t	48.43

Table 2 - Flotation Concentrate Predominant Minerals

Mineral	Chemical Formula	Units	Value
Pyrite	FeS_2	% w/w	23.9
Aresnopyrite	$FeAsS$	% w/w	5.8
Pyrrhotite	$Fe_{0.877}S$	% w/w	5.18
Chalcopyrite	$CuFeS_2$	% w/w	1.0
Quartz	SiO_2	% w/w	33.9
Calcite	$CaCO_3$	% w/w	2.23
Dolomite	$CaMg(CO_3)$	% w/w	2.18

Magnesium Oxide	MgO	% w/w	2.15
Magnesite	$MgCO_3$	% w/w	5.82
Talc	$3MgO*4SiO_2*H_2O$	% w/w	5.92
Other		% w/w	11.92

BASIS AND METHODOLOGY OF ESTIMATE

The cost estimates for this comparison are based on the following documents: process design criteria, mechanical equipment list (process and utilities equipment), material take offs (MTOs) (includes ducting, bins, etc.), block flow diagram, site plot plans and general arrangement drawings, project design basis (Aspen Capital Cost Estimator [ACCE] specifications only) and escalated prior quotations. The ACCE package generated pricing on a majority of the process and utilities equipment as well as most bulk materials with remaining equipment and bulk material pricing based on Jacobs in-house pricing (consistent with pricing on recent projects) and cost data from published estimating sources and other supporting engineering documents

The estimates were prepared per Jacobs' guidelines and standards for a Class 5 Capital Cost Estimate, accuracy range +50% / -50%. The estimates identify the capital costs associated with the flowsheets described below.

The estimates represent modelled quantities developed from the Jacobs-customised software program Aspen Capital Cost Estimator (ACCE) for all major direct cost accounts. Where specific data was not available, data was factored based on Jacobs historical data. Such quantification methods are suitable to support a Class 5 estimate.

A major mechanical equipment list was generated with conditions of service sufficient to support equipment pricing. Major equipment pricing was based predominantly on escalated previous quotes from similar projects and from ACCE system pricing. Engineering design data for the equipment specified was used to generate ACCE pricing for the quoted packages.

Approximately 73 - 78% of the purchased equipment costs (PEC) in the estimate represent escalated prior quotes and 22 - 36% represent ACCE pricing based on preliminary design data sheets, preliminary conditions of service, capacities and specifications. Approximately 1% of the PEC in the estimate represents in-house or historical pricing data.

Bulk MTOs were generated by ACCE for each of the relevant accounts based on the equipment sizing, capacities, specifications and conditions of service. Building costs are based on cost per square metre rates from recent purchases and in-house estimates. Design development allowances (DDAs) and MTO allowances for each relevant account were established by estimating and discipline engineering personnel, based on Jacobs' experience.

Jacobs prepared field construction installation hour estimates via ACCE based on Jacobs' standard estimating labour units, which represent U.S. Gulf Coast rates at ideal conditions for a mid-size project at a greenfield site. Productivity adjustments were then developed for the North American generic site and applied to those installation hours.

The remainder of the direct costs were developed using customised ACCE equipment cost modelling software. Construction indirect costs, engineering and other costs, as well as the overall area and total cost summaries were developed based on percentages and factors from historical project data based on the defined scope. Metrics and benchmarking analyses were then completed, along with the overall assembly and finalisation of the basis of estimate and review packages.

The operating costs development comprises: power consumption; reagents consumptions; operating and maintenance labour and materials. Power consumption estimates were factored from the equipment list. Reagent consumption rates have been developed based on the process design criteria. The costs of maintenance were based on data typical of the project location, factored from the installed capital and taking into account the service and operating conditions of the equipment. For the purposes of this comparison it has been assumed that the labour requirements for both circuits would be equivalent. This assumption is based on the availability of an experienced workforce in North America and may not hold for all geographies.

Prices for reagents, utilities and consumables have been sourced from Jacobs' North American database of recent pricing.

PROCESS FLOWSHEETS

Both processing routes treat a flotation concentrate containing predominantly pyrite, with minor arsenopyrite. The GPM Gold concentrate is relatively clean with no appreciable amounts of mercury, caesium, selenium or vanadium. The sulphur oxidation and gold liberation are completed using either the Albion Process™ or POx.

Albion Process™ Flowsheet

The flowsheet for the Albion Process™ circuit is based on the circuit installed at GPM Gold and incorporates learning based on the performance of this circuit.

In the Albion Process™ flowsheet, the flotation concentrate Thickener underflow concentrate is pumped to the IsaMill™ feed pumpbox where it is combined with media before being pumped to the M3000 IsaMill™. The IsaMill™ is specified to achieve a grind size of 80% mass passing 12 to 14 µm. The discharge slurry is then pumped to an agitated concentrate storage tank with over eight hours surge capacity to allow the leach to continue to operate when IsaMill™ maintenance activities are performed. The concentrate slurry is then transferred to one of the first three Albion Process™ leach reactors. The oxidative leach circuit flowsheet consists of six, 340 m^3 Albion Process™ leach reactors fabricated from lean duplex alloy steel (LDX2101) connected with launders allowing tank by-passing during maintenance events. Each reactor is fitted with a 160 kW dual impeller agitator, with oxygen delivered by a bank of six supersonic HyperSparge™ oxygen gas injection lances. The process is designed to run autothermally at or around 93°C.

The pH in each reactor is maintained between 5.0 to 5.5 through dosing of limestone slurry. The limestone slurry is produced in an on-site limestone milling plant, with a capacity of 6 t/h. The limestone is milled to an 80 % mass passing size of 75 µm in a ball mill operated in closed circuit with cyclones. The cyclone overflow reports to a 150 m^3 agitated distribution tank and circulates through the oxidative leach circuit by ring main.

Oxygen for the Albion Process™ and CIL is provided by two 60 t/d Vacuum Pressure Swing Adsorption (VPSA) plants with up to 98 t/d going to the Albion Process™ and the balance to the CIL plant. The turndown capability in the VPSA means the oxygen generating capacity can be reduced when less oxidation is required.

The leach discharge is around 30% solids and is mixed with flotation tailings before feeding the CIL circuit.

The Albion Process™ residue treated in the CIL circuit is characterised by low lime and cyanide consumption as a result of the continual neutralisation of iron and acid through the addition of the alkali limestone during oxidation. This process also prevents the formation of element sulphur. The cyanide consumption for the Albion Process™ residue is around 4kg cyanide per tonne of residue.

Pressure Oxidation (POx) Flowsheet

In the POx flowsheet, the Flotation Concentrate Thickener underflow concentrate is fed via a piston diaphragm feed pump in a duty standby arrangement into a single 5 compartment autoclave vessel. The autoclave vessel is equipped with a bottom-inlet sparge system to introduce gaseous oxygen, coolant water, and steam (for initial heat-up).

The oxidised slurry is then discharged from the last autoclave compartment to a flash tank where pressure reduction is accomplished by a ceramic lined control valve followed by a choke tube and fixed choke. The flash tank overheads are ducted to a venturi scrubber. The vent from the autoclave is reduced in pressure through a ceramic lined control valve to the POx vent spool before going to a dedicated venturi scrubber. The pressure safety salves (PSV's) included in the system discharge to a different vent spool and cyclonic separator, equipped with water addition points to clean the system when a PSV lifts. Gas and non-condensables discharge from the top of the cyclonic separators, with water and condensables flowing to the scrubber pump box, to be pumped to the decant thickener.

The autoclave support systems include a high pressure seal water system, demineralisation water package, stand-by / start up boiler unit, glycol handling system, coolant injection system and seal flushing system as well as process cooling. Oxygen for the autoclaves is provided by a VPSA plant.

From the autoclave circuit the hot flashed slurry enters the hot cure circuit. The hot cure circuit not only allows cooling of the slurry for the downstream CIL cyanidation circuit but it also aids in the conversion of basic iron sulphate to ferric sulphate by lowering the temperature and providing retention time for iron and arsenic precipitation.

After slurry cooling, the cooled slurry is pumped to the Counter-Current Decantation (CCD) circuit which consists of a decant thickener and two CCD's.

The purpose of the CCD circuit is to wash the POx slurry by adding wash water counter current to the slurry flow to remove acid and if applicable, the soluble copper from the gold bearing solids. The concentrated copper solution can report to a copper precipitation circuit for recovery (if economic) while the washed gold bearing solids are pumped to the neutralisation circuit prior to being combined with the flotation tail in the CIL circuit to recover the gold.

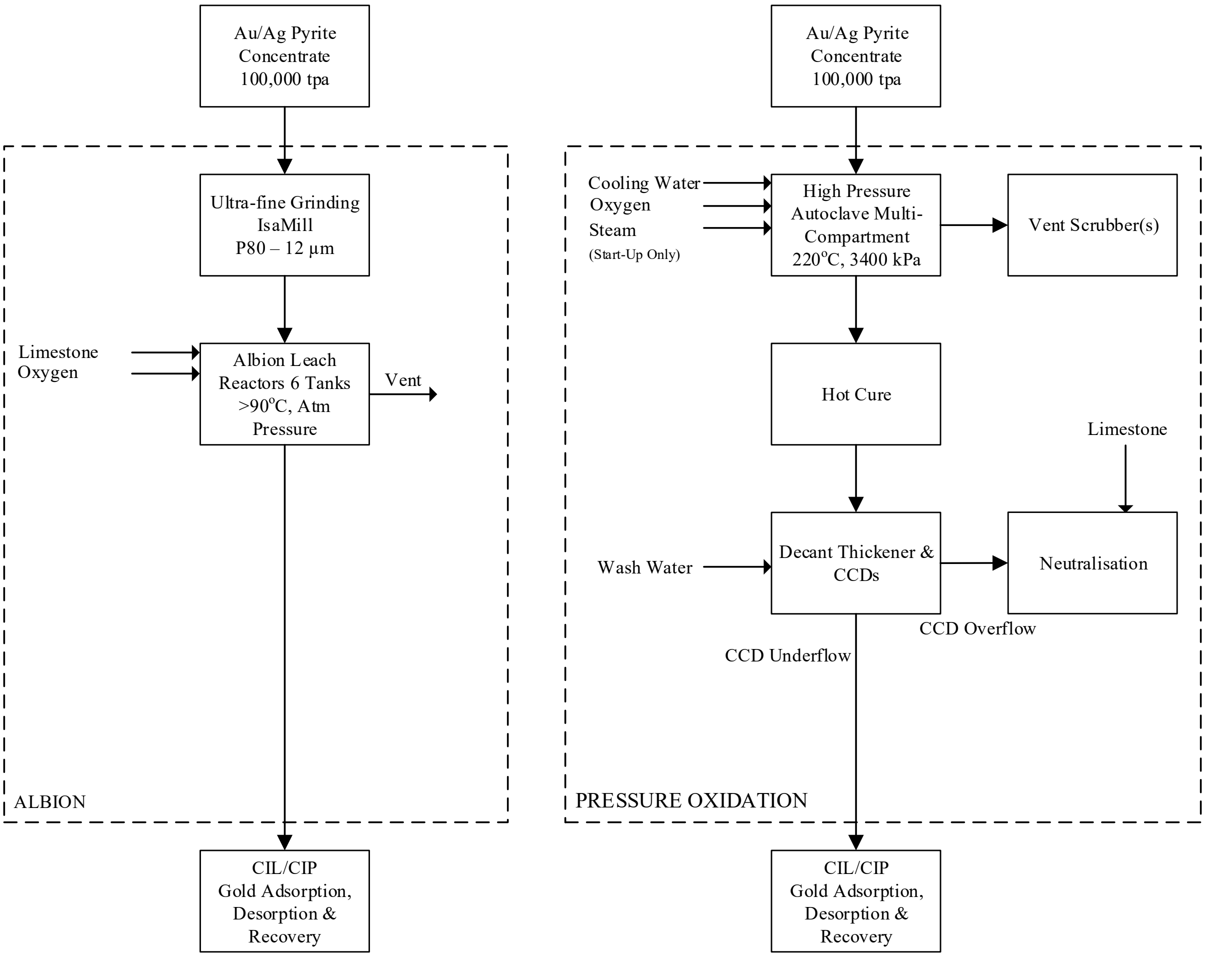


Figure 1 – Albion Process™ and POx Flowsheet Comparison

POX CHEMISTRY

The POx process was originally developed for the treatment of base metal concentrates before being adapted to treat gold containing pyrite and arsenopyrite ores and concentrates. The process uses an autoclave operating typically in a temperature range of 190 to 230°C with an oxygen over pressure of 350 to 700 kPa(g) and retention times that vary from 60 to 90 minutes. Sulphide oxidation in the POx autoclave is typically greater than >98% for most ores and concentrates. The feed density of the sulphide concentrate to the autoclave is regulated to provide sufficient heat, generated by the exothermic oxidation reactions, to maintain the desired operating temperature of the autoclave. The following oxidation and hydrolysis reactions occur in the autoclave when pyrite and arsenopyrite react with oxygen:

Pyrite: $2\,FeS_2 + 7\,O_2 + 2\,H_2O \leftrightarrow 2\,FeSO_4 + 2\,H_2SO_4$

$2\ FeSO_4 + H_2SO_4 + \frac{1}{2}\ O_2 \leftrightarrow Fe_2(SO_4)_3 + H_2O$

$Fe_2(SO_4)_3 + 3\ H_2O \leftrightarrow Fe_2O_3 + 3\ H_2SO_4$

Arsenopyrite: $2\ FeAsS + H_2SO_4 + 7\ O_2 + 2\ H_2O \leftrightarrow 2\ H_3AsO_4 + Fe_2(SO_4)_3$

$2\ H_3AsO_4 + Fe_2(SO_4)_3 \leftrightarrow 2\ FeAsO_4 + 3\ H_2SO_4$

ALBION CHEMISTRY

The Albion Process^{TM} comprises a patented combination of fine grinding and atmospheric leaching. The first step uses ultra-fine grinding IsaMill^{TM} technology to produce a sufficiently fine and narrow particle size distribution. The second step is sulphide oxidation achieved through supersonic oxygen injection into leach reactors operating at near neutral conditions (pH 5.5). The chemical reactions which form the basis of the Albion Process^{TM} require the addition of oxygen to oxidise the pyrite and arsenopyrite while limestone is added to continually neutralise acid generated by the oxidation reactions.

Pyrite: $4\ FeS_2 + 11\ O_2 + 18\ H_2O + 8\ CaCO_3 \leftrightarrow 4\ FeO.OH + 8\ CaSO_4.2H_2O + 8\ CO_2$

Arsenopyrite: $2\ FeAsS + 7\ O_2 + 4\ H_2O + 2\ CaCO_3 \leftrightarrow 2\ FeAsO_4 + 2\ CaSO_4.2H_2O + 2\ CO_2$

The exothermic chemistry results in autogenous reactions, operating at 93°C with no requirement for additional heating or cooling. For pyrite, the concentrate is ground to 80 % passing 10 to 12 µm (in the case of the GPM Gold plant, 12 to 14 µm) and the extent of oxidation can be varied depending on the level of oxidation required for certain feed types. In the case of GPM Gold, the design sulphide sulphur oxidation extent is 76 % to achieve over 93 % gold recovery, although the plant has been reported to operate at 60 % sulphide sulphur oxidation and achieved greater than 95% gold recovery [7]. To achieve the sulphide sulphur oxidation extents achieved in the GPM Gold plant requires a typical residence time of 30 hours.

CAPITAL AND OPERATING COST ESTIMATE COMPARISON

The capital and operating cost estimates prepared for the Albion Process^{TM} and POx flowsheets are based on the process design criteria, equipment list and other supporting documentation.

Table 3 - Capital and Operating Cost Estimate Comparison

	Albion Flowsheet	POx Flow sheet
Throughput	100,000 tpa	100,000 tpa
Gold Leach Extraction	>93%	>94%
Capital Cost USD	**66,200,000**	**95,900,000**
Direct Costs	30,200,000	44,500,000
Indirect Costs	9,700,000	13,500,000
EPCM	8,700,000	12,700,000
Contingency	15,300,000	22,100,000
Capital Spares & First Fills	2,300,000	3,100,000
Annual Operating Cost USD	**6,000,000**	**10,000,000**
Reagents	2,400,000	5,100,000
Power	2,300,000	2,500,000
Maintenance	1,200,000	2,300,000
Labour *	-	-

*For the purposes of this analysis it has been assumed that the Labour requirements are equal.

DISCUSSION

There are four commercially proven common, pre-treatment options for refractory gold ores: roasting; pressure oxidation; bio-oxidation; and ultrafine grinding, of which Albion Process^{TM} is one application. For the purposes of this direct comparison only POx is being evaluated against Albion Process^{TM} as POx is currently the process route often employed for refractory ores due to environmental permitting. In some countries it is difficult to obtain approvals to build a roaster circuit and in other countries arsenic materials can only be handled by pressure oxidation. Pressure oxidation for high arsenic ores and concentrates has been around for more than 30 years and can be said to be a mature technology.

The advantage of pressure oxidation is a higher oxidation (near complete breakdown of sulphides) allowing locked gold to be liberated and hence maximises gold recovery from any refractory ore or concentrate. Pressure oxidation can be used on a very wide range of feed materials from low sulphide whole ores to high grade flotation concentrates. Additionally, depending on acid levels and the iron to arsenic ratio, almost all of the arsenic forms a stable ferric arsenic complex currently considered to be environmentally safe. Ores and concentrates with significant copper levels can be treated for both gold and copper recovery; a significant portion of the copper in the ore is solubilised and available for recovery by precipitation of the Cu^{++} ion in a downstream process. For the purpose of this evaluation, only a high grade, relatively clean concentrate is being compared, as Albion Process^{TM} is not suitable for whole ore processing due to the high cost of grinding ore.

The disadvantages of pressure oxidation are the high capital cost for specialised equipment and the requirement for a more "technical" workforce. Pressure oxidation may also not be suitable for ores and concentrates containing high amounts of silver. Silver often reacts to form silver-jarosite which is resistant to cyanide leaching. Therefore, silver recovery may be lower than other treatment methods. If not well controlled, basic iron sulphate and iron jarosite (instead of hematite) can also form, making downstream processing difficult with the added possibility of some acid and metal release into the environment from tailings ponds. Basic iron sulphate causes problems in the cyanidation process mainly due to the fact that under certain conditions the compound breaks down, releasing acid.

As identified in the cost comparison above, the Albion Process^{TM} has a lower capital intensity and lower operating cost than the traditional POx process. The simplicity of the Albion Process^{TM} flowsheet results in a smaller number of unit operations, simple plant layout, moderate operating temperature and atmospheric leach conditions. In addition to the cost advantages of the Albion Process^{TM}, the chemistry associated with the process presents a number of advantages. The advantages associated with the Albion Process^{TM} chemistry include; fixing impurities in an inert residue with limestone with no production of sulphur dioxide, fixing of arsenic as ferric arsenate and the continual neutralisation of iron and acid in the process.

Less tangible advantages of the Albion Process^{TM} are based around the relatively simple unit operations. including the requirements for a less technically skilled workforce to operate and maintain the plant. In addition, the advantage of a fast ramp-up of the circuit to nameplate capacity avoids costly delays which have the potential to damage the business case.

During a project's initial development all ores and concentrates need to go through a rigorous test work program. In this particular instance the Albion Process^{TM} is the preferred process option both from a capital and operating cost basis. However, gold recovery is higher with POx than the Albion Process^{TM}. In this particular case study, high gold leach extractions were achieved (93%) with comparatively low oxidation (76% oxidation) compared to POx. It must be said that this performance is specific to the GPM Gold plant ore and concentrate; therefore, there is a need for a thorough metallurgical test program to fully evaluate the process options for other ores and concentrates.

SUMMARY AND CONCLUSIONS

Each refractory gold process selection is unique and should initially include the four current options of biological oxidation (BIOX), roasting, pressure oxidation and the Albion Process^{TM}. For this paper, only the latter two are considered on a comparative capital and operating cost basis. The metallurgical performance for POx has been implied. It is vital that project metallurgical process route selection be driven by adequate metallurgical testing as refractory gold ores are notoriously individual.

The Albion Process^{TM} process exhibits a shorter project implementation time, reduced technical complexity with resultant benefits of simplification, higher utilisation and reliable metallurgical performance. The higher recovery of POx must be balanced against these other project parameters in determining the optimum project economics.

The quantifiable cost benefits and the qualitative operating and maintenance benefits of the Albion Process^{TM} make it a viable alternative for processing refractory gold concentrates, warranting consideration in the flowsheet development for a refractory gold project.

ACKNOWLEDGEMENTS

It is acknowledged that there exists a considerable degree of prior knowledge and experience that has been built up and documented for gold POx and the practical application of the mature autoclave technology. Authors and contributors to this level of knowledge are recognised and thanked.

For the Albion Process^{TM} there is now emerging public access to detailed information able to be used for industry studies and the efforts in this regard from GPM Gold, Glencore Technology and Core Resources are acknowledged.

This paper was only made possible with the active support and generous provision of information from Glencore Technology and the Albion Process^{TM} marketer and testwork provider Core Resources. In particular, contributions were vital from Paul Voigt, Daniel Mallah, Mike Hourn and Peter Rohner.

The Denver and Brisbane teams of Jacobs collaborated to proactively work on this paper. The estimating effort from Denver is acknowledged as is the Jacobs expertise and database accessed for the POx information in particular. It should be emphasised that the theoretical POx flowsheet used in this comparative exercise requires assumptions and judgements that will not apply necessarily to a particular implemented project.

The permission of and support for this paper by Jacobs is acknowledged.

REFERENCES

[1] Aksenov A, Senchenko A, Seredkin Y, Vasiliev A. (2016) Technology for Processing of Refractory Gold-Containing Concentrates Based on Ultrafine Grinding and Atmospheric Oxidation. Paper presented at the proceedings of IMPC 2016, Quebec City, Canada, 11-15 September 2017.

[2] Aylmore M, Jaffer A. (2012) Evaluating Process Options for Treating Some Refractory Ores. Paper presented at the proceeding of ALTA 2012, Perth, Australia, 27 May – 1 June 2015.

[3] Bartsch P, Hourn M, Ngoviky K, Rohner P. (2005) Benefits of Using the Albion Process for a North Queensland project, and a Case Study of Capital and Operations Cost Benefits Versus Bacterial Oxidation and Pressure Oxidation. Paper presented at the Randol Gold Conference, Perth, Australia, 21-24 August 2005.

[4] Dunn G, Ibrado A, Parkison G, Mora N, Teo YY. (2014) Pressure Oxidation of Refractory Gold Ores: The Metates Gold Project Experience. Paper presented at the proceedings of ALTA 2014, Perth, Australia, 24-31 May 2014.

[5] Hourn M, Kloiber-Deane O, Voigt P, Walker D. (2018) Long Term Performance of the GPM Albion Process™ Plant. Paper presented at the CIM Convention, Vancouver, Canada, 6-9 May 2018.

[6] Hourn M, Mallah D, Turner D, Voigt P. (2015) Commissioning and Ramp-up of the Albion Process at the GPM Gold Project. Paper presented at the proceedings of ALTA 2015, Perth, Australia, 23-30 May 2015.

[7] Hourn M, Turner D. (2012) Commercialisation of the Albion Process. Paper presented at the proceedings of ALTA 2012, Perth, Australia, 27 May – 1 June 2015.

Выход на проектную мощность и производственные показатели за длительный период установки Albion Process™ на предприятии GeoProMining Gold, Армения,

П. Войт[1], Д. Уолкер[2], О. Клойбер-Дин[3] и А. Цветков[4]

1. Руководитель гидрометаллургического направления, «Glencore Technology», Брисбен, Квинсленд 4000. Адрес электронной почты: paul.voigt@glencore.com.au
2. Генеральный менеджер по комплексным решениям, «Core Resources», Брисбен, Квинсленд 4010. Адрес электронной почты: dwalker@coreresources.com.au
3. Главный технолог, «Core Resources», Брисбен, Квинсленд 4010. Адрес электронной почты: okloiber- deane@coreresources.com.au
4. Заместитель генерального директора по обогащению и металлургии, ООО «GeoProMining Gold», Ереван, Армения. Адрес электронной почты: atsvetkov@geopromining.com

АННОТАЦИЯ

На золотодобывающем руднике компании ГЕОПРОМАЙНИНГ в поселке Сотк (Зод) в Армении установка «Albion Process™» эксплуатируется с 2014 года. Впервые публикуются данные за длительный период эксплуатации установки. В настоящей работе представлены такие данные, как объем концентрата в виде питания установки Albion Process™ (тонн в сутки), содержание серы в концентрате (% S), общее окисление сульфидной серы в установке Albion Process™ (% SOx) и извлечение золота из осадков установки Albion Process™ (%). Данные служат основой для оценки эффективности запуска, выхода на проектную мощность и эксплуатации установки Albion Process™. Невзирая на задержку запуска обогатительной фабрики, по шкале МакНалти (McNulty), которая описывает наращивание мощности, установка относится к группе 1, что свидетельствует об успешном запуске и быстром наращивании производительности. В результате изучения данных и производственного опыта можно сделать заключение, что успешной эксплуатации удалось добиться благодаря гибкой, надежной и стабильной технологии, разработанной на основании результатов испытаний и эксплуатации опытных установок на этапе ТЭО, которые проводились совместно Glencore Technology и Core Resources, кроме того местные операторы прошли обучение по эксплуатации установки на действующей фабрике.

ОБЩАЯ ИНФОРМАЦИЯ

Проект ГЕОПРОМАЙНИНГ, Au

В собственности и под управлением ООО «GeoProMining Gold» (ГЕОПРОМАЙНИНГ) находится золотой рудник Сотк (Зод) и Араратская ЗИФ в Армении («ГЕОПРОМАЙНИНГ Голд»). Карьер находится недалеко от границы с Азербайджаном, золотая руда перевозится по государственной железной дороге на Араратскую ЗИФ, неподалеку от турецкой границы.

Запасы на руднике Сотк (оценка по состоянию на август 2011 года) составляли 14,2 млн тонн с содержанием 4,3 г/т Au. По оценкам, объем выявленных ресурсов составляет 28 млн тонн руды с содержанием 4,2 г/т Au, прогнозные ресурсы насчитывают 16 млн тонн при 4,2 г/т Au. Минералогический состав руды представлен преимущественно арсенопиритом и пиритом. Золото преимущественно связано с арсенопиритом и в меньшей степени с пиритом. Золото встречается в легко обогащаемой форме, в виде включений в сернистом мышьяке и теллуридах. Геологические характеристики месторождения рассматривались в других работах (Voigt, Hourn, Mallah and Turner 2014).

Проект отрабатывается на протяжении нескольких десятилетий, изначально добывались выветрелые окисленные руды, покрывавшие сульфидную часть. Компания ГЕОПРОМАЙНИНГ приобрела проект в 2007 и приступила к отработке габбро и низкосульфидных руд. Извлечение золота по результатам испытаний, проведенным на образцах, полученных при разведочном бурении подстилающих сульфидных зон, составило 20-30 процентов в традиционном процессе сорбционного выщелачивания «уголь-в-щелочи» (CIL). Были проведены дополнительные испытания для оценки окисления сульфидного материала перед цианированием. Успешно проведено окисление арсенопирита и пирита перед цианированием полученного осадка, извлечение золота повысилось с 20-30 процентов до 90 процентов и выше.

ГЕОПРОМАЙНИНГ провели оценку четырех технологий окисления: обжиг, автоклавное окисление (POX), биоокисление (BiOX) и процесс Albion Process™ (технология атмосферного окислительного выщелачивания). Было проведено экологическое-технико-экономическое обоснование, по результатам которого технология Albion Process™ оказалась оптимальным решением. Основными определяющими факторами стали низкие капитальные затраты и короткое время реализации.

Испытания на этапе разработки по проекту начались в 2009 году. Проведены испытания на четырех образцах сульфидной руды (по 600 кг) из рудного тела, которые увенчались программой непрерывного цикла на опытной установке окислительного выщелачивания. Окончательная технологическая схема состоит из измельчения и сульфидной флотации с получением сульфидного концентрата, который направляется на переработку в установке окислительного выщелачивания Albion Process™. Осадки, получаемые в установке Albion Process™, объединяют с хвостами флотации и подают в действующий цикл процесса CIL.

Паспортная производительность установки составляет 100 000 тонн концентрата в год с получением в среднем ~100 000 унций в год золота в виде сплава доре. Установка Albion Process™ была поставлена ГЕОПРОМАЙНИНГ по фиксированной цене в виде технологического комплекса, с привязкой к гарантиям производительности от Glencore Technology. Механическое проектирование было выполнено в декабре 2012 года. Строительство установки было завершено в апреле 2014 года, пуско-наладочные работы завершены в июле 2014 года.

Технология Albion Process™

Технология Albion Process™ является зарекомендовавшей себя технологией атмосферного окисления для переработки упорных сульфидных концентратов. В настоящее время технология установлена на шести предприятиях по всему миру, из них две установки Albion Process™ используются для выщелачивания золота, связанного с упорными сульфидами. Химические характеристики технологии Albion Process™ широко рассматривались в других работах (Voigt, Mallah and Hourn 2017).

Технология Albion Process™ для золотой руды состоит из нескольких этапов: измельчение сульфидного концентрата с последующим окислением в реакторах при атмосферном давлении в присутствии газообразного кислорода, который вводится на сверхзвуковой скорости для максимального повышения массового переноса кислорода и содействия протеканию реакции окисления. Окисленный остаток процесса Albion Process™ далее перерабатывается в стандартном цикле цианистого выщелачивания для извлечения золота.

Реакция окисления протекает при немного кислотном pH (5,5) и поддерживается за счет добавления известняка. Реакторы автотермические, рабочая температура обычно находится в диапазоне 85-98°C. Типовым коэффициентом использования кислорода является 80-90 процентов. Технология Albion Process™ для золота не предусматривает растворения цветных металлов и получения элементарной серы или ярозитов, то есть пульпа не требует промывки и расход цианида в цикле выщелачивания остается низким. Осадкам свойственны хорошие фильтрационные характеристики, а требования к материалам изготовления обычно ниже, чем для циклов кислотного автоклавного выщелачивания или биоокисления.

РАЗРАБОТКА И ПРОЕКТИРОВАНИЕ ТЕХНОЛОГИЧЕСКОЙ СХЕМЫ ALBION PROCESS™ ДЛЯ ГЕОПРОМАЙНИНГ

Общая схема технологического процесса установки на ГЕОПРОМАЙНИНГ показана на Рис. 1.

До внедрения технологии Albion Process™ в состав производства ГЕОПРОМАЙНИНГ входили переделы измельчения, процесс CIL и золотоизвлекательная фабрика для обогащения руды непосредственно с рудника Сотк. На предприятии был цех флотации, однако он выведен из эксплуатации. Для внедрения Albion Process™ потребовалось провести реконструкцию цеха флотации и установить новое оборудование Albion Process™.

Состав технологической схемы Albion Process™ по переработке упорных концентратов для извлечения золота типичные стадии: тонкое измельчение с последующим окислительным выщелачиванием и иногда сгущением. На ГЕОПРОМАЙНИНГ производство состоит из установки тонкого измельчения IsaMill™ M3000, девяти (9) реакторов выщелачивания Albion Process™ объемом 270 м^3, далее установлен высокоскоростной сгуститель диаметром 10 м, откуда осадок выщелачивания направляется в процесс CIL при целевом содержании твердого 42 процента. Технологическая установка подробно описывается в других работах (Voigt, Hourn, Mallah and Turner, 2015).

Ключевые критерии проектирования для проекта ГЕОПРОМАЙНИНГ представлены в Таблице 1.

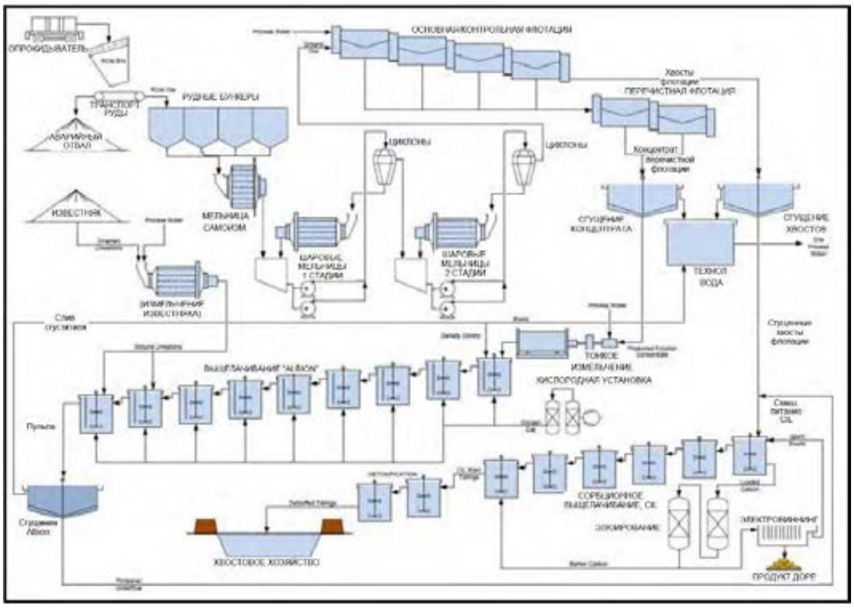


Рисунок 1: Технологическая схема ГЕОПРОМАЙНИНГ

ПРОИЗВОДСТВЕННЫЕ ТЕХНОЛОГИЧЕСКИЕ ДАННЫЕ ГЕОПРОМАЙНИНГ

В настоящей работе приводятся технологические данные за период с 2014 года по 2017 год, предоставленные ГЕОПРОМАЙНИНГ, с целью оценки наращивания мощности и производительности процесса Albion Process™. Это крупнейший блок производственных данных по технологии Albion Process™, опубликованный на сегодняшний день. Данные описывают работу процесса Albion Process™ в нескольких режимах работы, некоторые из них считаются отрицательными для циклов гидрометаллургической переработки сульфидов.

Ключевые технологические данные собирали на протяжении срока эксплуатации установки с последующим анализом. К ним относятся:

1. Объем концентрата в виде питания установки Albion Process™ (тонн в сутки);
2. Содержание серы в концентрате (% S^{2-});
3. Общее окисление сульфидной серы в установке Albion Process™ (%); и
4. Извлечение золота из осадков установки Albion Process™ (%).

На предприятии не собирают данные по использованию кислорода, эти сведения подавались с других объектов на основании исследовательских работ (Voigt, Mallah and Hourn, 2017). Так как договор на поставку электроэнергии заключен на условиях полной оплаты при отказе от поставки, полученное избыточное количество кислорода направляется в процесс CIL.

Месячные данные приводятся на Рисунках 3-6. Графики технологических данных составлены с использованием среднемесячных значений, столбики на диаграммах обозначают одно стандартное отклонение, вычисленное из отдельных суточных данных по каждому месяцу. Представленные данные основаны на среднемесячных значениях суточных данных, без обработки данных - то есть, простои на техобслуживание и прочие обстоятельства не вычитались из данных перед выдачей среднемесячных значений.

Изучение подробных технологических данных в совокупности с производственным опытом свидетельствует о том, что кроме достижения проектных параметров в части производительности и извлечения, успешная реализация проекта также выражается в некоторых ключевых факторах:

1. Быстрое наращивание мощности – процесс Albion Process™ был запущен и вышел на проектную мощность в течение трех месяцев с даты строительства;
2. Гибкость, надежность и стабильность – установка Albion Process™ эффективно перерабатывает подаваемый концентрат при варьируемой производительности и непостоянном качестве питания;
3. Изначально безупречное проектирование - проект был разработан и исполнен на основании фундаментальных испытаний и инженерно-технического проектирования, благодаря чему

результаты работы промышленной установки превзошли лабораторные испытания (Voigt and Walker, 2018); и

4. Передача технологии – технический опыт по процессу Albion Process™ был эффективно передан персоналу заказчика, чтобы они обладали всеми необходимыми знаниями по эксплуатации и обслуживанию установки.

В настоящей работе данные факторы рассматриваются более подробно.

Параметр	**Ед. изм.**	**Значение**
Номинальная производительность по питанию установки Albion	т/ч	12,1
	т/сут	290
Проектная производительность по питанию установки Albion	т/ч	13,5
	т/сут	334
Концентрация S^{2-} во флотационном концентрате	%	17,6
Степень окисления S^{2-}	%	76
Проектный коэффициент использования кислорода	%	80
Проектное извлечение золота в цикле флотации	%	88
Проектное извлечение золота в процессе CIL (осадок Albion)	%	92
Проектное извлечение золота в процессе CIL (хвосты флотации)	%	40
Проектное общее извлечение золота	%	85,8
Годовая производительность по золоту	унц.	100 000
Годовая производительность по руде	т/год	1 000 000

Таблица 1: Ключевые критерии технологического проектирования

Быстрый выход на проектные показатели

Ввод в эксплуатацию установки Albion Process™ на предприятии ГЕОПРОМАЙНИНГ был завершен в июле 2014 года. Запуск произошел за четыре месяца до завершения пуско-наладочных работ на фабрике, что привело к сокращению поставок концентрата и повлияло на скорость наращивания мощности установки Albion Process™. Кроме того, вскоре после запуска произошла поломка одной из двух воздуходувок кислородной станции, однако в тот период установка Albion Process™ не испытывала недостатка в кислороде и производительность по концентрату не пострадала.

Данные по производству золота на заводе ГЕОПРОМАЙНИНГ были нанесены на график МакНалти для оценки показателей наращивания мощности, как показано на Рисунке 2.

На Рисунке 2 показано, что показатели установки на заводе ГЕОПРОМАЙНИНГ находятся между группами 1 и 2 в течение первых двух лет, по прошествии двух лет, показатели относятся к группе 1. Показатели, соответствующие группе 1, свидетельствуют об успешной реализации проекта, обычно подразумевается высокая степень «тщательного подхода на этапе разработки проекта». (McNulty, 2014). Основанием для анализа послужило производство золота, которое было принято в качестве основного мерила успешности всего проекта в целом.

На ранних стадиях проекта показатели соответствуют группе 2, что отчасти обусловлено задержкой запуска фабрики. Однако вскоре характеристики наращивания мощности проекта быстро повысились, и показатели установки вышли на уровень группы 1 в течение первых трех лет эксплуатации.

Это является серьезным достижением данной технологии. МакНалти и соавторы утверждают, что «проекты, показатели которых соответствуют группе 1, обычно применяют апробированные технологии» (McNulty, 2014) и использование лицензированной технологии «без предшествующего опыта или с небольшим опытом» (McNulty, 2014) является одним из факторов риска, который может способствовать замедлению наращивания мощности с переходом из группы 2 в группу 4. Опыт установки на производстве ГЕОПРОМАЙНИНГ опровергает данное наблюдение, где ограничения по фабрике оказались более существенными, чем ограничения, связанные с процессом Albion Process™.

Наращивание мощности обогатительной фабрики

В течение 2015 года показатели производительности и извлечения на обогатительной фабрике оставались на низком уровне по причине проблем с наращиванием мощности и переработкой переходных руд с неудовлетворительными результатами по флотации. Начиная со второй половины 2015 года в цех флотации стали подавать первичную сульфидную руду, что привело к соответствующему повышению извлечения, содержания и производительности. В середине 2016 года произошло качественное повышение производительности после реализации проекта по устранению «узких мест».

На Рисунке 3 показано, что производительность обогатительной фабрики существенно повысилась с запуска в категориях абсолютного содержания, извлечения и производительности, а также стабильности. Извлечение золота и производство концентрата в цикле флотации остается ниже номинального и проектного уровня (88 процентов, 282 тонны в сутки и 324 тонны в сутки, соответственно), однако содержание золота и серы выше проектных, что позволяет достичь проектной переработки золота. Единицы золота, которые не флотируются, обычно представлены несульфидным типом и лучше поддаются переработке в цикле прямого цианирования, поэтому данные хвосты флотации направляются непосредственно в цикл CIL.

Наращивание мощности установки Albion Process™

В течение периода после запуска эксплуатационные показатели установки Albion Process™ оставались на стабильном уровне выше проектного, производилось окисление всех видов питания, загружаемого в процесс, несмотря на существенные колебания в качестве и количестве питания, как упоминалось в предыдущем разделе. Ограничивающим фактором для установки Albion Process™ является загрузка концентрата, установка работает на проектной скорости в почасовом выражении до сработки всего питания. Это способствует эффективной и стабильной работе мельницы IsaMill™.

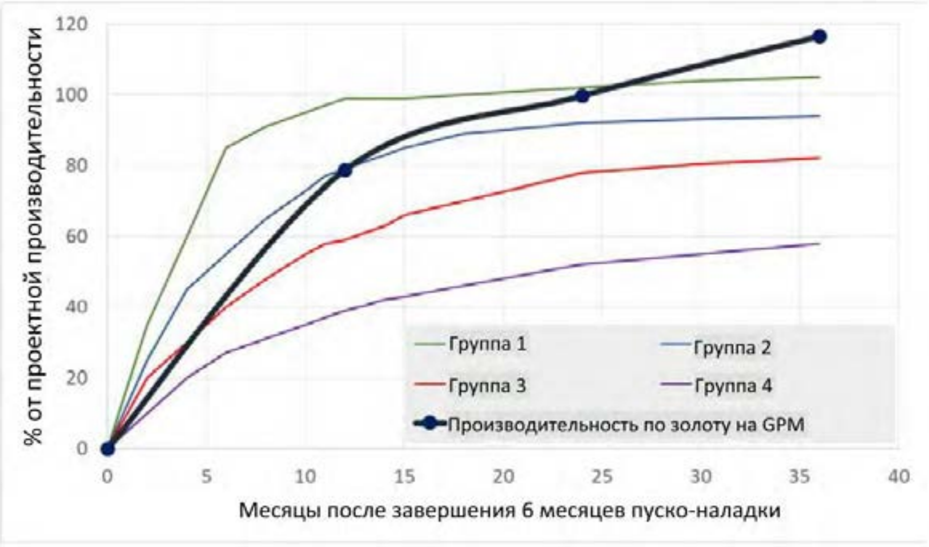


Рисунок 2 – Шкала МакНалти по технологической установке ГЕОПРОМАЙНИНГ.

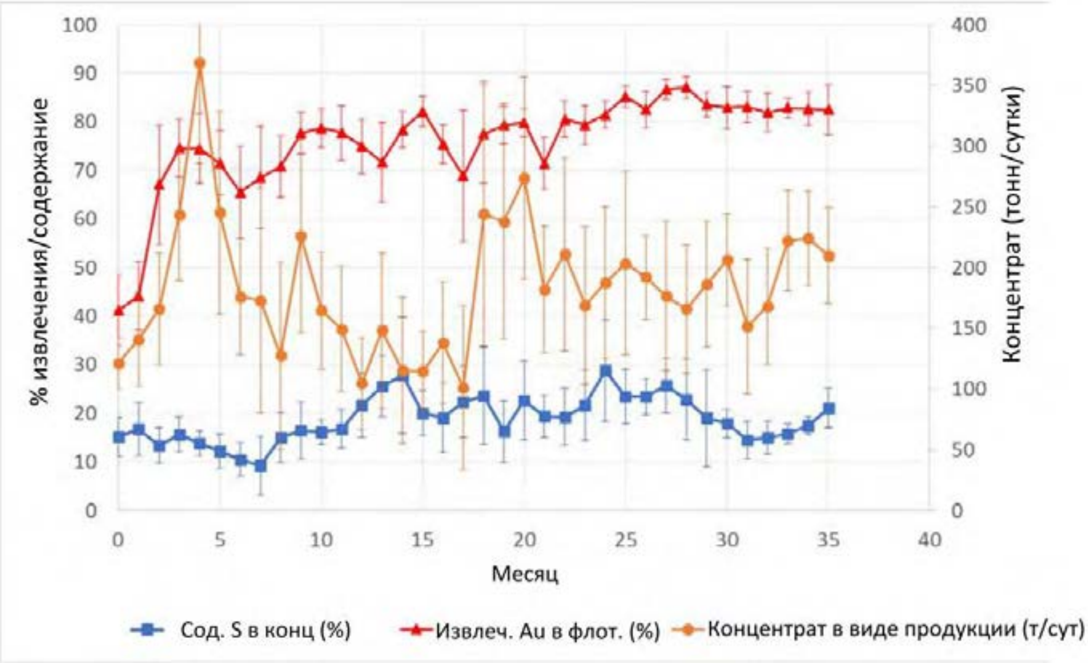


Рисунок 3: Показатели работы ОФ.

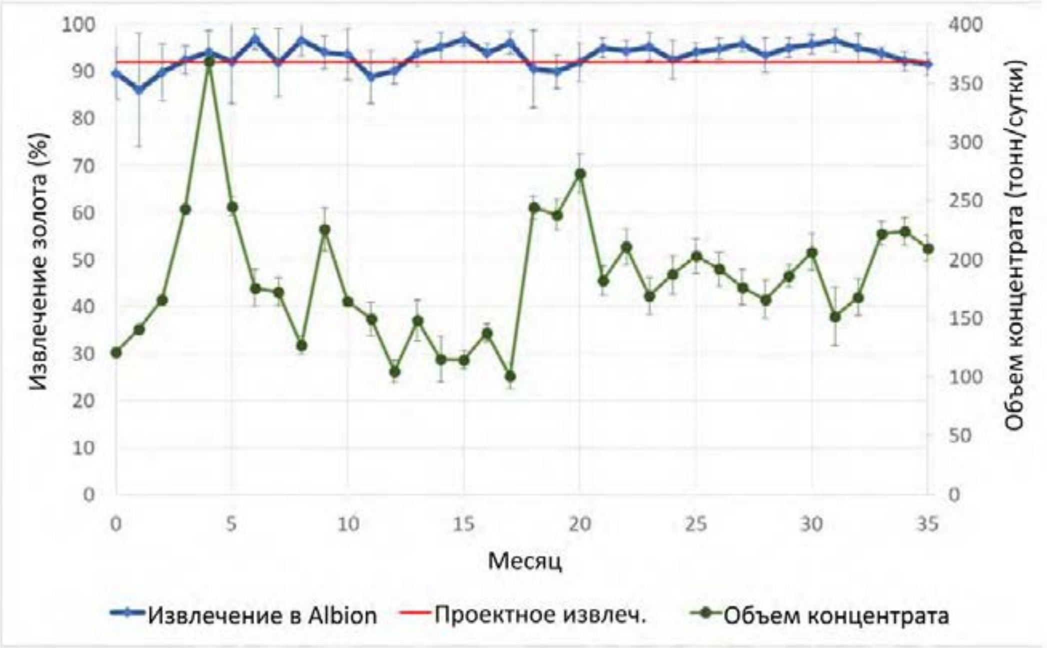


Рисунок 4 - Зависимость извлечения золота от производительности.

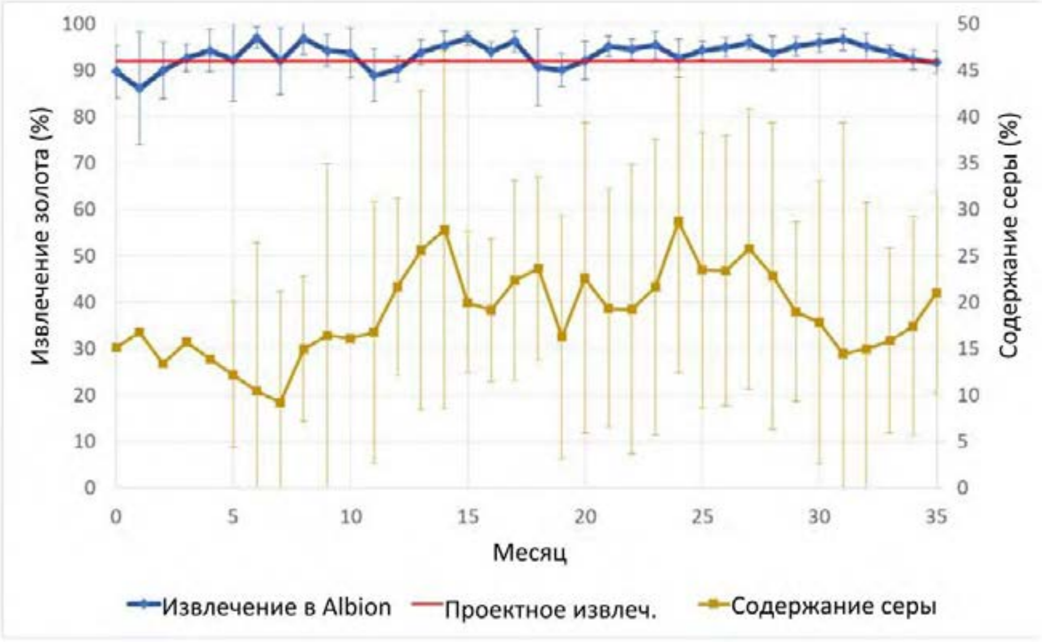


Рисунок 5 - Зависимость извлечения золота от содержания серы в концентрате.

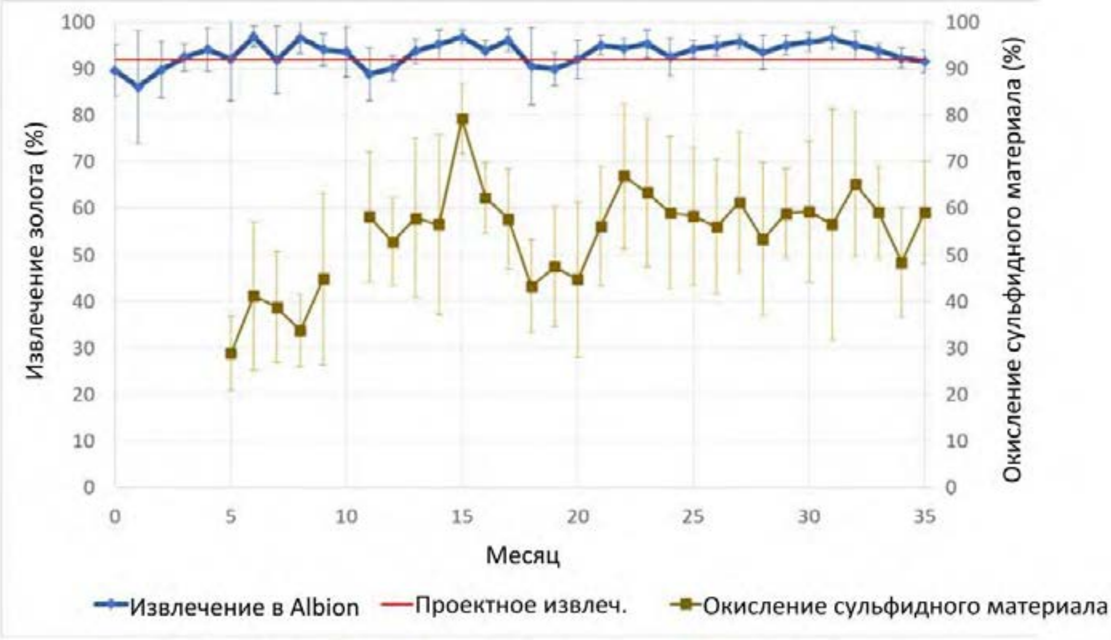


Рисунок 6 - Зависимость извлечения золота от степени окисления сульфидов.

ГИБКОСТЬ, НАДЕЖНОСТЬ И СТАБИЛЬНОСТЬ

Ключевой характеристикой данных является стабильность показателей процесса Albion Process™ при переработке широкого диапазона объемов и качества питания в плане содержания серы и золота. Показатели по извлечению золота оставались стабильными с момента запуска, преимущественно на уровне выше проектного.

Стабильные значения извлечения золота поддерживались невзирая на непредсказуемые колебания в объемах питания установки Albion Process™ и существенные колебания в содержании серы в питании. Показатели ниже служат графическим представлением постоянства показателей извлечения золота в процессе Albion Process™ относительно изменчивости данных параметров.

Колебания в объемах сократились по мере отработки производственного процесса. Тем не менее, извлечение золота оставалось на стабильном уровне даже в начале эксплуатации установки, что дает преимущество оператору установки, так как с ранних этапов ведется работа по совершенствованию управления технологическим процессом.

Способность установки Albion Process™ перерабатывать сырье с непостоянным содержанием серы на производстве ГЕОПРОМАЙНИНГ обусловлена автотермическими свойствами реакторов Albion.

При работе с сырьем с минимальным содержанием сульфидной серы, рабочая температура реакторов находится в диапазоне 85-98°C. Если на установку подается питание с низким содержанием сульфидов, обычно существует возможность несколько часов отработать без существенных последствий по причине большой тепловой инерции в процессе.

К этому времени проблемы, вызвавшие низкое содержание серы на обогатительной фабрике, обычно решаются. При сохранении низкого уровня серы и падении температуры в первых нескольких реакторах, первый реактор может быть частично опустошен за счет слива пульпы в обвалованный участок под сгустителем, после чего она возвращается в технологическую схему насосами пролива. Это позволяет заполнить реакторы материалом с более высоким содержанием с повышением температуры для восстановления нормального режима работы. Реакторы на ГЕОПРОМАЙНИНГ открытого исполнения, поэтому при повышении подачи сульфидной загрузки, дополнительное экзотермическое тепло, вырабатываемое в ходе окисления, приводит к повышению давления пара пульпы, что в свою очередь способствует повышению скорости испарения, снимая излишки тепла. Реакторы на ГЕОПРОМАЙНИНГ функционируют при температуре воды, близкой к точке кипения.

В этом состоит отличие от циклов автоклавного окисления, которые обычно чувствительны к колебаниям в содержании серы ввиду необходимости регулировать тепловой эффект реакции в автоклаве. На этапе разработки экологической части технико-экономического обоснования проекта были приняты более стабильные объемы и содержание серы для сравнения процессов автоклавного выщелачивания и Albion Process™. Выбор технологии Albion Process™ частично обусловлен этой способностью справляться с непредсказуемыми изменениями в питании, вытекающими из состояния рудного тела или показателей фабрики; выбор автоклавного цикла переработки мог привести к существенному снижению объемов для решения вопроса изменчивости руды, с которым производство столкнулось после запуска проекта.

Кроме того, при разработке проекта для разных рудных тел требовалось установить разные уровни окисления сульфидной серы. Выбор кислородной установки по технологии вакуумной короткоцикловой абсорбции (VPSA) позволяет понижать уровень производства кислорода для обеспечения необходимого уровня окисления и последующей потребности системы в кислороде. Так как потребность в кислороде ниже проектной, реакции выщелачивания обычно завершаются в реакторе выщелачивания Albion №6, без необходимости использовать все 9 реакторов.

Производительность по окислению

Следует также отметить, что пересмотрена степень окисления сульфидной серы для достижения и превышения проектного извлечения, которая обычно составляет примерно 55 процентов, намного ниже принятого проектного значения в 76 процентов. Установка Albion Process™ в целом более приспособлена для регулирования степени окисления, чем цикл автоклавного окисления. Наблюдается сокращение эксплуатационных расходов на кислород ввиду более низкого потребления кислорода.

Основной функцией процесса Albion Process™ является окисление серы для высвобождения золота, вкрапленного в сульфидную матрицу, с возможностью извлечения в следующем цикле цианидного выщелачивания. Окислительная способность установки Albion Process™ ограничивается либо временем пребывания в технологическом контуре, либо наличием кислорода, этот фактор отдельно учитывался при инженерно-техническом проектировании установки на этапе разработки. Номинальная и проектная производительность по окислению составляет 1,56 и 1,80 тонн сульфидной серы в час, соответственно, принимая коэффициент использования кислорода равный 80 процентам.

По проекту установка ГЕОПРОМАЙНИНГ Albion Process™ должна обеспечить 76 процентов окисления сульфидной серы при номинальном и проектном окислении серы в 37,6 и 43,2 тонны в сутки. Проектный расход кислорода принят на уровне 336 кг/тонну концентрата при 80% коэффициенте использования кислорода. Фактический расход составил 215 кг/тонну концентрата исходя из опыта эксплуатации установки, по причине более высокой степени использования кислорода и низкой степени окисления (Voigt, Hourn, Mallah and Turner, 2015). Судя по предоставленным технологическим данным установки, степень окисления регулировалась для обеспечения извлечения золота, что свидетельствует о гибкости процесса.

В 2016 и 2017 годах наблюдались периоды, когда степень окисления превысила проектные максимальные значения. Нереализованная производительность окисления в технологии Albion Process™ свидетельствует о том, что в проекте был заложен резерв, который достигается за счет превосходных показателей сверхзвуковых газовых эжекторов HyperSparge™. Сверхзвуковое введение газа позволяет максимально повысить механическое воздействие на пульпу, в результате снижается сопротивление массовому переносу кислорода из газа в жидкость, что в свою очередь повышает степень массового переноса кислорода. Фактический коэффициент использования кислорода был определен на уровне 90 процентов или выше, исходя из проведенных исследований, эти данные описываются в работе Войта, Малла и Хорна (2017 год).

ИЗНАЧАЛЬНО БЕЗУПРЕЧНОЕ ПРОЕКТИРОВАНИЕ

Фундаментом быстрого наращивания мощности и достижения проектной производительности и извлечения служит успешно реализованный процесс инженерно-технического проектирования, от лабораторных испытаний до полномасштабной промышленной установки. Установка на заводе ГЕОПРОМАЙНИНГ была спроектирована и поставлена компанией Glencore Technology на основании результатов испытаний в периодическом и непрерывном режиме работе, проведенных в рамках банковского ТЭО под началом Core Resources в течение 2011 и 2012 годов. Цель проведения испытаний заключалась в получении следующих ключевых параметров проектирования, которые привели к определению критериев массового и энергетического баланса и технологического проектирования:

- оптимизация крупности измельчения для сокращения энергопотребления IsaMill™,
- составление графиков энергопотребления IsaMill™ для определения необходимой энергии измельчения для достижения определенной крупности частиц, чтобы рассчитать размер мельницы IsaMill™ (Проектная удельная энергия для обеспечения крупности 80% класса 10мкм составила 59,0 кВт.ч/тонну, по последним фактическим данным это значение составляет 59,7 кВт.ч/тонну, хотя крупность 10 мкм требуется не на постоянной основе),
- зависимость извлечения золота от степени окисления сульфидов для оптимизации извлечения золота и снижения потребности в окислении;
- измерение степени поглощения кислорода в пульпе при различных условиях для проектирования системы массового переноса кислорода,
- реологические измерения для определения типоразмеров насосов,
- определение чувствительности кинетики сульфидного выщелачивания к ключевым параметрам, таким как рабочая температура и плотность пульпы,
- данные по скорости осаждения для расчета типоразмера сгустителя, и
- испытание технологических показателей на образцах, которые являются представительными в отношении возможных колебаний в питании в течение срока службы установки.

Испытания строились на определении металлургических показателей типов руды и их смесей исходя из графика отработки рудника. При проведении испытаний были получены разные результаты по флотации разных типов руды, потребовалось отрегулировать окисление для обеспечения извлечения золота в 90 процентов в процессе цианирования.

Подход к проектированию заключался в принятии консервативного целевого значения окисления сульфидной серы в 76 процентов, чтобы предусмотреть возможность переработки различных сочетаний типов руды с достижением 90 процентов извлечения золота в процессе CIL. Объем чанов и типоразмер кислородной установки подбирались исходя из целевой степени окисления. Так как на кислород приходится существенная доля эксплуатационных затрат, для производства кислорода были выбраны две установки вакуумной короткоцикловой адсорбции (VPSA), с возможностью снижения выработки кислорода при отсутствии потребности и сокращения эксплуатационных затрат (как показано на Рисунке 6). Технология VPSA предпочтительнее технологии PSA (короткоцикловой адсорбции) ввиду более низкой стоимости энергии на единицу производимого кислорода. Степень чистоты кислорода составляет 93% (об.).

Важным элементом при проектировании окислительных процессов является массовый перенос кислорода в пульпе. Для успешного функционирования процесса особенно важно масштабирование массового переноса кислорода, этот аспект хорошо изучен и отработан посредством применения технологии сверхзвуковой подачи газа HyperSparge™ (Voigt, 2017). Консервативный коэффициент использования кислорода принимается на уровне 80 процентов, однако измерения на установке свидетельствуют о том, что данный показатель приближается к 90 процентам (Voigt, 2015).

На Рисунках 2-6 можно увидеть, что в месячной перспективе не наблюдается падения в производительности. Стабильность производительности говорит о продуманном проектировании установки. К числу отдельных факторов, которые способствовали высокой эксплуатационной готовности установки, относятся:

1. Не потребовалось проведение крупных работ или технологических модификаций после запуска, деятельность направлена преимущественно на вывод процессов измельчения и флотации на полную производительность;
2. На обогатительной фабрике работает две линии, возможно сохранение производства

на одной линии на период перефутеровки мельницы; при проектировании технологии Albion Process™ предусматривалось исключение длительных простоев за счет следующих факторов:

- буферные емкости вокруг IsaMill™,
- возможность демонтировать, осматривать и устанавливать элементы HyperSparge™ без необходимости остановки или опустошения рееактора выщелачивания,
- возможность обхода секции выщелачивания, и
- использование изученного оборудования (реакторы, насосы, сгустители, мешалки).

В результате подобного проектирования достигается очень высокая эксплуатационная готовность (более 95%) и стабильный выпуск продукции.

ПЕРЕДАЧА ТЕХНОЛОГИИ

Albion Process™ представляет собой зарекомендованную технологию, но ранее данный процесс не применялся в Армении или странах СНГ. Это означало необходимость обучения производственного персонала и инженеров по техническому обслуживанию для обеспечения безопасного и своевременного выхода на проектные показатели. Обучение проводилось в рамках аудиторных занятий на последних этапах строительства и впоследствии на установке при проведении пуско-наладки и ввода в эксплуатацию. Обучение проводилось экспертами Glencore Technology и представителями субпоставщиков.

Проведение обучения до ввода в эксплуатацию позволило Glencore Technology (GT) выполнять функции надзора за пуско-наладочными работами с минимальным присутствием персонала (три сотрудника), в то время как сотрудники ГЕОПРОМАЙНИНГ выполняли управление установкой. Благодаря этому достигнута передача ответственности за управление технологическим процессом владельцу проекта на самых ранних стадиях.

После запуска и выхода на проектную мощность, представители Core Resources и Glencore Technology посещали проект дважды в год, в результате обеспечено регулярное рассмотрение возникающих технических вопросов.

В рамках процесса передачи технологии и ответственности за технологический процесс, ГЕОПРОМАЙНИНГ внедрили несколько технологических модификаций и улучшений:

1. Перепуск в обход сгустителя

Осадок выщелачивания из процесса Albion Process™ направляется на сгуститель для повышения содержания твердого с 30 до 42 процентов, приемлемого для процесса CIL. Так как осадок установки Albion Process™ смешивается с хвостами флотации и направляется на процесс CIL, был предусмотрен обход сгустителя Albion Process™, а необходимые характеристики плотности питания CIL обеспечивались за счет повышения плотности песков сгустителя хвостов флотации.

2. Сокращение количества реагентов и замкнутый контур слива на реакторах

Конфигурация реакторов выщелачивания Albion Process™ включает связанные между собой желоба для переноса пульпы, каждый реактор оснащен системой перелива, в которой предусмотрен сливной порог для сохранения герметичности свободного пространства под крышкой, чтобы в вентиляционную трубу выводились только газы. В случае высокого содержания карбонатов в питании или скопления флотационных реагентов в технологической воде, возможны случаи перелива из реакторов по причине избыточного пенообразования. Предусматривается перенос перелива между реакторами по желобам, но зачастую он следует по пути наименьшего сопротивления через сливные патрубки реактора. Хотя материал возвращается в процесс за счет системы обвалования и сбора проливов, иногда это приводит к снижению скорости подачи кислорода в проблемном реакторе и повышению в другом реакторе, вызывает пульсацию газа и дополнительные проблемы с пенообразованием. Вторичная переработка пролива также вызывает технологические всплески.

Производственный персонал реализовал проект, который предполагает постепенное сокращение флотационных химических реагентов (преимущественно вспенивателей) и закрытие сливных порогов. Пенообразование существенно сократилось за счет уменьшения количества добавляемых реагентов на обогатительной фабрике. Сливные пороги также были заварены, таким образом вся пена направляется на соединенные желоба чанов. Внедрение этих двух мероприятий позволило устранить проблему перелива пены.

3. Система управления

Технология Albion Process™ поставлена с распределенной системой управления (РСУ), управление установкой осуществляется из центральной диспетчерской. Существующие циклы измельчения и CIL управлялись при минимальной оснащенности КИП и полевыми пультами управления, поэтому переход к централизованной системе управления оказался новой концепцией. При реализации проекта Albion Process™, установки измельчения и флотации были интегрированы в РСУ для полного централизованного управления начиная с подачи руды до выпуска продукции с установки Albion Process™.

4. Теллуриды

На начальных этапах испытаний было выявлено наличие золота, вкрапленного в теллуриды, и судя по данным разведочного бурения, через несколько лет добычи преимущественно сульфидов, золото будет встречаться в этой форме. В связи с этим, при проектировании технологии Albion Process™ GT заложили возможность повышения pH до 9,0 в последнем реакторе выщелачивания, что позволяет выполнять окисление при высоком pH с целью окисления и высвобождения золота, заключенного в теллуриды. Эта функция также имеет дополнительное преимущество, связанное с подготовкой продукта к процессу CIL, рабочий уровень pH которого составляет примерно 10,5. Так как процесс Albion Process™ функционирует при pH 5,5, подобное технологическое изменение выполняется просто. Кольцевая система подачи известняка от цикла CIL была расширена до реактора выщелачивания Albion Process™ №9, обеспечивая работу при более высоком pH и извлечение золота и серебра из теллуридов. В загружаемом концентрате наблюдается присутствие теллуридов в объеме от 80 частей/млн до 1200 частей/млн. Изначально на установке наблюдалось существенное понижение извлечения золота (примерно на 2-3 процента). Данная проблема была решена повышением уровня pH в последнем реакторе выщелачивания.

СОГЛАШЕНИЕ О СОВМЕСТНОЙ ДЕЯТЕЛЬНОСТИ

В 2017 году ГЕОПРОМАЙНИНГ и Glencore Technology подписали соглашение о совместной деятельности с целью продвижения технологии Albion Process™ на российском рынке. Это значимая веха в сотрудничестве двух компаний так как позволит новым компаниям, приобретающим технологию Albion Process™, эффективно использовать опыт сотрудничества по внедрению процесса Albion Process™ на заводе ГЕОПРОМАЙНИНГ. Положения соглашения предусматривают посещение сотрудниками новых проектов завода ГЕОПРОМАЙНИНГ для изучения технологии Albion Process™ и обучения персонала эксплуатации и техническому обслуживанию.

ПРИЗНАТЕЛЬНОСТЬ

Авторы выражают благодарность владельцам ГЕОПРОМАЙНИНГ в Армении за предоставленные данные.

СПИСОК ЛИТЕРАТУРЫ

Hourn, M, and Turner, D, 2012. Commercialisation of the Albion Process, in *Proceedings ALTA Gold Conference 2012, pp 231-248* (ALTA Metallurgical Services: Melbourne). [Промышленное внедрение технологии Albion]

McNulty, T, 2014. Plant Ramp-Up Profiles: An update with emphasis on process development, in *COM 2014 – Conference of Metallurgists Proceedings* (Canadian Institute of Mining, Metallurgy and Petroleum: Westmount) [Профили выхода установок на проектную мощность: Актуализация информации с упором на разработку технологии]

Voigt, P, Hourn, M, Mallah, D and Turner, D, 2014. Development of the Albion Process plant to treat refractory concentrates from the ГЕОПРОМАЙНИНГ Gold Project, in *Proceedings – Hydroprocess Conference,* 2014, Chapter 5 (Gecamin: Santiago). [Разработка установки по технологии Albion для переработки упорных концентратов на золотодобывающем проекте ГЕОПРОМАЙНИНГ]

Voigt, P, Hourn, M, Mallah D and Turner D, 2015. Commissioning and Ramp-Up of the Albion Process at the ГЕОПРОМАЙНИНГ Gold Project, in *Proceedings MetPlant 2015*, pp 207-219 (The Australasian Institute of Mining and Metallurgy: Melbourne). [Ввод в эксплуатацию и наращивание мощности установки Albion на золотодобывающем проекте ГЕОПРОМАЙНИНГ]

Voigt, P, Mallah, D and Hourn, M, 2017. Oxygen Mass Transfer in the Albion Process™: From the Laboratory To The Plant, in *Proceedings ALTA 2017*, pp 135-144 (ALTA Metallurgical Services, Melbourne, Australia). [Массовый перенос кислорода в процессе Albion Process™: От лаборатории до промышленной установки]

Voigt, P and Walker, D, 2017. The Albion Process™ at the ГЕОПРОМАЙНИНГ Gold Project – The Success of a Technology, in *Proceedings CIM Convention 2018*, (Canadian Institute of Mining, Vancouver, Canada). [Установка Albion Process™ на золотодобывающем проекте ГЕОПРОМАЙНИНГ – Успешное применение технологии]

The Albion Process™ at the GPM Gold Project – The success of a technology

Paul Voigt, Glencore Technology (paul.voigt@glencore.com.au)
David Walker, Core Resources (dwalker@coreresources.com.au)
CIM Conference 2018,
Stream: A Vision For The Future - Mineral Processing In Canada

www.albionprocess.com

GPM Gold – The success of a technology

Outline

1. Albion Process™ technology

2. Performance of the Albion Process™ at GPM

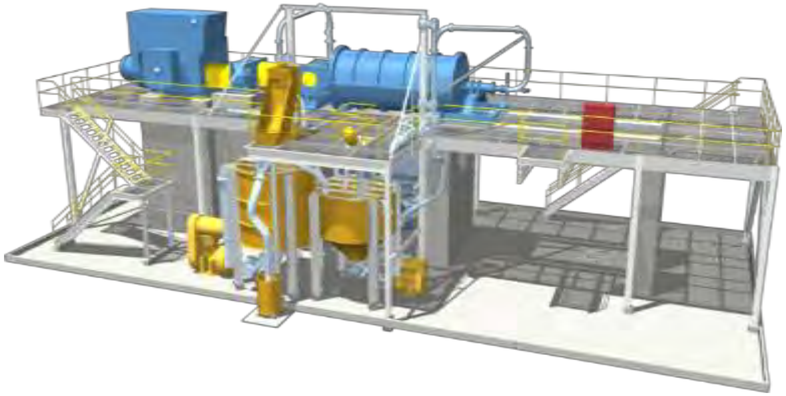
3. Review of global installations

4. The Future for Refractory Gold Projects

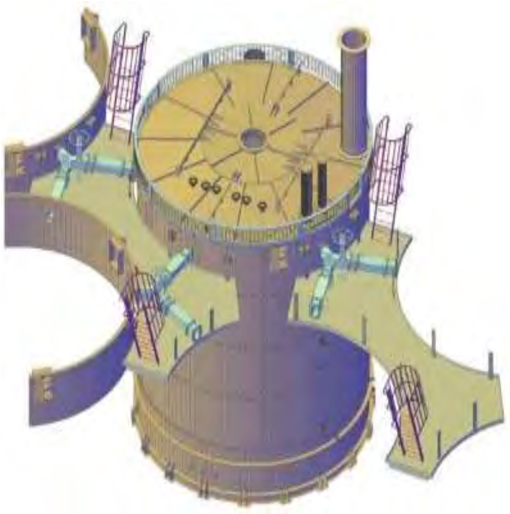
1. Albion Process™ Technology

Albion Process™

The Albion Process is a combination of mechanical and chemical liberation

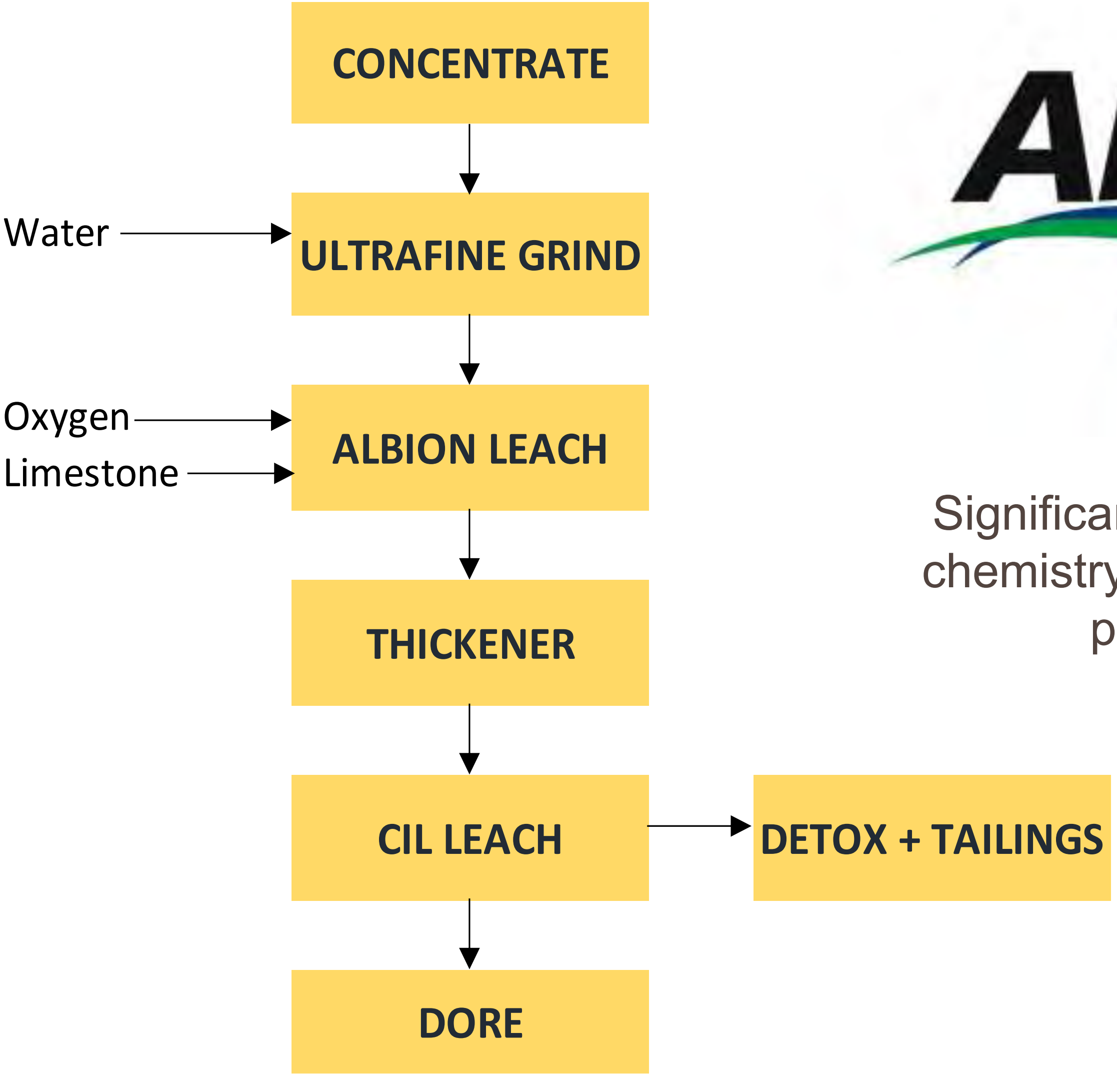


- **Ultrafine grinding:**
  - *IsaMill^{TM} stirred mill:*
  - *FeS_2 = 80 % passing 10 microns*
  - *$CuFeS_2$ = 80 % passing 12 – 18 microns*
  - *$Ni_9Fe_9S_{32}$= 80 % passing 10 – 14 microns*
  - *ZnS = 80 % passing 16 – 20 microns*



- **Oxidative Leaching:**
  - *Atmospheric pressure leach*
  - *Gold Applications – pH = 5.5 (“Neutral Albion Leach”)*
  - *Conventional baffled tank (Modular)*
  - *Sulphate solutions - no chlorides*
  - *Supersonic oxygen injection*

Gold Albion Process Flowsheet



CONCENTRATE

Water → ULTRAFINE GRIND

Oxygen → ALBION LEACH

Limestone → ALBION LEACH

THICKENER

CIL LEACH → DETOX + TAILINGS

DORE

ALBION PROCESS™

Significant detail of Albion Process chemistry and design is available in published literature.

Albion Process™ - Low Process Risk

All components of the technology are well demonstrated:

IsaMill

- > 130 IsaMills in operation globally
- 17 mills in gold re-grind applications

Atmospheric Leach

- 6 operating Albion Process Oxidative Leach Pants
- Zinc, lead, gold and copper plants
- 700,000 tpa of sulphide concentrate currently processed

Oxygenation System

- >520 HyperSparge units installed in Albion Process and other oxidative leach applications
- Extremely successful at high utilisation of oxygen of 90+%, above design

2. GPM Project - Albion Process™ Plant Performance

GPM Gold – The success of a technology

The GPM Gold Project

- Armenian gold project, owned by GeoProMining LLC
- Open cut mine - 1 Mtpa ROM, 14.5 Mt reserves
- Historical grinding & flotation plant (1976) + CIL plant (1997)
- Oxide ores exhausted 2012, gold now refractory in pyrite

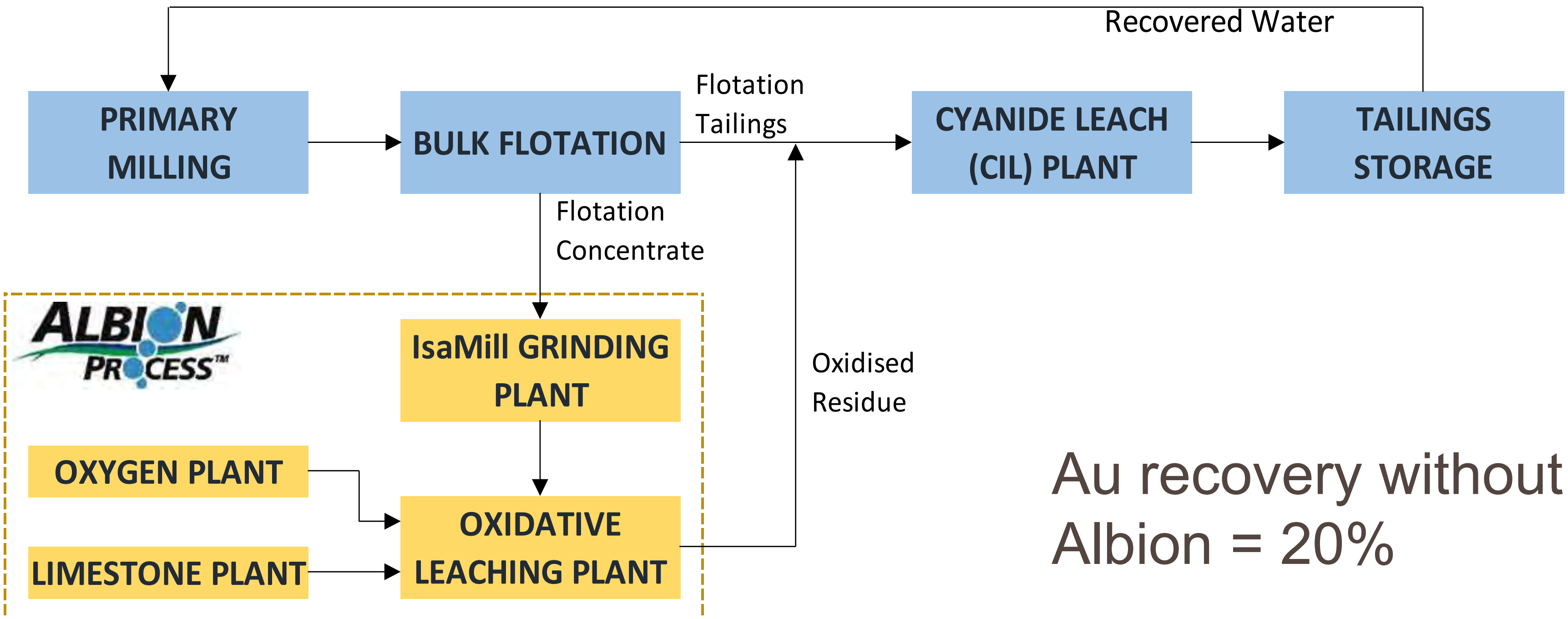
GPM Gold – The success of a technology

Refractory gold plant installed 2012

- Albion Process™ technology installation
- BFS and Pilot Plant completed 2011 by Core Resources
- Gold recoveries of 95%+ (up from 20% recovery without Albion Process™ treatment).
- Low skill workforce
- Plant tolerates highly variable throughput, sulphur grades and climate
- 100,000 tpa concentrate, producing 120,000 ozpa gold
- Plant production at up to 130% of nameplate design

GPM Albion Process™ Plant Performance

Process Plant Overview

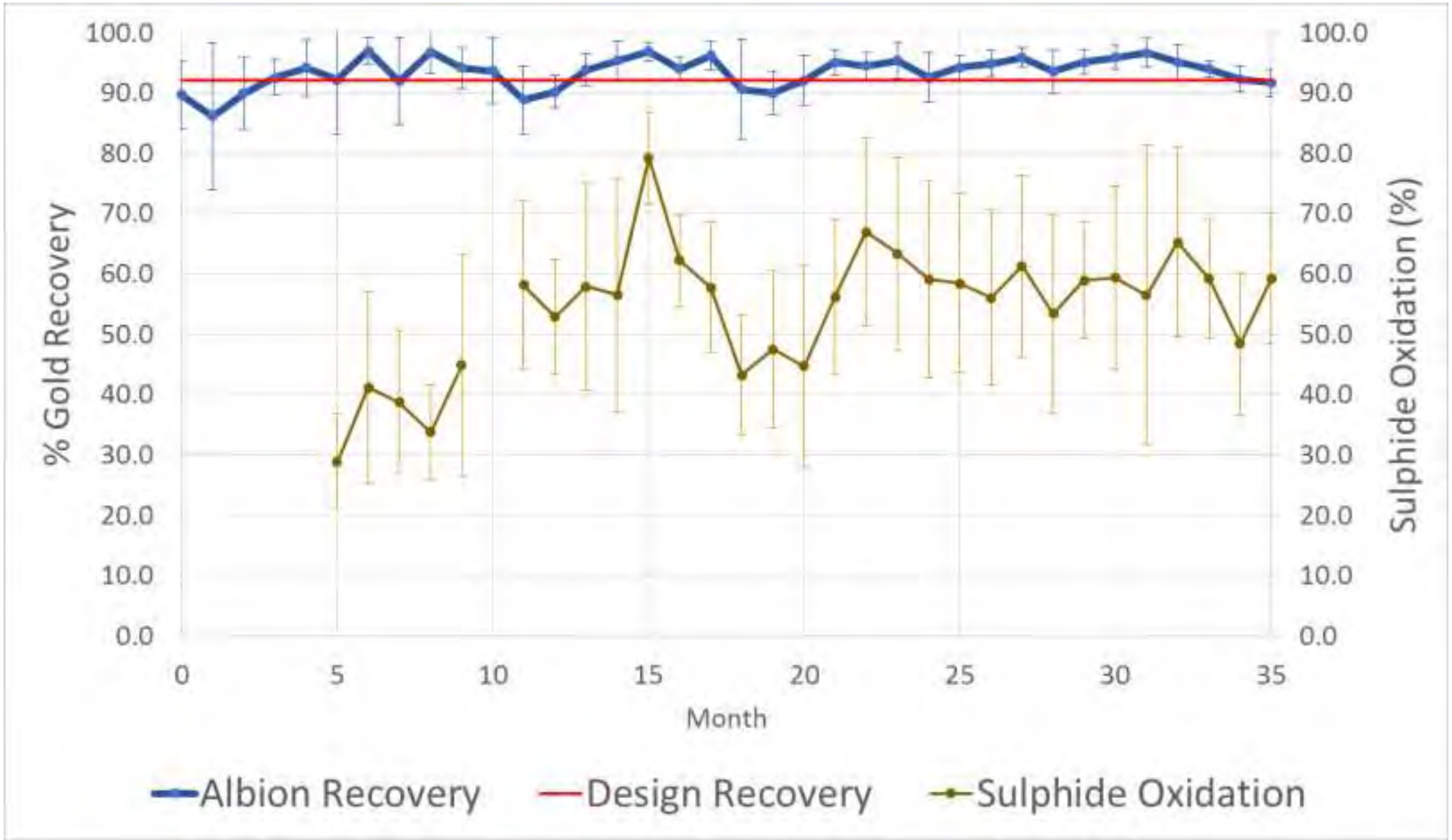


Au recovery without Albion = 20%

Au recovery with Albion = 95-98%

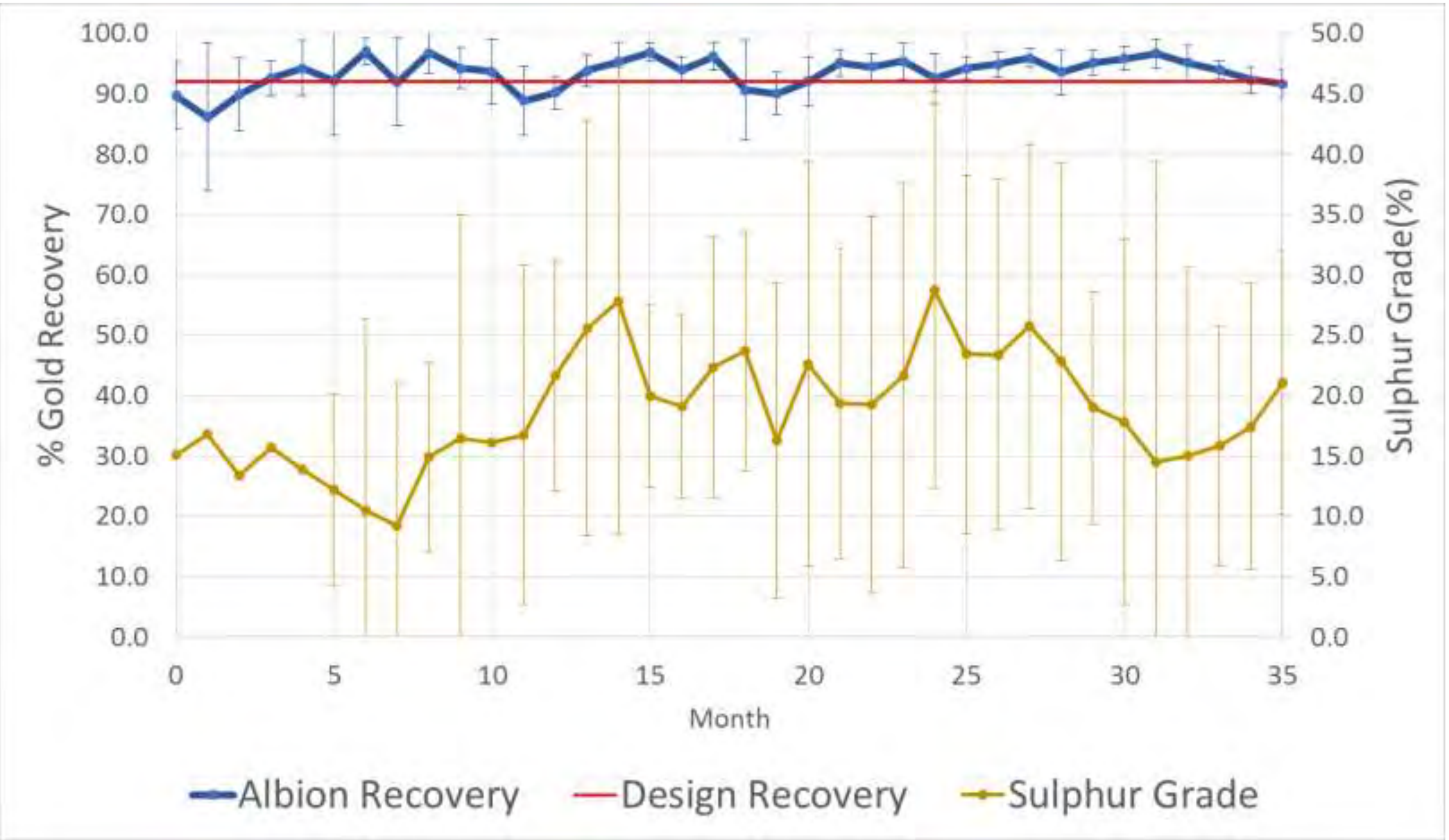
GPM Albion Process™ Plant Performance

3 years of stable gold recoveries



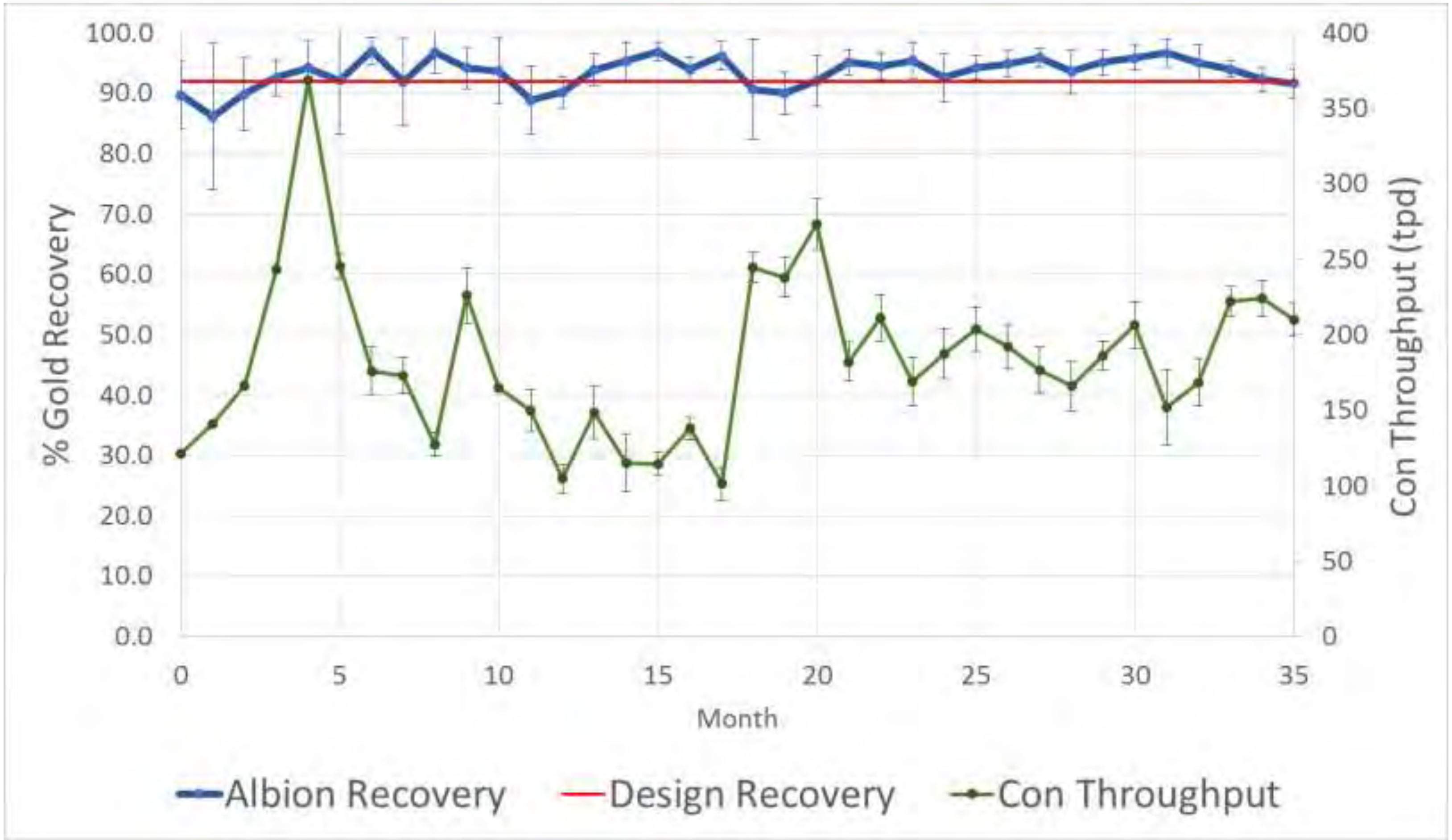
GPM Albion Process™ Plant Performance

Variable Sulphur Grades (<10% to >30%)



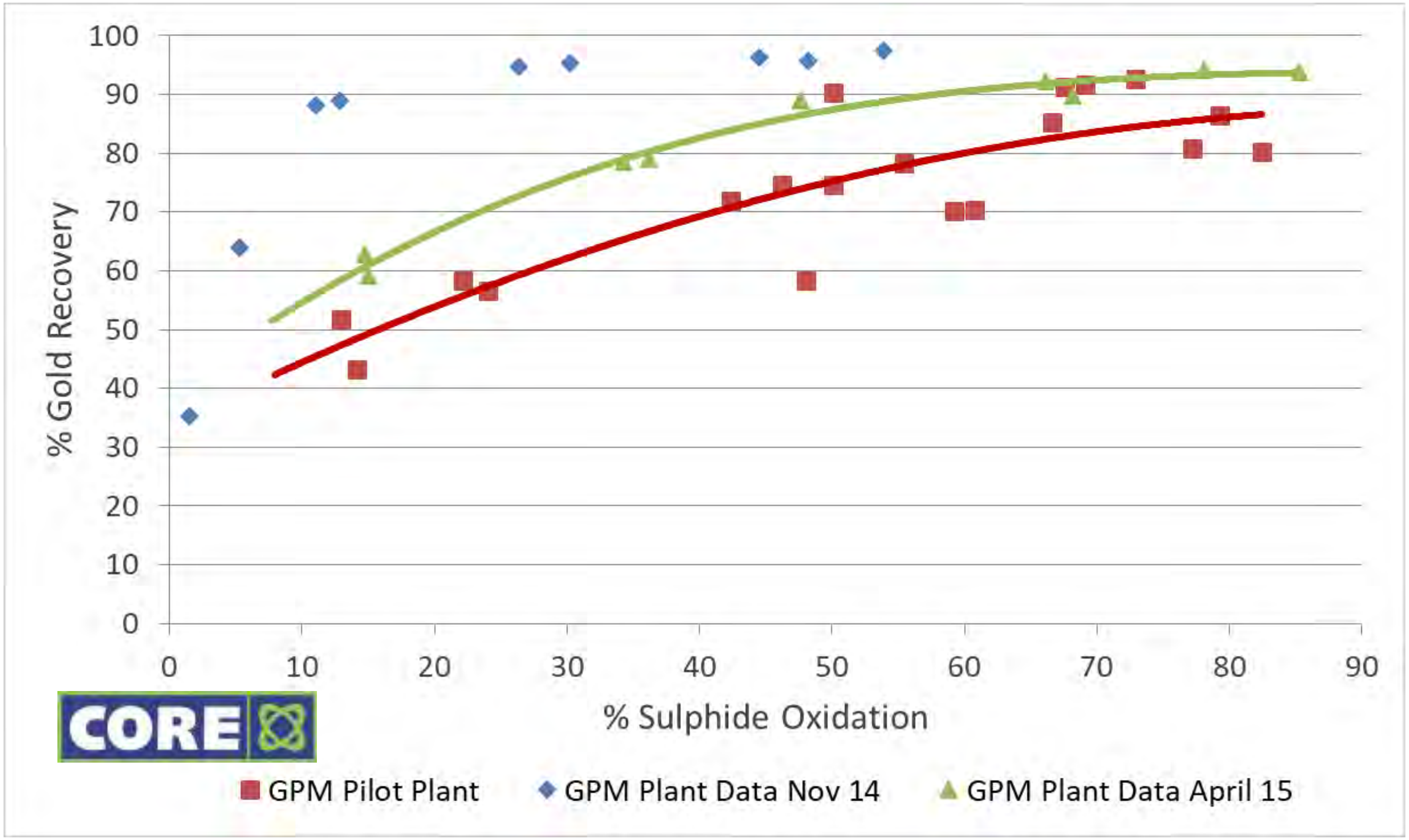
GPM Albion Process™ Plant Performance

Variable Throughput (100 – 350 tpd)



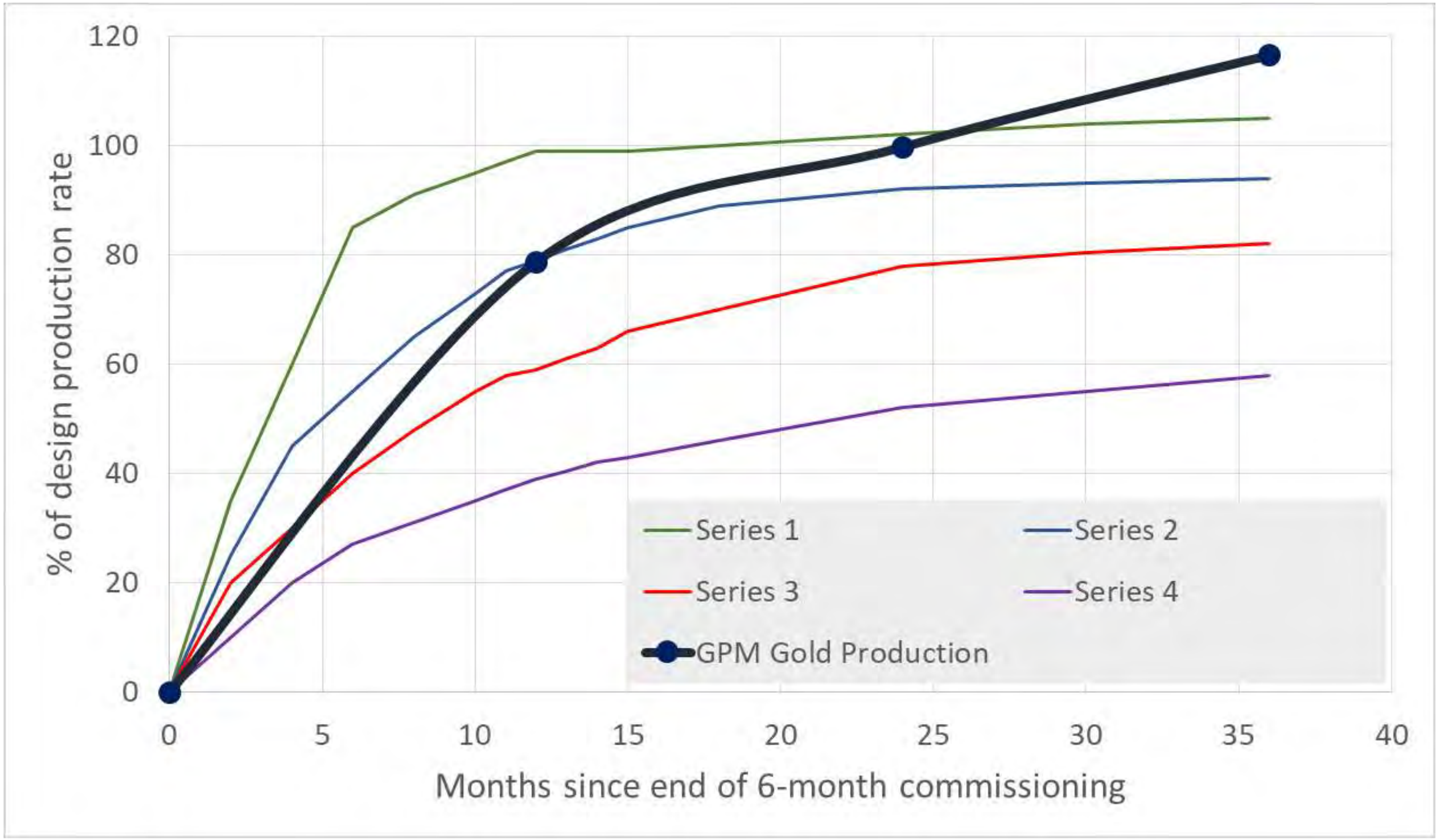
GPM Albion Process™ Plant Performance

Better than Pilot Plant



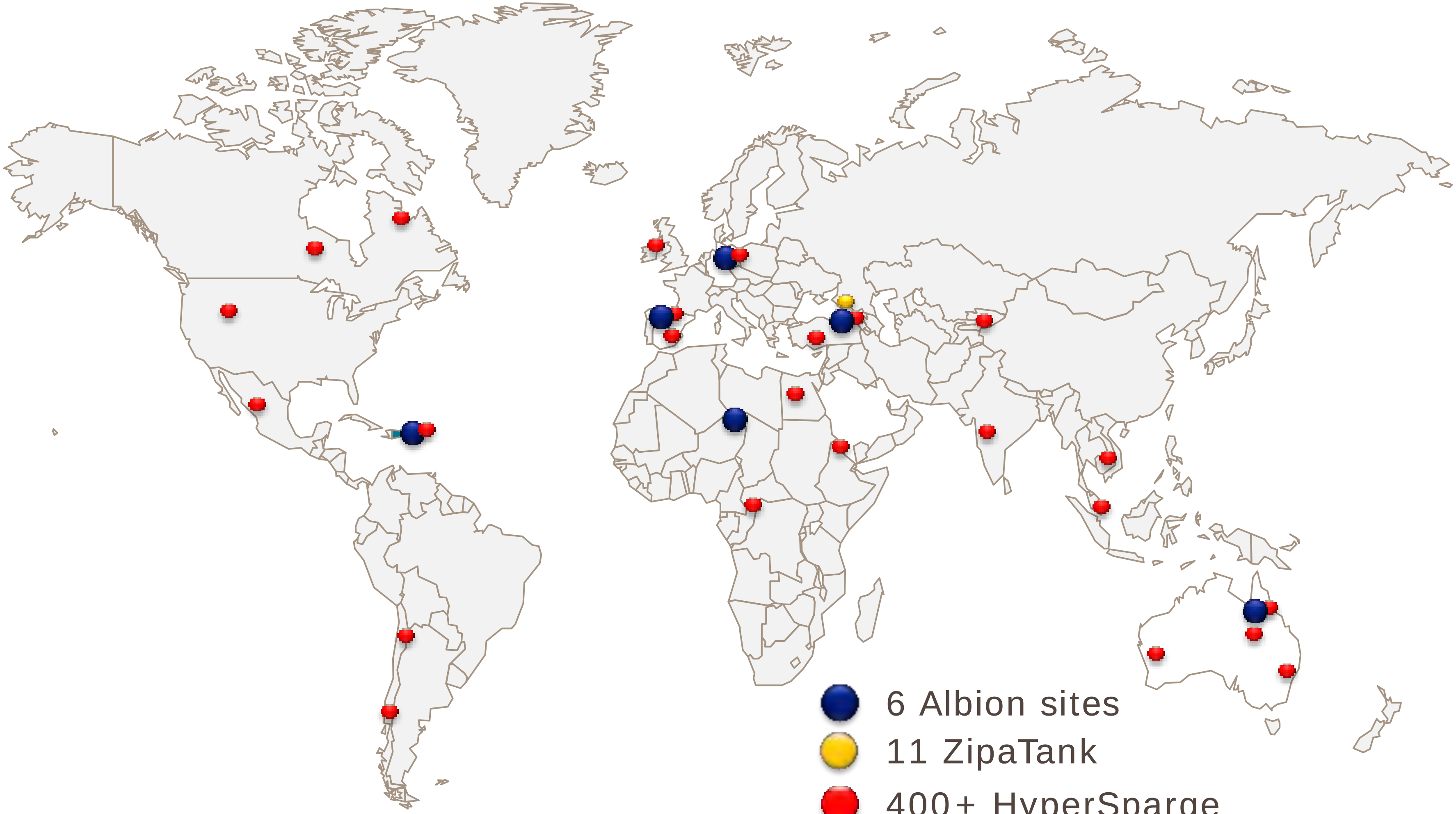
GPM Ramp Up – Relative McNulty Curve

Series 1-2 Performance



3. Albion Process™ - Global Installations

Albion Process™ Installations



● 6 Albion sites

● 11 ZipaTank

● 400+ HyperSparge

GPM Gold Project – Albion Process™ Plant

Oxidative Leach Circuit

M5000 IsaMill™

Limestone Grinding Plant

Commodity – Gold

Location – Armenia

Client – GeoProMining

Refractory pyrite concentrate

120,000 ozpa gold

Commissioned March 2013

Application: Recovery of precious metals from a refractory arsenic bearing deposit within the setting of a soviet era mining complex

Las Lagunas Tailings – Albion Process™ Plant

M5000 IsaMill™ being installed

HyperSpargers™ oxygen addition

First gold pour from Albion

Commodity – Gold

Location – Las Lagunas, Dominican Republic

Client – Panterra

Complex arsenopyrite/gold tailings

80,000 ozpa gold

Commissioned in 2012

Application: Albion Process required to recover gold from complex matrix in tails dam (80% recovery, up from 35%), and leave arsenic minerals inert

Copper Project – Albion Process™ Plant

Oxidative Leach Circuit

First Copper Cathode Production

Commodity – Copper

Location – Africa

Client – Confidential

Copper Concentrate

10,000 tpa copper cathode

>99 % copper recovery

Commissioning late 2017

Application: Recovery of copper and cobalt from low and medium grade concentrates in the African region

Asturiana de Zinc – Albion Process™ Plant

Oxidative Leach Circuit

Sparging systems

Commodity - Zinc

Location – Spain

Client – Glencore

Bulk lead/zinc concentrate

4,000 tpa zinc cathode

>99 % zinc recovery

Commissioned 2010

Application: Recovery of zinc from a bulk concentrate as electrowon cathode with lead and silver in residue for smelting

Nordenham – Albion Process™ Plant


Oxidative Leach Circuit


Sparging system

Commodity - Zinc

Location – Germany

Client – Glencore

Bulk lead/zinc concentrate

35,000 tpa zinc cathode

>99 % zinc recovery

Commissioned 2011

Application: Recovery of zinc from a bulk concentrate as electrowon cathode with lead and silver in residue for smelting

MRM – Albion Process™ Plant

Oxidative Leach circuit

Off Gas Scrubber

HyperSparge system

Commodity - Zinc

Location – Australia

Client – Glencore

Bulk lead/zinc concentrate

150,000 tpa of cleaned zinc concentrate

Commissioned 2014

Application: Selective oxidation of galena in a bulk concentrate to chemically liberate lead from zinc

4. The Future for Refractory Gold Projects

Demonstrated Alternative

	Albion Process™	POx
Demonstrated high recoveries	✔	✔
Demonstrated in current operations	✔	✔
Guaranteed by technology provider	✔	✔/✕
Lower capital costs	✔	✕
Simple equipment + low skills requirement	✔	✕
Short commissioning and ramp up period	✔	✕
Can treat high carbonate material	✔	✕
Tolerates variable feed rate and quality	✔	✕
High availability and low maintenance	✔	✕

Albion Process – Project Development

Study phase well defined and understood

- Scale up now well understood, less sample and testwork required to define process.
  - Phase 1 – Amenability testwork and Class 5 Engineering Study (+/- 40%)
  - Phase 2 – Further batch testwork and Class 4 Engineering Study
  - Phase 3 – Feasibility study
- Piloting can be conducted if client requires, but not required for process guarantees.
- Study management can be provided by Core Resources (GT's laboratory and marketing partner).
- Basic engineering conducted by Glencore Technology.

Flexible project delivery model

- Can work direct to client or through engineering companies

Albion Process – Technology Access

Information and contacts:

www.albionprocess.com

GLENCORE TECHNOLOGY

Paul Voigt
Manager – Hydrometallurgy, Glencore Technology
paul.voigt@glencore.com.au
+61 7 3833 8500

CORE

David Walker
General Manager – Integrated Solutions, Core Resources
dwalker@coreresources.com.au
+61 7 3637 8100

Thank You.

MOUNT ISA MINES NECESSITY DRIVING INNOVATION

*V. Lawson, H. DeWaal, G. Heferen, N. Aslin, P. Voigt, and M. Hourn

Glencore Technology
Level 10, 160 Ann St
Brisbane, Australia, 4000
*(*Corresponding author: virginia.lawson@glencore.com.au)*

ABSTRACT

Mount Isa Mines (MIM) acquired a reputation for the successful application of R&D to develop break-through technologies for the mining industry starting in the 1978's through until the early 2000's. The ISAPROCESS™ tank-house technology has been licensed to copper refineries throughout the world, and a significant per cent of the world's copper is refined using this technology. Since development in the late 1980's more than 20 ISASMELT™ copper and lead smelting furnaces are now installed in countries around the world. Jameson Cell flotation technology developed jointly by Mount Isa Mines and Professor Graeme Jameson is widely used in the Australian coal mining industry and increasingly in the base-metal and gold industry. The IsaMill™'s developed at Mount Isa and McArthur River made it possible to develop the McArthur River and George Fisher orebodies and has been successfully implemented into base metal fine grinding applications around the world. The most recent commercialised innovation is the atmospheric leach Albion Process™ with its supersonic HyperSparge™ gas sparger, is being adopted as a solution to the increasing complexity of orebodies.

MIM's contribution to the industry was significant given the size and the remote location of its operations with Townsville Copper Refineries more than 1350 km and Mount Isa 1800 km from the nearest state capital of Brisbane. This paper will briefly discuss the development of each of these technologies and why MIM – now owned by Glencore - was so successful innovating and developing such technologies over a period of nearly 40 years.

KEYWORDS

Innovation, Mount Isa Mines, ISAPROCESS™, IsaKidd™, ISASMELT™, IsaMill™, Jameson cell, Albion Process™, HyperSparge™, ZipaTank™

INTRODUCTION

Mount Isa is located in the Gulf Country region of Queensland about 1800 kilometers North West of Brisbane (see Figure 1). It came into existence because of the world class mineral deposits found in the area. In 1923 the orebody containing lead, zinc and silver was discovered by the miner John Campbell Miles. Mount Isa Mines Limited (MIM) was founded in 1924 to develop the minerals discovered by Miles, but production did not begin until May 1931. It paid its first dividend in 1947 after 16 years of troubled production. In 1954 the 1100 copper orebody was discovered and with rapidly rising reserves during the 1950's and 1960's led to the construction of new concentrators to treat lead/zinc/silver ores in 1966 (#2 concentrator) and copper ore's in 1973 (#4 concentrator). The difficult nature of the Mount Isa lead-zinc orebodies has meant that the company had always needed to be at the forefront of mining technology. In the 1970's through to the 1990's, it became a world leader in developing new mining techniques and processing technologies as a response to declining metal prices and rising costs. Mount Isa has been smelting copper since 1953 and lead since the early 1930's. Copper Refining at Mount Isa's fully owned subsidiary of Copper Refineries Proprietary Limited (CRL) had commenced operations in 1959.

Figure 1 – Location of Mount Isa and Townsville relative to Brisbane the nearest Capital City

Technologies to come out of Mount Isa include the ISAPROCESS^{TM} copper refining technology, the ISASMELT^{TM}, The Jameson Cell, the IsaMill^{TM}, the Albion Process^{TM} and the Hypersparge^{TM}. Mount Isa Mines Ltd was acquired by Xstrata in 2003 and Xstrata was then merged with Glencore in 2015. The level of innovation achieved at Mount Isa Mines is unsurpassed and was the result of the difficult nature of the Mount Isa ore bodies and its response to declining metal prices and rising operational costs in the 1970's and 1980's. By the 1990's, Mount Isa had become a world leader in innovative mining techniques and state of the art processing technologies. The processing technologies are discussed below.

INNOVATIONS

Each of the innovations developed at Mount Isa Mines had a driver but the overarching desire was to make technology more efficient and cost effective. Each of these process developments will be discussed separately.

ISAPROCESS™

The development of the ISAPROCESS™ tank house technology had its beginning in the zinc industry. During the mid-1970s, MIM was considering building a zinc refinery in Townsville to treat the zinc concentrate produced by its Mount Isa operations. As a result, MIM staff visited the zinc smelters using the best-practice technology and found that modern electrolytic zinc smelters had adopted permanent cathode plate and mechanised stripping technology. MIM realised that the copper refineries performance was constrained by the conventional practice of copper starter sheets. The preparation of these copper starter sheets was labour intensive and the overall cycle was several weeks in duration.

MIM initiated a research program aimed at developing similar permanent cathode technology for copper refining. CRL, a subsidiary of MIM, had been operating in Townsville since 1959, using conventional starter-sheet technology and treating blister copper produced in the copper smelter at Mount Isa. Permanent cathode technology was developed and adapted over many years of in-plant experimental work and successfully introduced to the Townsville refinery in 1978. . The fundamental difference between the new ISAPROCESS™ and the conventional starter sheet technology is the use of a permanent reusable cathode blank instead of a non-reuseable copper starter sheet and the introduction of mechanised and automated electrode handling machines replacing labour-intensive manual operations. The vertical edges had plastic strips and the bottom cased in wax to prevent copper cathode from growing around the edges of the cathode plate during, stripping and allowing two separate copper sheets from each cathode plate. This technology led to major advances in the electrode handling systems and automation in copper tank houses. The improved geometry of the cathode plates and the significantly shorter cathode cycle times allowed for increased intensity and efficiency of the refining process. Introduction of permanent cathode technology resulted in higher capacity, better copper cathode quality with less defects, safer operation and a four-fold improvement in productivity. Considerable development work was required to modify the original stripping machines from their zinc cathode origins due to the heavier cathodes. The stripping capacity of the machines has increased from 250 plates per hour to 600 plates per hour in the latest designs. More recent developments include the elimination of wax masking from the cathode plate, robotic electrode handling machines, and the introduction Duplex Stainless Steel cathode plates giving greater durability and corrosion resistance. Through the use of ISAPROCESS™ user forums, to exchange ideas and developments in the technology and to share operational experiences, the technology has enjoyed continued improvement with higher productivity and improved quality at low cost.

Figure 2 – The IsaKidd process

In mid 1981 Falconbridge Limited commissioned a copper smelter near Timmins to treat concentrate from its Kidd Mine. The original copper cathode produced at Kidd suffered from the presence of higher concentrations of lead and selenium and could not meet customer specifications. It was determined that the use of copper starter sheets was preventing the Kidd refinery from meeting its cathode quality targets. Testwork began with the use of permanent stainless steel cathodes after preliminary tests showed a significant reduction in deleterious elements. The Kidd Process cathode used a solid copper header bar welded onto stainless steel resulting in a lower voltage drop than the ISAPROCESS™. Falconbridge began marketing the Kidd Process technology in 1992 providing competition between the two suppliers of permanent cathode technology. Between 1992 and 2006, 25 Kidd technology licenses were sold and 52 ISAPROCESS™ licenses.

The development of the ISAPROCESS™ and Kidd Process set the scene for a run of technology developments that continued until the mid 2000's. Xstrata took over MIM in 2003 and then Falconbridge in 2006. The Kidd Process technology consequently became part of the tank house package and together they have since been marketed as IsaKidd™ representing the dual heritage of the technology. The current robotic stripping machine (Figure 2) is based on over 30 years of copper refining and winning technology. Today over 100 licensees are using IsaKidd™ technology.

ISASMELT™

The sinter plant/blast furnace combination was the dominant technology for lead smelting throughout the 20th century. In the early 1970's companies using this technology came under sustained political and economic pressure as tighter environmental regulations were introduced, and energy costs increased, leading to higher capital and operating costs (Fewings 1988). It was in this environment that Mount Isa Mines sought a process that would improve the performance of the operations at their lead smelter in Mount Isa. After investigating the various processes under development, researchers turned their attention to the Sirosmelt lance. It had recently been developed on a laboratory scale at the Commonwealth Scientific and Industrial Research Organisation (CSIRO) in Melbourne. Following initial investigations Mount Isa Mines recognised the potential of the novel concept for smelting of lead concentrates and embarked on an extensive development program.

In 1978 a joint project was initiated between Mount Isa Mines and CSIRO to investigate the application of the Sirosmelt submerged-combustion technology to the smelting of Mount Isa lead concentrates. The ISASMELT™ process, as it became known, was developed to maturity for smelting copper, nickel, lead and zinc feeds by Mount Isa Mines through the 1980's and 1990's using incremental scale up. Commercialization only occurred once the process had been proven on laboratory, pilot and demonstration scale over many years. Approximately ten years were required for development of the lead and copper ISASMELT™ from crucible to demonstration scale (refer to Figure 4). During this decade the core know-how that was accumulated enabled the development team to reach the point where they were much better equipped to design and construct a full scale commercial plant – the final stage of the scale up process. Key aspects in this process were the selection of the scale up factors and the systematic design, development and re-engineering of several components of the technology. Figure 3 shows a comparison for the scale up stages for the lead and copper ISASMELT™ processes. Pilot scale was defined as unity for scale up comparison.

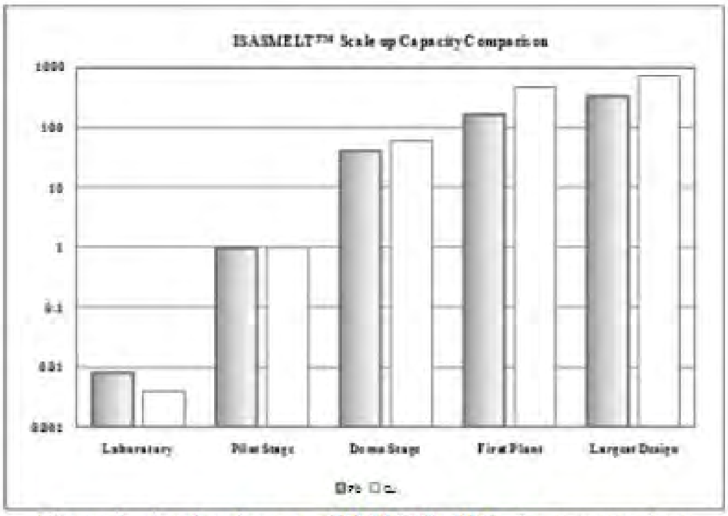


Figure 3 – Lead and copper ISASMELT™ Scale up comparison

During the scale up process, refer to Table 1, several aspects of the technology were developed to a high standard that allowed the ISASMELT™ technology to become a commercial success. As a result, ISASMELT™ technology now operates successfully at numerous plants around the world. The methodical approach to development of the technology has allowed owners to modernise their existing operations or create new businesses with significantly reduced technical risk.

An important parameter in the evolution of the ISASMELT™ technology has been the refractory campaign life. Figure 5 shows the history of the refractory campaigns at the commercial copper ISASMELT™ plant at Mount Isa since commissioning. At the time Mount Isa Mines management considered the installation of water cooling on the furnace refractories undesirable because of the potential for fatal incidents and increased operating costs. As a result the commercial scale furnaces were constructed with minimal water cooling. Although this led to shorter campaign lives initially, a development program was begun that focussed on optimising refractory materials selection and installation methodology. When coupled with process control strategies and continuous on-line monitoring of the bath temperature using systems developed over more than 10 years of operation, it allowed Mount Isa Mines to achieve campaign lives of more than 3 years without using any water cooling of the furnace refractories.

Figure 4 – Tapping matte from the copper ISASMELT at Kazzinc

Table 1 – Key Indicators of ISASMELT™ Plants from pilot to commercial scale

Topic	Unit	Pilot Scale		Demo Scale		First Full Scale		Current Design[1]	
		Pb	Cu	Pb	Cu	Pb[3]	Cu	Pb	Cu
Furnace ID	m	0.4	0.4	1.8	2.3	2.5	3.75	3.6	4.4
Lance Diameter	mm	38	38	150	250	250	350	250	500
Lance Control	-	Manual		Semi Automatic		Semi Automatic		Automatic	
Oxygen Enrichment	%	21	21	21	28	35	45	70	90
Nominal Feed Rate	tph	0.12	0.25	5	15	20	101	40	183
Offgas Treatment	-	Flue System / Baghouse		Gas cooler/ Baghouse		WHB		WHB[2]	

Notes:
ID: Internal Diameter; WHB: Waste Heat Boiler
[1] Refers to maximum throughput
[2] Some of the plants use a combination of radiation section and evaporative cooler for offgas treatment
[3] Refers to the smelting furnace from the two stage lead ISASMELT™ process

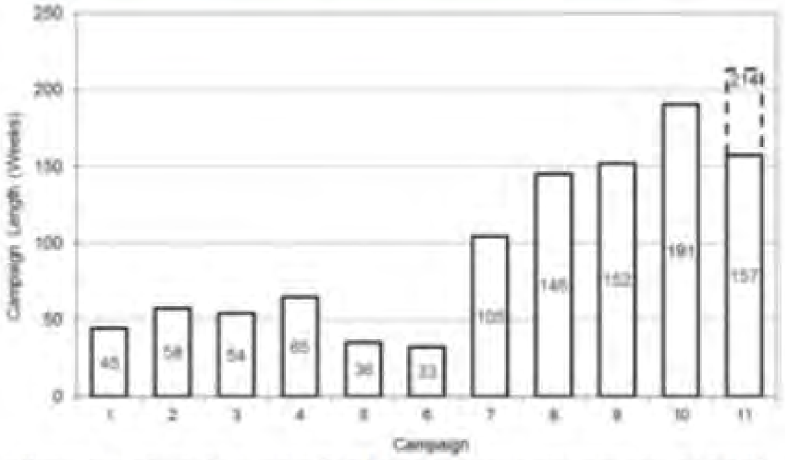


Figure 5 – Mount Isa copper ISASMELT™ plant campaigns (as of 2013)

Jameson Cell

The Jameson Cell (Figure 6) was jointly developed by Mount Isa Mines and Laureate Professor Graeme J Jameson (AO) of the University of Newcastle. Mt Isa had commenced operations with conventional flotation cells but was installing columns in cleaning duties in the mid 1980's. The columns had the benefit of froth washing that was likely to allow significant grade benefits in the very fine lead-zinc circuit. The first observations of the columns was that the collection process was slow necessitating long residence times and large volumes which remains a limitation of columns even today. In 1985 Professor Jameson was commissioned to undertake a project to improve the column sparger design.

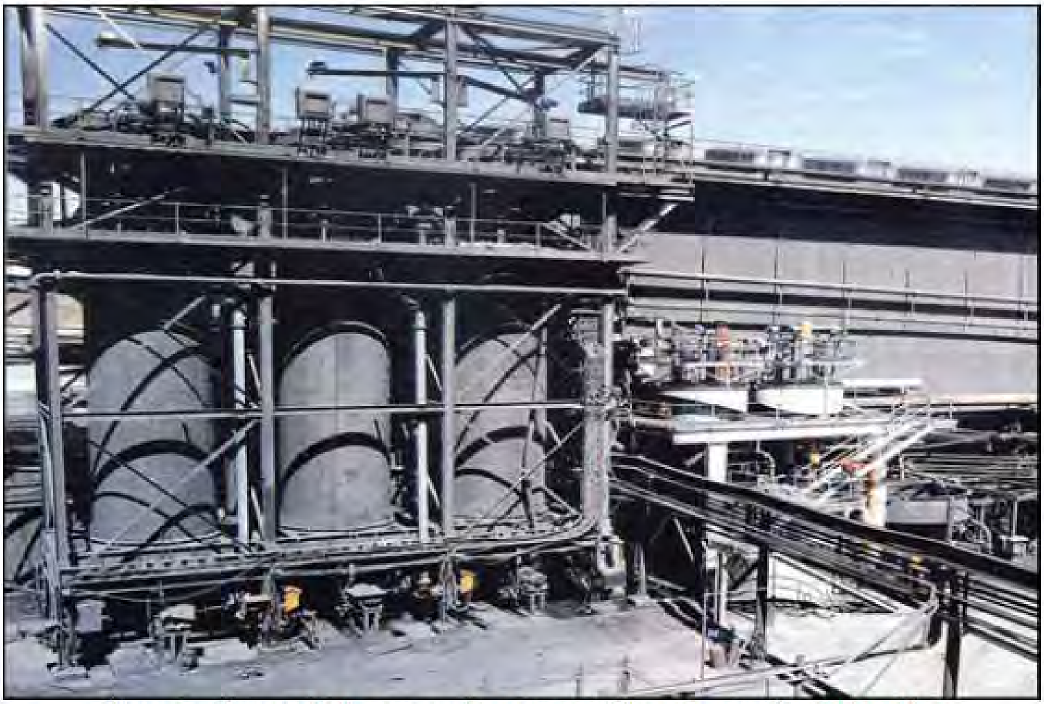


Figure 6 – Jameson Cells compared to columns of the same capacity at Mount Isa

Following initial work to provide an alternate method to bring together bubbles and particles, the downcomer was created. In the downcomer the air and the slurry are co-current with the air being entrained into the plunging jet under vacuum. Investigation showed that all of the bubble particle contact took place in the downcomer and thus the flotation tank could be much smaller. The first application at an industrial scale was in the lead zinc concentrator on the heavy media plant (HMP) lead slimes circuit. The initial improvement in performance were attributed to the very short residence time that allowed the minimisation of oxidation of galena fines. The cells were significantly smaller than the columns and there is no doubt the performance was superior as shown in Figure 3.

The testwork and trials in the early applications showed improved metallurgical performance when operated correctly. The challenge was operating them correctly. The technology hadn't been sufficiently developed to be successfully adopted into plant operations. The cell fell out of favour in base metals and in the 1990's was adopted into the Australia Coal industry and into niche SXEW applications where the main design challenges were resolved. The operability was improved by the introduction of a partial recycle to maintain constant flow and the maintainability of the cell was improved through various design modifications in operating plants. It was a period of continuous improvement. The result was a robust, low maintenance, easy to operate cell with the original features of excellent bubble particle contact.

The final obstacle was overcome when its adaption into the flowsheet was recognised to enable successful installations at the head of cleaner circuits and as low cost brownfield expansions. It is clear that the fast failures have had a significant effect on the success of the cell limiting its adoption into the industry. It is interesting that a significant proportion of sales are to return customers. Once you get over the hurdle of getting a Jameson Cell into your plant then seeing is believing. 2016 was the best year for Jameson cells into base metals and include the first sales back into South America where the cell had been abandoned after the difficulties of operations and maintenance of the Alumbrera installation. The metallurgical performance in Alumbrera was never the issue but the operators and maintainers hated the cells and they failed fast and hard.

The Jameson Cell celebrates its 30th birthday this year and has finally been adopted into mainstream base metals concentrators mainly as cleaner scalper at the head of the cleaner circuit. The cells generally recover up to 80% of the cleaner feed at high grades enabling much lower capital expenditure on the entire circuit. Process performance can be predicted from laboratory and pilot plant testing with demonstrated direct scale-up. It may have taken 30 years but the Jameson Cell is finally a success story. There are many lessons that can be learned from the implementation of innovation into industry from this case study.

IsaMill^{TM}

Unlike the developments of some of the other technologies at Mount Isa where efficiency was the main driver, the IsaMill was developed based on necessity. Figure 7 shows photomicrographs with the same scale of 40 micron demonstrating the increased complexity of Mount Isa ore over Broken Hill ore and the very difficult McArthur River ore. Although McArthur River was discovered in 1955 it was not able to be economically processed until the successful development of ultrafine grinding. McArthur River processing began in 1995 – 40 years after discovery when the IsaMill^{TM} made it technically and economically feasible to grind all of the rougher concentrate to 7 micron to facilitate the rejection of non-sulphide gangue. Even at 7 micron galena liberation is not possible and a bulk zinc-lead concentrate is produced.

Figure 7 – Photomicrograph of a) Broken Hill ore b) Mount Isa ore c) McArthur River ore

Investigations into fine-grinding started at Mount Isa started in the 1970s using conventional grinding technology to increase mineral liberation by grinding to fine sizes. These technologies were not only found to have high power consumption but also proved to be detrimental to flotation performance as a result of pulp chemistry and iron contamination from steel media. These poor results were revisited during pilot plant and tower mill testwork in the 1980s which also showed an inability of tower mills to economically achieve the required sizes. When it became clear that the solution to efficient fine-grinding did not exist in the minerals industry, MIM looked for ideas to "crossover" from other industries that also ground fine particles – pigments, pharmaceuticals, foodstuffs (e.g. chocolate). While these mills operated at a much lower scale and treated high value products they demonstrated the principle that stirring fine media at high speed was highly efficient. The challenge was transferring this concept to continuous, high tonnage and lower-value streams in the minerals industry.

In 1991 the introduction of a Netzsch laboratory stirred mill to the Mount Isa site was a turning point in fine-grinding and ultrafine grinding. The ½ litre bench scale mill resembled a milk shake maker and used fine copper smelter slag as grinding media. Testwork on McArthur River ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the Mount Isa pilot plant. The testwork showed that high speed, inert, horizontal mills could efficiently grind to 7 microns at laboratory scale providing major improvements in metallurgical performance. To make ultrafine grinding applicable to full-scale production a program of development was undertaken between Mount Isa Mines Limited and NETZSCH-Feinmahltechnik GmbH.

After 7 years of development and testing of prototypes in the Mount Isa operations, the IsaMill™ evolved. It was large scale, continuous, and most importantly robust because it was developed by operators. The crucial breakthrough was the perfection of the internal product separator – this allowed the mill to use cheap natural media (sand, smelter slag, ore particles) and to operate in open circuit. These are significant advantages for operating cost and circuit simplicity. Scale-up was tested using trial installations at the Hilton and Mount Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill™ (1.1MW) was installed in the Mount Isa concentrator. Improvements to the technology were continually made by the operators, maintainers and engineers working with the technology.

In 1998 the rights for commercialisation of the IsaMill™ where transferred from Mount Isa Mines Limited to MIM Process Technologies (now Glencore Technology) and under an exclusive agreement with Netzsch. In December 1998, the IsaMill™ technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. The IsaMill™ is now a mainstream fine grinding machine with over 130 installations around the world.

The Albion Process™

In the 1990's, MIM were studying options for the development of the large Frieda River/Nena project in PNG through its subsidiary Highlands Pacific. The Nena ores were not amenable to smelting, due to the elevated arsenic content, and several hydrometallurgical options were examined. Out of this work, MIM developed the Albion Process™, named after the suburb in Brisbane where MIM's development laboratory was located. The Albion Process™ is a combination of ultrafine grinding using Glencore Technology's IsaMill™, followed by oxidative leaching at atmospheric pressure in a series of reactors designed to achieve high oxygen mass transfer efficiency. The HyperSparge™ was also developed to deliver oxygen to the reactors efficiently.

Various small scale continuous pilot plant campaigns were conducted in 1994 and 1995. A larger pilot plant (120kg zinc cathode/day) was constructed in 1997 to conduct testwork as part of a feasibility study on the zinc/gold resources of Pueblo Viejo in the Dominican Republic. Extensive piloting was also conducted on lower grade chalcopyrite concentrates for Cyprus Amax in 1998, and for Mount Isa Mines in 2000. Pre-feasibility and feasibility pilot testing was conducted on the zinc/lead bulk concentrates from McArthur River and Mount Isa in Australia between 2001 and 2005. During this time the Albion Process™ was successfully tested on over 70 different ores and concentrates. The process is designed to recover gold and base metals from refractory ores. The key to the process is the ultrafine grinding stage followed by a hot oxidative leach at atmospheric pressure.

In the period from 1994 until 2004, the Albion Process™ (see Figure 8) was seen as strategic to the MIM/Xstrata group, and was not marketed externally. In 2005, a decision was made to offer the technology to external clients under licence, and a marketing agent – Core Resources, was appointed to market the technology globally. Interest in the technology has been very strong in the subsequent period, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Certej Project. The technology moved into commercial production in 2010 with the commissioning of Glencore's Albion Process™ plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Glencore of a second plant in Germany (16,000 tpa zinc metal). The Las Lagunas refractory gold project commissioned in 2012, and the GPM Gold refractory gold project commissioned in 2013.

Figure 8 – The Albion Process oxidative leach plant in Armenia

The major scale up risk with any oxidative leaching technology is oxygen mass transfer. High agitator power demands are common to achieve the shear rates in the vessel required for effective mass transfer at a commercial scale. A different approach was taken in the design of the Albion Leach Reactor to lower the agitator power demand. Glencore developed the HyperSparge supersonic gas injection lance to provide gas injection velocities of the order of 500 $m.s^{-1}$ within the leaching vessel, compared to the 4 – 8 $m.s^{-1}$ achieved with a typical agitator. Supersonic oxygen injection is a far more efficient method of generating shear than conventional agitation, allowing the total power input into the vessel to be significantly reduced, and greatly reducing the scale up risk for the oxidative leach.

The Albion Process™ was enabled by the fine grinding of the IsaMill™ and the process was designed to deliver a lower cost processing option for treating refractory mineral resources. There are now six operating Albion Process™ plants and the process has now an extensive database of potential applications.

CONCLUSIONS

MIM developed a significant number of processing innovations that are technical and economic successes. The ability to innovate at MIM was enabled by very challenging orebodies and the need to process efficiently to remain economically viable. The success has been attributed to the development of these technologies on an operating site with the R&D group solving the technical issues on small scale. Each subsequent scale up was completed in the operating plants where the operators, maintainers, engineers and metallurgists were required to achieve production goals at each step of the scale up to ensure funding for the next step.

The number of innovations, at MIM, was disproportionate to the scale of operations and may have been enabled by the remoteness of the site and the researchers and operators working collaboratively to solve economic and technical problems. The research group were not capital city based but worked on the same site and were required to assist with installation, commissioning and operation of the various stages. This co-operation led to adoption into the plant and a fast feedback loop for improvements. The ultimate success of the innovations has been their widespread adoption into the mainstream industry where feedback from operating sites based on a user group model has enabled continuous improvement of each of the technologies.

ACKNOWLEDGEMENTS

The authors would like to acknowledge Glencore Copper and Glencore Technology for permission to publish and to all the research and production personnel who enabled and improved the

technologies in their plants. The success of these developments continues with the input from end users in the ongoing development.

REFERENCES

Armstrong, W. "The Isa Process and its contribution to electrolytic copper," paper presented at the Rautomead Conference, Scotland, August 1999.

Fewings J.H., Management of Innovation – The IsaSmelt Process. Presented at AMIRA "Innovation in Metal Production" Technical Meeting at Mount Isa, October 3-4, 1988.

Fountain C. Isasmelt and IsaMills – Models of successful R&D – AusIMM Young Leaders conference, 19-21 March 2002, Kalgoorlie, W.A.

Nihill D.N., Stewart C M. and Bowen P. the McArthur River Mine – The First Years of Operation. The AusIMM '98 Mining Cycle. Mount Isa 19-23 April 1998.

Pease, Joe (August 2005), "Complex leaching becomes much simpler" (PDF), Australian Mining, pp. 26–32, archived from the original (PDF) on 12 September 2009, retrieved 6 January 2010

Pease, Joe September 2016, "Crossing the innovation valley of death" Presented to JKMRC.

Hydroprocess · ICMSE · 2017 | 9th International Seminar on Process Hydrometallurgy
International Conference on Metal Solvent Extraction

Oxygen Mass Transfer in the Albion Process™: from the Laboratory to the Plant

Paul Voigt, Mike Hourn and Daniel Mallah
Glencore Technology, Australia

ABSTRACT

The successful commissioning and ramp up of the Albion Process™ at the GPM Gold Project relied on the successful scaling up of the process from batch and continuous pilot plant campaigns (Voigt, 2016) Critical information about reaction kinetics and residence time, grind size and pulp density were determined at the laboratory scale and successfully applied to the commercial scale. A limitation of small scale testwork, is that some parameters cannot be measured reliably and scaling up is a function of the physical size of the equipment which isn't possible to test with laboratory scale equipment. Oxygen mass transfer rate is one such parameter since this is a complex interaction of many factors including slurry temperature, solution and slurry chemistry, slurry viscosity, agitator type, dimensions and power, oxygen bubble residence time, oxygen purity, tank geometry and oxygen injection technique. Oxygen generation represents an important operating cost for the Albion Process™. Pivotal to the Albion Process™ operating economically at atmospheric pressure is the capability to efficiently transfer oxygen while utilising as much oxygen injected to the process as possible. To respond to this Glencore Technology developed the HyperSparge™ supersonic gas injector. This paper compares the HyperSparge™ against other sparging techniques to quantify the benefits of oxygen injection via a supersonic gas jet on scale up of the oxygen mass transfer system. The paper then examines plant survey data from the GPM Gold Project to demonstrate the very high oxygen utilisation that can be achieved with a correctly designed oxygen mass transfer system.

INTRODUCTION

One of the main challenges in chemical reactor design is the scale-up of processes from the laboratory to industrial scale. One particular challenge are those processes which cannot be faithfully simulated with an experiment at the laboratory due to the physical dimensions and complexity of the real system. One example of this is the design of oxygen mass transfer systems in atmospheric oxidative leaching such as the Albion Process™. Some of the problems with the small size of equipment are the specific agitator power input is artificially high, the bubble residence time in the vessel is artificially low, the oxygen partial pressure at the base of the vessel is artificially low and the way oxygen is injected into the process may not be the same as the industrial process. Some of the problems of the complexity of the system is the dynamic nature of the process including the presence of solids, recycle streams and minor elements plus the variances within the reality of a process plant such as disruptions or variance of feed quality.

It is critical that for such processes the fundamentals are understood so that key data can be collected from appropriately designed experiments at the laboratory scale for input into a proven design approach to render the industrial scale process effective and fit-for-purpose.

To maximise oxygen mass transfer efficiency, operational availability, simplicity and safety, Glencore Technology (GT) developed the HyperSparge™ supersonic gas injector. Originally for injecting oxygen to the Albion Process™, the HyperSparge™ has found application in other processes using air, oxygen, sulphur dioxide and gas mixtures. Over 400 HyperSparge™ units are installed globally in duties from in-line slurry conditioning, fermentation, waste-water treatment, CIL/CIP processes and oxidative leaching.

This paper compares the HyperSparge™ to other gas injection techniques such as ring spargers and converging nozzles. The key data are identified for scale-up from the laboratory to industrial scale. Survey data are examined from the GPM Gold plant in Armenia where the oxygen mass transfer system was successfully designed and comments are made about the operational advantages of the HyperSparge™ compared to other systems such as sintered spargers.

OXYGEN MASS TRANSFER

The Albion Process™ is an atmospheric oxidative leaching process developed by GT in 1994 and is described extensively in the literature (Hourn & Turner, 2010; Hourn & Turner, 2012; Hourn et al., 2014; Voigt et al., 2015; Senshenko et al., 2016). Oxidation of pyrite is one of the key reactions in the Albion Process™ chemistry as shown in Equation 1.

$$FeS_2 + 15/4O_2 + ½ H_2O = ½ Fe_2(SO_4)_3 + ½ H_2SO_4 \qquad [1]$$

The first step of the oxidation mechanism is the dissolution of oxygen in the liquid phase since electron transfer occurs in part through the action of the ferric and ferrous ion couple and the action of dissolved oxygen reacting directly with pyrite. Hence the rate at which oxygen is transferred is critical to the process efficiency and can be simplified for a reactor as shown in Equation 2 (Middleton & Smith, 2004).

$$O_2 \text{ Transfer Rate} = k_La.V.(C_{sat} - C) \qquad [2]$$

Where k_L is the liquid film transfer co-efficient ($m.s^{-1}$), a is the specific gas surface area (m^{-1}), V is reactor volume (m^3), C_{sat} is the oxygen solubility at saturation ($g.m^{-3}$) and C is the steady state oxygen level ($g.m^{-3}$). The terms k_L and a are normally combined to represent the oxygen mass transfer co-efficient which is the parameter to be maximised.

For practical purposes the k_La needs to be maximised by determination of a relationship with power input through the sparger and the agitator and then selection of equipment that satisfies the oxygen transfer rate requirements of the system. Middleton (1992) simplified the equation relating k_La to power input is shown in Equation 3.

$$k_La = K . (U_s)^a . (P_g/V)^b \qquad [3]$$

Where K is a coalescence constant, U_s is superficial gas velocity ($m.s^{-1}$) and P_g is absorbed power (W). A correction is also applied for temperature since increasing temperature will reduce slurry viscosity and surface tension increasing the mass transfer interfacial area. The coefficients a and b are empirical constants that are system specific and are determined through testwork.

There are many theoretical correlations between k_La and energy input reported in the literature and the majority tend to be based on data in biochemical reactors meaning the extension to oxidative leaching can be problematic with limited applicability because the agitation intensity, system chemistry, operating temperature and oxygen injection methods are very different (Van't Reit, 1979; Vasconcelos et al., 1998; Oguz et al., 1987; Moo-Young & Blanch, 1981). Most of these correlations are also derived at very small scale, with laboratory vessels of 10 litres or less in size.

The best way to achieve successful industrial scale up is to determine the a and b empirical constants for equation 3 for a given system through specific experiments that reflect the actual system conditions. Once determined for a specific system, Equation 3 is then used along with other reactor design equations and experience to calculate the tank geometry and power input required from the agitator and gas injector to achieve the required k_La. GT has developed an experimental procedure to determine the a and b empirical constants for a given system which will result in a successful scale up with minimal error. An example of using this method to compare the experimental and calculated k_La for a copper leach system is shown in Figure 1.

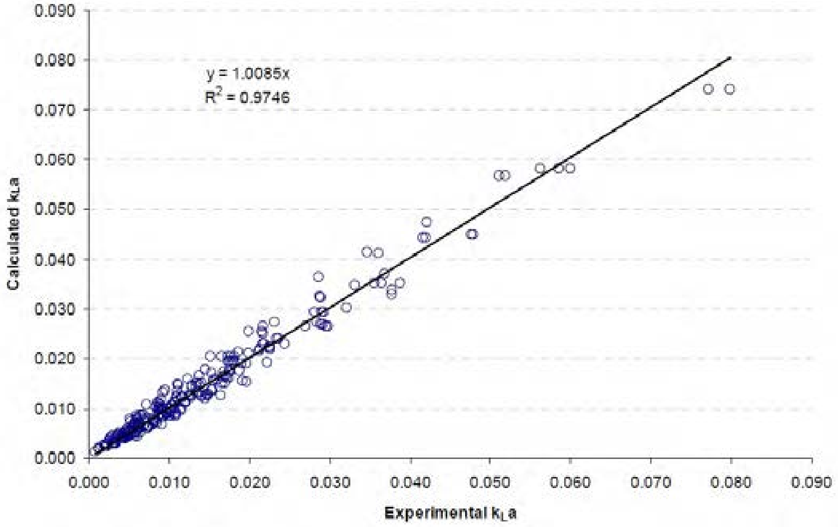


Figure 1 Experimental and calculated k_La for a system using GT experimental methodology

INJECTION OF OXYGEN

Importance of oxygen injection method

The method in which oxygen is injected has a large bearing on the efficiency of the oxygen mass transfer. Referring to Equation 2, the liquid film co-efficient (k_L) is proportional to the oxygen diffusion coefficient and inversely proportional to the liquid film thickness. The consequence for oxygen injector selection is that high pressure gas injection maximises shear and significantly erodes the liquid film thickness resulting in increasing k_L. Referring to Equation 3, the surface area (a) is proportional to power absorbed into the system relative to system volume and superficial gas velocity. Surface area is also inversely proportional to bubble size. The consequence for oxygen injector selection is to maximise shear and minimise bubble size but also to select an agitator that provides sufficient complimentary power input.

Options for oxygen injection

There are many gas injectors available on the market. The converging nozzle type and ring spargers (essentially open pipes) and among the most common from an operational perspective. Sintered spargers find application in flotation machines and chemical reaction systems and are well suited to the laboratory but are plagued with operational issues at the industrial scale, particularly in processes with chemical reactions due to the fast accumulation of reaction products and the continual need to clean them.

GT performed a number of experiments comparing the performance of an open pipe sparger, a converging nozzle type sparger and a newly developed supersonic sparger by GT, the HyperSparge™.

Open pipes and sparge rings do not transfer a significant amount of energy into the leach vessel, with all the energy required for oxygen mass transfer provided by the impeller. For these spargers, the gas is compressed to just above the hydrostatic head level in the slurry tank and the pressure drop across the tip of the injection point is low, resulting in low gas speeds.

Convergent nozzles, such as the CPT Slamjet and Minnovex gas spargers, were developed for flotation column aeration duties, and have found some use in slurry oxidation systems. These nozzles converge to a narrow opening, and the gas is compressed to several atmospheres above the hydrostatic head level in the slurry tank. The pressure drop across the tip of the injection point is 3 – 5 atmospheres, resulting in higher gas speeds at the point of injection. This results in more energy being put into the gas from the sparging system, and so less energy is required at the impeller.

Testwork carried out by GT has found that for oxygen mass transfer purposes, energy input to the system via the gas sparging system is more efficient than energy input via the impeller, as long as the majority of the energy released from the gas pressure drop across the sparger is converted to kinetic energy and not lost as heat. As such GT developed a supersonic gas injector, the HyperSparge™, to further improve the efficiency of the gas injection system. The HyperSparge™ used a converging-diverging nozzle to maximise energy recovery from the gas pressure drop by accelerating the gas to supersonic velocities.

Efficiency of oxygen injection methods

Energy is put into the gas sparging system by pressurising the gas prior to the sparger, to store energy in the gas stream. The theoretical energy input to compress the gas is defined in Equation 4.

$$W = (n/(n-1))mRT[(P_2/P_1)^{(n-1)/n} - 1 \quad [4]$$

Where n is isothermal efficiency, W is work (J), P_1 is outlet pressure ($N.m^{-2}$) and P_2 is inlet pressure ($N.m^{-2}$), T is temperature (K), R is the ideal gas constant (8.314472 $J.K^{-1}.mol^{-1}$) and m is moles.

The pressure energy is then recovered as work into the system, ideally the efficiency of this conversion is maximised. High speed gas injection maximises turbulence at the injection point, which reduces the liquid film boundary layer around the gas bubble, increasing the liquid film diffusion rate, and so increasing the k_L value. The higher kinetic energy of the gas will translate to a higher interfacial area, which also assists mass transfer.

In a conventional pipe or ring sparger, the nozzle diameter is relatively large, resulting in a low velocity gas flow, which remains constant along the length of the nozzle. In this configuration, higher gas flows can lead to larger bubble sizes and consequently smaller interfacial areas (a). As a general rule, the size of the bubble generated from a conventional sparger will be close to the size of the orifice.

The formation of fine bubbles requires a specific type of nozzle to be fitted to the end of the sparger. Nozzles can normally be described as *convergent* (narrowing down from a wide diameter to a smaller diameter in the direction of the flow) or *divergent* (expanding from a smaller diameter to a larger one in the direction of the flow).

Convergent nozzles accelerate subsonic gases, with the dynamic pressure decreasing from the nozzle opening to the narrowest point as the gas is accelerated. Once the ratio of the gas pressure at the nozzle throat to the feed pressure reaches a critical value, the flow will reach sonic velocity. This forms a shock wave in the nozzle throat. The nozzle is then said to be choked, and the gas flow is at Mach 1. Gas flow cannot be accelerated beyond Mach 1 in the throat of the nozzle.

Increasing the nozzle feed pressure further will not increase the velocity of the gas at the throat of the nozzle, and so any additional pressure energy is then stored in the gas in the form of heat.

Divergent nozzles decelerate subsonic gases; the gas expands in the nozzle and slows down. However, when the gas enters the divergent section at Mach 1, the divergent nozzle will accelerate the gas, as the stored heat energy is converted to velocity. Combination convergent-divergent nozzles, such as those used in the HyperSparge™ can, therefore, accelerate gases that have choked in the convergent section to supersonic speeds. These nozzles develop thrust by converting the stored heat that builds up in the gas in the throat into velocity when in the divergent section of the nozzle.

When gas is accelerated to velocities above the speed of sound in the divergent section of the nozzle, the local pressure drops further, and does not recover to ambient pressure until the gas passes through the resulting shock wave as illustrated in Figure 2.

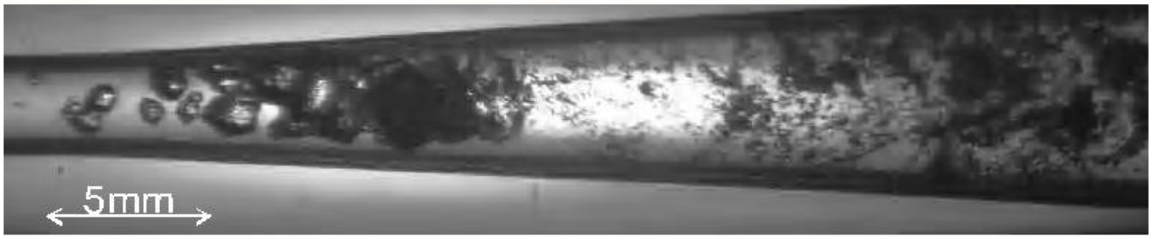


Figure 2 Collapsing bubbles crossing the shock wave of a supersonic nozzle (Kawamura et al., 2004)

The low pressure that is generated in the divergent section of the nozzle leads to collapse of the gas bubble into a fine mist, and results in a dramatic increase in the gas surface area. This is shown in the figure above, where the gas bubble leaves the throat of the nozzle at a size that is fairly similar to the size of the throat diameter, and then collapses to a fine mist of micro-bubbles in the low pressure zone in the divergent section. This collapse is triggered by the gas crossing the shock wave that is set up once the gas is choked in the throat of the nozzle. This shock wave can extend anywhere from the nozzle throat to the end of the nozzle and beyond, depending on the pressure applied to the nozzle.

A convergent-divergent nozzle has two important roles. The design of the nozzle determines the exit velocity for a given pressure and temperature. And because of flow choking in the throat of the nozzle, the nozzle design also sets the mass flow rate through the nozzle. Therefore, the nozzle design determines the thrust of the gas injection system. By changing the shape of the nozzle and the flow conditions upstream and downstream, you can control both the amount of gas that passes through the nozzle and the thrust generated by the nozzle. A convergent-divergent nozzle is represented below.

Figure 3 Cross section of a convergent-divergent nozzle (NASA, 2015)

EXPERIMENTAL

A number of experiments were performed to compare the nozzle types evaluating on the variables of thrust and oxygen mass transfer co-efficient (k_La). Experiments were performed in a jacketed, 1750 litre tank fitted with four baffles and a lid to minimise evaporative loss. The process solution employed was a copper leach solution at pH 1, with 15 g/l^{-1} copper. High purity gaseous oxygen and nitrogen was used. The nozzles employed for the experiments were:

- 10mm open pipe
- Converging high pressure sparger, 4mm exit diameter

- Converging-diverging HyperSparge™ nozzle, Mach 3 (MV), 1mm throat and 5mm exit diameter
- Converging-diverging HyperSparge™ nozzle, Mach 1 (MT), 1mm throat and 1.7mm exit diameter

Thrust

Thrust is a direct measure of the efficiency of conversion of pressure energy in the incoming gas stream into momentum. The higher the thrust measured, the more efficient the nozzle will be in converting pressure energy into mass transfer without loss to heat dissipation. The nozzles were positioned perpendicular and centred to a balance surface. The gas plume impinged the balance's surface and the mass imparted recorded. For an identical energy input in the sparging gas (ie fixed flow and delivery pressure), each nozzle generated a different degree of thrust. The data are shown in Figure 4.

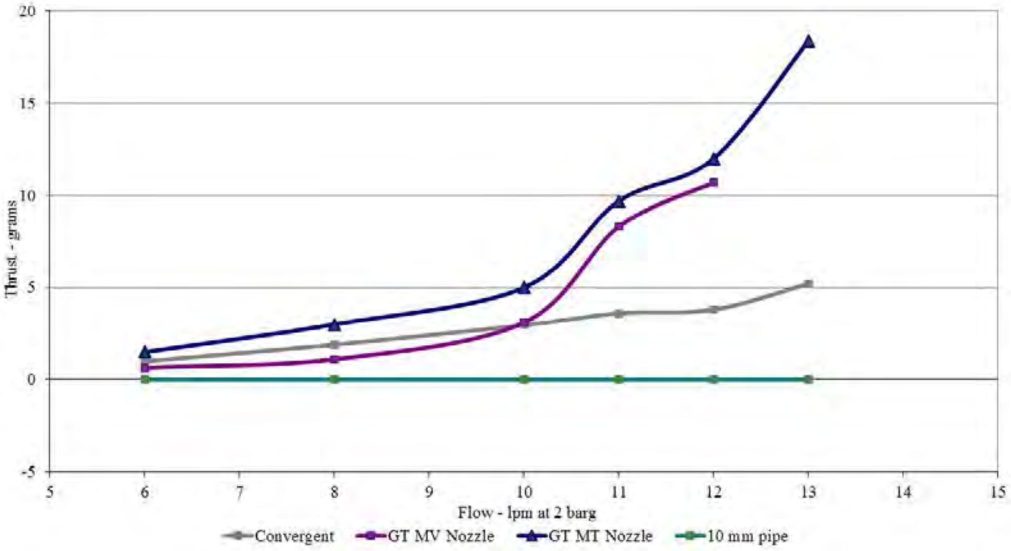


Figure 4 Developed thrust measured at 200kPa

The results above show that at low oxygen flow all of the profiled nozzles exhibited similar thrust values; less than 3 grams (g) of thrust recorded by the scale. However, at higher flows, the nozzles have a very different behaviour. The open pipe sparger did not develop any thrust at all across the entire range of gas flowrates, and so did not exert any force on the balance. The convergent nozzle showed an almost linear response, with thrust increasing with flowrate, up to a maximum of 5 g at 13 lpm of gas flow.

Both GT HyperSparge™ supersonic nozzles generated significantly more thrust than the open pipe or convergent nozzle. In general, the nozzle designed to produce maximum thrust outperformed the high velocity nozzle across all flow rates, achieving a thrust of 18 g at a gas flowrate of 13 lpm.

This compared well to the maximum theoretical thrust for a nozzle discharging 13 lpm of gas flow of 21 g. This meant that 85 % of the pressure energy in the gas stream was converted to momentum for the HyperSparge™ MT nozzle profile. Both supersonic nozzles also displayed a considerable steepening of the thrust curve at gas flowrates above 11 lpm, close to the point where the nozzles were choked, indicating superior thrust as the gas speeds exiting the convergent-divergent nozzles approached supersonic speeds.

Oxygen mass transfer co-efficient (k_La)

For temperatures up to 50°C oxygen mass transfer co-efficient was determined using the Dynamic Gassing Out Method (Taguchi and Humphrey, 1966). Above 50°C the combined sulphite method was used (Puskeiler, 2005). Experiments were conducted with and without solids in the reactor. The results are shown in Figure 5.

The GT MV and MT nozzle designs resulted in a consistently higher k_La value across the entire range of gas flow rates tested relative to the open pipe and the convergent nozzle. As the gas flow rate approached choked flow in the GT HyperSparge™ MV and MT nozzles, the difference in k_La was more pronounced, and was up to 300 % of the value observed for the conventional open pipe and 50% compared to a converged nozzle.

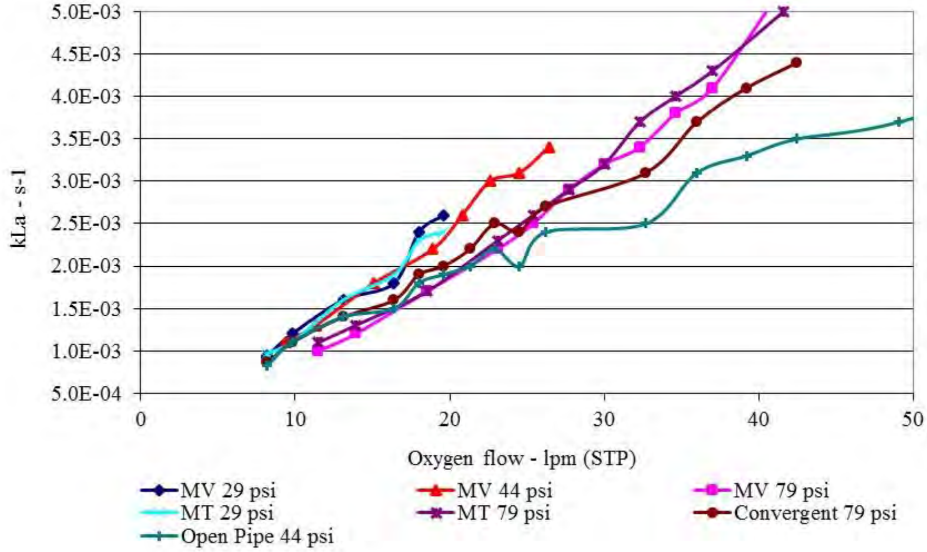


Figure 5 k_La at different supply pressure

The GT HyperSparge™ MV and MT nozzles showed similar behaviours because both are convergent-divergent type nozzles. Of the two HyperSparge™ nozzles, the MT nozzle provided the highest k_La, gas utilisation and thrust values, typically by 25 % relative to the MV nozzle at a set gas flow rate. The higher thrust generated in the MT nozzle at the expense of velocity explains this difference.

SCALE UP TO INDUSTRIAL SCALE

Through the scientific process, GT established that supersonic gas injection was technically a superior way to inject gas into a slurry requiring oxygen. The next step was to develop the converging-diverging nozzle into a product that can be used at the industrial scale.

The HyperSparge™

The HyperSparge™ was developed by GT as a complete gas injection solution. At the heart of the device is a converging-diverging nozzle fabricated from a specially selected wear-resistant ceramic which accelerates gases to supersonic velocities for injection into solutions and slurries. The nozzle is designed to minimise pressure energy loss. A schematic of the HyperSparge™ and close up of the nozzle is shown in Figure 6.

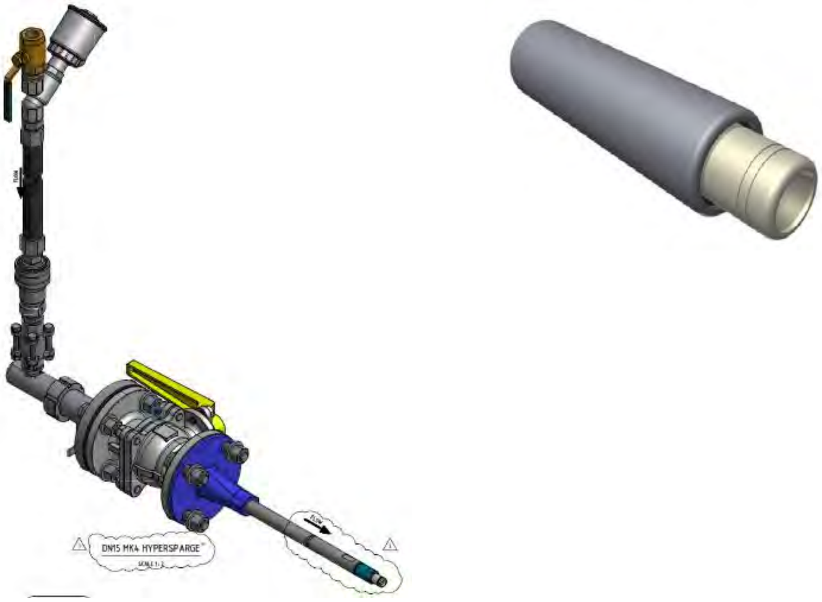


Figure 6 HyperSparge™ supersonic gas injector and nozzle

Apart from the advantages of superior oxygen mass transfer, the HyperSparge™ is safe and easy to use. The HyperSparge™ can operate over a range of pressures which can be adjusted to regulate flow. The nozzles can be easily interchanged with the threaded nozzle holder allowing the use of different nozzle diameters and hence flow operating envelopes. One attractive feature is that the sparger can be inserted and removed from a vessel while it is full of slurry spillage free without using any additional removal device plus with a throat diameter of typically 7 mm, the sparger does not become blocked and self-cleans by the action of the gas flow. This means less operational downtime and process interruption.

Oxygen Mass Transfer System Design

During the GPM Gold Albion Process™ plant development, key oxygen mass transfer experiments were conducted to determine the k_La under conditions as close as practicable for temperature and slurry conditions. The data was used by GT to then correctly size the gas injectors, agitators and select ideal tank dimensions. The design criteria for the GPM Gold Plant is shown in Table 1.

Table 1 GPM Gold Albion Process™ Plant Design Criteria

Parameter	Units	Nominal	Design
Throughput	tph	13.1	14.5
S^{2-} concentration	%	17.6	20.0
S^{2-} oxidation	%	76	76
Oxygen Utilisation	%	90	80
Required k_La	$m.s^{-1}$	0.12	0.14
CIL Gold recovery	%	92	90

The economics of the industrial process rely on the efficiency of oxygen transfer to the system to maximise transfer rate (achieving design residence time) and maximise transfer efficiency (utilisation of the oxygen that is injected.

A survey was completed to collect plant data and assess the transfer rate and transfer efficiency. Since an industrial plant was the subject of the survey, the plant couldn’t be taken off-line to replicate laboratory experiments. Instead, the oxidation extent and oxygen consumption from the operating plant were compared against the design. The conditions of the survey were a throughput rate of 14.0 tph concentrate and a sulphide concentration of 17.0%.The survey data for oxidation extent and gold recovery are shown in Figure 7.

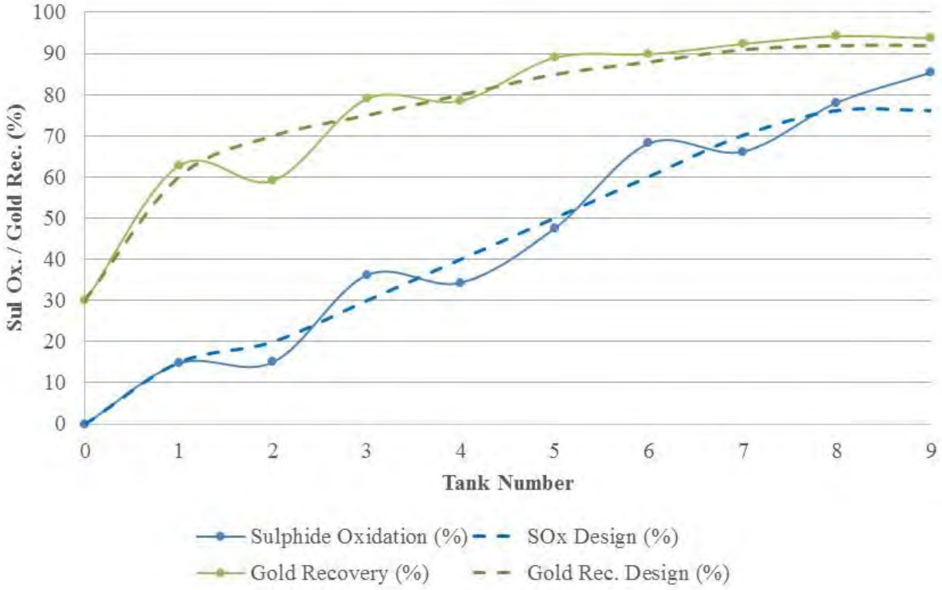


Figure 7 Sulphide Oxidation survey of GPM Gold Albion Process™ Plant.

Figure 7 shows that at the design throughput the sulphide oxidation is achieved along with corresponding gold recovery.

The oxygen utilisation was determined by comparing the oxygen injection to the process during the survey period against the modelled values. The results are shown in Figure 8.

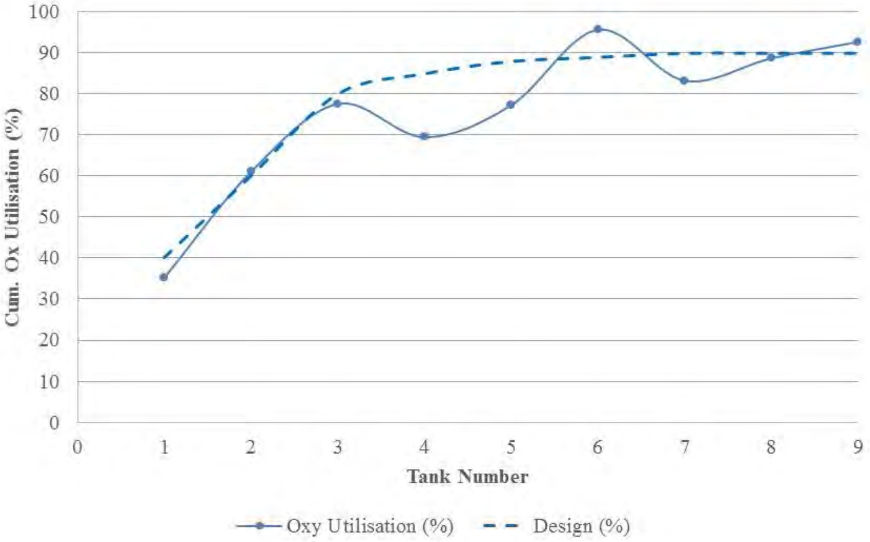


Figure 8 Oxygen utilisation at the GPM Gold Albion Process™ Plant

Figure 8 shows that the cumulative oxygen utilisation is in excess of the design value of 90% achieving 93%. This observation is consistent with other multi-tank Albion Process™ plants where it appears oxygen that is dissolved in the first few tanks of the leaching train is utilised later in the process.

Both Figures 7 and 8 illustrate that the oxygen mass transfer system was correctly and successfully designed.

CONCLUSIONS

Oxygen injection using convergent-divergent nozzles generate superior thrust and oxygen mass transfer compared to other gas injection techniques. Power delivered to the system is more efficient through gas injection rather than mechanical agitation. The HyperSparge™ is a development of the convergent-divergent nozzle and offers additional advantages over other sparging technologies contributing to a safer work environment, maximising process run-time and optimising energy input through the agitator. GT successfully scaled up the oxygen mass transfer system from the laboratory to the industrial scale at the GPM Gold Albion Process™ plant which achieves greater than design performance in terms of oxygen mass transfer and oxygen utilisation.

REFERENCES

Hourn, M., & Turner, D.W. (2010). Albion Process for treatment of refractory ores, *Proceedings of ALTA Conference 2010*, Perth, Australia.

Hourn, M., & Turner, D.W. (2012). Commercialisation of the Albion Process, *Proceedings of ALTA Conference 2012*, Perth, Australia.

Hourn, M., Voigt, P., & Turner, D.W. (2014). Development of the Albion Process plant to treat refractory concentrates from the GPM Gold Project, *Proceedings of Hydroprocess Conference*, Vina del Mar, Chile.

Kawamura, T., Fujiwara A., Takahashi, T., Kato, H., Matsumoto, Y. & Kodama, Y. (2004). The effect of the bubble size, dispersion and skin friction reduction. *Proceedings of On smart control of turbulence*, Tokyo, Japan.

Middleton, J.C. (1992). Gas-liquid dispersion and mixing. In N. Harnby, M.F. Edwards, A.W. Nienow (Ed.), *Mixing in the Process Industries* (pp. 322–363).

Middleton, J.C. & Smith, J.M. (1992). Gas-liquid mixing in turbulent systems. In E.L. Paul, V. Atiemo-Obeng, S.M. Wiley (Ed.), *Handbook of industrial mixing: Science and practice* (pp. 585–638).

Moo-Young, M. & Blanch, H.W. (1987). Design of biochemical reactors: Mass transfer criteria for simple and complex systems, *Adv. Biochem. Eng.*, 19, 1–69.

NASA (2015). Nozzle design – converging/diverging (CD) nozzle, https://www.grc.nasa.gov/www/k-12/airplane/nozzled.html, retrieved January 2017.

Oguz, H., Brehm, A. & Deckwer, W.D. (1987). Gas/liquid mass transfer in sparged agitated slurries, *Chemical Engineering Science*, 42, 1815–1822.

Puskeiler, R. & Weuster-Botz, D. (2005). Combined sulfite method for the measurement of the oxygen transfer coefficient k_La in bioreactors, *Journal of Biotechnology*, 120(4), 430–438.

Senshenko, A.Y., Aksenov A.V., Vasiliev A.A., & Seredkin, Y.G. (2016). Technology for processing of refractory gold-containing concentrates based on ultrafine grinding and atmospheric oxidation, *Proceedings of IMPC Conference*, Montreal, Canada.

Taguchi, H. & Humphrey, A.E. (1966). Dynamic measurement of volumetric O2 transfer coefficient in fermentation systems, *Journal of Fermentation Technology*, 44, 881–889.

Van't Riet, K. (1979). Review of measuring methods and nonviscous gas-liquid mass transfer in stirred vessels, *Ind Eng Chem Process Des Dev*, 18, 357–364.

Vasconcelos, J.M.T., Alves, S.S., Nienow, A.W. & Bujalski, W. (1998). Scale-up of mixed gassed multi-turbine agitate vessels, *Canadian Journal of Chemical Engineering*, 76, 398–403.

Voigt, P., Hourn, M., Mallah, D. & Turner, D.W. (2015). Commissioning and ramp up of the Albion Process at the GPM Gold Project, *Proceedings of World Gold Conference*, Johannesburg, South Africa.

Voigt, P., Hourn, M. & Mallah, D. (2016). Treatment of low grade materials, *Proceedings of MINEX Conference*, Moscow, Russian Federation.

MetPlant 2015

Metallurgical Plant Design & Operating Strategies – World's Best Practice

7–8 September 2015, Perth, Western Australia

Image courtesy of Outotec. Atlas Mining's state-of-the art copper concentrator plant upgrade at its Carmen Copper Mining Complex in Toledo City, Cebu, Philippines.

Principal Sponsor

Outotec

Proceedings

Proceedings Sponsor

MAGOTTEAUX

AusIMM
THE MINERALS INSTITUTE

Commissioning and Ramp-up of the Albion Process at the GPM Gold Project*

P Voigt[1],M Hourn [2], D Mallah[3] and D Turner[4]

1. Manager - Hydrometallurgy, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, paul.voigt@glencore.com.au
2. General Manager - Technology, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, mike.hourn@glencore.com.au
3. Metallurgist - Hydrometallurgy, Glencore Technology, Level 10, 160 Ann St Brisbane QLD 4000, daniel.mallah@glencore.com.au
4. General Manager Technology, Core Resources, 44 Corunna St Albion QLD 4010

* This paper was originally published in the proceedings of ALTA 2015.

ABSTRACT

The GPM Gold Project is located in Armenia, and consists of an open cut mine at Zod, near the Azerbaijan border, and a CIL processing plant at Ararat near the Turkish border. Mining at Zod commenced in 1976, and focused on near surface oxide ores, which overlay deeper refractory sulphides. Historical mining has now almost depleted the oxide ores, and the sulphide content of ore delivered to the processing plant at Ararat is increasing. Gold and silver recoveries through the Ararat plant were declining steadily.

GeoProMining, the owners of the project, have expanded the Ararat facility to deal with the increasing sulphide content in the ore. In 2014, GeoProMining refurbished an existing concentrator on site to recover a sulphide concentrate from the ore, and have constructed an Albion Process^{TM} plant to oxidize the refractory concentrate. Glencore Technology (GT) provided the Albion Process^{TM} plant as a technology package.

In July 2014, the progress of the GPM Gold Project Albion Process^{TM} Plant was reported (Hourn, Voigt and Turner, 2014). At the time of writing, the construction of the plant was nearly complete. This paper presents an update of project progress, covering commissioning and ramp up of the GPM Gold Project.

The commissioning phase occurred over June to August 2014, which included an M3,000 IsaMill^{TM} fine grinding plant, a 6 tph limestone milling plant, a 60 tpd vacuum pressure swing adsorption oxygen plant, a 10 m residue thickener and a 12 tph Albion Process^{TM} oxidative leach plant. Since commissioning was completed, ramp up occurred over the following three months with downstream gold recoveries from cyanide leaching reaching over 98%.

INTRODUCTION

The GPM Gold Project

The GPM Gold Project is owned by GeoProMining Gold LLC and is located in Armenia. The project consists of an open cut mine at Zod, near the Azerbaijan border, and a processing plant at Ararat near the Turkish border. The Ararat plant has a milling and flotation facility built during the soviet era, with a capacity of 1 million tonnes annually, and a CIL plant built in 1997, with a capacity of 1.5 million tonnes annually. The gold bearing ore, mined at the Zod Mine, is transported to the process plant at Ararat via a state owned rail link.

The Zod deposit originally consisted of weathered oxide ores overlying deeper sulphides. Arsenopyrite and pyrite are the major sulphide minerals. Historical mining has depleted the oxide ores, and the processing plant at Ararat currently treats significant quantities of sulphide ore with increasing amounts of gold locked within refractory sulphides.

The mineral reserves for the Zod mine, at August 2011, were estimated to be 14.2 million tonnes at a gold grade of 4.3 g/t. The mineral resources were estimated to contain 28 million tonnes of ore in the indicated category, at a gold grade of 4.2 g/t and 16 million tonnes of ore in the inferred category, at a gold grade of 4.2 g/t.

GeoProMining are expanding the Ararat plant by re-commissioning an existing flotation concentrator to recover a sulphide concentrate from the ore, and constructing an Albion Process (Hourn & Turner, 2012) plant to oxidize the refractory concentrates. The Albion Process plant will convert the sulphides to oxides, breaking down the sulphide matrix and liberating gold and silver for recovery. Tailings from the concentrator and the Albion Process plant will be combined and transferred to an existing CIL plant to recover the gold and silver as bullion.

Refurbishment of the concentrator and construction of the Albion Process plant commenced in 2013, with commissioning completed in August 2014. The Albion Process plant will has a design capacity to process up to 110,000 tonnes per annum of refractory concentrate from the concentrator.

Deposit Geology

The Zod deposit is located in the Vardenis District of Western Armenia within a setting of volcanogenic and volcanogenic-carbonate sequences, with gabbro-peridotite intrusions that have metamorphosed to serpentinite (Konstantinov & Grushin, 1970).

Gold mineralization is associated with carbonate alteration of ultramafic rocks and is commonly hosted within hydrothermal alteration zones, represented by talc carbonate and quartz-carbonate assemblages. The ore is moderately hard with a medium level abrasion index.

Gold occurs as native free gold, finely dispersed gold in arsenical sulphides, gold tellurides and secondary native gold remaining after oxidation of sulphides and tellurides. Silver occurs in its native form in quartz, chalcopyrite and pyrite, and as silver tellurides.

The deposit has an average sulfur grade of 1.4 % w/w, with an average gold and silver grade of 4.54 g/t and 4.65 g/t, respectively. The arsenic grade across the deposit is 0.3% w/w. The majority of the sulphides occur as relatively coarse mineral grains. The dominant gangue minerals are quartz, talc and chlorite, with minor magnesite, dolomite and calcite.

Development Testwork

Development testwork for the GPM Gold Project began in 2009 initially with batch testwork, and culminated in a continuous flotation and Albion Process pilot plant run over the months May and June 2010. Approximately 4,600 kg of sulphide ore samples were collected from across the Zod ore body to support the testwork program. The samples were classified by ore type, spatial location and sample type and blended into 163 composites. The composites were then grouped into the four major orebodies identified in the primary sulphide resource – orebodies 1, 4, 16 and 23.

Diagnostic leaching and ore characterization testwork (Rohner & Andreatidis, 2010) confirmed that an average gold recovery of only 48% w/w was possible from the blended Life of Mine ore adopting conventional carbon in leach (CIL) leaching methods, and that the majority of the refractory gold was present in arsenical minerals, such as arsenopyrite and arsenical pyrite. Laser ablation work showed that the majority of the pyrite had levels of arsenic in the lattice, averaging 0.9% w/w.

Comminution testwork focused on generating comminution modeling parameters to determine the capacity of the existing crushing and grinding circuit at the Ararat plant. The ore displayed an average Bond Crushing Index of 10 kWh/t, an abrasion index of 0.085 and an unconfined compressive strength of 59 kN. The Bond Ball Mill work Index was 16.5 kWh/t and the Bond Rod Mill work index was 15.8 kWh/t. Modeling work by SMMC (Morrell, 2010) confirmed the milling circuit at the Ararat plant would be capable of processing between 0.9 and 1 Mt/a of ore from the Zod deposit, with minor refurbishment.

Batch and locked cycle flotation campaigns were completed on the testwork samples and a flow sheet consisting of a bulk roughing and single cleaning stage was developed and taken forward into a continuous pilot run. The continuous pilot plant testwork proved a sulphide recovery of 93% could be achieved from the Zod ores, at a mass recovery of 9 – 10%. Gold recovery to the sulphide concentrate was 87%, at a silver recovery of 91%. The sulphide grade of the concentrate was in the range 16 – 18%.

The analysis of the blended pilot plant cleaner concentrate is presented in Table 1.

Table 1 – Cleaner concentrate XRD data.

Mineral	**Chemical Formula**	**Conc (%w/w)**
Unidentified		20.03
Arsenopyrite	$FeAsS$	6.78
Chalcopyrite	$CuFeS_2$	2.43
Clinochlore	$(Fe,Mg)_3Fe_3AlSi_3O_{10}(OH)_8$	1.58
Dolomite	$CaMg(CO_3)_2$	1.30
Galena	PbS	0.40
Magnesite	$MgCO_3$	2.80
Plagioclase (Albite)	$NaAlSi_3O_8$	4.50
Pyrite	FeS_2	36.63
Pyrrhotite	FeS	5.60
Quartz	SiO_2	3.90
Talc	$Mg_3Si_4O_{10}(OH)_2$	17.58

The flotation tailings contained 13% of the gold, and CIL testwork indicated a gold recovery of 60% could be achieved from the flotation tailings at modest reagent demand. The CIL plant at Ararat has a capacity well in excess of the 1 million tonnes per annum treatment rate for the project, and so co-treatment of both the oxidized flotation concentrate and the flotation tailings was incorporated in the design.

IsaMill signature plot testwork on the pilot plant composite concentrate sample returned a specific grinding energy of 59 kWh/t to grind the concentrate to the target 80 % passing size of 10 µm. The current plant is operating at a coarser grind of around 12µm with a specific energy of 45 kWh/t. To date the coarser grind has not impacted precious metals recovery.

Extensive testwork was carried out to determine the best oxidative leach pH for the finely ground concentrates. The testwork examined oxidative leaching under mildly acidic conditions for selective oxidation of the arsenopyrite minerals, and leaching at a more neutral pH. All tests were carried out under atmospheric pressure, with oxygen gas as the oxidant.

Leaching at near neutral pH was ultimately chosen for the oxidative leach. Leaching at near neutral pH allowed lower cost materials of construction to be used in the leaching circuit, and resulted in a final residue with more stable arsenic phases when tested in accordance with the USEPA TCLP protocol. Cyanide and lime demands were lowest for the residue generated under near neutral pH, and the gold and silver recoveries were higher.

The two major oxidative leach reactions observed under the near neutral oxidative leaching conditions were:

Pyrite: $FeS_2 + 15/4O_2 + 9/2H_2O + 2CaCO_3 = FeO.OH + 2CaSO_4.2H_2O + 2CO_2$

Arsenopyrite: $FeAsS + 7/2O_2 + 2H_2O + CaCO_3 = FeAsO_4 + CaSO_4.2H_2O + CO_2$

Confirmatory bench scale oxidative leaching testwork was then carried out on flotation concentrates from the four main orebodies at the Zod deposit. Economic modeling work that compared the capital and operating costs for the Albion Process plant at varying levels of sulphide oxidation was carried out using the batch leach test results. The modeling work showed that a sulphide oxidation of 70% returned the highest Net Present Value for the project, at a hurdle rate of 10%. The gold recovery at this level of oxidation was 93%.

Continuous pilot plant oxidative leaching testwork was then carried out on a blended concentrate. The continuous pilot testwork confirmed a sulphide oxidation of 70% was required to achieve an average gold recovery from the blended feed of 93%. The average silver recovery was 80%. A design oxidation target of 75% was taken forward into detailed design of the oxidative leach circuit. The oxygen demand for the concentrate to achieve the design oxidation of 75% was 336 kg/tonne, and the limestone demand was 326 kg/tonne. Mass and heat balance modeling indicated an average operating temperature in the oxidative leach circuit of 96^{o}C.

PROCESS PLANT DESCRIPTION

Plant General

The Ararat Process Plant experiences a hot, dry summers and cold winters, with an absolute maximum temperature of +42°C; and a minimum of -30°C. The region is classified as semi-desert with an average rainfall of 238 mm. The maximum 10 days depth of snow mantle is 35 cm, at a design snow load pressure of 70 kg/m². The area is seismically active, and the plant is designed to survive a magnitude 7.2 earthquake. A flow sheet for the GPM Gold Project is shown in Figure 1.

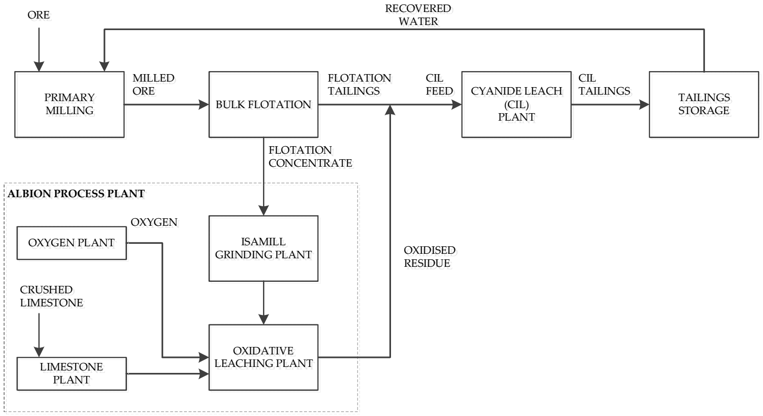


Figure 1 – GPM Gold Project overall flow sheet.

Comminution and Flotation

Ore is mined at the Zod Mine by open cut methods and delivered to a run of mine stockpile for blending ahead of crushing. Crushed ore is trucked to a stockpile at a rail siding and loaded into rail cars for transport to the Ararat Plant. Ore is recovered at the plant by tippler, into ore storage bins. Ore is recovered from the bins by apron feeder and conveyed to the comminution circuit.

The comminution circuit consists of two parallel milling lines. Each line has a primary 1600 kW Autogenous Grinding mill, operating in closed circuit with spiral classifiers, followed by two secondary 630 kW Ball Mills, each operating in closed circuit with cyclones. The Ball Mills are configured in parallel. Cyclone overflow from the Ball Mills are directed to the flotation feed tank.

The flotation circuit had not been operated for over 10 years and refurbishment works were part of the overall project. The existing flotation plant equipment consisted of two banks with five cells in each, and two banks with four cells in each. Each cell is fitted with dual 45 kW agitators, with a cell volume of approximately 32 m^3. The detailed design, procurement and installation was conducted by GPM and their nominated engineering company separate from Glencore Technology.

Slurry is fed to an agitated 25 m^3 flotation feed tank which overflows to the first cell for conditioning with frother. Conditioned slurry then gravitates to Pre-flotation cells, which will be used for pre-flotation of talc and carbonaceous slimes. The pre-flotation concentrate is gravitated to the tailings pumpbox and then to the 15 m diameter Flotation Tailings Thickener.

Tailings slurry from the pre-flotation stage gravitates to a Rougher Conditioner cell, which is used for conditioning of the slurry with copper sulfate and collector prior to rougher flotation. Conditioned slurry flows to the Rougher/Scavenger cells to produce a rougher concentrate for cleaning. The tailings slurry from Rougher/Scavenger flotation is transferred to the Flotation Tailings Thickener, with the thickened underflow slurry pumped to the CIL circuit.

Rougher concentrate will be transferred to the Cleaner Flotation Bank. The two cleaner banks operate in series with both cleaner concentrates combined as the final concentrate. The final Cleaner Concentrate is pumped to the 10 m diameter Flotation Concentrate Thickener, with the thickened underflow concentrate pumped to the IsaMill circuit for fine grinding. Cleaner tails are recycled to the scavenger circuit.

Albion Process

The Albion Process is a combination of ultrafine grinding and oxidative leaching at atmospheric pressure. The Albion Process™ technology was developed in 1994 by Glencore Technology and is patented worldwide. There are five Albion Process plants currently in operation.

The first stage of the Albion Process is fine grinding of the concentrate. Most sulphide minerals cannot be efficiently leached under normal atmospheric pressure conditions. The process of ultrafine grinding results in a high degree of strain being introduced into the sulphide mineral lattice. As a result, the number of grain boundary fractures and lattice defects in the mineral increases by several orders of magnitude, relative to un-ground minerals. The introduction of strain lowers the activation energy for the oxidation of the sulphides, and enables leaching under atmospheric conditions. The rate of leaching is also enhanced, due to the increased mineral surface area.

Fine grinding also prevents passivation of the leaching mineral by products of the leach reaction. Passivation occurs when leach products, such as iron oxides and/or elemental sulfur, precipitate on the surface of the leaching mineral. These precipitates passivate the mineral by preventing the access of oxidants to the mineral surface.

After the concentrate has been finely ground, the slurry is then leached in agitated vessels with oxygen to oxidize the sulphide minerals. Leaching is carried out under atmospheric pressure, and autothermally. Excess heat generated from the oxidation process is removed through humidification of the vessel off gases.

The average nominal throughput for the Albion Process Plant is 94,007 t/a of cleaner concentrate, with a design factor of 15 % applied to the average rate to achieve a design rate of 108,108 t/a. The average gold and silver throughput is 127,000 and 131,000 ounces per annum, respectively.

The feed rate to the IsaMill Fine Grinding circuit is 12.1 t/h, with a design feed rate of 13.9 t/h of concentrate and a final 80% passing size of 11 µm. Since commissioning, the IsaMill energy demand to achieve this grind size is typically 800 kW. The available drawn power for an M3000 IsaMill is 1,120 kW and this mill was chosen for the ultrafine grinding circuit. The completed IsaMill circuit at Ararat is shown in Figure 2. The current grind size is typically 80% passing size of 11 – 12 µm with a specific energy of 45 kWh/t which results is above target downstream gold recovery.

Finely ground slurry is then pumped to an agitated ground concentrate storage tank. The oxidative leach circuit consists of nine 240 m^3 Albion Leach Reactors, each with a live height of 9.4 meters and a diameter of 5.4 meters. Each reactor is agitated by a 160 kW dual impeller agitator, with oxygen delivered by a bank six HyperSparge oxygen injection lances in each reactor. The HyperSparge units are shown in Figure 3.

The slurry pH is maintained at 5.0 – 5.5 in each reactor by limestone slurry dosing.

Figure 2 – IsaMill grinding circuit.

Figure 3 – HyperSparge units.

The design rate of sulphide oxidation within the oxidative leach is 1800 kg/h. Under the near neutral pH conditions employed in the oxidative leach, sulfate is the reaction product of sulphide oxidation, with a design oxygen requirement of 3750 kg/h. The Albion Leach Reactors have all been designed to achieve an oxygen transfer rate of 4700 kg/h. The design oxygen capture efficiency in the leach train was 80 %. Site survey data collected to date suggests that the oxygen capture efficiency currently being achieved exceeds 90%.

The oxygen mass transfer rate for the oxidation of the sulphide minerals is defined by the following equation (Shuler and Kargi, 2002):

$$\text{Oxygen Transfer Rate} = K_La\,(C_{sat} - C) \qquad (1)$$

where:

K_L = liquid film mass transfer coefficient for oxygen into the slurry, in units of $m.s^{-1}$

a = the specific gas surface area, in units of $m^2.m^{-3} = m^{-1}$

C_{sat} = the solubility of oxygen in the slurry at saturation, in units of $g.m^{-3}$

C = the steady state oxygen level in the slurry, in units of $g.m^{-3}$

The "K_L" and "a" terms are typically combined in the form of a mass transfer coefficient for the system. The design K_La for the Albion Leach Reactors is 0.12 s^{-1}. Oxygen gas has poor solubility in water, and so mechanical devices such as agitators and spargers are required to assist the mass transfer. In the Albion Leach Reactor, oxygen gas is sparged into the vessel using the HyperSparge supersonic gas injection lances. The HyperSparge oxygen injection system achieves very high oxygen mass transfer rates at the interface between the supersonic gas jet and the impinging slurry, reducing the amount of power required from the agitation system.

The agitator drawn power required to achieve the design mass transfer coefficient was determined using an empirical correlation of the form (Nielsen and Villadsen, 1994):

$$K_L a = A * U_s^{\alpha} * (P_g/(p_{SL}\ V))^{\beta} \qquad (2)$$

where:

A = a constant specific to the ionic strength of the leach solution

U_s = the gas superficial velocity in the reactor, in units of $m.s^{-1}$

P_g = the agitator drawn power under gassed conditions, in units of Watts

p_{SL} = the density of the slurry, in units of $kg.m^{-3}$

V = the volume of the slurry, in units of m^{-3}

α, β = dimensionless empirical constants

The A, α and β parameters used for sizing the agitator were determined based on over 900 laboratory and pilot mass transfer tests. This correlation has been used successfully in the scale up of all operating Albion Process plants to date. A drawn power requirement of 120 kW per Albion Leach Reactor was determined using the correlation.

The residence time for the oxidative leaching circuit was designed based on the specific rate constant for pyrite leaching measured in the batch and continuous leaching testwork. Pyrite oxidation under near neutral pH conditions is first order (Singer and Stumm, 1970), allowing a simple scale up. The residence time scale up was based on the method of Henein and Beigler (Henein & Beigler, 1988). A design residence time of 40 hours was calculated for the oxidative leach circuit.

Each Albion Leach Reactor was fabricated from lean duplex alloy steel having a diameter of 5460 mm and a live height in the range 9100 – 8100 mm. The Albion Leach Reactors were supplied in modular sections for rapid assembly on site. Each Reactor was constructed from 15 panels, each with a height of approximately 2.0 m and an arc length of 5.9 m. These panels were all fabricated off site and imported to the plant site in shipping containers. Baffles, slurry risers, leach tank lids, agitator support platforms and off gas stacks were all provided as part of the modular Glencore Technology equipment supply. Assembly of the oxidative leach train was rapid, with all nine leach reactors and two slurry storage tanks complete within 8 weeks. The final two tanks were erected in approximately three days each. The completed oxidative leach train is shown in Figure 4.

Overflow slurry from the oxidative leaching circuit will gravitate via a slurry sampler to a 10 m diameter thickener and be thickened to 45 %w/w solids prior to transfer to the CIL circuit. Thickener overflow is returned to the leach circuit to compensate for evaporative losses in a density control loop.

Figure 4 – Oxidative leach circuit.

A limestone plant with a capacity of 6 t/h was installed to generate limestone slurry for neutralizing duty. Limestone for the oxidative leach will be milled to an 80 % passing size of 75 microns in a 132 kW overflow ball mill operating in closed circuit with cyclones. Cyclone overflow will report to a 150 m^3 agitated distribution tank and be circulated through the oxidative leach train by a ring main. Individual dosing lines will add limestone slurry to each Albion Leach Reactor. The limestone distribution tank was a 150 m^3 ZipaTank zip join tank – the first of its kind the in the world. The tank was erected in approximately 5 days and was internally sealed with specially selected paint. The joins sealed on the first filling. The limestone distribution tank is shown in Figure 5.

Figure 5 – Limestone storage tank.

Two 60 t/d VPSA oxygen plants will operate in parallel to provide oxygen to the Albion Process Plant. Oxygen will be delivered from each plant at a maximum flowrate of 1,745 Nm^3/h, at a purity of 93% v/v.

The thickened oxidative leach residue and thickened flotation tailings will report to a 100 m^3 mixing tank and be blended prior to feed to the CIL plant. The CIL Plant will process 137.5 t/h of feed comprised of oxidized residue and flotation tailings. All six existing CIL tanks will be utilized, providing a total residence time in the CIL circuit of 41 hours. The CIL Plant is expected to consume

5.3 kg/t of sodium cyanide and 10 kg/t of lime. Carbon levels in the CIL Plant will be 10 – 15 kg/m^3, with a design carbon loading of 2,500 g/t. Carbon movements will total 7.5 t/d, and the existing dual AARL elution circuits will be used for carbon processing.

CIL Plant tailing gravitates to a cyanide destruction plant prior to being pumped to tailings. The tailings will be deposited within the existing tailings impoundment, approximately 6 km from the Ararat plant site.

Plant control is achieved through a Distributed Control System (DCS) located in a centralized control room between the concentrator and Albion Process plants. Training for field and control room operators was provided by GT and sub-contractors as part of the commissioning process. The control room is shown in Figure 6.

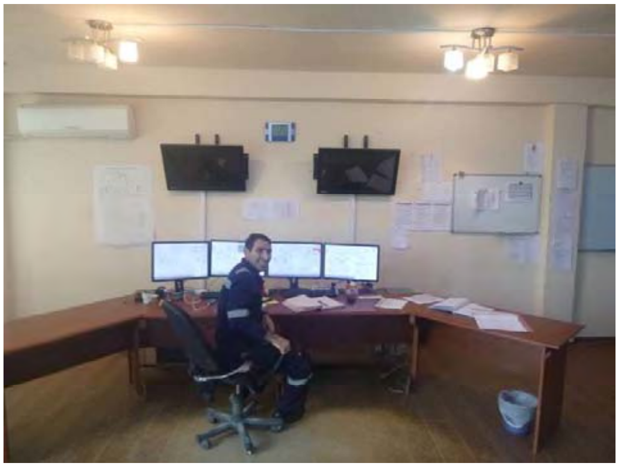


Figure 6 – Central control room for concentrator and Albion Process^{TM} plant.

PROJECT STATUS AND PLANT PERFORMANCE

The Albion Process plant was provided to the GPM Gold Project as a Lump Sum technology package by GT. The package included all detailed design, mechanical equipment, electrical, instrumentation and control equipment, structural steel, flooring, handrails, piping and valves. The scope of supply includes the fine grinding plant, oxidative leaching and thickening plant and the supporting limestone, oxygen, flocculent and caustic reagent plants.

Mechanical design was completed in December 2012, with the majority of mechanical equipment and fabricated components delivered to site by May, 2013. Site civil works were completed in March 2013. Construction was completed in April 2014.

The pre-commissioning phase was conducted during April/May 2014 with wet commissioning commencing in May/June 2014 and completed in August 2014. The commissioning was managed and coordinated by Glencore Technology. The commissioning team comprised three permanent Glencore Technology personnel (manager/process engineer, mechanical engineer and instrumentation engineer) supported by equipment specialists brought to site during crucial commissioning events. The GPM site team provided all other support.

The main setback during commissioning was the failure of an oxygen plant blower which had a lead time of 12 months to replace. Dual oxygen plants were supplied to the project, each with the capacity to oxidise 70% of the design sulphide feed, and so the blower failure has not impacted on plant throughput to date. The blower will be repaired and the second oxygen plant will be in service by May 2015.

The main setback for ramp-up has been the lack of feed quantity and quality from the refurbished concentrator. A project is in place to improve concentrator performance with expected results by October 2015.

As of March 2015 the refurbished concentrator was running at around 60% capacity with recent assistance from GT during March increasing throughput by 30%. The Albion Process plant performance has not been impacted by the slow ramp up of the concentrator, with the plant regularly achieving 95% gold recovery with around 50% sulfur oxidation.

A survey of the nine Albion Leach Reactors was collected to determine tank-by-tank sulfur oxidation and resulting gold extraction. The sulfur oxidation was determined using a Leco sulfur analyzer. Gold extraction was determined by subjecting each collected sample to a bottle roll test at the GPM laboratory and cross checked with an agitated CIL test at hrltesting laboratory in Brisbane. The sulfur oxidation against gold recovery from the plant and compared to pilot plant results is shown in Figure 7.

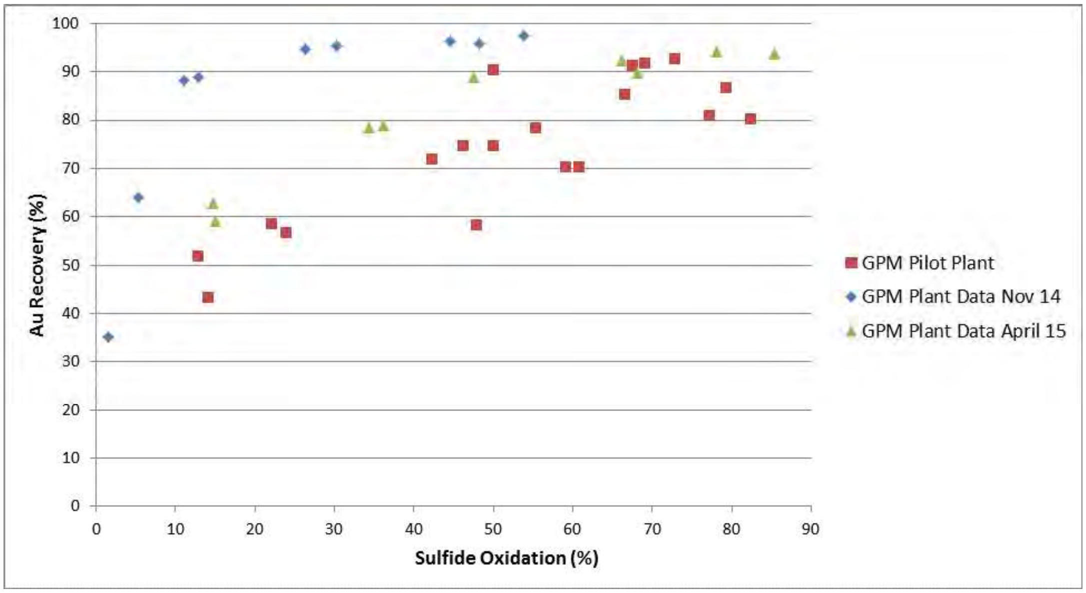


Figure 7 – Sulfide oxidation against gold recovery comparing pilot plant data and actual operating data.

Testwork done on individual components of the pilot plant feed are consistent with this level of oxidation and corresponding gold recovery.

A profile of sulfur oxidation and gold recovery was collected down the nine leach reactors. The data are presented in Figure 8.

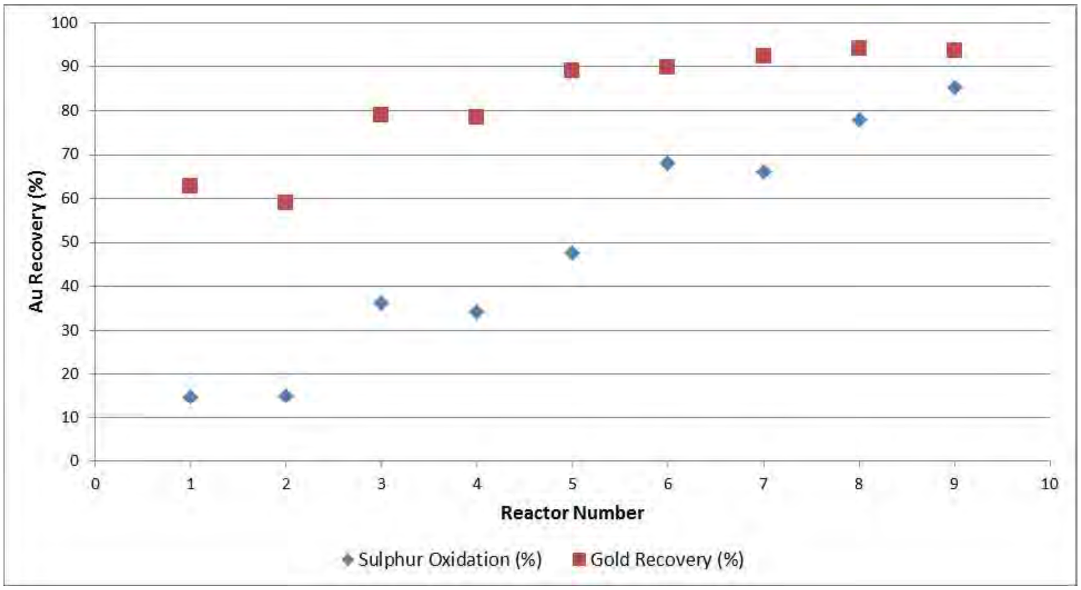


Figure 8 – Profile of sulfide oxidation and gold recovery down the leaching train. Plant is on reduced rates.

Figure 8 shows the profile of the sulphide oxidation and gold recovery down the leach train. Although one oxygen plant is operating, there is sufficient oxygen available to increase the throughput rate if feed was available from the concentrator.

The current oxygen consumption is operating at around 215 kg / t concentrate which is below the design value of 336 kg / t owing to higher than design oxygen utilization and lower oxidation levels for this feed material compared to the design case.

Current limestone consumption is very low owing to higher entrainment of acid consuming gangue with the flotation concentrate. Steady feed composition at full production rates will allow limestone consumption to be better analysed.

Cyanide consumption in the CIL plant is within the range expected during pilot testwork. The plant is currently operating at 1.8 to 2.2 kg sodium cyanide per tonne of feed to the CIL plant which is the combination of leach residue and flotation feed.

CONCLUSIONS

The GeoProMining Albion Process^{TM} plant was commissioned successfully over a 14 week period. The plant is achieving greater than 95% gold recovery in the cyanide leaching plant, consuming 1.8 – 2.2 kg cyanide / t combined leach residue and flotation tails.

The plant is running on reduced rates due to concentrate feed availability, as mine development has been slower than planned. GT continues to work with GPM to improve the concentrator performance. At the end of April 2015, GPM achieved a 30% increase in concentrator throughput with the assistance of GT. The second oxygen plant commissioning is scheduled for June 2015, and full plant throughput should be achieved by July 2015. An update will be provided after the plant is at full capacity with more comprehensive performance data including oxygen and limestone consumption as well as sulfur oxidation, oxygen utilization and resulting gold recovery.

ACKNOWLEDGEMENTS

GT wish to thank GeoProMining Gold LLC for the permission to publish this paper. This paper was first published in proceedings of ALTA Conference in 2015, Perth, Australia.

REFERENCES

Henein H, & Biegler, L, 1988. Optimization Of Reactor Volumes For Topochemical Reactions. *Trans. Inst. Min. Metall.*, 97:215-223.

Hourn, M, and Turner, D, 2012. Commercialisation of the Albion Process, in *Proceedings ALTA Gold Conference 2012*, pp 231-248 (ALTA Metallurgical Services: Melbourne)

Hourn, M, Voigt, P, and Turner, D, 2014. Development of the Albion Process plant to treat refractory concentrates from the GPM Gold Project, in *Proceedings Hydroprocess Conference 2014*, (Gecamin; Santiago)

Konstantinov, M and Grushin, V, 1970. Geologic Position of the Zod-Agduzdag Gold-ore Node in Transcaucasia. *International Geology Review*, 12(12):1447-1453.

Morrell, S, 2010. Estimates of the Throughput of the Ararat Comminution Circuit. Personal communication. June.

Nielsen, J and Villadsen, J, 1994. *Bioreaction Engineering Principles*, pp 27-32 (Plenum Press: New York).

Rohner, R and Andreatidis, J, 2010. Bankable Feasibility Report, Geoprominng Gold LLC, Zod Upgrade Project Volume 3, Metallurgy And Processing. Personal communication. June.

Shuler, M and Kargi, F, 2002. *Bioprocess Engineering: Basic Concepts*, 2nd ed. pp171 (Prentice-Hall: New Jersey).

Singer, P and Stumm, W, 1970. Acidic Mine Drainage: The Rate Determining Step. Science, 167(3921):1121-1123.

EVALUATING PROCESS OPTIONS FOR TREATING SOME REFRACTORY ORES

By

[1]Mark Aylmore and [2]Ashraf Jaffer

[1]Bateman Engineering Pty Ltd Australia

[2]Bateman Engineering Pty Ltd South Africa

Presenter and Corresponding Author

Mark Aylmore
Mark.aylmore@bateman.com

ABSTRACT

A significant contemporary challenge for gold mining companies is to define economic process options for treating more complex ore bodies as the less refractory ores become depleted. Demonstration of economic treatment options allows gold companies to maintain and increase their ore reserves.

The selection of options to treat a specific ore is significantly impacted by factors such as mineralogy, precious metal grades and deportment, gold to sulphur ratios and hazardous impurities. Bateman Engineering has carried out a number studies to help mining companies evaluate further the potential for developing ore bodies.

This paper provides two examples of work to evaluate the potential for treating a low grade ore with high carbonate and arsenic content and an ore with high silver grade. Capital and operating cost estimates for different methods of pretreatment were prepared and these estimated costs and ore sulfide grades then used to compare the process economics.

INTRODUCTION

A significant contemporary challenge for gold mining companies is to define economic process options for treating more complex orebodies as the less refractory ores become depleted. Demonstration of economic treatment options allows gold companies to maintain and increase their ore reserves.

Refractory gold ores are those that do not yield high gold recoveries in conventional cyanide-leaching circuits, even when the ore is finely ground. The cause of low gold recoveries can be either through the presence of naturally occurring 'preg robbing' carbonaceous materials or gold physically 'locked' in sulphide minerals. The majority of gold ores are classified refractory because the gold mineralogy is contained in sulphides. This type of gold is typically very fine gold or locked in solid solution within sulphide minerals pyrite and arsenopyrite (FeAsS). The arsenic bearing pyrite is commonly known as arsenian pyrite and can contain gold in its structure. Rimming of pyrite with arsenic-rich assemblages is often common as well as arsenopyrite-marcasite-pyrite associations. The gold-pyrite association is dependent on ore grade and textural features of the ore. When present with pyrite, arsenopyrite commonly will have significantly higher gold grades than the pyrite associations. Enargite (and Luzonite) is a common copper-arsenic mineral that can be a significant gold carrier. Other less common are chalcopyrite and other sulphides and sulphosalts such as pyrrhotite, covellite, arsenides, and antimonite

To render gold 'locked' in the sulphide mineral amenable to cyanidation a chemical pre-treatment is required. No matter what process option is used it generally involves the oxidation of sulphide minerals which produces acid. The reaction products and overall reaction stoichiometry is determined by the reaction conditions which range from mild in alkaline leaching and biological leaching to the aggressive conditions found in pressure oxidation. The gold and sulphur (sulphides and elemental sulphur) grades relate directly to the economics of the process.

Much has been written about the suitability of particular pretreatment processes, such as pressure oxidation, bio oxidation and roasting, for specific ores bodies(1-4). More recent technologies such as the Albion Process can be potentially comparable or lower cost compared with the conventional technologies(5). The most significant item affecting the economics is the amenability of the ore to flotation where gold can be concentrated in a small mass pull and reduce size and cost of a plant(6). In addition the deportment/management of the acid produced in the oxidation process is a major factor. In BIOX and pressure oxidation all of the acid must be neutralized resulting in a large operating cost (limestone and lime) and a significant volume of anhydrite precipitate that must be disposed of whereas in roasting acid is a saleable by-product.

Bateman Engineering has carried out a number of comparative capital and operating cost studies to allow mining companies to develop the potential for further investigations. This paper provides two examples of work to evaluate the potential for treating two different ore bodies located in different parts of the world. Due to confidentiality the identity and location of these two ores cannot be given so they have been designated as Ore A and Ore B. Capital and operating cost estimates for different methods of pretreatment were prepared and these estimated costs and ore sulfide grades then used to compare the process economics.

The Study involved the following activities:

- Review existing mineralogy and testwork data.
- Review possible refractory gold flowsheets.
- Conduct a preliminary comparative assessment of the options.
- Develop conceptual level Capital and Operating Cost estimates (+/- 40%) for each viable processing flowsheet option.

The battery limits for this scope of work was ROM feed through to plant discharge. Utilities such as power and water were assumed to be available on site. It should be pointed out that this level of study is restricted to the processing plant and that development costs associated with mining, access to site and access to power lines and water are not included but can be considerable.

Based on the mine models ore A will be a low throughput plant at 0.6 Mtpa containing high arsenic and carbonate content, whereas ore B will be treated at higher throughput of 3.65 Mtpa, containing high silver with low arsenic and carbonate contents. Although more geometallurgical studies are required to define the variability of the ore bodies the concept studies carried out on these ores allowed determination of the potential viability of different processes to treat these ores.

ORE CHARACTERISTICS

It is essential that the ore body be defined and characterised in terms of mineralogical composition and deportment of gold and silver to allow for efficient targeting of metallurgical test work and subsequent process design.

Mineralogy

Mineralogical analyses were carried out on both ore type composites. The expected average composition of the two ore bodies is presented in Table 1.

Table 1: Average Mineralogy

Minerals	Ore A	Ore B
	g/t	g/t
Gold grade	2.25	1.04
Silver grade	-	32
	wt%	wt%
Arsenopyrite	2	0.01
Pyrite	1.8	13.8
Copper Sulphides	-	0.15
Quartz and feldspars	66.2	72
Dolomite	20.0	
Muscovite	5.0	5.84
Iron Oxides	5.0	1.44
Jarosite/Alunite	-	2.62
Other Gangue	-	4.14

Studies on the low throughput A ore indicate that the mineral assemblages comprise of pyrite and arsenopyrite minerals as the main gold carriers with quartz, dolomite, muscovite, albite and goethite the main gangue components. A gold deportment study carried out indicated that the majority of the gold is present mainly as very fine inclusions (<10 µm) or in solid solution associated with the arsenian pyrite and arsenopyrite. The distribution of gold based on size is illustrated in Figure 1. Some pyrrhotite intergrowths with other sulphide minerals such as chalcopyrite and sphalerite also occurred but are minor. Most of the work indicated that there is little gravity free-gold available. Most of the particulate gold occurs as inclusions in both pyrite and arsenopyrite grains with only about 17% occurring peripheral to the sulphides.

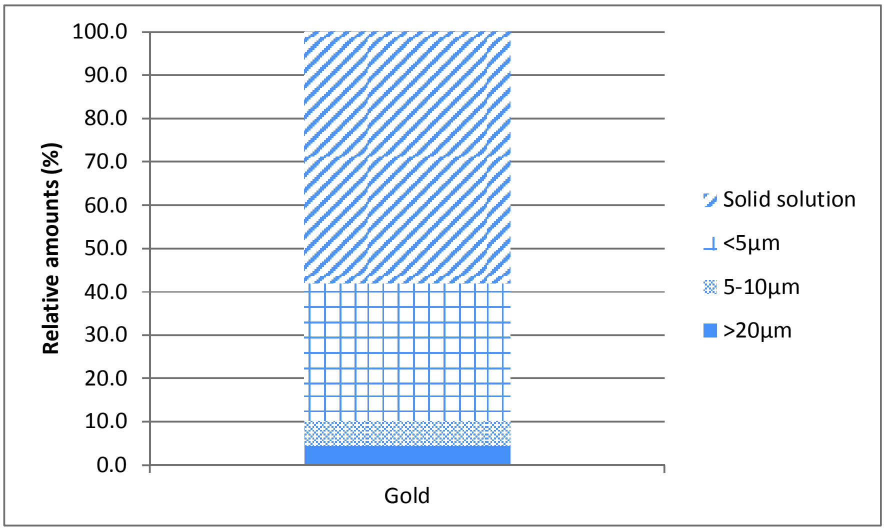


Figure 1: Gold deportment in Ore A

The high throughput ore B comprise predominantly of pyrite and small copper sulphide minerals as the main gold carriers with quartz, dolomite, muscovite, feldspars the main gangue components. Dynamic SIMS analysis studies indicate that the majority of the gold in the sample is sub-microscopic, present as finely disseminated colloidal size inclusions (<0.5 µm) or in solid solution associated with the pyrite (namely arsenian pyrite, FeS_2 containing As). Four different morphological types of pyrite were identified in the sample as coarse, porous, fine and disseminated pyrites. No deportment data on silver is available but some electrum and tellurides have been reported in the geological reports. Notably the low recovery of silver observed in the flotation concentrate for high sulphide recovery discussed below indicated that the silver minerals do not correlate directly with the sulphide minerals. The slow leaching kinetics of at least half the silver could be due to the association of silver with the mineral acanthite (Ag_2S) which is slow leaching.

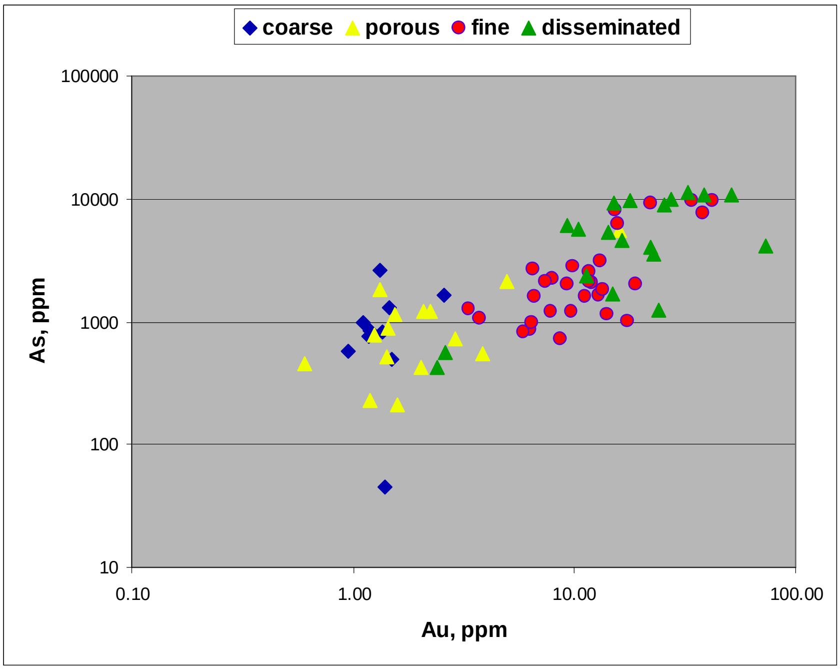


Figure 2: Correlation between measured concentrations of sub-microscopic gold and arsenic in different morphological types of pyrite (from 74 point analyses)

Comminution

As mining of both ores are in the early stages of development comminution data is limited for determining the mill sizes and power for the ore grinding circuits. For Ore A a Bond ball mill work index value of 17 kWh/t was reported. For ore B a Bond ball mill work index value of 15 kWh/t was used. Further comminution tests will be required in the future to confirm this is representative of the ores to be processed. The bond index values generally indicate that grinding is moderate power consumer.

Liberation and Concentration

Flotation test work was conducted on both samples and the optimum conditions presented in Table 2. For the low throughput Ore A flotation test work indicated that the majority of gold can be concentrated by concentrating the sulphide component in the ore to mass pulls of 9 to 10 wt%, with gold recovery of 85 to 90%. In addition the majority of carbonate present in the ore is significantly reduced in the concentrates which will reduce acid consumption issues for oxidative pretreatment processes.

Flotation test work on high throughput Ore B indicated that gold can also be concentrated to an extent by concentrating the sulphide component in the ore to mass pulls of 15 to 30 wt%, with gold recovery of 60 to 70%. Maximum Au, Ag and sulphide recoveries of 70%, 45% and 89% respectively were obtained at a grind size of P_{80} of 70µm at a considerably high mass pull to concentrate (30 wt%). The bulk rougher concentrate Au grade averaged about 2 to 3g/t is considered low. Attempts to improve gold recovery and reduce mass in the concentrate with NaSH and or copper sulphate addition were unsuccessful. Further work investigating fine grinding and a cleaning may improve recoveries, but the fine nature of some of the gold bearing pyrite may restrict the effectiveness of such an approach.

Table 2: Flotation recoveries from test work

	P_{80} (µm)	**Reagents**	**Au (g/t)**	**Ag (g/t)**	**S (%)**	**As (%)**	CO_3^{2-} (%)	**Recovery (%)**	**Mass pull (%)**
Ore A									
Concentrate	75		22.5	-	12.9	6.96	3.3	Au 90% S 95% As 90% CO_3^{2-}	9%
Ore B									
Concentrate	70		2.6	48	18.5	0.002	-	Au 70% Ag 45% S 87% As 57%	30%

Cyanide Recoverable gold

Average gold and silver extraction reported in test work for both Ore A and B are illustrated in Figure 3.

Direct cyanidation of Ore A yielded poor gold extractions in the mid twenty percentile for ore ground to a P_{80} of 53 micrometres. Some test work investigated cyanidation of tails but the grade of the gold in the tails was significantly low to be of value (0.2 g/t). Average cyanide consumption was 2 kg/t.

Direct cyanidation of the bulk ore and concentrate for Ore B yielded poor Au (27%) and Ag (56%) recoveries. Cyanide consumption ranged from 1.6 to 2.2 kg/t whereas lime ranged from 2.5 to 3.2 kg/t. Cyanidation of the tails recovered 50% Au and 40% Ag, with Sodium cyanide and lime consumption 0.7 kg/t and 1.5 kg/t respectively. In all cases, the gold and silver leaching kinetics were very fast with most of the extractable gold recovered to solution within two hours. Silver continued to leach at a slower rate, with some evidence of gold loss from solution with time. The copper minerals present in ore B did not appear to significantly impact cyanide consumption.

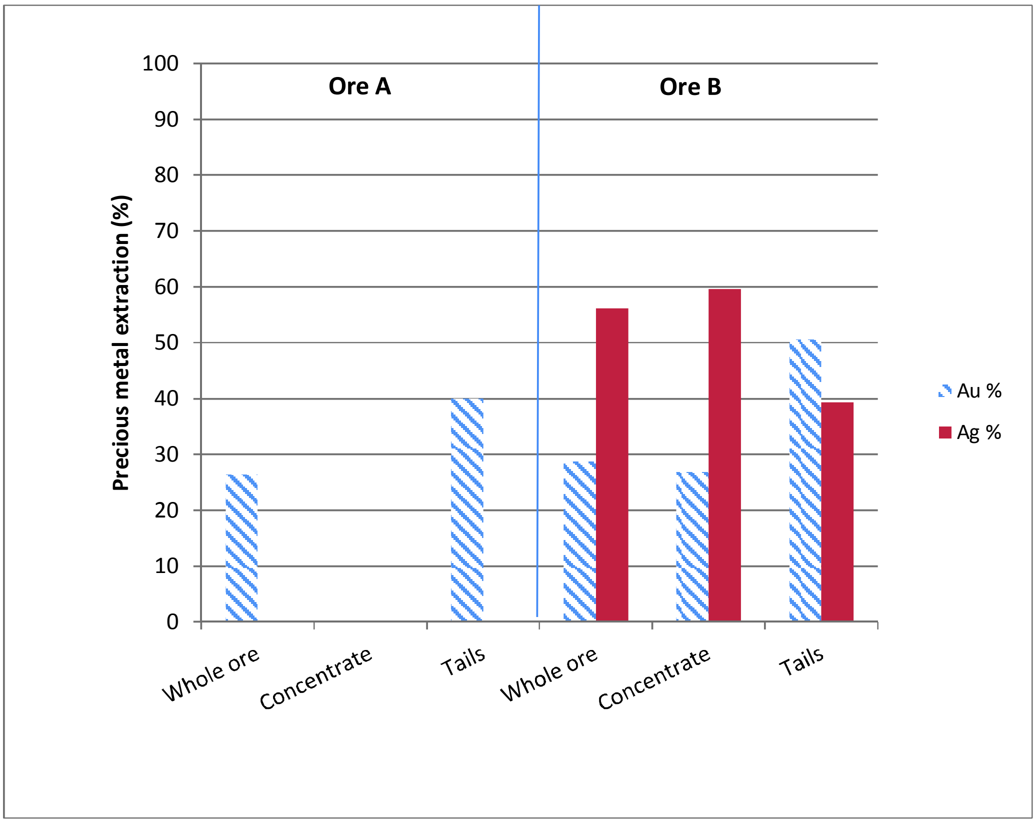


Figure 3: Precious metal extraction following cyanidation on both Ore A and Ore B samples

Oxidation of Sulfides – Liberation of Gold and Silver

Most refractory gold ores and sulphide concentrates require some degree of oxidation of the sulphides to liberate gold locked as 'solid solution' or as very fine inclusions in the gold bearing sulphide minerals. The amount of oxidation required to expose gold for subsequent recovery by cyanidation varies for different ores and concentrates.

For Ore A the limited biooxdation test work indicated that about 80% of the sulphide needs to be oxidised to obtain a maximum gold extraction of 92%. The relationship between gold recovery and sulphide oxidation is shown in Figure 4. The high carbonate content (12%) in the ore A indicates that acid consumption will be high for acidic oxidative leaching processes. Bio oxidation test work on whole ore indicated that extra acid was required to neutralise the carbonates, either by adding acid or providing an extra sulphide source. Cyanide consumption was high at 6 kg/t for ore pretreated by biooxidation.

No test work was carried out on any oxidative process option on the Ore B. The fact that direct cyanide yielded poor gold recovery and gold associations in arsenian pyrite indicated some oxidation of sulphide present would be required. Approximately 57% of the silver is leachable without any oxidation. For this study it was assumed that full oxidation of the sulphide is required for maximum gold recovery.

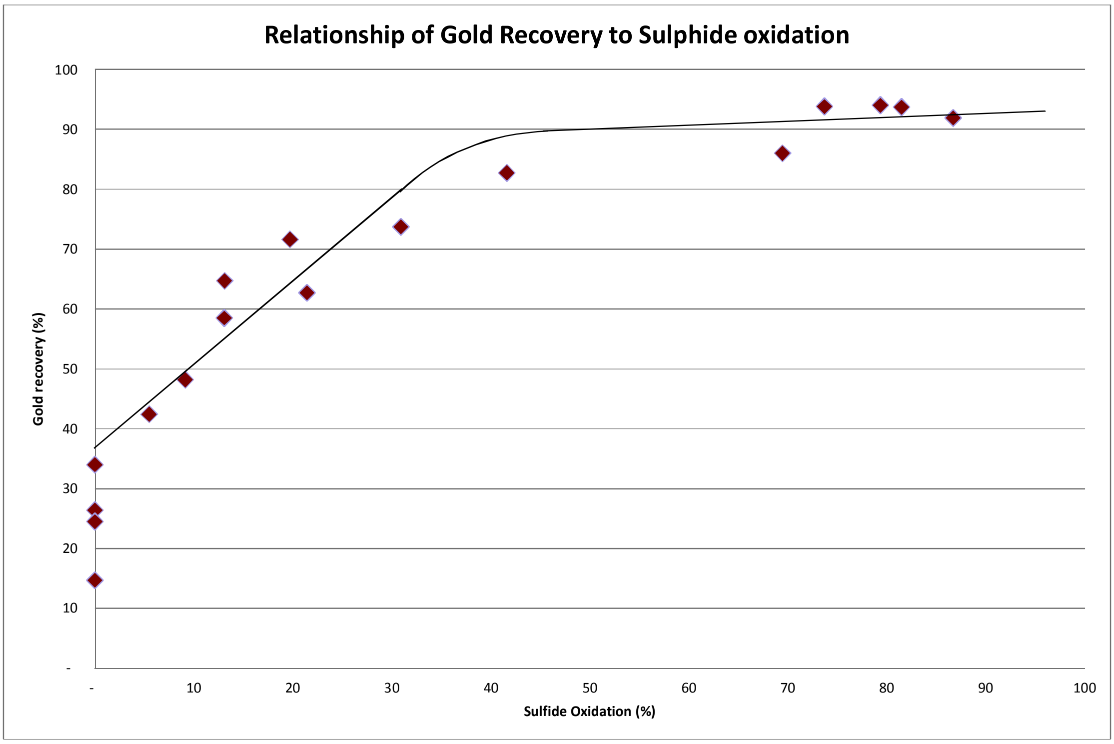


Figure 4: Relationship of gold recovery to sulfide oxidation required for Ore A and concentrate to be amendable to cyanidation

Sulphur is the element the pre- oxidation process converts to sulphate in the forms of sulphuric acid and to a minor extent iron sulphosalts. As such oxygen usage, limestone consumption, and lime consumption used for pH control are directly related to the sulphur grade of the ore or sulphide concentrate being treated. These three reagents/consumables represent a significant portion of the process operating cost. The cost of supply for these three reagents/consumables must be sufficiently offset by the gold grade (plus recovery) and its related revenue in order to make the process commercially economic. Hence the gold and sulphur (sulphides and elemental sulphur) grades relate directly to the economics of the process. Clearly if sulphide oxidation can be reduced oxygen requirement and power cost will be lower.

Figure 5 compares the gold to sulphur ratio of the whole ore and concentrate for Ores A and B with existing and previous plant operations. Notably for Ore A Au:S ratio is mid-range whereas ore B is at the bottom. Even taking the silver value into account ore B is still marginal.

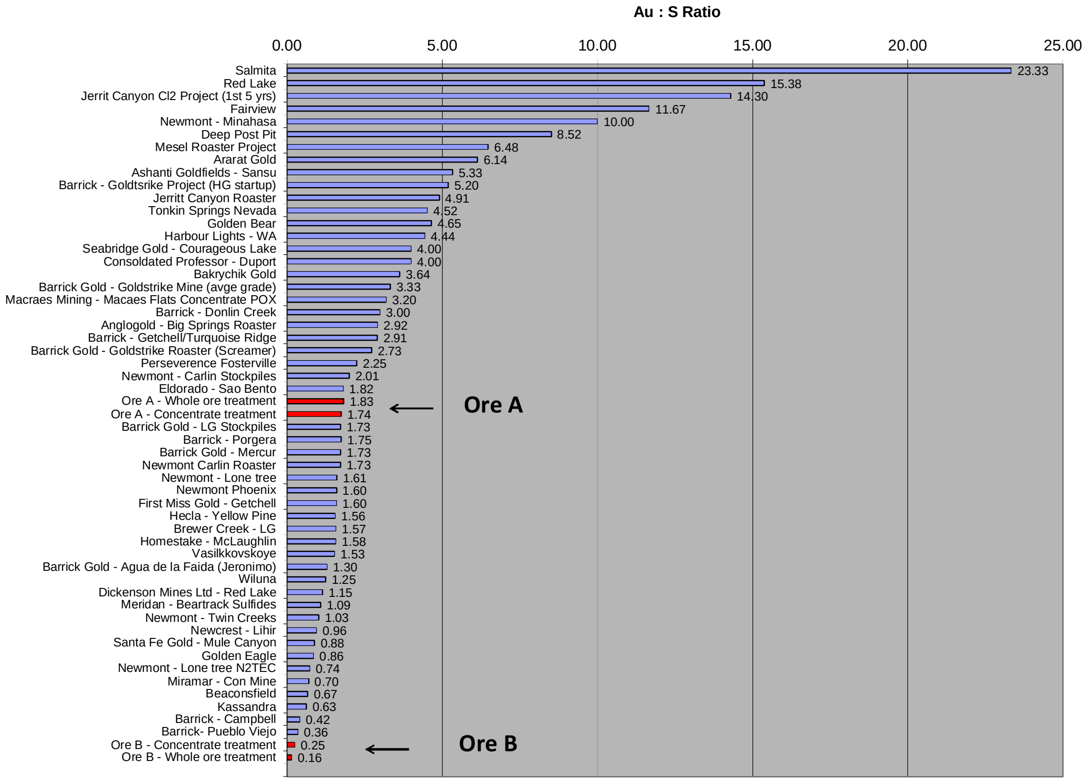


Figure 5: Comparison of gold to sulfur ratio for the two ore types and various existing and previous process operations around the world

With increasing restrictions on environmental disposal and emission the handling of arsenic containing residues is an important consideration in developing process flowsheet. Methods for removal and fixation of arsenic, current industrial practise and current understanding on the stability of arsenic containing residues has been extensively reviewed(6-11). Scorodite ($FeAsO_4.H_2O$) and type II ferric arsenate ($Fe_4(AsO_4)_3$-$(OH)_x(SO_4)_y$) are routinely produced at elevated temperature under pressure in autoclaves. Production of stable ferrihydrite, historically used in atmospheric leach operations, requires limestone or lime addition and a Fe:As ratio of > 4:1 in the precipitation reactor. Current practice is to dispose as a ferric arsenate waste product and store residue at the plant site in a suitably designed containment facility. Arsenic emission and storage of arsenolite (As_2O_3) formed from oxidation during roasting of arsenical ores possess a challenge. The arsenic has to be oxidised to the arsenic (V) state with hydrogen peroxide before adding ferric sulphate and limestone to produce a scorodite product for waste disposal.

Notably both Ore A and B contain arsenic which has to be removed to liberate gold from sulphide minerals. Ore A contains 6% arsenic in the concentrate whereas Ore B is low at 0.002%.

PRE-TREATMENT PROCESSING OPTIONS

The technologies considered for evaluation are listed in the following two tables. In determining which processes are appropriate the following points need to be considered:

- Economics - Capital costs, Operating (non-power) costs, power costs, gold recovery performance
- Risk – Status and availability of technology
- Operability – Scalability, operability and maintainability
- Impacts – community, environmental and security
- Project implementation – infrastructure and constructability

With the high arsenic content in the feed concentrate for Ore A the roasting process was considered likely to be uneconomical. Fixing the arsenic in calcine product is possible but would reduce gold recovery and increase reagent consumption in the subsequent cyanidation stage. Test work would be required to determine viability of an approach. Hence the roasting option was removed from the assessment. Ulfrafine milling the ore or concentrate was considered as a low capital option even though gold recovery would not be expected to be high. Historical test work data ensured that biooxidation heap leach option be considered. The Albion process was considered as a possible viable approach given its development in recent years towards full commercial scale.

For Ore B only the three conventional processes were considered. Notably the high silver to gold ratio in the proposed feed favours the recovery of precious metals via the Merrill Crowe process discussed below.

Table 3: Process Options Considered for ore A

Processing options	Crushing and grinding	Flotation	Pre-treatment	Cyanidation and recovery	Expected Au Recoveries (Overall)
Bacterial Oxidation Heap leach – Whole ore	Tertiary crushing, stacking	None	Heap bio oxidised and then reclaimed	Product processed through a conventional CIL plant	70%
Ultra fine milling – Whole Ore	Ore is crushed and milled to 75µm	None	Ore fined milled to 10µm		50%
Ultra fine milling – Concentrate		Floated for a 9% mass pull	Concentrate fined milled to 10µm		45%
Roasting – Concentrate					
Albion - Concentrate			Fine milled to 10µm Atmospheric oxygenated leach		83%
Bacterial Tank Leach – Concentrate			Bacteria oxidation Residue neutralised and washed		83%
Pressure Oxidation - Concentrate			High pressure oxidation Residue neutralised and washed		85%

Table 4: Process Options Considered for ore B

Processing options	**Crushing and grinding**	**Flotation**	**Pre-treatment**	**Cyanidation and recovery**	**Expected Au & Ag Recoveries (Overall)**
Roasting – Concentrate	Ore is ground to 75µm	floated for a 30% mass pull	Roasted in air and residue quenched	Oxidised concentrate combined with flotation tails and cyanide leached. Solid/liquid separation. Au & Ag in PLS recovered by the Merrill Crowe process.	81.5% Au 62% Ag
Bacterial Tank Leach – Concentrate			Bacteria oxidation Residue neutralised and washed		81.5% Au 62% Ag
Pressure Oxidation - Concentrate			High pressure oxidation Residue neutralised and washed		81.5% Au 62% Ag

DESIGN PARAMETERS

The overall process design based on input from clients and derived from average values obtained in the test work programme are summarized in Table 5.

Expected throughputs, battery limits, expected available process infrastructure and unit costs associated with power and reagent supply were supplied by clients or derived from other work carried out by Bateman.

A basic mass balance for all the process options was generated to determine the stream flow rates and compositions. Sizing of key equipment was based on the technical specifications derived from the mass balance. Pricing of equipment was derived from vendor quotations, together with in-house database costs.

Table 5: Physical and Plant Data

Ore		A	B	
Time Data				
	Life of Mine	8	10	yrs
	Overall Plant availability	91.3	91.3	%
Plant Data				
	Resource	-	50	Mt
	Basis for Design	600,000	3,650,000	t_{ore}/yr
	Flotation Plant production	54,000	1.095,000	t_{concs}/yr
Ore Mineralogical Composition				
	Arsenopyrite FeAsS	2.0	0.01	%
	Pyrite FeS_2	1.75	12.3	%
	Copper Sulphides (Covellite, chalcopyrite,tetrahedrite)	-	0.07	
	Carbonates	20.0	-	%
	Gangue	76.25	86.84	%
	Au	2.25	1.04	g/t
	Ag	na	32	g/t
Concentrate Mineralogical Composition				
	Arsenopyrite FeAsS	20.0		%
	Pyrite FeS_2	18.5	34.7	%
	Copper Sulphides (Covellite, chalcopyrite,tetrahedrite)	-	0.06	
	Carbonate	3.3	-	%
	Gangue	58.2	87.5	%
	Au	22.5	2.5	g/t
	Ag	-	45	g/t
Bond work data				
	Ore	17	15	kWh/t
	Concentrate	30	20	kWh/t

A description of the ore preparation, cyanidation and tails detoxification processes which are applicable for the majority of the process treatment options are presented in the following sections. Details of the pre-treatment processes are described later. The design of tailings deposition facility has not been considered in the study.

Crushing and Grinding

For Ore A crushing will comprise a primary jaw crusher only, the product from which will feed the SAG mill.

For Ore B a primary and secondary crusher system will have been previously installed for treating oxide ore for a heap leach operation. For the purpose of this study the crushed feed from the secondary crusher is transported by conveyor to be screened and cone crushed to yield a P80 of 2000 micron particles.

For both ores the grinding circuit were configured as a SAG Mill / Ball Mill circuit to grind the material to 80% passing 75 microns. The SAG mill will be operated in open circuit and the ball mill in closed circuit with hydro-cyclones. The cyclone overflow will feed either into the flotation circuit or thickener for whole ore processing.

For the whole of ore heap bioleach option for treating Ore A the ore is crushed to 100% passing 9 mm prior to bioleaching. The residue reclaimed from the bioleaching is ground to 80% passing 75 microns in a 5 x 5.5m ball mill.

Flotation Circuit for Concentrate Options

The trash screen underflow from the cyclone overflow gravitates to the conditioning tank where lime, collector and fresh water, if required, are added. The slurry in the conditioning tank flows to rougher flotation tank cells in series with an assumed total retention time of 24 minutes. Flotation air to each tank cell is provided by a dedicated external blower which ensures each cell receives a controlled air flow. The concentrate from the rougher cells is pumped to a high rate concentrate thickener. The thickened flotation concentrate is then pumped to one of the refractory treatment processes.

For Ore A the flotation circuit has been considered to be a rougher-scavenger circuit with a single cleaner which will upgrade the concentrate by a ratio of 10:1. The cleaner tailing will be recycled back to the scavenger feed. The overall recovery is estimated at 90% and the weight recovery 9% based on the results of the flotation test work. The flotation tails with the high residual carbonate content will be used as neutralising material in acidic oxidation processes before being disposed in a tailings dam.

For Ore B the gold and silver recovery is estimated at 70% and 45% respectively, with the weight recovery 30% based on the results of the flotation test work. The thickened tails is pumped to the cyanidation circuit to be treated along with the oxidized residue from the refractory process plant to extract the gold and silver.

Cyanidation and Gold Recovery Circuits

Ore A residue

The gold cyanide leaching circuits were designed on a 24 hour residence time. The same circuit configuration will be used for either whole or concentrate, with the whole ore sized for treating 75 t/h, whereas the concentrate is sized for treating 7 t/h.

Feed from the refractory pre-treatment plant is pumped to a slurry neutralising tank where applicable and on to the Carbon in leach (CIL) circuit for gold recovery to carbon. Gold will be recovered from the leach pulp using granular carbon before stripping in a spilt AARL elution circuit.

Gold loaded to carbon will be recovered in batches from CIL Tank #1 before acid washing with dilute hydrochloric acid and rinsing with water. The loaded carbon will be stripped using the AARL elution process. The pregnant strip solution flows from the elution column into the pregnant solution tank from where it is pumped at a controlled rate to the EW circuit.

The stripped carbon is thermally reactivated in a regeneration kiln at a temperature of 700°C. The reactivated carbon is screened to remove carbon fines before being returned to the last CIL tank to replace the forwarded carbon. The fine carbon is forwarded to a settling pond and periodically recovered and bagged for sale.

Gold is plated onto the electrowinning cell stainless steel wool cathodes. The barren electrolyte flows to a collection tank and pumped to the barren solution storage tank for recycling to the elution circuit. Cells are periodically cleaned of gold sludge. The sludge will be calcined and smelted to produce gold doré bullion for transport.

Ore B Residue

The carbon in leach or carbon pulp process conventionally used in recovering precious metals is not conducive to recovering high silver content in ore and concentrate. Ores with high silver to gold content generally favour Merrill-Crowe recovery process. This is largely because of the very large carbon stripping and electro winning systems required for processing large quantities of silver in a conventional CIL plant(12). Examples of operations in Peru where the Merrill Crowe approach has been used are Newmont's Minera Yanacocha mine and Barrick's Pierina Mine. Ore B has a silver to gold ratio of 30:1 and consequently the Merrill Crowe process is favoured.

The gold and silver cyanide leaching circuit is designed on a 24 hour residence time. The circuit configuration has been sized to treat the oxidised concentrated as well as the tails from the flotation circuit.

Oxidised feed from the refractory pre-treatment plant is pumped and combined with the flotation tails in a conditioning tank where lime is added, if required, to bring the pH up to 10.5, and cyanide added prior to being pumped to a series of six cascading tanks.

Following leaching the pregnant cyanide liquor solution is separated and washed from the solids via a three stage CCD thickener circuit consisting of 35m diameter high rate carbon steel thickeners. The pregnant cyanide liquor overflow from the final CCD thickener is treated in the Merrill Crowe plant to recover silver and gold.

The 940 m^3/hr Modular Merrill Crowe and refinery plant consists of the following steps following cyanidation:

- Solid liquid separation using three 35 m high rate counter current decantation thickeners to remove the pregnant precious metal solution as described in the above section.
- The pregnant solution is clarified in a polish leaf filter to reduce the suspended solids to approximately 1 mg/L.
- The clarified solution is de-aerated using packed tower (Crowe) under vacuum.
- Powdered zinc is added to the de-aerated clarified solution inline or tank to precipitate precious metals from solution.
- The resultant precipitate is pumped to a filter press and washed.
- The filtered precipitated is collected and acid washed to remove the zinc.
- The acid washed precipitated product is refiltered, washed and air dried.
- The product is smelted to form silver/gold dore.

Detoxification Circuits

Following the requirements of the International Cyanide Management Code tailings will be detoxified before storage in a lined tailings facility. Weak acid dissociable (WAD) cyanide in the leach tailings will be neutralised to below 0.5 ppm CN using an Inco system. The resultant detoxified slurry residue is pumped to a tailings storage facilities.

TREATMENT OPTIONS FOR ORE A

Ultra Fine Milling – Whole Ore or Concentrate

The general concept flow sheet for whole ore and concentrate is shown in Figure 6. Concentrate or whole ore feed from the grinding circuit is further ground to a particle size of less than 10 µm to physically liberate gold using either a vertical or horizontal stirred mill consisting of rotating stirrers within a stationary mill shell to impart kinetic energy to a fine media charge (usually sand). Xstrata Technology owns the rights for the commercialisation of the IsaMill™, a horizontal stirred mill. The mill product is then treated in a conventional cyanidation CIL plant to recover the gold. The presence of very fine inclusions of gold in the order of 10 µm and the possibility of some surface oxidation of the sulphide minerals may improve in gold recovery from coarser ground feed.

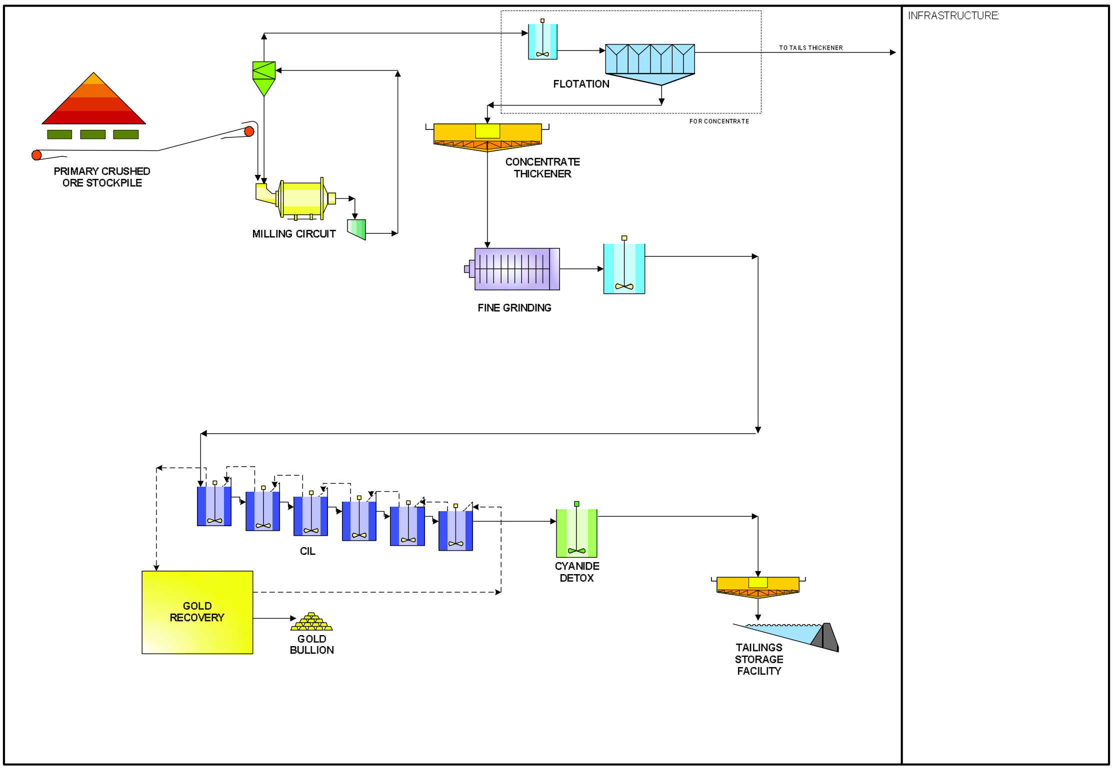


Figure 6: Ultra fine grinding circuit for whole ore and concentration (flotation circuit inclusive)

Bacterial Heap Oxidation/Leaching

Biological heap leach technology can be used to oxidise whole ore sulphides in open stockpiles before treatment in a conventional oxide gold treatment plant. Biological heap leaching is often used for base metal recovery but in that case the solutions are recovered rather than the solids. The most successful biological heap refractory leaching has been conducted by Newmont in their operations in Nevada(13). Heaps of low grade ore are treated at a rate of 3.5 Mtpa. The schematic diagram of the bacteria whole ore heap is shown in Figure 7.

In the process refractory ore is crushed to -9 mm before conveyor stacking on a heap leach type leaching pad constructed on re-usable lined leach pads. A bio-solution (containing bacterial inoculum and nutrients) is then added to the ore on the stacker conveyor and recirculated through the heap. Air is percolated through the heaps over a nominal total cycle time of 400 days, including 165 days bio oxidation. A mixture of bacteria consisting of mesophiles, moderate thermophiles and thermophiles are propagated with the heap with heap temperatures reaching up to 80 C. The bio-solution is recycled to the heap and a bleed from the bio-solution recycle is neutralised using limestone before discharge or re-use to maintain moderate pH and neutralise the acid generated during oxidation. Arsenic is precipitated as stable ferric arsenate during neutralisation. At the completion of the oxidation cycle, the ore is washed and allowed to drain before reclaiming.

The oxidised ore is reclaimed from the oxidation heaps and processed through a conventional cyanide leaching plant. The process has a low capital cost but has a relatively high technology risk. Notwithstanding this, it can be a cheap simple technology.

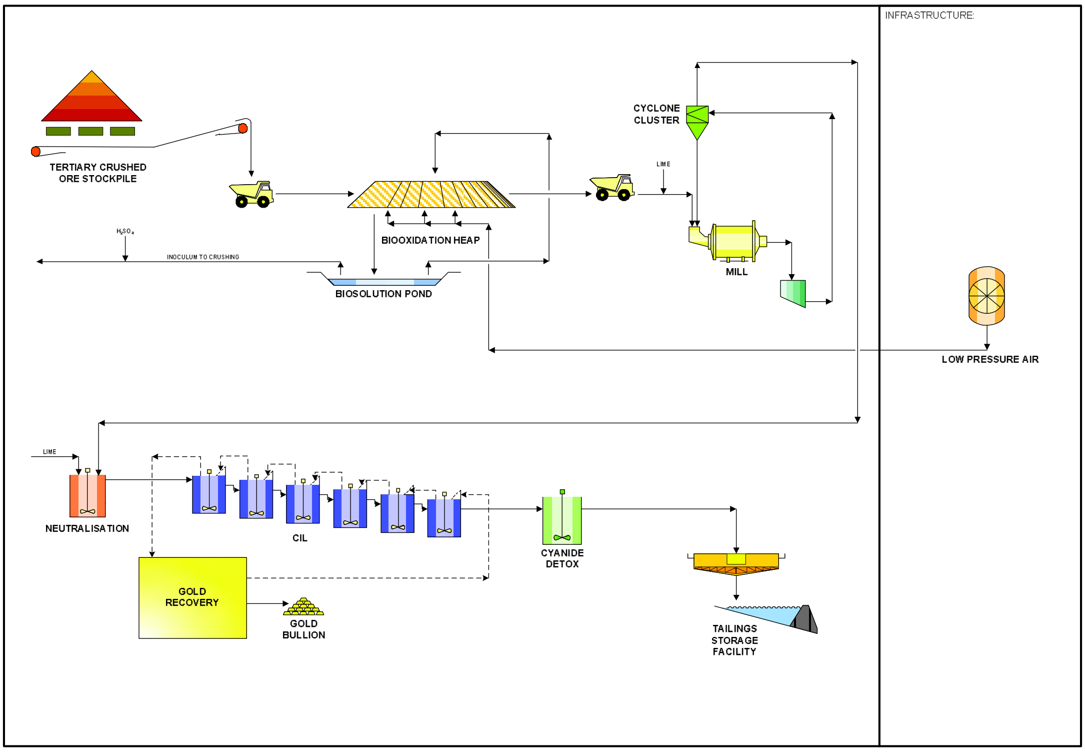


Figure 7: Biological Oxidation Heap Leaching – Whole Ore

BIOX® Agitated Tank Reactors

The Biological oxidation (BIOX®) process uses bacteria to oxidise sulphides in large agitated tanks before treatment in a conventional oxide gold treatment plant. The technology is well proven in particular for concentrates and in use in a number of similar operations(14). The process is illustrated in Figure 8.

The concentrate with a P_{80} of 75 µm is fed to the biological oxidation circuit comprising a series of primary bioleach reactors in parallel. Partially oxidised product exiting the primary reactors are recombined and enter a series of secondary reactors. The configuration (normally three primary reactors in parallel and three secondary in series) allows for at least double the residence time in the primary reactors to promote growth and division of the bacteria and prevent washout. The temperature is controlled to between 40-45 C at ambient pressure. Cooling and aeration are significant contributors to cost and power demand.

The pulp pH is maintained at less than 2 with a total residence time of 3 to 5 days. A CCD circuit separates solids and liquids. Iron and arsenic and any base metals report to solution and are neutralised with limestone followed by lime. Arsenic is precipitated as the stable ferric arsenate phase. The residue is washed and neutralised for cyanidation. Gold and silver are leached from the residue, recovered on carbon and electrowon in the conventional way.

Biological oxidative leaching is easily used to control oxidation levels of the ore for optimum recovery. BIOX® technology requires significant capital although the technology is relatively simple and well understood. This technology has the registered trade mark BIOX® and is now owned by Gold Fields of South Africa. Similar technology developed by Mintek of South Africa and BacTech Mining Company has been installed commercially.

Biooxidation testwork has been conducted on Ore A. Gold recovery from the oxidised residue was 92% at an 80% oxidation level. It is probable that there was some preferential oxidation of the arsenopyrite / arsenian pyrite to achieve the recovery at the lower oxidation level. Tank biooxidation is particularly suited to partial oxidation of the sulphides. Hence some costs associated with power and reagent consumption can be reduced compared with full sulphide oxidation processing.

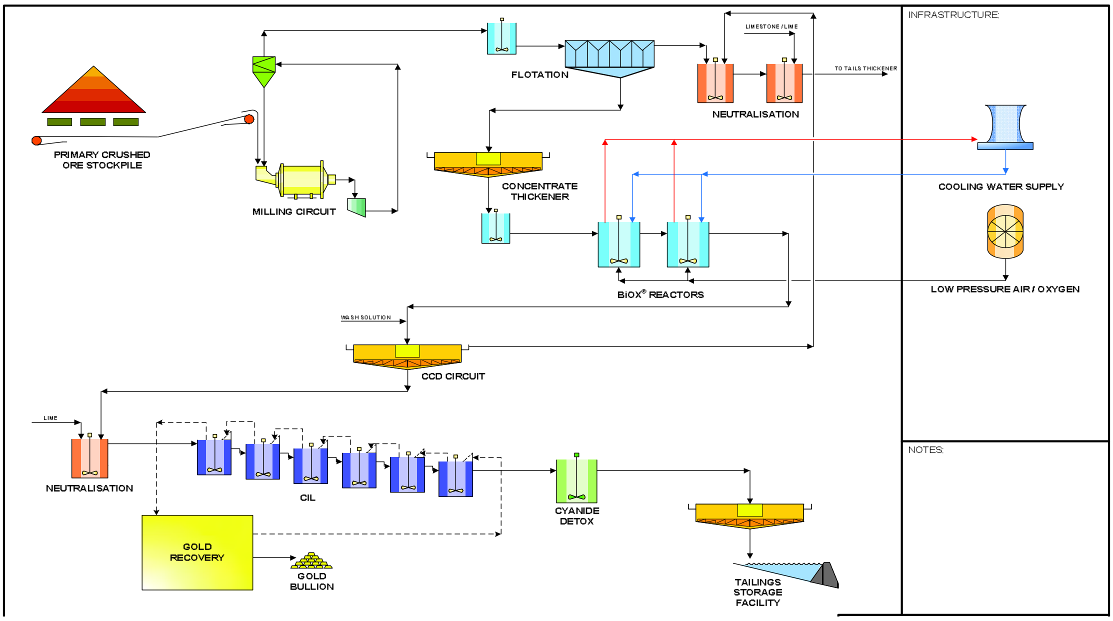


Figure 8: Agitated Tank Bioleaching – Concentrate

Albion Process

The Albion process uses ulfrafine grinding in stirred mills to produce a highly activated, finely ground concentrate at low energy inputs. The finely ground concentrate is then leached with oxygen at atmospheric pressure in conventional agitated tanks. The flow sheet of the process is illustrated in Figure 5.

The Albion Process was developed by a MIM subsidiary, Hydrometallurgical Research Laboratories in the mid 1990s, and a US patent was obtained in late 1999. Xstrata acquired control of MIM in June 2003. Numerous piloting campaigns at the Hydromet Research Laboratories have been run since 1995, on copper-gold and gold applications at up to 50 kg/day. In recent years development to full commercialisation is proceeding. The Certej Project owned by European Goldfields is under development in Romania. This project uses an Albion Leach to leach gold from a flotation concentrate produced from 3 Mtpa ore. The process has been subjected to bench scale and pilot testing at Xstrata's Hydrometallurgical Research Laboratories in Australia. A total of 57 operations have tested the Albion process at HRL and 11 have proceeded to full pilot plant trials.

The sulphide concentrate produced from flotation will be fed to a fine grinding M1000 Isamill to yield a product P_{80} of 10 µm. The ground ore is then transferred to three Albion oxidation reactors where low pressure oxygen is sparged into the slurry. In the process the sulphide matrix containing the gold is broken down to release ferric ions and sulphuric acid. The temperature is not controlled and rises to close to 100^{o}C. The breakdown of the sulphides liberates the gold. The ferric ion and acid are neutralised in situ by the continual addition of limestone slurry. The limestone is dosed to maintain the leach pH within the range of 5 to 5.5. Arsenic is precipitated as the stable ferric arsenate phase. The neutralised iron oxides (mainly goethite) and gypsum formed in the process is then processed through the CIL circuit.

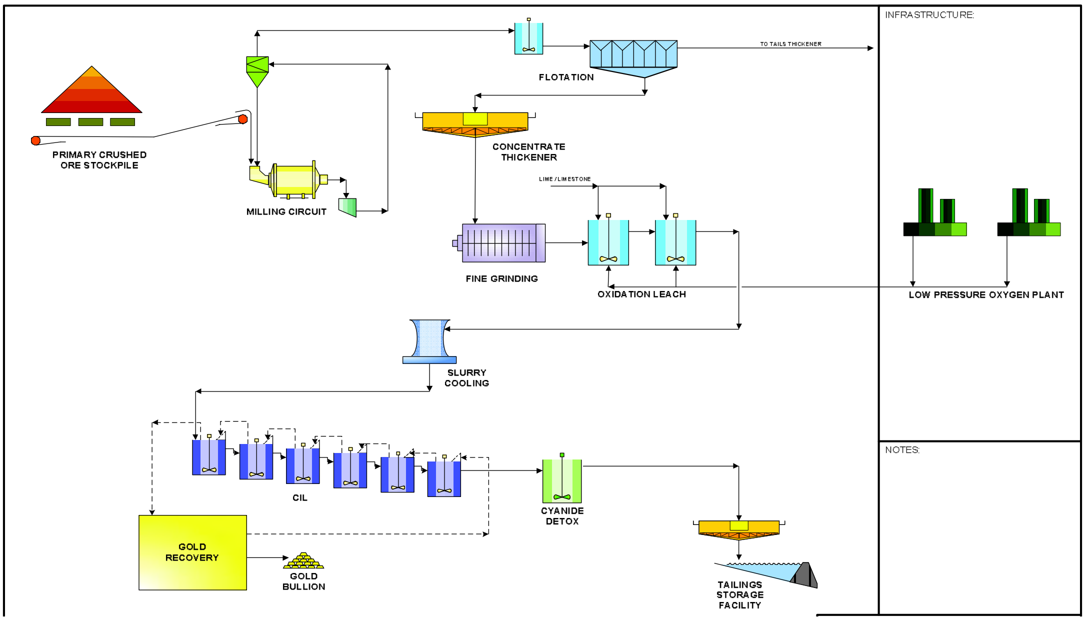


Figure 9: Albion Process – Concentrate

Pressure Oxidation

Pressure Oxidation is a high intensity process to oxidise sulphides in autoclaves before treatment in a conventional oxide gold treatment plant. Over the last 20 years around 15 pressure oxidation plants were commercialised specifically for the gold industry. Of these, eight were designed for treating concentrate. Original processes followed the designs developed in the Sherritt Gordon patents. Treatment approaches for overcoming operational issues have been patented largely by Newmont (15).

The plant layout is illustrated in Figure 10. Flotation concentrate is fed to the autoclave feed tank and then into the autoclave. The autoclave circuit includes a high pressure slurry feed system, autoclave vessel and agitators, flash vessel, and gas handling system. The autoclave will operate at 200ºC and 3,100 kPa and a retention time of approximately 30 – 60 min. Close to full oxidation of the sulphide is expected as partial oxidation is not likely to be viable. The process is expected to be autothermal operation not require any pre heating of slurry to achieve operating temperature. The sulphide minerals are oxidised to haematite with the arsenic precipitated as stable ferric arsenate.

The quenching of the excess flash steam and autoclave vent gas is accomplished with CCD overflow solution. The acidic CCD overflow solution is neutralised with the tails from the flotation circuit which contain high carbonate content. By using the tails to neutralise the effluent will enable a reduction in limestone consumption and costs.

Slurry from the let down tank is treated in a three-stage CCD circuit. Each thickener will be constructed from 316 L stainless steel walls, floor and rakes. The nominal wash ratio in the CCD circuit is maintained with wash efficiency of 99.0%.

The washed oxidation product will be pumped to the CIL tanks for conventional gold recovery. No test work on the performance of pressure oxidation on the concentrate has been carried out. However, from experience it is expected that gold recovery may be slightly better (~2%) than bio-oxidation.

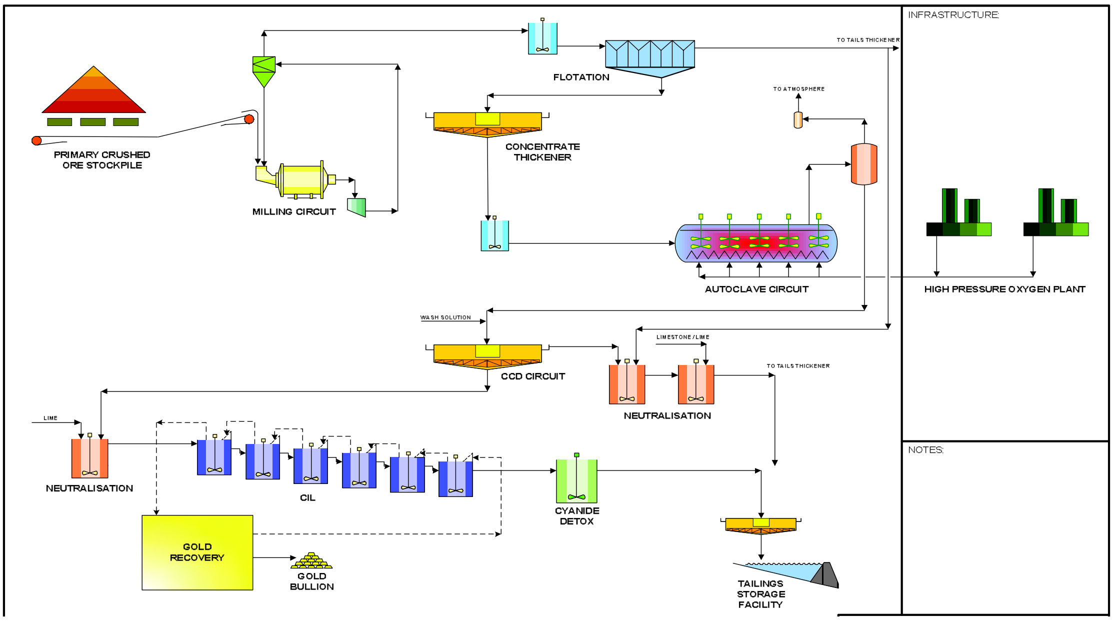


Figure 10: High Temperature Pressure Oxidation

TREATMENT OPTIONS FOR ORE B

BIOX® Agitated Tank Reactors – Concentrate

The process is illustrated in Figure 11. The concentrate with a P_{80} of 75 µm is fed to the biological oxidation circuit comprising a series of primary bioleach reactors in parallel. Partially oxidised product exiting the primary reactors are recombined and enter a series of secondary reactors. The configuration (5 modules each containing four primary reactors in parallel and four secondary in series) allows for at least double the residence time in the primary reactors to promote growth and division of the bacteria and prevent washout. The temperature is controlled to between 40 and 45 ºC at ambient pressure. Cooling and aeration are significant contributors to cost and power demand.

The pulp pH is maintained at less than 2 with a total residence time of 3 to 5 days. A CCD circuit separates solids and liquids. Iron and arsenic and any base metals report to solution and are neutralised with limestone followed by lime. Arsenic is precipitated as the stable ferric arsenate phase for disposal as tailings.

The washed oxidation product is pumped to the precious metals recovery plant. The oxidised residue is treated in the Lime boil circuit where silver containing jarosite formed during the oxidation process is decomposed to liberate the silver. The lime boil circuit consists of four cascading agitated tanks with the temperature maintained at 90˚C using steam generated a boiler or refractory process. Lime is added to obtain a pH of 10.5 (at ~25˚C). To conserve heat in the process a heat exchanger transfer heat from the outgoing slurry to preheat fresh oxidation product feeding into the lime boil circuit.

To recover silver and gold, the neutralised oxidised residue, combined with the flotation tails, is cyanide leached to extract the precious metals. The pregnant liquor is separated from the solids via counter current decantation to recover the precious metals by a Merrill Crowe plant.

No biooxidation testwork has been conducted on ore B concentrate. Tank biooxidation is particularly suited to partial oxidation of the sulphides. Hence some costs associated with power and reagent consumption can be reduced compared with full sulphide oxidation processing, if test work was to show the precious metals could be recovered from a partial sulphide oxidation product.

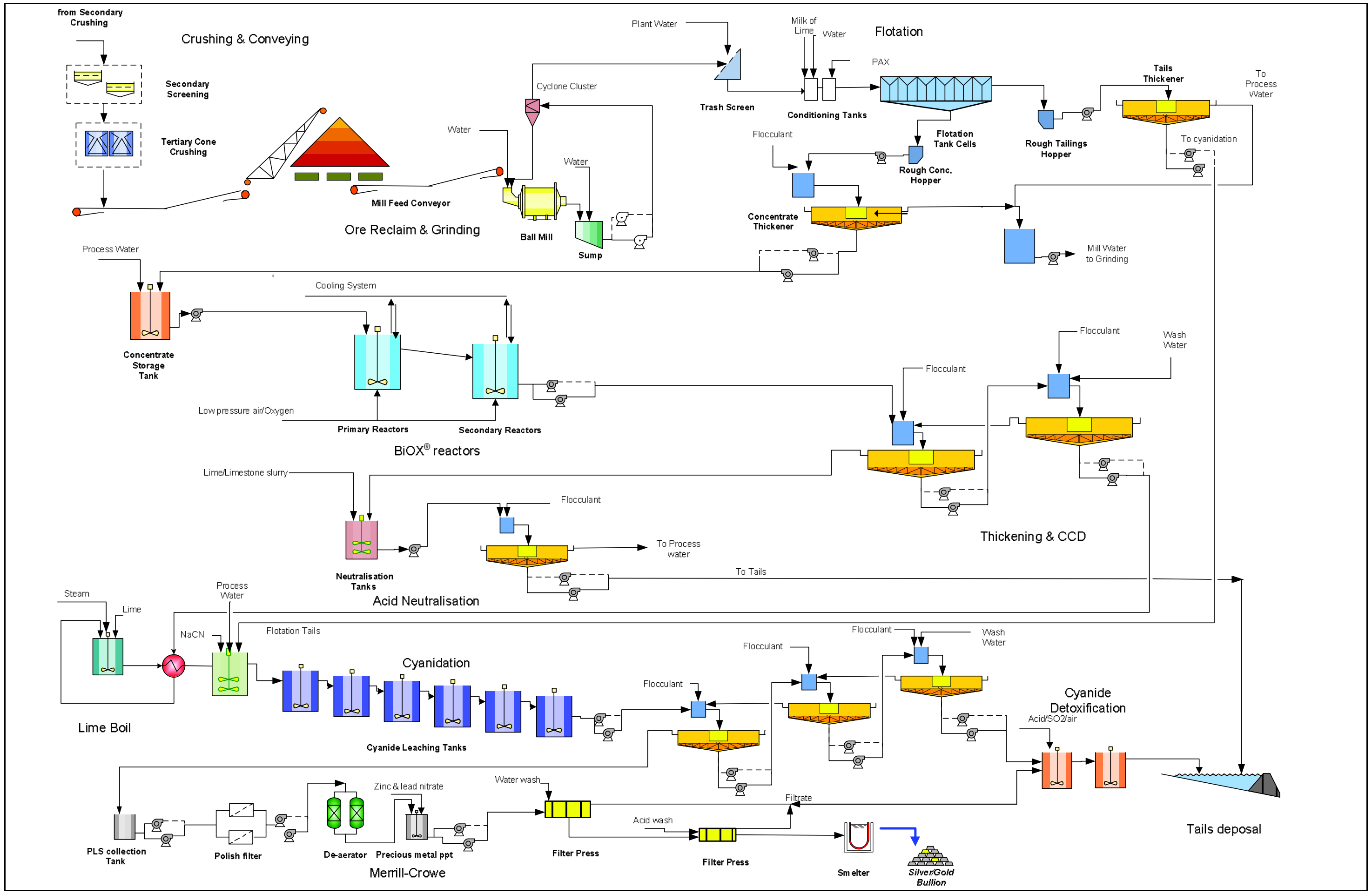


Figure 11: Agitated Tank Bioleaching – Concentrate

Pressure Oxidation – Concentrate

The Pressure Oxidation plant layout is illustrated in Figure 12. Flotation concentrate is fed to the autoclave feed tank, diluted and then fed into one of two autoclaves. The autoclave circuit includes a high pressure slurry feed system, autoclave vessels and agitators, flash vessels, and gas handling system. The autoclave will operate at 210-220ºC and 3,100kPa-3,300KPa and a retention time of approximately 30 – 60 min. Full oxidation of the sulphide is expected as partial oxidation is not likely to be viable. The process is expected to be autothermal and not requiring any pre heating of slurry to achieve operating temperature during operation. The sulphide minerals are oxidised to haematite with the arsenic precipitated as stable ferric arsenate. Slurry from the let-down tank is treated through a hot cure stage consisting of four tanks in series. This process is required to convert all basic ferric sulphate formed during the pressure oxidation process to ferric sulphate so as to reduce lime consumption during neutralisation.

The quenching of the excess flash steam and autoclave vent gas is accomplished with CCD overflow solution. Slurry from the hot cure stage is treated in a three-stage CCD circuit. Each thickener will be constructed from 316 L stainless steel walls, floor and rakes. The nominal wash ratio in the CCD circuit is maintained with wash efficiency of 99.0%. The overflow wash liquor from the CCD circuit is collected and neutralised in a series of eight tanks. The neutralised water is stored and reused in the process plant.

The washed oxidation product is pumped to the lime boil circuit where silver containing jarosite formed during the pressure oxidation process is decomposed to liberate the silver. Similar to the Biox plant the lime boil circuit consists of four cascading agitated tanks with the temperature maintained at 90˚C using steam generated by a boiler. Lime is added to obtain a pH of 10.5 (at ~25˚C). To conserve heat in the process a heat exchanger transfers heat from the outgoing slurry to preheat fresh oxidation product feeding into the lime boil circuit. The lime boiled slurry is pumped and combined with the flotation tails in a conditioning tank for cyanidation to extract gold and silver. Following liquid-solid separation step silver and gold are recovered from the pregnant cyanide liquor solution via the Merrill Crowe plant.

No test work on the performance of pressure oxidation on the concentrate has been carried out.

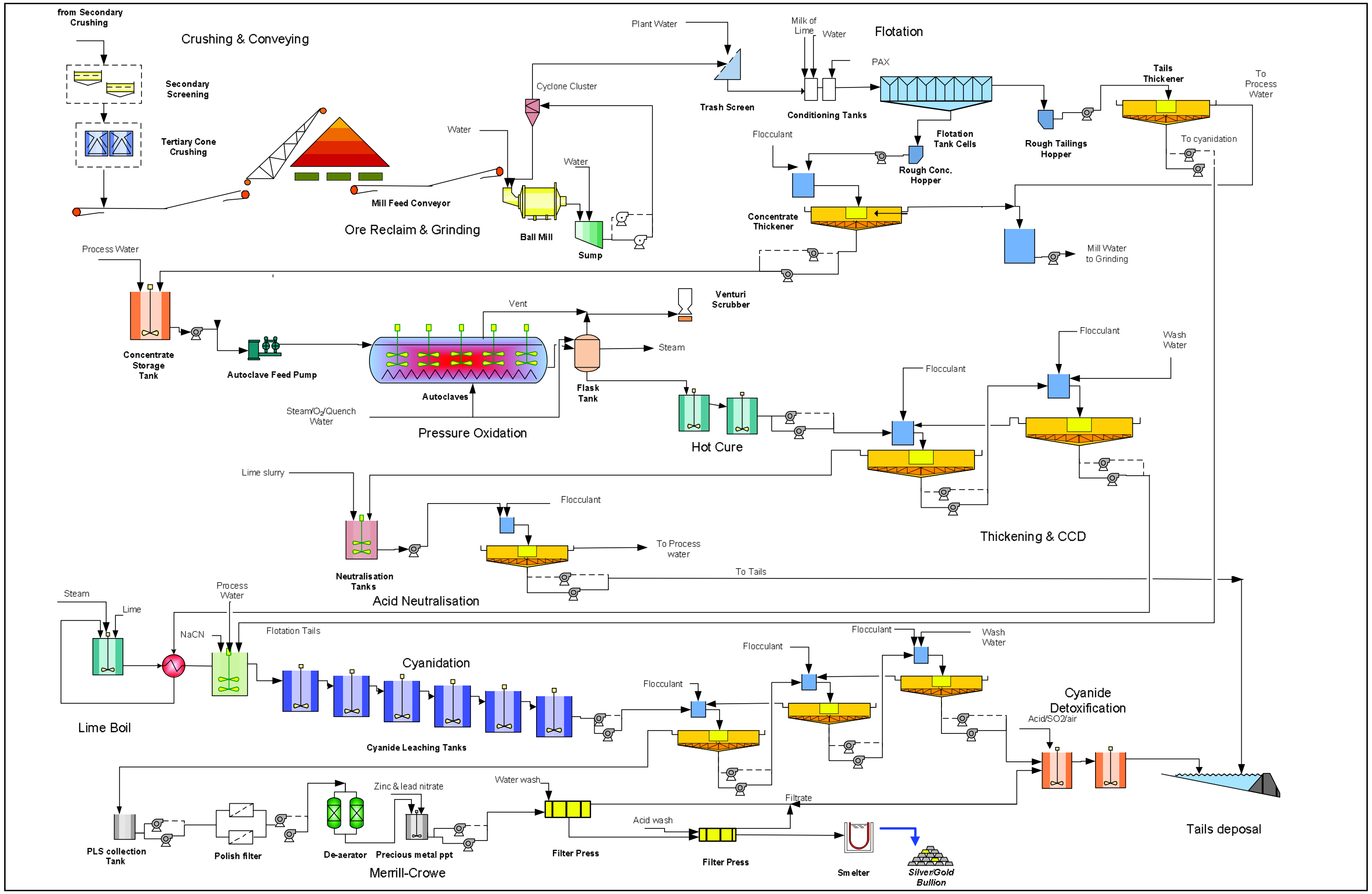


Figure 12: High Temperature Pressure Oxidation – Concentrate

Air Roasting – Concentrate

Roasting in an oxidising atmosphere, to produce a cyanide leach amenable porous calcine, has traditionally been the most commonly applied oxidative pre-treatment process for refractory gold ores. The concentrate from the flotation circuit will be thickened and filtered to produce a feed to the roasters. Oxidation by roasting uses air or oxygen enriched combustion to oxidise sulphides before quenching and treatment in a conventional oxide gold treatment plant. Three roasting technologies are available; circulating fluid bed (CFB) roasting under air and oxygen enriched atmosphere and conventional fluidised (or bubbling bed) technology. A single stage roasting step should be sufficient to fix arsenic at the levels encountered in the ore. This is subject to confirmation by testwork.

The concentrate is fed to a rotary drier, where it contacts hot roaster flue gas to reduce the moisture content to 5%. The dried concentrate is then distributed by the roaster rotary valve to the fluidised roasting bed chambers where the sulphide minerals react with the fluidising air. The bed temperature will be maintained at around 600 to 650ºC, to achieve 95% oxidation of total sulphide. For the treatment of concentrate the process design is based on two roasters supplied by Sinopec Nanjing Design Institute (SDNI) with one acid plant supplied by Noram.

The off-gases will be passed through a waste heat boiler at approximately 10% SO_2, and 6-7% O_2, before passing through the electrostatic precipitator (ESP). The ESP will reduce particulates in the gas to ~0.2g/Nm^3 before entering the gas purification section of the sulphuric acid plant. Arsenic trioxide will be captured in the ESP and returned to the calcine.

The gas is diluted with air to give the optimum SO_2 concentration for conversion in the double conversion and double absorption process. Nominally 99.75% of the SO_2 is removed and converted to acid, with a tail gas exhausted to atmosphere at an SO_2 concentration of 250 ppm. The acid is adsorbed in a circulating acid solution and 98-98.5% sulphuric acid is cooled in the product acid cooler before going to the acid storage tank for storage and export.

A water-tube boiler will produce saturated steam for use in electricity generation. It is expected that a nominal amount of electricity will be generated in the roaster/acid plant for potential use in the process. Excess LP saturated steam will be available for heating duties. For the purposes of this study, the benefit arising from potential power generation has not been included as a credit in the overall cost of the process.

The quenched roaster calcine will be slurried and pumped to the cyanide leach circuit and gold and silver recovered from the pregnant liquor by the Merrill Crowe plant.

In the absence of tests, the Au and Ag recoveries across the roasting process have been chosen from a range that would be typically expected for pyritic feedstocks. While roasting can produce a relatively inert calcine leading to lower cyanide consumption by complete oxidation of sulphur, the control of emissions of As and SO_2 is difficult and adds to the capital cost and complexity of the operation. Fixing the arsenic in calcine product is possible but would reduce gold recovery and increase reagent consumption in the subsequent cyanidation stage. Furthermore issues such as sintering and lock-up in hematite may also lower recoveries. Test work would be required to determine viability of an approach.

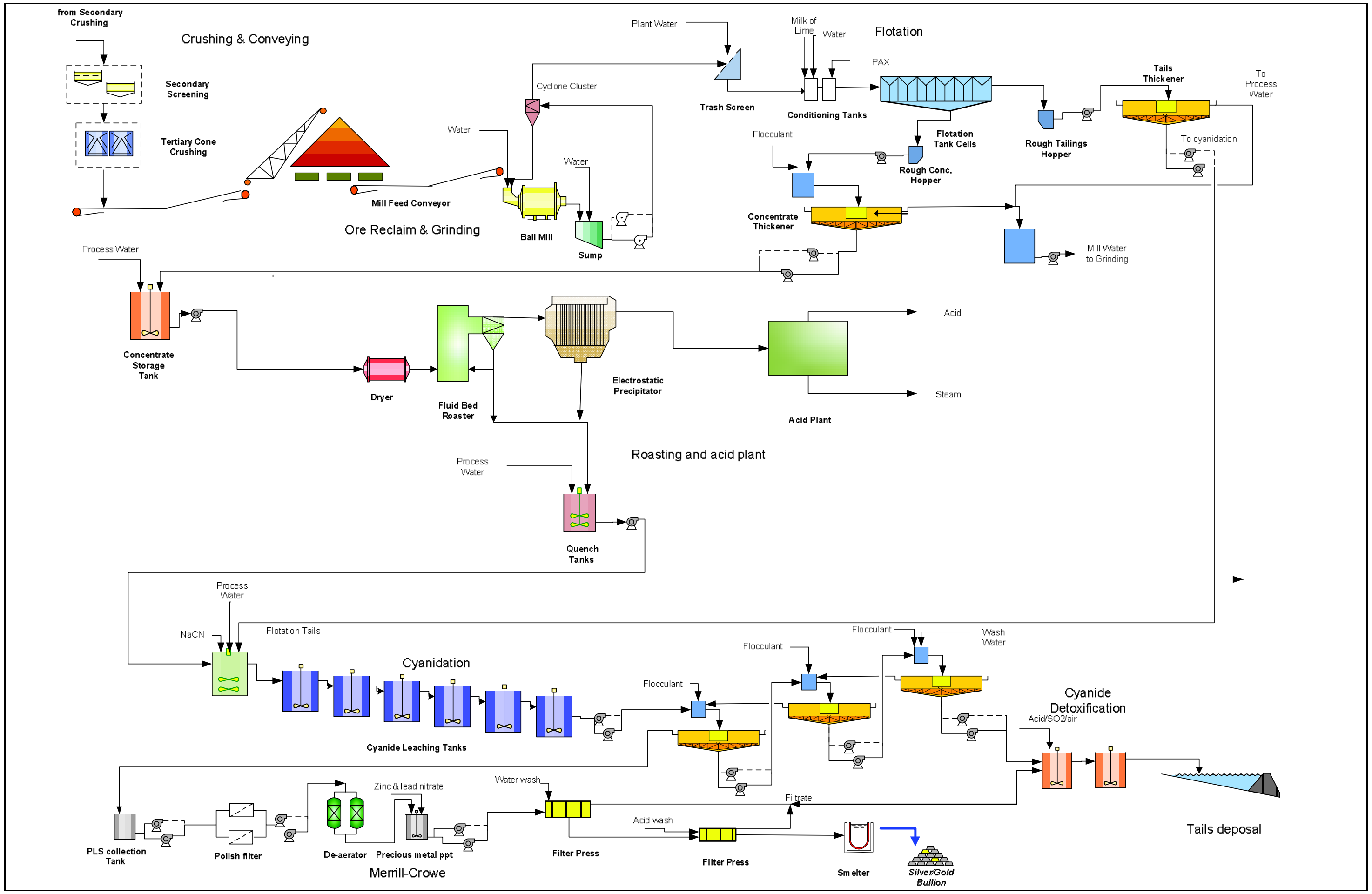


Figure 13: Roaster – Concentrate

CAPITAL AND OPERATING COST ESTIMATES FOR EACH PROCESS OPTION

Basis for Estimate

Specific factors that were considered in the economics of the process include the following:

- Concentrate grade
- Mineralogical composition
- Process route adopted
- Power costs
- Labour rates
- Transport rates
- Accessibility to existing infrastructure and land

These are broken down and taken into account in the capital and operating cost estimates, with similar methodology taken for both Ore A and B.

Capital Cost Estimate

The capital cost estimates have been defined as a "Scoping/conceptual Level Study". This corresponds to a Class 4 estimate in the Bateman guidelines and is deemed to have accuracy in the range of +/- 30%.

Vendor estimates were obtained for proprietary technologies. The bio-oxidation tank processes were sourced through Gold Fields Ltd and the Albion process through Xstrata. Capital costs for crushing, grinding, pressure autoclave, CCD, CIL, gold refinery, thickeners and tails detoxification plants were based on recently completed works of similar operational throughputs. Price escalation was factored in where applicable to allow for current costs. The Merrill Crowe plant was sourced from FLSmidth Salt Lake City.

Indirect costs were calculated based on erection of Temporary Facilities, First Fills and Lubricants Commissioning and Start-Up Spares, Engineering, Procurement and Construction Management and Contingency.

No allowance has been made for Infrastructure, Environmental studies, investigations, permits or liabilities, Costs of permits. Legal fees, Land acquisition or rights of way and any resettlement Costs.

Operating Cost Estimate

The operating cost estimates have been prepared to an accuracy of ± 40%. The operating labour costs were developed from the schedule of required labour prepared by Bateman, with costs for each labour resource based on typical wages expected for the regions.

The consumption rates of reagents and consumables were based on assumptions from test work data and predicted calculated values from mass balance models. Pricings including power were supplied by the clients. Reagent and power costs are listed in Table 6.

Table 6: Local Reagent and Power Costs Assumed

Local Reagents	**Ore A**	**Ore B**	**Unit**
Lime	130	121	$/t
Limestone	25	75	$/t
Sulphuric acid	390	125	$/t
Sodium cyanide	2660	2250	$/t
Power			
Power Cost	0.08	0.10	$/kWh

Power consumption was calculated based on unit processes. Maintenance costs for each plant area were estimated as a percentage of the total direct capital cost of mechanical equipment for that area. A contingency of 10% was included to cover unforeseen operating costs and is meant to cover the normal level of unknowns that are inherent in early design.

Comparative Cost Estimates for Ore A

A summary of plant design data, operating and capital costs for all six process treatment options are presented in Table 7. The capital and operating cost estimates for each of the process options is shown graphically in the following figures.

Table 7: Summary Comparison of selected Refractory Gold processes for Ore A

Summary Comparison of Selected Refractory Gold Processes for Ore A

OPTION		**1**	**2**	**3**	**4**	**5**	**6**
PROCESS		**Bacterial Heap Oxidation Whole Ore**	**UFM Con Concentrate**	**UFM Ore Whole Ore**	**Albion Concentrate**	**Bacterial Tank Leaching (BIOX) Concentrate**	**POX Concentrate**
Plant Design Data							
Ore Processed	t / yr	600,000	600,000	600,000	600,000	600,000	600,000
Concentrate Production	t / yr	0	54,000	0	54,000	54,000	54,000
Au in Feed	oz / yr	43,403	39,062	43,403	39,062	39,062	43,403
S in Feed	t / yr	7,401	6,942	7,401	6,942	6,942	6,942
CO3 in Process Feed	t / yr	71,946	1,080	71,946	1,080	1,080	1,080
Final Grind Size (P_{80})	µm	150	10	10	10	75	75
Sulphur Oxidation	%	60	2	2	80	80	95
Overall Recovery of Au	%	70	45	50	83	83	85
Acid produced	t/yr	13,599				1,057	1,057
Net Acid required by process	t/yr	56,859			1,200		
Power	kWh/t ore	43.9	31.6	67.5	41.7	74.8	96.6
Annual Operating Cost							
Labour	US $ million / annum	2.6	2.9	2.2	2.5	2.5	2.7
Reagents	US $ million / annum	28.2	3.9	5.4	3.8	3.7	4.5
Power	US $ million / annum	2.1	1.5	3.2	2.0	3.6	4.6
Sundry Other	US $ million / annum	3.2	1.5	1.7	2.0	2.0	3.0
Total	***US $ million / annum***	***36.1***	***9.8***	***12.5***	***10.3***	***11.9***	***14.9***
Contingency	US $ million / annum	3.6	1.0	1.2	1.0	1.2	1.5
Total Operating Cost	***US $ million / annum***	**39.7**	**10.8**	**13.7**	**11.3**	**13.0**	**16.4**
Capital Cost							
Direct Capital Cost - Front End and Gold Recovery	US $ million	20.32	28.45	29.38	31.67	27.00	32.17
Direct Capital Cost - Refractory	US $ million	15.42	3.50	7.80	5.00	12.67	14.91
Total Direct Capital Cost	**US $ million**	**35.74**	**31.95**	**37.18**	**36.67**	**39.67**	**47.07**
Total Indirect Capital Cost	**US $ million**	**15.55**	**13.90**	**16.17**	**15.95**	**17.26**	**20.48**
Total	**US $ million**	**51.3**	**45.9**	**53.3**	**52.6**	**56.9**	**67.6**
Cash Flow							
Gold Produced	oz / annum	30,382	17,578	21,701	32,344	32,344	36,719
Operating Cost Relative to Ounces of Au	**US $ / oz**	1,308	612	633	350	403	447

The whole ore Bacteria heap leach and ultrafine milling ore processes will treat 600,000 tpa feed whereas the other processes will treat 54,000 tpa flotation concentrate. The mass pull from the flotation circuit will be 9% by weight recovering 90% of the gold in the ore.

The final particle size of the bacteria heap will be coarser than the other processes for cyanidation. The ultrafine milling processes, including Albion, will be ground to 10 micrometre, while for bacteria tank and pressure oxidation a grind to 75 micrometres is considered.

For heap leaching only 60% of sulphide minerals is assumed to be oxidised due to percolation and surface passivation issues through the heap. While the ultrafine milling process is considered a physical reduction in particle size process some allowance has been made for surface oxidation. Both the Albion and bacteria tank leach processes sulphide oxidation can be controlled to 80% to allow for maximum gold recovery without excess oxidation. In pressure oxidation partial sulphide oxidation is not easy to control and 95% of the sulphide has been considered in this study.

Gold recoveries are estimated based on the limited test work on bio-oxidation work. No gold recovery in tails from the concentrate treatment processes have been considered in this study. The tails are expected to be too low grade and refractory to be treated economically. Hence overall gold recovery is based on the 90% recovered in the flotation circuit for processes treating concentrate. The gold recovered from pressure oxidation is expected to be 2% higher than bio-oxidation tank process.

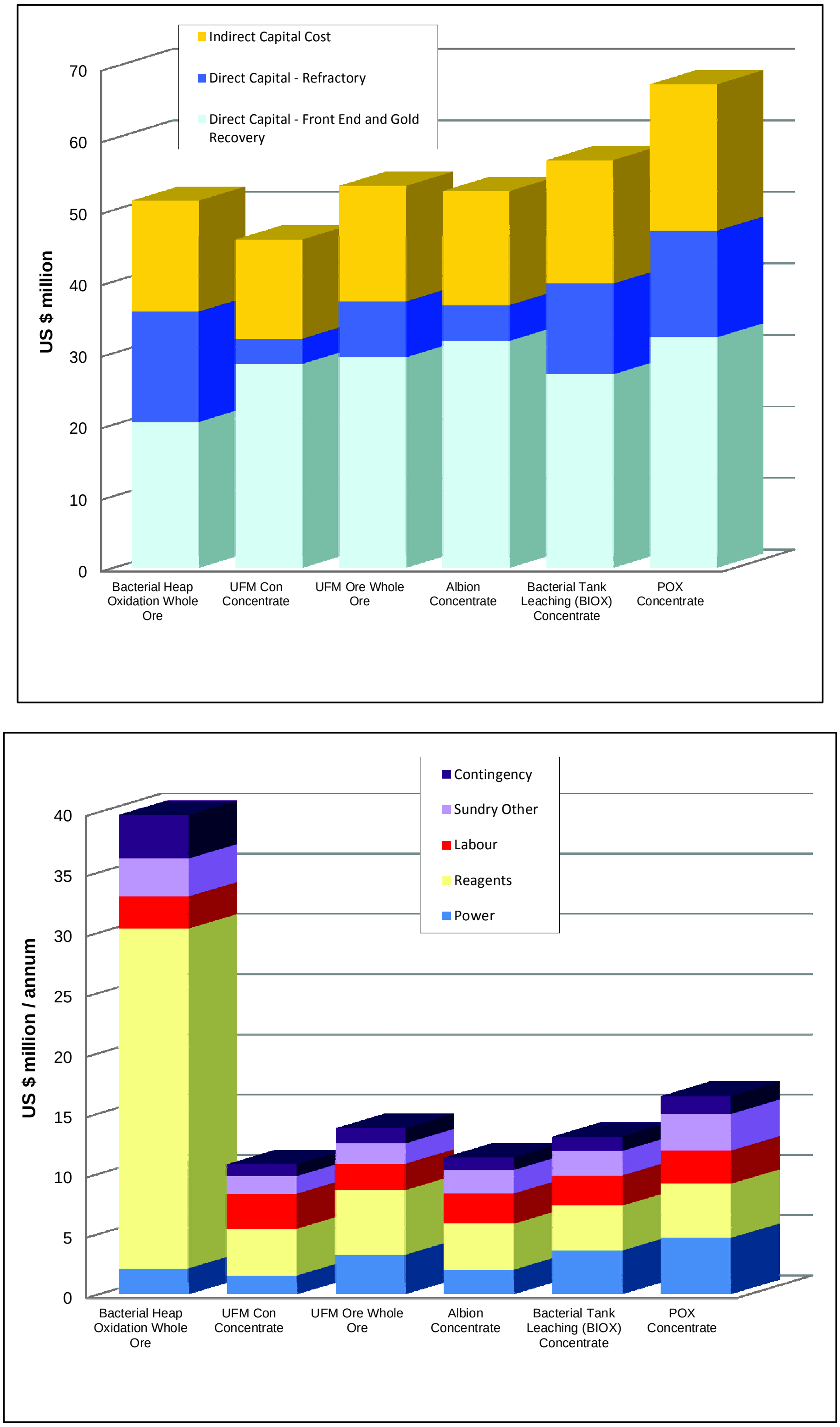


Figure 14: Capital (a) and Operating (b) costs shown graphically for Ore A

For Ore A the high operating costs for the bacteria heap leach is associated with high acid demand for neutralisation of the carbonate in the ore. Early test work data alluded to the requirement of either acid or excess pyrite being added to the heap to neutralise the carbonate component of the ore. Utilisation of the flotation tails (containing carbonate) as a neutralising material reduces the costs of limestone in the concentrate treatment processes. The bacteria tank leaching and pressure oxidation are slightly more energy intensive than the other processes.

A comparison of the operating cost per ounce of gold recovered versus process capital costs is presented in Figure 15. For the ultrafine milling processes to be cost effective in line with the other concentrate process options a gold recovery of 81.2% for concentrate or 78.4% for whole ore is required. Test work however will be required to validate actual gold recoveries.

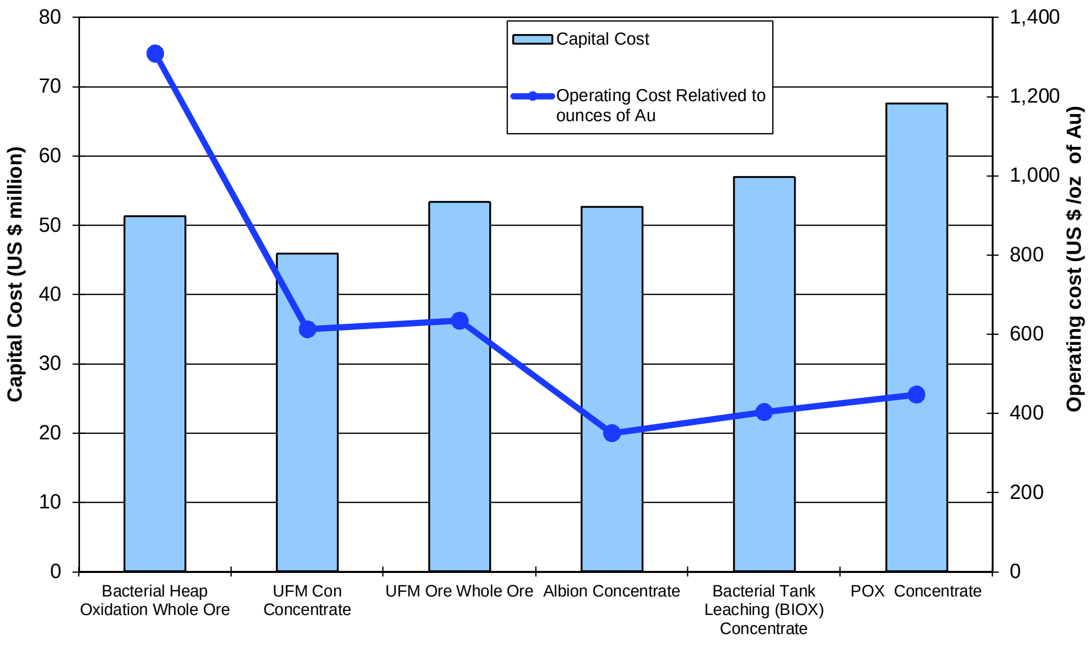


Figure 15: Comparison of conceptual Operating cost to ounces of gold produced versus capital costs for the six process options

Comparative Cost Estimates for Ore B

A summary of plant design data, operating and capital costs for all six process treatment options are presented in Table 8. The capital and operating cost estimates for each of the process options is shown graphically in the following figures.

Table 8: Summary Comparison of selected Refractory Gold processes for Ore B

Summary Comparison of Refractory Gold Processes for Ore B				
OPTION		**1**	**2**	**3**
PROCESS		**Air Roasting Concentrate**	**Bacterial Tank Leaching Concentrate**	**POX Concentrate**
Ore Processed	t / yr	3,650,000	3,650,000	3,650,000
Au in ore	oz / yr	124,388	124,388	124,388
Ag in ore	oz / yr	3,755,120	3,755,120	3,755,120
Concentrate Production	t / yr	1,095,000	1,095,000	1,095,000
Au in Conc	oz / yr	87,072	87,072	87,072
Ag in Conc	oz / yr	1,689,804	1,689,804	1,689,804
S in Conc	t / yr	202,985	202,985	202,985
Final Grind Size (P_{80})	µm	75	75	75
Sulphur Oxidation	%	95	95	100
Overall Recovery of Au	%	81.5	81.5	81.5
Overall Recovery of Ag	%	62.0	62.0	62.0
Acid produced	t/yr	590,558	590,558	621,640
Power	kWh/t ore	56.4	115.1	84.7
Annual Operating Cost				
Labour	US $ million / annum	4.5	4.0	4.1
Reagents	US $ million / annum	12.5	79.2	75.9
Power	US $ million / annum	19.9	42.0	30.9
Sundry Other	US $ million / annum	36.3	33.6	38.2
Total	***US $ million / annum***	***73.1***	***158.8***	***149.2***
Contingency	US $ million / annum	7.6	15.9	14.9
Total Operating Cost	***US $ million / annum***	**80.8**	**174.7**	**164.1**
Capital Cost				
Direct Capital Cost - Front & End	US $ million	115.7	119.4	135.8
Direct Capital Cost - Refractory	US $ million	124.4	136.0	136.6
Direct Capital Cost	**US $ million**	**240.0**	**255.4**	**272.4**
Indirect Capital Cost	**US $ million**	**108.3**	**115.4**	**122.8**
Total	**US $ million**	**348.3**	**370.8**	**395.2**
Revenue				
Gold Produced	oz / annum	101,377	101,377	101,377
Silver Produced	oz / annum	2,326,297	2,326,297	2,326,297
Operating Cost Relatived to Ounces of Precious Metals Produced [#]				
	US $ /oz of Au equivalent (with Ag)	**580**	**1,205**	**1,132**
#Silver value converted to Au equivalent based on ratio of current price of Au and Ag				

All three process options considered will treat 1,095,000 tpa of flotation concentrate. Based on the limited test work data the mass pull from the flotation circuit will be 30% by weight recovering 70% of the gold and 45% of the silver into a flotation concentrate. The oxidised residues from the refractory circuit will then be combined with the flotation tails for cyanidation.

It is assumed that gold and silver extraction on oxidised concentrates will be 95% and 90% respectively for all three process options. Gold and silver recoveries from the flotation tails will be 50% and 39% respectively as observed in the test work. This yields overall gold and silver recovery of 81.5% and 62.0% respectively. Sulphide oxidation is assumed to be 95% except for pressure oxidation which is set at 100%.

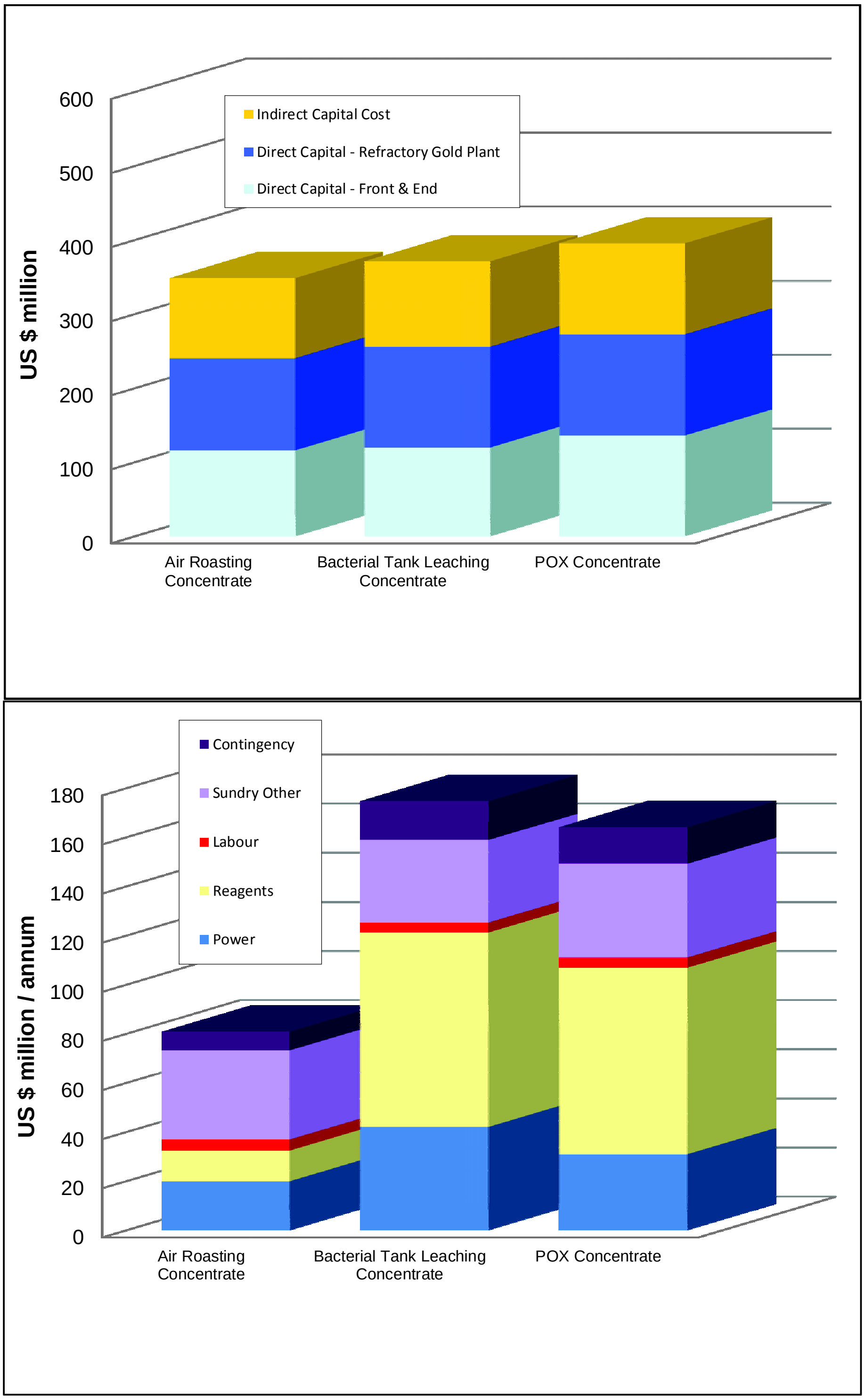
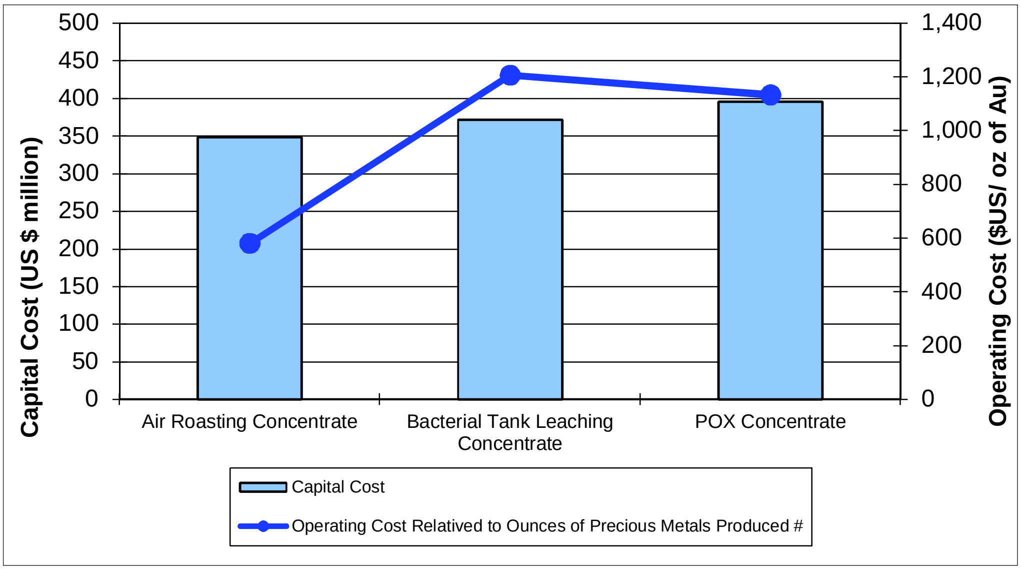


Figure 16: Capital (a) and Operating (b) Cost Estimate for Ore B

For ore B the operating costs for both bacterial tank leaching and pressure oxidation is approximately double the roasting costs. The bacterial tank leaching and pressure oxidation are more energy intensive than roasting and incur high reagent costs associated with limestone use for neutralisation.

To allow comparison between the three refractory process costs the total operating costs were expressed in terms of operating cost per ounce of equivalent gold produced taking into account silver. The ounces of silver produced were converted to ounces of equivalent gold by ratio of the average price of gold at $US 1600/oz and silver at $US 30/oz. A comparison of operating costs per ounce of gold versus process capital costs is presented in Figure 17. The operating costs are significantly high and higher grade ores would appear to be required to make this ore body profitable and sustainable. The capital costs of all three process options are not significantly different with roasting having the lowest cost. The capital cost comparisons are consistent with early studies(6).



.# Silver value converted to Au equivalent based on ratio of current price of Au and Ag

Figure 17: Comparison of conceptual Operating cost to ounces of equivalent gold produced (Au + Ag) versus capital costs for the three process options

Roasting is predicted to provide the lowest operating cost per ounce of gold followed by pressure oxidation. In the roasting option if the acid produced is neutralised this increases the overall operating cost from $US80.8M to $US 128.5M. The resultant loss of revenue from the acid selling increases the operating cost per ounce of gold to $909M, which is still predicted to be marginally better than pressure oxidation at $US1132M. However the sulphide sulphur in the concentrate (19%) is considered on the borderline to run a roaster autogenously without some extra heat source. Hence, sulphide grades need to be optimised during flotation to increase sulphide grade. Alternatively to operate the roaster at lower sulphide content the feed will have to be further dried to remove moisture and would incur extra capital and operating costs. Biooxidation option provides the highest expenditure and this is largely due to the high power costs and limestone usage.

Test work is required to determine what actual precious metal recoveries are expected for the different process options, and the degree of sulphide oxidation required to liberate and extract the gold and silver metals from the concentrate.

For the roasting option it is assumed that the arsenic level in the feed is low based on the sample on which flotation tests were conducted and will be captured in the gas and subsequently fixed with the calcine product. Arsenic can be fixed as ferric arsenate in the calcine during roasting but generally results in lower recovery of gold (and silver) during the subsequent cyanidation step. To treat high arsenic content concentrates may require a two stage roasting process with an approximate 25% increase in capital cost. The roasting option assumes the acid can be sold at $US 125/t.

In bacterial tank leach processes sulphide oxidation can be controlled to allow for maximum gold and silver recovery without excess oxidation. Reducing the oxidation requirement would significantly improve the operating costs. In roasting and pressure oxidation partial sulphide oxidation is not easy to control.

Extend of silver being liberated is sensitive to the amount of lime added in the lime boil stage (16). Hence the degree of recovery of silver from pretreatment processes would need to be tested.

Sensitivity

Using the process cost analyses model developed for the study the sensitivities of power, and limestone cost to the overall process costs were examined for Ore B. The base case values (Power \$0.10 kWh, Limestone \$75/t) were varied separately up to ±50% and the operating costs to ounces of gold produced calculated. The results of these analyses for all three process options are presented in Figure 18. In addition the impact of the variability of gold and silver extraction expected from cyanide leaching the oxidised concentrate separately on operating margin was assessed. The gold and silver extraction from the tails was left fixed at the test work data values. The variability of gold and silver extraction results are presented in Figure 19.

Overall limestone cost has no impact on roasting unless the acid produced is neutralised. In which case the operating costs for the roaster is closer to the pressure oxidation and biooxidation processes. Limestone cost has more of an impact than power cost for pressure oxidation. Whereas for biooxidation power and limestone costs yield similar trends. The availability and cost of limestone will have a major impact on the viability of pressure oxidation and biooxidation process.

As expected variation in the gold extraction in the refractory processes have significant influence on operating cost margin. Roasting has been reported in some studies to yield lower gold recoveries than pressure oxidation and biooxidation(17). Even for allowing a decrease in gold extraction by 10% (-13% from base case) the roaster option still yields lower operating cost per ounce of gold than the other two process options. From experience it may be expected that gold recovery may be slightly better (~2%) for pressure oxidation than bio-oxidation.

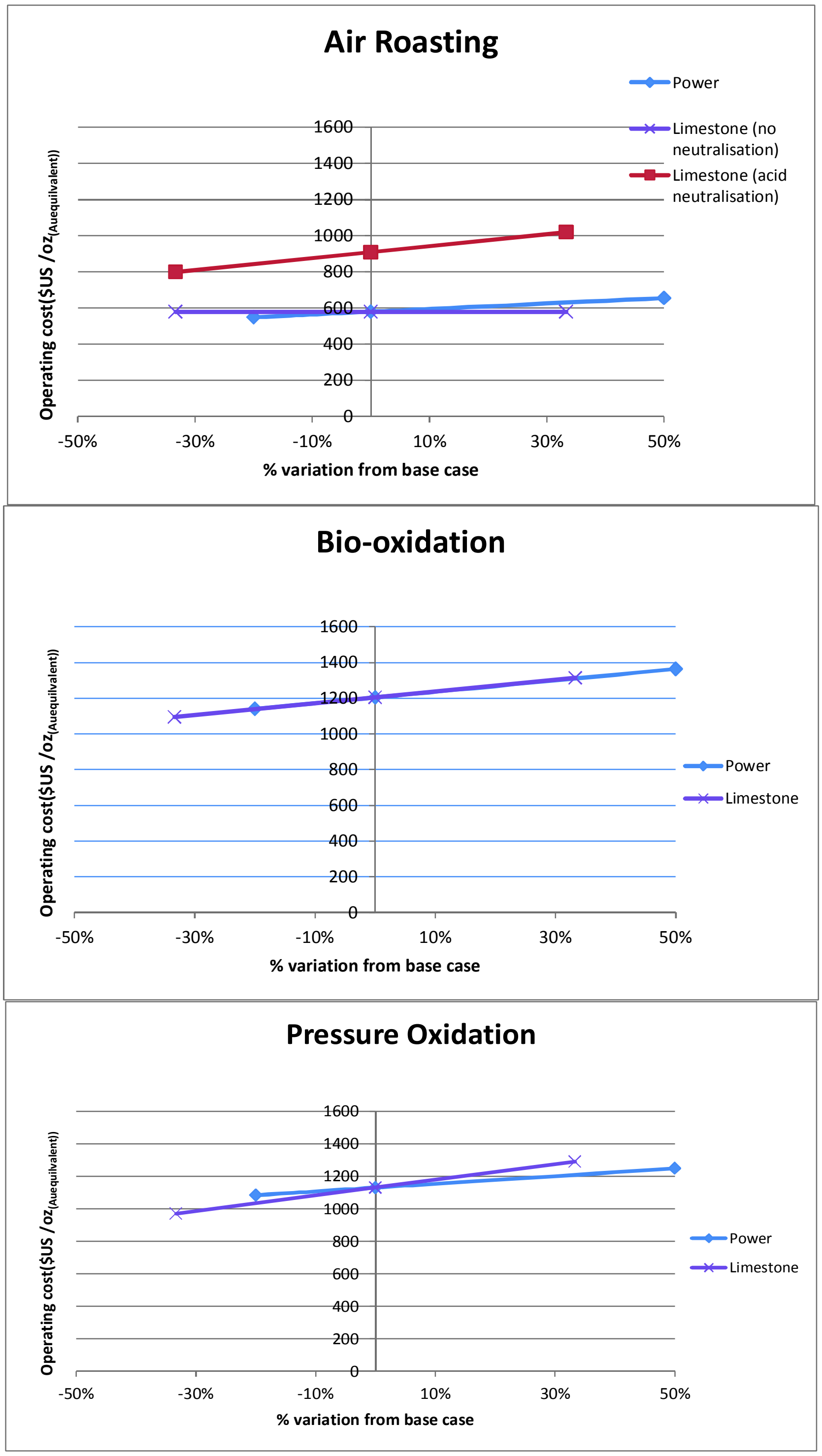


Figure 18: Sensitivity analyses of limestone and power costs for three process options (base case, Power $0.10 kWh, Limestone $75/t)

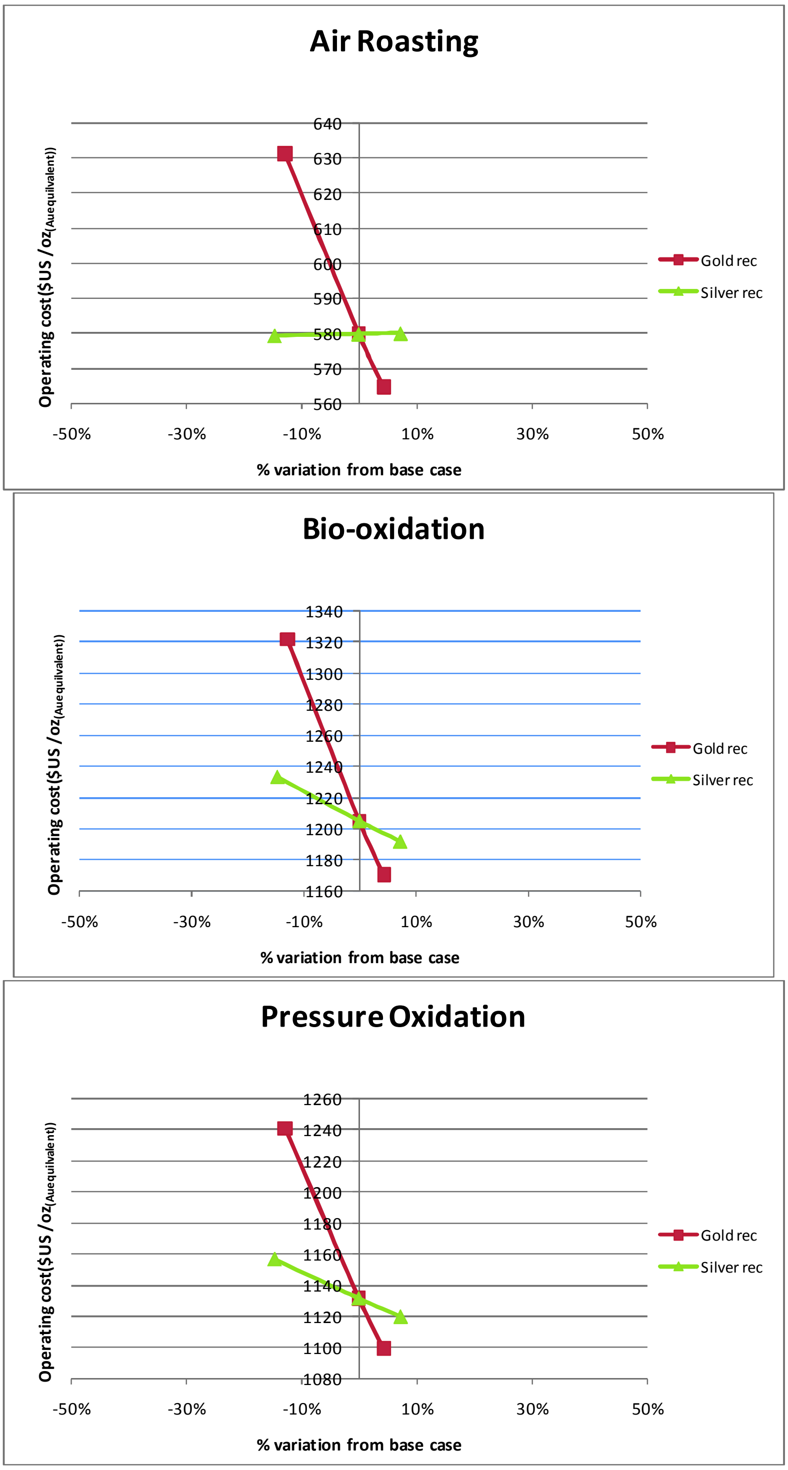


Figure 19: Sensitivity analyses of gold and silver recovery for all three process options (assuming gold and silver extraction variability from oxidised concentrate only, base case Au 95% and Ag 90% in extraction)

SUMMARY AND CONCLUSIONS

Conceptual level studies looking at various process treatment routes were carried out on two different ore bodies. The gold in both ore deposits is locked in arsenic bearing sulfide minerals and require a pre-treatment process prior to recovery.

Five different process options were examined for treating Ore A which has high arsenic and carbonate content at 0.6 Mtpa. The high carbonate content will make heap bio oxidation leach option not viable. The capital cost for alkaline Albion, bacteria tank leaching and POX were not significantly different.

The three conventional technologies were considered for treating the high silver Ore B at 3.65 Mtpa. The significantly high operating costs associated with the low gold to sulphur ratio will make treating Ore B a challenge. The capital costs of all three processes are not significantly different. Roasting operating costs appeared about half that of biooxidation and pressure oxidation. However air roasting assumes acid generated from SO_2 can be sold. In addition it is assumed the arsenic in the concentrate can be contained and stabilised within the roasted calcine.

These estimates were produced to provide the clients with sufficiently detailed options to assess the viability of the project. Further geometallurgical & gold deportment studies are required on both ore bodies. Extensive test work programme on all options is required before further engineering studies can be carried out..

ACKNOWLEDGEMENTS

Permission of two mining companies to present data from studies is greatly appreciated. Information provided by Goldfields and Xstrata is also greatly appreciated.

REFERENCES

1. Thomas, KG (2005) Pressure Oxidation Overview, Developments in Mineral Processing Vol 15, MD Adams, Elsevier, p 346-369.

2. Thomas, KG and Cole AP (2005) Roasting Developments – especially oxygenated roasting, Developments in Mineral Processing Vol 15, ed MD Adams.

3. Jones, D.L. Knoerr, S. and Mansanti J. (1990) Processing Options for Buckhorn Sulphide Ore, Gold 90, p261-273.

4. Fraser KS (1991) Processing of Refractory Gold Ores, KS, Walton, Wells, Minerals Engineering Vol 4 1991 Adams.

5. Hourn, M., Rohner, P., Bartsch, P., Ngoviky, K., (2005) Benefits of using the Albion process for a North Queensland project, and a case study of capital and operating cost benefits versus bacterial oxidation and pressure oxidation" Xtrata Technology Report (2005).

6. Filippou, D. & Demopoulos, G.P. (1997): Arsenic Immobilization by Controlled Scorodite Precipitation. Journal of Metals, December, pp 52-55.

7. Fujita, T., Taguchi, R., Abumiya, M., Matsumoto, M., Shibata, E. & Nakamura, T. (2008): Novel atmospheric scorodite synthesis by oxidation of ferrous sulphate solution. Part I. Hydrometallurgy, Vol 90, pp 92-102.

8. Fujita, T., Taguchi, R., Abumiya, M., Matsumoto, M., Shibata, E. & Nakamura, T. (2008): Novel atmospheric scorodite synthesis by oxidation of ferrous sulphate solution. Part II. Effect of temperature and air. Hydrometallurgy. Vol 90, pp 85-91.

9. Harris, B. (2003): The removal of arsenic from process solutions: theory and industrial practice. Hydrometallurgy 2003, Vol 2, pp 1889-1902, TMS, Canla.

10. Litz J. F. and Carter, R. W. (1990) Economics of Refractory Gold Ore Processes, 119th TMS Annual Meeting February 18-22, 1990.

11. Riveros, P.A., Dutrizac, J.p. & Spencer, P. (2001): Arsenic disposal practices in the metallurgical industry. Canadian Metallurgical Quarterly, Vol 40, No 4 pp 395-420. Canada.

12. Singhania, S., Wang, Q., Filippou, D. & Demopoulos, G.P. (2005): Acidity, valency and third ion effects on the precipitation of scorodite from mixed sulfate solutions under atmospheric pressure conditions. Met and Mat Trans. B 37B, 189-197.

13. Millard M (2005) Processing of High Silver Ores Developments in Mineral Processing Vol 15, MD Adams, Elsevier.

14. Logan T. C., Seal, T. Brierley. (2007) Whole ore heap Biooxidation of sulphidic Gold Bearing Ores,.Biomining (ed D E Rauling and Johnson D B) Springer-Verlog Berlin, Heidelberg, 2007 p112-138.

15. Aswegen PC, van Niekerk J and Olivier W (200&) The Biox process for the treatment of Refractory Gold Concentrates, Biomining (ed D E Rauling and Johnson D B) Springer-Verlog Berlin, Heidelberg, 2007 p

16. Simmons, G.L. (1996) Pressure oxidation process development for treating refractory carbonaceous ores at Twin Creeks. Randol Gold Forum, 1996.

17. Collins, M.J. Yuan, D. . Masters, M. Kalanchey R. and Yan L. (2011) Pilot Plant Pressure Oxidation Of Refractory Gold-Silver ConcentrateWorld Gold 2011 Proceedings of the 50th Annual Conference of Metallurgists of CIM Montreal, QC, Canada (Edited by G. Deschenes, R. Dimitrakopoulos, J. Bouchard).

18. Ketcham V. J, O'Reilly J.F and Vardillt W.D. (1993) The Lihir Gold Project; Process Plant Design Minerals Engineering, Vol. 6, Nos. 8-10, pp. 1037-1065.

COMMERCIALISATION OF THE ALBION PROCESS

Mike Hourn
Manager Hydrometallurgy
Xstrata Technology
Level 4, 307 Queen St, Brisbane, QLD 4001
Ph +61 7 3833 8539
Fax +61 7 3833 8555
Email mhourn@xstratatech.com.au

Duncan W. Turner PhD
General Manager - Albion Process
Core Resources Pty Ltd
44 Corunna Street, Albion, QLD 4010, Australia
Ph +61 7 3637 8105
Fax +61 7 3637 8199
Email dturner@coreresources.com.au

Website www.albionprocess.com

Presented by

Mike Hourn

ABSTRACT

The Albion Process™ is a combination of ultrafine grinding using Xstrata Technology's IsaMill™, followed by oxidative leaching at atmospheric pressure in a series of reactors designed to achieve high oxygen mass transfer efficiency. The feed to the Albion Process™ is a sulphide concentrate containing base or precious metals, and the Albion Process™ is used to oxidise the sulphide minerals in the concentrate and liberate these metals for recovery by conventional means. The oxidative leach circuit is operated at near neutral pH for treatment of refractory gold and silver concentrates, simplifying plant layout and reducing capital costs. For base metal concentrates, the oxidative leach is operated under acidic conditions.

The Albion Process™ has had a long and varied road from concept to commercialisation. The Albion Process™ technology was originally developed in 1994 by MIM Holdings/Xstrata and is patented worldwide. The idea for the Albion Process™ followed the successful commissioning of the first M3000 IsaMill™ at Mt Isa in 1994. The technology had been seen as strategic by MIM Holdings for the first 12 years of its development and was not offered to clients outside the MIM Holdings group. In 2005, after the Xstrata takeover of MIM Holdings, there was a change in strategic direction, and the technology was offered to external clients, through Xstrata Technology.

Interest in the technology has been very strong, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Certej Project. The technology moved into commercial production in 2010 with the commissioning of Xstrata's Albion Process™ plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Xstrata of a second Albion Process™ plant in Germany (16,000 tpa zinc metal). The Las Lagunas refractory gold project will be commissioned in 2012, and the GPM Gold refractory gold project will be commissioned in 2013. A fifth Albion Process™ plant for the Certej refractory gold project in Romania is in final Permitting stages.

THE ALBION PROCESS™

The Albion Process™ is a combination of ultrafine grinding and oxidative leaching at atmospheric pressure. The feed to the Albion Process™ is a sulphide concentrate containing base or precious metals, and the Albion Process™ is used to oxidise the sulphide minerals in the concentrate and liberate these metals for recovery by conventional means.

The first stage of the Albion Process™ is fine grinding of the concentrate using Xstrata's IsaMill™ technology. Most sulphide minerals cannot be leached at acceptable rates at atmospheric pressure. The process of ultrafine grinding introduces a high degree of strain into the sulphide mineral lattice. As a result, the number of grain boundary fractures and lattice defects in the mineral increases by several orders of magnitude, relative to un-ground minerals. This introduction of strain lowers the activation energy for the oxidation of the sulphides, and enables leaching under atmospheric conditions. The rate of leaching is also enhanced, due to the increase in mineral surface area.

Fine grinding also prevents passivation of the leaching mineral by products of the leach reaction. Passivation occurs when leach products, such as iron oxides and elemental sulphur, precipitate on the surface of the leaching mineral. These precipitates passivate the mineral by preventing the access of chemicals to the mineral surface. Passivation is normally complete once this precipitated layer is 2 – 3 μm thick. Ultrafine grinding of a mineral to a particle size of 80% passing 10 – 12 μm will prevent passivation, as the leaching mineral will disintegrate prior to the precipitate layer becoming thick enough to passivate the mineral.

After the mineral has been finely ground, the slurry is then leached in agitated tanks specially designed by Xstrata, known as the Albion Leach Reactor. In the Albion Leach Reactor oxygen is introduced to the leach slurry for oxidation at supersonic velocity to improve mass transfer efficiency. The Albion Leach Reactor is designed to operate at close to the boiling point of the slurry, and no cooling is required. Leaching is carried out autothermally, and the temperature of the leach slurry is set by the amount of heat released by the leaching reaction. Heat is not added to the leaching vessel from external sources, and excess heat generated from the oxidation process is removed through humidification of the vessel off gases.

THE ISAMILL™ TECHNOLOGY

The IsaMill™ is a large-scale energy efficient continuous grinding technology specifically developed for rugged metalliferrous applications. Xstrata Technology supplies the IsaMill™ to mining operations around the world, with over 100 mills installed in 9 countries worldwide. The IsaMill™ uses a very high energy intensity of 300kW/m^3 in the grinding chamber, resulting in a small footprint and simple installation. A diagram of the typical components of the IsaMill™ Grinding Plant is shown in Figure 1 and Figure 2.

The grinding media size for the IsaMill™ is within the range 1.5 – 3.5 mm. Media can come from various sources, such as an autogenous media screened from the feed ore, silica sands or ceramic beads.

The IsaMill™ will contain up to eight discs on the shaft, with each disc acting as a separate grinding element. The slurry residence time distribution through the mill approaches perfect plug flow with virtually no short circuiting. This allows the IsaMill™ to be operated in open circuit without the need for cyclones.

The IsaMill™ is available in the following models:

- M1000 (500 kW), capable of throughputs in the range 10 – 16 tonnes per hour
- M3000 (1100 kW), capable of throughputs in the range 20 – 35 tonnes per hour
- M5000 (1500 kW), capable of throughputs in the range 30 – 55 tonnes per hour
- M10000 (3000kW), capable of throughputs in the range 60 – 100 tonnes per hour

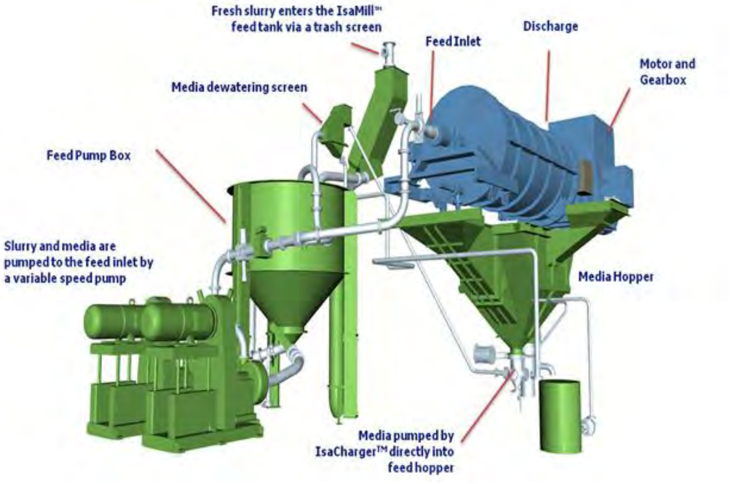


Figure 1
IsaMill^{TM} Key Components

The IsaMill^{TM} produces a sharp size distribution in open circuit, as the feed must pass through multiple distinct grinding zones in series before reaching the Product Separator. This plug flow action ensures no short circuiting, and efficiently directs energy to the coarser feed particles. The Product Separator is a centrifugal separator at the end of the mill shaft that spins at sufficient rpm to generate over 20 "g" forces, and this action is responsible for the sharp classification within the mill. The IsaMill^{TM} can be operated in open circuit at high slurry density, which is a key advantage for the leaching circuit, as the entry of water to the leach is limited, simplifying the water balance.

The IsaMill^{TM} uses inert grinding media that produces clean, polished mineral surfaces resulting in improved leaching kinetics. A steep particle size distribution is produced in the mill, with very little coarse material. The 98 % passing size in the mill is typically less than 2.5 times the 80 % passing size, and very little coarse material enters the leaching circuit, resulting in very high leach recoveries.

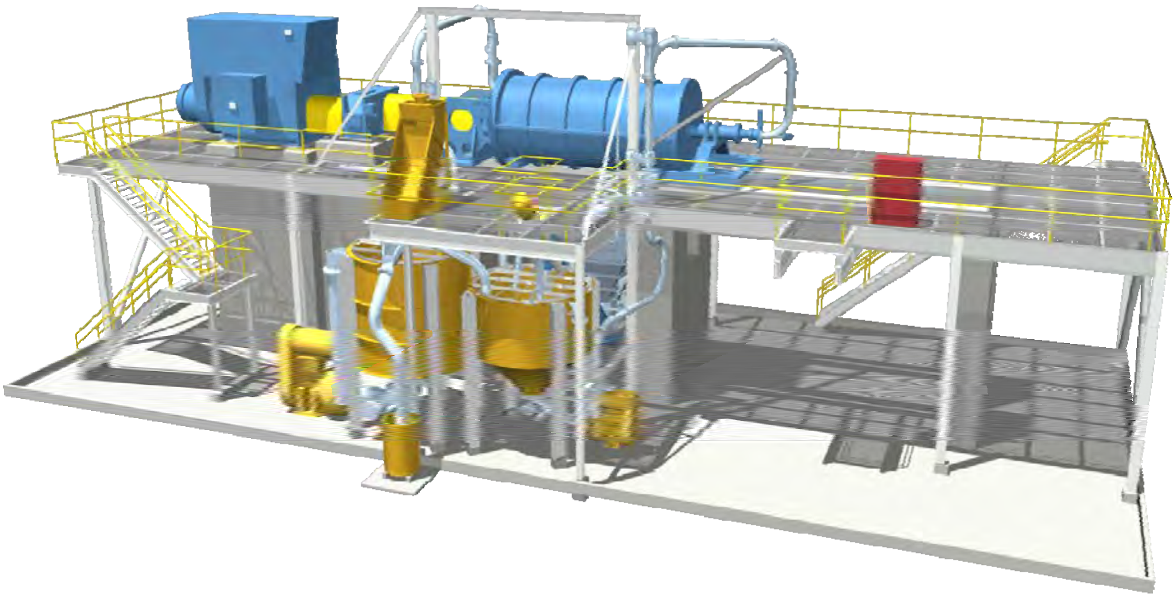


Figure 2
IsaMill^{TM} Grinding Plant Layout

The IsaMill^{TM} is the highest intensity grinding technology available (>300kW/m^3), meaning it is also the most compact, with a small footprint and low profile. The IsaMill^{TM} is oriented horizontally, with the grinding plant accessed by a single platform at an elevation of approximately 3 m. Access to the mill and maintenance is simplified by the low operating aspect of the IsaMill^{TM} and the associated grinding plant. Maintenance of the IsaMill^{TM} is similar to routine maintenance for a slurry pump. The internal rotating shaft is counter-levered at the feed inlet end so the discharge end flange and grinding chamber can be simply unbolted and slid off using hydraulic rams. A shut down for inspection and replacement of internal wear parts takes less than 8 hours. Availability of 99% and utilisation of 96% are typical of the IsaMill^{TM}.

Scale-up of the IsaMill^{TM} is straight forward. Laboratory test results are directly scaled to commercial size with 100% accuracy. The IsaMill^{TM} has a proven 1:1 direct scale-up to reduce project risk.

OXIDATIVE LEACHING

After the sulphide mineral has been finely ground, it is then leached under atmospheric conditions in a oxidative leach consisting of interconnected Albion Leach Reactors. The Albion Leach Reactor is an atmospheric leaching tank that has been designed by Xstrata Technology to achieve the required level of oxygen mass transfer to facilitate oxidation of the sulphide feed at low capital and operating cost[1].

The oxidative leaching circuit in an Albion Process^{TM} leach plant is similar to a conventional cyanide leach plant, with the Albion Leach Reactors connected in series with a launder system that allows gravity flow of the slurry through the leach train. All Albion Leach Reactors are fitted with bypass launders to allow any reactor to be removed from service for periodic maintenance. This is a low cost leaching system that is simple and flexible to operate, and the overall availability of the leach train is 99%.

Oxygen is injected into the base of the Albion Leach Reactors using a series of HyperSparge supersonic injection lances. The design of the HyperSparge injection system is carried out in conjunction with the design of the agitation system to ensure high oxygen mass transfer rates are achieved in the reactor. The agitator unit power is moderate, and the reactor is typically designed to achieve a blend time of 1 minute in the tank. The impeller tip speed is chosen in combination with the HyperSparge injection velocity to provide the required mass transfer.

The Albion Leach Reactor has a corrosion resistant alloy steel shell and base, supported on a ring beam or raft foundation. The tank aspect ratio is designed to achieve high oxygen transfer rates and capture efficiencies. Xstrata Technology has developed fully modular tank shell systems, which can be rapidly installed on site in one third the time of a field welded tank and at much lower costs. The Xstrata modular reactor designs require no site welding. The modular Albion Leach Reactor is shown in Figure 3.

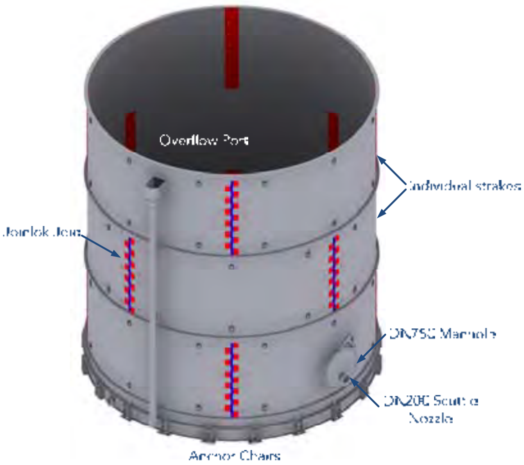


Figure 3
Modular Albion Leach Reactor

The reactor is fitted with a centrally mounted agitator consisting of one or more hydrofoil impellers. The agitator sizing and impeller geometry is chosen by Xstrata Technology using in house design correlations and testwork data to provide sufficient power to meet the oxygen mass transfer requirements in the leach vessel, as well as provide adequate solids suspension and gas dispersal. Impeller arrangements and spacing are also designed to assist in foam control within the vessel. The agitator is mounted off the tank shell, and modular maintenance platforms and structural supports are provided as part of the Albion Leach Reactor. Key design aspects of the agitator, such as the solidity ratio, the impeller diameters and tip speeds and the overall pumping rate are determined in combination with the design of the oxygen delivery system to provide the optimum mass transfer rates in the reactor.

The oxygen delivery system on the Albion Leach Reactor consists of Xstrata's Technology's HyperSparge supersonic oxygen injection lances, which are mounted circumferentially around the reactor, close to the base. The HyperSparge is mounted externally to the tank, and penetrates through the tank wall using a series of sealing assemblies. This novel design means that no downtime is incurred for maintenance of the oxygen delivery system, as all HyperSparge units can be removed live for inspection. The high gas capture efficiencies achieved by the HyperSparge system results in low gas rates entering the Albion Leach Reactor, and so there is little or no correction to the drawn power for the agitator. Aeration numbers are typically less than 0.025.

The HyperSparge injects oxygen at supersonic velocities, typically in the range 450 – 550 m.s^{-1}. The supersonic injection velocities result in a compressed gas jet at the tip of the sparger that incorporates slurry via shear resulting in very high mass transfer rates within the Albion Leach Reactors. The unique design of the HyperSparge means that the agitator power required for the Albion Leach Reactors is much lower than is required in a conventional system. Oxygen capture efficiencies of 85 % or higher are achieved in Albion Plants within the Xstrata group using the HyperSparge system. A typical HyperSparge assembly is shown in Figure 4. The high jet velocities at the tip of the HyperSparge keep the nozzle clean and eliminate blockages.

The HyperSparge can be incorporated in an overall oxygen control system, consisting of in stack off gas monitoring and control of the HyperSparge delivery pressure. The oxygen control system is used to maintain high oxygen capture efficiencies within the Albion Leach Reactor.

Exhaust gas from the oxidative leach is inert, and so the Albion Leach Reactor is fitted with sectional lids and an off gas stack to vent steam from the vessel to a safe working height. As the Albion Leach Reactors operate at close to the boiling point of the slurry, significant water vapour is released from the vessel with the exhaust gas, which assists in overall process water balance. The off gas stack is designed as a natural chimney to vent this exhaust gas to a safe working height. The exhaust gas it typically vented, however condensers can be fitted if required to recover the evaporated water.

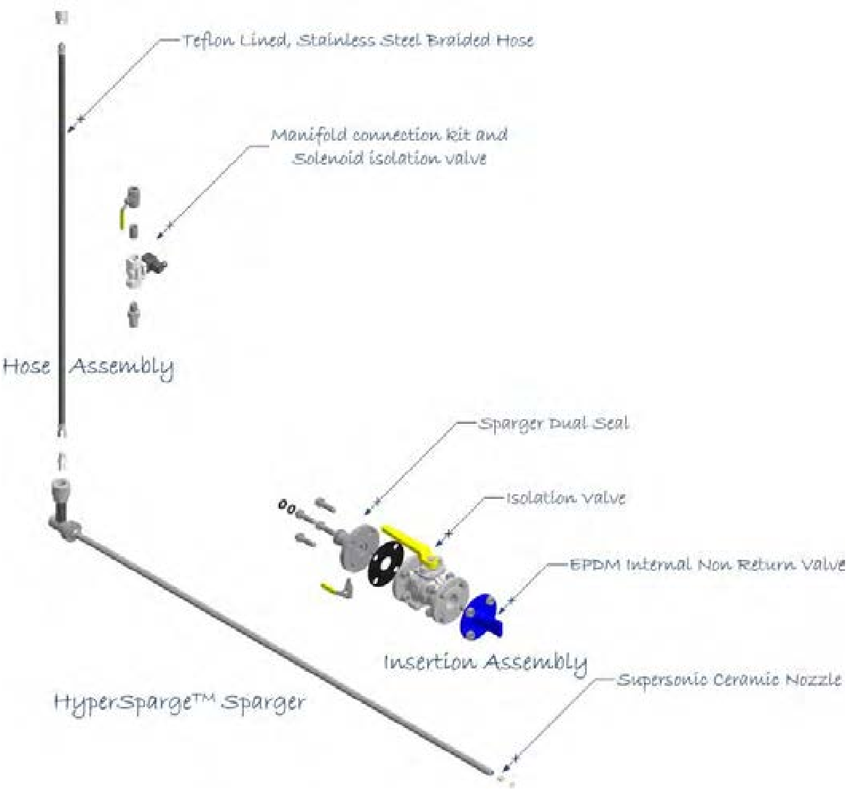


Figure 4
HyperSparge System

Each Albion Leach Reactor has modular Internal baffles to assist mixing and prevent slurry vortexing, as well as a modular slurry riser to prevent slurry short-circuiting and assist in transport of coarser material through the leaching train.

The Albion Leach Reactors are connected to each adjacent reactor via a launder system to transport slurry between the reactors. The launders are designed according to sound hydraulic principles and accomodate both slurry and foam transport. Xstrata Technology's launder design accommodates froth, preventing a build up of foam in the leach train.

No internal heating or cooling systems are required in the Albion Leach Reactors. The vessel is allowed to operate at its equilibrium temperature, which is typically in the range 90 – 98 ^{o}C, depending on site elevation. Heat is provided by the oxidation of the sulphide minerals, with heat lost from the vessel by humidification of off gas. No direct or indirect temperature control is required, simplifying tank construction and maintenance. No external cooling towers or flash vessels are required.

The Albion Process^{TM} oxidative leach has three primary control loops. The major control loop is the leach pH, with either acid or limestone dosed into the reactor to maintain pH to a set point. The pH set point is determined in testwork, and is set by the desired extent of oxidation of sulphide sulphur in the leach. For refractory gold circuits, full oxidation of sulphide to sulphate is desired, and the pH is held at near neutral with the addition of limestone slurry. Limestone is dosed off a ring main.

The slurry density in the reactor is maintained at the optimum level for viscosity and mass transfer control by the addition of process water. Oxygen addition is the third primary control loop, with oxygen gas addition rates through the HyperSparge oxygen delivery system varied based on off gas composition and flow rate. Temperature monitoring is provided, however no temperature control is necessary in the Albion Leach Reactor, and the leach operates autothermally.

Reagent dosing ports are located in the lid of the Albion Leach Reactor with all dosing valves and instruments located in easily accessible positions on the leach tank top platform.

OXIDATIVE LEACH CHEMISTRY FOR REFRACTORY GOLD

The Albion Process^{TM} oxidative leach oxidises sulphide minerals to either elemental sulphur or sulphate. This process liberates significant heat, and the oxidative leach is allowed to operate at a temperature close to the boiling point of the slurry. Typical operating temperatures are in the range 93 – 98 ^{o}C. At these operating temperatures, mineral leaching in the Albion Process^{TM} will occur in two steps. In the first step, the mineral sulphide is oxidised to a soluble sulphate and elemental sulphur. In the second step, the elemental sulphur is then oxidised to form sulphuric acid:

Step 1 $MS + H_2SO_4 + ½\ O_2 = MSO_4 + S^o + H_2O$ (A)

Step 2 $S^o + H_2O + 3/2\ O_2 = H_2SO_4$ (B)

These reactions can be catalysed by the action of ferric iron under acidic conditions.

The oxidative leach can be operated under a range of pH conditions, varying from acidic to neutral. The control pH will set the amount of elemental sulphur oxidation desired via reaction B. The extent of elemental sulphur oxidation can be varied from virtually nil to 100 % by control of the leach pH in the range 1 – 6.

When the oxidative leach is operated under acidic conditions, employed for copper, zinc or nickel leaching, some elemental sulphur oxidation is required to provide acid for the leach. In these systems, the background acidity is held in the range 5 – 15 gpl, and the leach acidity is maintained by either the addition of raffinate, or by allowing Reaction (B), the oxidation of elemental sulphur, to proceed. Elemental sulphur oxidation will proceed readily under the conditions found in the Albion Process^{TM} oxidative leach at acidities below 10 gpl, and slows significantly as the acidity approaches 15 gpl. In this way the Albion acidic leach is self regulating, oxidising elemental sulphur as required to maintain the required acidity.

The acidic leach is a two stage process, where economic metals are first leached in oxygenated acidic solution, with the acidic leach slurry then neutralised to precipitate iron and other deleterious elements such as arsenic, prior to separation of the leached solids and recovery of the economic metals from the neutralised leach solution. Metal recovery can be via conventional processes. Iron removal by goethite precipitation is the preferred neutralisation circuit for Albion Process^{TM} acid leach circuits.

Neutral leaching is used to oxidise pyrite and arsenopyrite concentrates, as well as concentrates containing tellurides and sellenides. The neutral leach is carried out at a pH in the range 5 – 7, with the continuous addition of an alkali to neutralise acid and iron sulphates generated by oxidation of the sulphide metals. All elemental sulphur is converted to sulphate, ultimately in the form of gypsum in the neutral leach. The neutral leach product is suitable for direct feed to a cyanide leach plant, without any filtration, counter current decantation or neutralisation stage, resulting in substantial capital savings relative to an acidic oxidation process.

Under the neutral pH operating conditions, the overall pyrite leach reaction is:

$FeS_2 + 15/4O_2(g) + 9/2H_2O + 2CaCO_3 = FeO.OH + 2CaSO_4.2H_2O + 2CO_2$ (C)

The iron precipitate, goethite, is very stable and has no solubility in cyanide. Albion Process^{TM} oxidative leach residues do not generate ferro or ferri cyanide species in the cyanide leach, simplifying cyanide destruction.

Arsenopyrite is a common gold carrier in many refractory gold concentrates. Depending on the level of arsenopyrite present in the concentrate, arsenopyrite is readily oxidised under the neutral leaching conditions, to form a stable ferric arsenate product.

The overall arsenopyrite leach reaction is:

$FeAsS + 7/2O_2(g) + 4\ H_2O + CaCO_3 = FeAsO_4.2H_2O + CaSO_4.2H_2O + CO_2$ (D)

Some refractory gold concentrates can contain a range of telluride bearing phases, such as AgAuTe, AgTe, PbTe, Pb(Bi)Te, PbAu(Sb)Te. All of these telluride phases contain high levels of gold and silver. Telluride leaching in an oxidative system is enhanced by ultrafine grinding, and is also accelerated under alkaline conditions. Tellurides break down quickly at elevated pH, with oxidation of telluride to $HTeO^{3+}$ and Au^+. The gold and tellurium then precipitate as oxides.

Telluride breakdown occurs rapidly in the neutral leach, according to the following general reaction:

$AgAuTe + 2O_2(g) = TeO_2 + AuO + AgO$ (E)

No elemental sulphur is formed under the neutral leaching conditions, and the oxidised residue will have a low cyanide consumption, as thiocyanate formation is avoided.
At high levels of oxidation, the final oxidised residue may be inert, with no residual acid generating components. The neutral pH operating conditions prevent the formation of jarosite, and so silver recoveries from the oxidised residue are high.

The neutral operating pH in the oxidative leach also results in very low background salt levels in the leach solutions. This improves oxygen solubility and significantly reduces the formation of scale in the Albion leach, simplifying operation. Gypsum scale formation is a major concern for neutralisation circuits where the initial tanks operate under acidic conditions. At the operating pH of 5 - 6 the stability of the HCO_3^- ion in solution is very low relative to acidic neutralisation circuits. When neutralising acid, limestone reacts according to the following general reaction:

$CaCO_3 + H^+ = HCO_3^- + Ca^{2+}$ (F)

Calcium will precipitate liberated sulphate to form gypsum. High bicarbonate activities seen under more acidic conditions will lower the calcium solubility and push the system closer to super-saturation with respect to gypsum. Bicarbonate activity follows the acidity level inversely, and so the higher the pH, the lower the bicarbonate activity, and hence the lower its effect on gypsum solubility. Ideally, the operating pH for the oxidative leach would be as high as the limestone quality will allow, to minimise scale, however limestone utilisation then ultimately becomes limiting. For this reason, a control set point in the range pH 5.5 – 6.5 is used for the oxidative leach circuit.

DEVELOPMENT HISTORY

As is the case with all new technologies, the Albion Process^{TM} has had a long and varied road from concept to commercialisation. The Albion Process^{TM} technology was originally developed in 1994 by MIM Holdings (MIM) and is patented worldwide [2]. The idea for the Albion Process^{TM} followed the successful commissioning of the first M3000 IsaMill^{TM} at Mt Isa in 1994. The IsaMill^{TM} was developed to provide efficient grinding down to 80 % passing sizes of 5 – 7 microns prior to flotation for Xstrata's complex M^{c}Arthur River and George Fisher lead-zinc deposits.

At that time, MIM were studying options for the development of the large Frieda River/Nena project in PNG through its subsidiary Highlands Pacific. The Nena ores were not amenable to smelting, due to the elevated arsenic content, and several hydrometallurgical options were examined. The option chosen originally was pressure oxidation, as MIM had experience in this operation through its holding in the Porgera joint venture, also in PNG. Feasibility studies indicated that the cost of the pressure oxidation circuit was prohibitive, and so lower cost options were examined. MIM had experience in atmospheric leaching through its joint venture zinc operations in Germany, and the atmospheric leach was seen as a low cost option that was also robust and simple to operate.

Initial batch testwork examining the combination of fine grinding and oxygenated atmospheric leaching began in early 1994, and the testwork was very successful with copper recoveries in excess of 98 % from the chalcopyrite, covellite, enargite mineralogy. Techno-economic studies were carried out with Davy John Brown in late 1994, and the economics of the process compared very favourably with pressure oxidation. The original patents were then filed in 1995

MIM decided to divest its interests in Highlands Pacific in 1996/7 and sold out of the Frieda River joint venture. MIM made a strategic decision at the time to focus future development of the Albion Process^{TM} on refractory gold. The major project targeted was the large Pueblo Viejo project in the Dominican Republic. MIM Holdings evaluated the Pueblo Viejo project between 1997 and 2002, during which time they constructed a 1 tonne per day Albion Process^{TM} demonstration plant in Brisbane that operated continuously for 18 months. This led to the development of a new variation of the Albion Process^{TM}, which involved oxidation of pyrite under near neutral pH conditions. Patents for this new process were lodged in 1999.

MIM Holdings competed in the tender process for the Pueblo Viejo project in 2001, however were not successful, with the project being awarded to Placer Pacific, later Barrick GoldCorp, who are now developing this project.

Following the unsuccessful bid for the Peublo Viejo Project, MIM then turned its focus for the Albion Process^{TM} to zinc, and started to develop a project around the large M^{c}Arthur River mine in the Northern Territory. The M^{c}Arthur River mine produces a low grade zinc concentrate containing elevated levels of lead, copper and iron. The deposit is complex, with lead and iron present as fine, sub 3 micron intergrowths within the sphalerite lattice. The IsaMill^{TM} had been instrumental in allowing MIM Holdings to develop the M^{c}Arthur River project, however the ongoing closures of imperial smelting furnaces in the late nineties caused concern about long term placement of the concentrate.

MIM began pilot testing of an Albion Process^{TM} circuit to recover zinc as cathode from the M^{c}Arthur River concentrate in 2001. In the flowsheet developed, zinc was leached in the Albion Process and recovered as cathode using a conventional purification/electrowinning circuit, with lead and silver reporting to a residue that could be sent to a smelter for lead recovery. Development of this project was continued by Xstrata following the takeover of MIM Holdings in 2003, and this has now led to the construction of two Albion Process^{TM} plants in Zinc and Germany. These plants will be discussed in the next section of this paper.

In the period from 1994 until 2004, the Albion Process^{TM} was seen as strategic to the MIM/Xstrata group, and was not marketed externally. In 2005, a decision was made within Xstrata to offer the technology to external clients under licence, and Xstrata appointed a marketing agent – Core Resources, to market the technology globally. Interest in the technology has been very strong in the subsequent period, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Certej Project. The technology moved into commercial production in 2010 with the commissioning of Xstrata's Albion Process^{TM} plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Xstrata of a second plant in Germany (16,000 tpa zinc metal).

The Las Lagunas refractory gold project will be commissioned in 2012, and the GPM Gold refractory gold project will be commissioned in 2013. A fifth Albion Process^{TM} plant for the Certej Project in Romania is in final Permitting stages. These projects will be reviewed briefly.

XSTRATA ZINC PROJECTS – SAN JUAN DE NEIVA AND NORDENHAM

An Albion Process^{TM} plant was constructed at the San Juan de Neiva zinc refinery in July 2010. San Juan de Neiva is the world's largest zinc refinery, with an annual production of over 450,000 tonnes of zinc. The refinery is owned by Xstrata, and the feed to the Albion Process^{TM} plant is finely ground lead/zinc concentrate from the McArthur River mine in the Northern Territory. The Albion Process circuits employed by Xstrata Zinc are unique, in that the fine grinding stage is located at a different site to the oxidative leach, with the finely ground concentrate transported by sea freight.

Figure 5
Albion Leach Reactor at San Juan de Neiva

The San Juan de Neiva Albion Process^{TM} plant has a throughput of 9000 tonnes per annum of concentrate, ground to a size of 80 % passing 7 microns. The plant produces 4,000 tonnes per annum of cathode zinc from the concentrate, at a recovery of 98.6 % w/w. The leach is an oxygenated sulphuric acid leach, using spent electrolyte from the electrowinning plant.

The McArthur River Lead-zinc concentrate is first re-slurried in spent electrolyte from the electrowinning cellhouse, and the slurry pumped to the oxidative leaching circuit. The oxidative leach consists of 280 m^3 Albion Leach Reactors, with a live height of 9.5 m and a diameter of 6 m.

One of the Albion Leach Reactors, showing the HyperSparge insertion points, is shown in Figure 5.

Concentrate slurry and spent electrolyte are transferred to the Albion Leach Reactors along with residue slurry from the Neutral Leach circuit, which introduces ferrite residues from the Neutral Leach into the Albion leach circuit for oxidation. This allows the Albion Process^{TM} leach circuit to act as a traditional Hot Acid Leach for the recovery of zinc from ferrite residues.

Oxygen is injected into the base of the reactors using 8 HyperSparge supersonic oxygen injection lances, and the slurry is agitated by a centrally mounted dual hydrofoil impeller. Residence time in the oxidative leach circuit is approximately 28 hours.

Sphalerite is the main zinc mineral in the McArthur River concentrate, and is oxidised to form soluble zinc sulphate and elemental sulphur. The reaction consumes acid and oxygen:

$$2\ ZnS + O_2(g) + 2\ H_2SO_4(a) = 2\ ZnSO_4(a) + 2\ S^o + 2\ H_2O \qquad (G)$$

Lead is present in the concentrate as Galena, and is oxidised almost exclusively to lead sulphate, which remains in the leach residue:

$$2\ PbS + O_2(g) + 2\ H_2SO_4(a) = 2\ PbSO_4(s) + 2\ S^o + 2\ H_2O \qquad (H)$$

Some plumbojarosite is formed in the early stages of the leach, however this is converted to lead sulphate as the leach progresses.

The McArthur River concentrate is a relatively low grade zinc concentrate, and contains elevated levels of iron, present as pyrite. Pyrite is slow to leach at the acid levels in the oxidative leach, and will not start to oxidise at a significant rate until over 90 % of the sphalerite is oxidised, due to galvanic effects. The leach reaction for pyrite under the acidic conditions in the oxidative leach is:

$$2\ FeS_2 + O_2(g) + 2\ H_2SO_4(a) = 2\ FeSO_4(a) + 4\ S^o + H_2SO_4(a) \qquad (I)$$

The elemental sulphur is slow to react under the 40 – 160 gpl acid concentrations in the leach, and less than 5 % oxidation is typical:

$$S^o + 3/2\ O_2(g) + H_2O = H_2SO_4(a) \qquad (J)$$

The oxidative leach circuit is also used to recover zinc from ferrite phases that do not leach in the existing calcine leaching circuit. The high operating temperature of the leach circuit results in rapid breakdown of the ferrites, according to the reaction:

$$ZnFe_2O_4 + 4\ H_2SO_4(a) = ZnSO_4(a) + Fe_2(SO_4)_3(a) + 4\ H_2O \qquad (K)$$

The oxidised slurry from the oxidative leach plant processed to recover both the elemental sulphur and the lead as separate products. The slurry is first thickened, and the thickener underflow then treated to recover elemental sulphur in a series of 10 m^3 flotation cells. The flotation concentrate, containing up to 70 % w/w elemental sulphur, is filtered in a plate and frame filter assembly, and the filtrate returned to the thickener circuit. The sulphur concentrate is transferred to the roasting plant and burnt to produce sulphuric acid.

The flotation circuit tailings are filtered in a horizontal belt filter, with the high grade residue then sold to secondary lead recyclers. The filtrate and wash from the horizontal belt filter is returned to the thickener. Thickener overflow reports to the existing jarosite precipitation circuit for iron removal ahead of the zinc dust precipitation circuit and the cellhouse.

Figure 6
Albion Leach Reactor at Nordenham

Following the success of the first Albion Process^{TM} plant at San Juan de Neiva, Xstrata then constructed a second Albion Process plant at the Nordenham zinc refinery. This plant was commissioned in March 2011. The feed to the plant is also finely ground lead/zinc concentrate from the McArthur River mine, and the process flowsheet is similar to that employed at San Juan de Neiva.

The Nordenham Albion Process^{TM} plant has a throughput of 36,000 tpa of concentrate, and the plant produces 16,000 tpa of cathode zinc from the concentrate, at a recovery of 98.8 % w/w.

The oxidative leach consists of an 800 m^3 and a 280 m^3 reactor in series. The 280 m^3 reactor is similar in geometry to the San Juan de Neiva leach reactor, and the 800 m^3 has a height of 13.6 m and a diameter of 9 m. Oxygen is injected into the base of the both reactors using HyperSparge supersonic oxygen injection lances, and the slurry is again agitated by a centrally mounted dual hydrofoil impeller. A picture of the Nordenham Albion Process^{TM} oxidative leach is shown in Figure 6.

The lead residue from the Nordenham Albion Process^{TM} plant is also sold locally to secondary lead producers.

A summary of the key operating data for the two Xstrata Zinc Albion Process^{TM} plants is presented in Table 1.

Table 1
Summary Data for the Xstrata Zinc Albion Process^{TM} Plants

Parameter	**San Juan de Neiva**	**Nordenham**
Feed Rate - tph		
MRM Concentrate	1	4
Neutral Leach Residue	0.25	1
Concentrate Composition		
Zinc - %	47.2%	47.2%
Lead - %	8.17%	8.17%
Iron - %	5.4%	5.4%
SiO_2 - %	4.3%	4.3%
Copper - %	0.8%	0.8%
Sulphur - %	47.2%	47.2%
Tank Size – m^3	280	800, 280
Leach Recovery - % w/w	98.6	98.8
Conversion to elemental sulphur	<5	<5
Leach acid demand – kg/tonne	488	470
Leach oxygen demand – kg/tonne	209	214
Residue Composition		
Sulphur - %	43	42
Zinc - %	1.3	0.8
Lead - %	15.5	16.2
Iron - %	8.4	8.1
SiO_2 - %	7.5	7.1
Copper - %	0.20	0.20

PANTERRA GOLD – LAS LAGUNAS

The Las Lagunas project is a refractory gold and silver project owned by Panterra Gold Limited. The project involves the reprocessing of high grade gold/silver pyritic CIL tailings from prior operation of the Rosario mine in the Dominican Republic. The Project is 15 km west of the provincial capital of Cotui and approximately 100 km northwest of the national capital of Santo Domingo. The lease is adjacent Barrick Gold Corp’s Pueblo Viejo gold mine.

The tailings dam was constructed in 1991, and was filled with sulphide tailings from open pit operations at the Rosario mine between 1992 and 1999. The Rosario mine was operated by Rosario Dominicana S.A, a mining corporation owned and operated by the Dominican government. Oxide reserves at the mine were depleted in the late eighties, with the feed to the plant containing increasing levels of sulphides, present primarily as pyrite, with minor enargite. The host rock was predominantly lithic tuff, with minor carbonaceous shales.

Operations at the Rosario mine were put on care and maintenance in 1999 due to the increasingly refractory nature of the ore. At the time of closure of the mine, gold and silver recoveries in the carbon in leach plant had fallen to below 30 %, with the unrecovered gold present as sub-microscopic inclusions in pyrite. Treatment of these increasingly refractory ores in the early nineties resulted in significant tonnages of tailings with +3.5g/t gold being stored in the Las Lagunas dam.

PanTerra Gold's subsidiary, EnviroGold (Las Lagunas) Limited, won an international tender and signed a Contract with the Dominican State in 2004 giving it the right to reprocess the tailings under a profit sharing agreement with the Government. The project has a JORC Indicated Resource of 5.137mt of ore grading 3.8g/t gold and 38.6g/t silver.

Annual production from the project will be 65,000 ounces per annum of gold and 600,000 ounces per annum of silver. The project contains significant silver revenue, and the neutral Albion Process^{TM} oxidative leach was chosen for the project, due to the high gold and silver recoveries from the oxidised residue. The process flowsheet was developed by Xstrata Technology with testwork carried out Xstrata's research facilities in Brisbane in 2004 – 2006.

Construction at the project was completed in the first quarter of 2012, and the project is currently being commissioned. Capital costs for the project were approximately $US 82.7 million, and the operating cost is projected at $US 30.68 /tonne, resulting in a cash cost of about $US 307 per ounce of gold equivalent produced.

The process flowsheet consists of an electric dredge to recover 100 tph of tailings from the dam. The dam is also used to re-deposit treated tailings and is mined in cells. Slurry from the dredge is stored in a 223 m^3 stock tank prior to being pumped to a 700 kW ball mill to freshen the mineral surfaces prior to flotation. The slurry is then processed through a series of flotation cells to recover a sulphide concentrate grading on average 14 g/t gold and 125 g/t silver, with 16 % w/w sulphide sulphur. A single bulk rougher stage is used, consisting of five flotation cells with a volume of 40m^3 each.

Figure 7
Las Lagunas IsaMill Grinding Plant

The flotation concentrate is produced at a rate of 28 tph and is thickened in a high rate thickener prior to being fed to the Albion Process^{TM} Plant. The Albion Process^{TM} plant consists of an M3000 IsaMill^{TM}, with an installed power of 1.5MW and a series of five 600 m^3 stainless steel Albion Leach Reactors. Each Albion Leach Reactor is fitted with 12 HyperSparge supersonic oxygen injection lances, and operates at a pH of 5.5 with continual dosing of limestone slurry. Limestone is dosed into each Albion Leach Reactor off a central ring main. Limestone for the Albion Process^{TM} circuit is quarried from an historic limestone quarry.

The Albion Process Plant operates under near neutral pH conditions, with the pyrite oxidised to goethite, and all sulphide reporting as gypsum. Gold recoveries from the oxidised residue will be above 90 %, and as no jarosites are formed in the leach circuit, silver recoveries are also above 90 %. The leach operates autothermally at a temperature of 98 ^{o}C, and exhaust gas is vented from the reactors via exhaust stacks.

Figure 8
Las Lagunas Oxidative Leaching Plant

The oxidative leach plant was sized to oxidise 80 - 85 tonnes per day of sulphide to sulphate. Oxygen is supplied to the Albion Process^{TM} Plant from a 220 tpd VPSA oxygen plant owned and operated by Panterra Gold. A photograph of the M3000 IsaMill Grinding Plant for the Las Lagunas project is shown in Figure 7. A photograph of the Albion Leach Reactors, showing the reactor shell, HyperSparge insertion nozzles and the off gas vent stacks is shown in Figure 8.

Oxidised residue from the Albion Process^{TM} Plant is thickened to approximately 45 % w/w and the thickener underflow cooled in a slurry cooling tower from 80 ^{o}C to 45 ^{o}C prior to the cyanide leach. The cooled underflow is transferred to a conventional Cyanide Leach circuit. The CIL circuit consists of one 496 m^3 neutralization tank and 6 x 441 m^3 Carbon-in-Leach (CIL) tanks and an AARL elution circuit. Tailings from the CIL circuit are returned to a sectioned cell within the tailings dam.

GPM GOLD PROJECT

The GPM Gold Project is owned by GeoProMining Gold LLC (GPM) and is located in Armenia. The Project consists of an open cut mine at Zod, near the Azerbaijan border, and a processing plant at Ararat near the Turkish border. The gold bearing ore, mined at the Zod Mine, is transported to the Ararat Process Plant via a state owned rail link.

The metallurgical facility at Ararat was originally constructed by the Soviets in 1973 and acted as a central processing facility to treat pyritic gold ores, with flotation concentrates produced at the site and sent to Russia for gold recovery. The project was purchased by First Dynasty Mines in the early 1990's and a CIL plant was constructed and commissioned in 1997 to recover gold from the concentrator tailings. All plant and equipment are still operating at the Ararat site, and will undergo refurbishment as part of the project. The project was briefly owned by Sterlite Gold in the early 2000's and was taken over by GPM in late 2007. GPM have been treating low sulphide ores and gabbro from the Zod mine in the intervening period.

The Zod deposit originally consisted of free milling weathered oxide ores overlying deeper sulphides. Historical mining has depleted the oxide ores, and the Ararat Process Plant now treats sulphide ore with declining gold recoveries. To improve the gold recovery from sulphide ores, GPM will install an Albion Process™ Plant at the Ararat Process Plant.

The remaining deposit has an average sulphur grade of 1.4 % w/w, with an average gold grade of 4.54 g/t and silver grade of 4.65 g/t. Of the remaining minable reserves, less than 30 % of the gold is free milling, with the remainder housed within the sulphides and telluride phases. Within the sulphide phases, gold is preferentially associated with arsenopyrite, and to a lesser extent pyrite. The arsenic grade across the deposit is 0.3 % w/w.

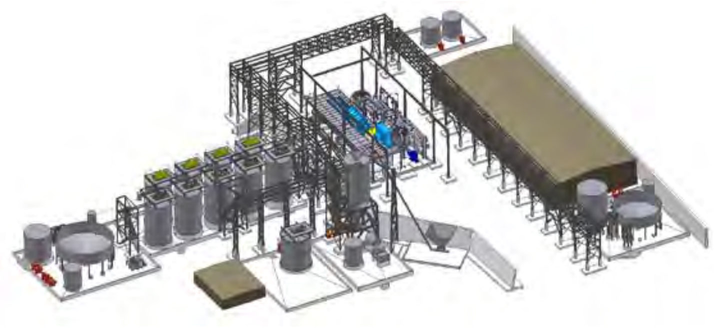


Figure 9
GPM Gold Albion Process™ Plant

Run of mine ore will be ground through parallel milling lines consisting of one AG and two ball mills, with the ground slurry then processed through a flotation circuit to recover a sulphide concentrate. The flotation circuit will consist of a bulk rougher/scavenger float followed by a cleaner stage. Cleaner Concentrate will then be transferred to the Albion Process™ Plant for oxidation ahead of the CIL circuit.

Mill tailings will also report to the CIL plant, and be blended with the oxidised residue prior to processing. The comminution circuit already exists at the project site, having been initially installed in 1976, and will undergo modernisation as part of the project. Components of the flotation circuit exist at the site, and will be refurbished as part of the project as well. Xstrata Technology is providing the detailed design services for the refurbishment of the flotation circuit. The CIL circuit will also undergo partial modernisation as part of the project.

The average throughput for the Albion Process™ Plant will be 94,007 tpa of Cleaner Concentrate, with a design factor of 15 % applied to the nominal rate to achieve a design rate of 108,108 tpa. The cleaner concentrate will grade 20 % w/w iron, 18 - 20 % w/w sulphide and 2.5 – 3.5 % w/w arsenic. The concentrate will grade approximately 40 g/t gold and 35 g/t silver. An Isometric of the Albion Plant for the GPM Gold Project is shown in Figure 9.

Cleaner Concentrate will be ground to an 80 % passing size of 11.1 μm in an M3000 IsaMill™, with an installed power of 1,120 kW. Ground concentrate will be transferred to the oxidative leaching circuit. The oxidative leach circuit volume will be based on a residence time of thirty (30) hours and the Leach circuit will consist of nine Albion Leach Reactors with a volume of 225 m^3 each.

Each Albion Leach Reactor will be fabricated from LDX 2101 alloy and will be agitated by a centrally mounted agitator fitted with dual hydrofoil impellers. The impeller assembly will be designed to provide the required gas hold up and solution pumping rate to ensure efficient reaction within the tank. The Albion Leach Reactors will be identical, and will all be supplied by Xstrata Technology using the ZipaTank^{TM} modular construction system. The ZipaTank^{TM} will be supplied in fully fabricated modular sections for assembly on site. Skilled artisans are rare in Armenia, and so modular construction techniques will be widely used across the plant.

Oxygen will be added to each leach tank via banks of 9 HyperSparge™ oxygen spargers from dual 60 tpd VPSA oxygen plants. The pH in each Albion Leach Reactor will be maintained by dosing of limestone slurry into the reactor. Limestone will be dosed into the Albion Leach Reactors off a central ring main. The project will have a 6 tph limestone milling and distribution circuit that will mill minus 20 mm limestone to 80 % passing 75 microns for dosing into the oxidative leach. Coarse limestone will be sourced from a cement plant located 3 km to the east of the plant site.

Overflow slurry from the Albion Process^{TM} oxidative leach circuit will gravitate by launder to a 10 m diameter free standing thickener, and thickened slurry, at 41 %w/w solids will be pumped to the CIL plant and blended with mill tailings. Gold recovery from the oxidised residue will be 93 – 96 % w/w, with a silver recovery of 90 - 93 % w/w.

Xstrata Technology is providing the full design and equipment supply for the Albion Process^{TM} plant to the GPM Gold project as a fixed price package. The scope of work also includes the limestone milling and distribution circuit as well as the oxygen plant. Civil and structural works have begun at the plant site, with plant commissioning scheduled to begin in March, 2013.

CERTEJ PROJECT - ROMANIA

The Certej project is located in Western Romania, in the “Golden Quadrilateral” area of the Apuseni Mountains of Transylvania. The project is 12km from the regional town of Deva, and close to the town of Certej. European Goldfields owns 80% of the project through its subsidiary Deva Gold S.A. There is an existing open pit mine at the project site that was operated by the Romanian State mining entity Minvest, until 2006. Ore from the mine was processed in a comminution and flotation facility located at Certej. Deva Gold holds a valid operating permit for the mine.

The Certej deposit is a low to medium sulphidation epithermal deposit. The major sulphide minerals are pyrite with minor levels of arsenical pyrite. Copper and zinc sulphides are present in the ore and report to the flotation concentrate. The gold is present mostly as sub-microscopic inclusions in the pyrite and arsenical pyrite matrix.

The project involves the open cut mining and processing of 3 million tonnes of ore per annum over a project life of eleven and a half years. The project will yield approximately 160,000 oz of gold and 820,000 oz of silver per year in the form of doré, reflecting an average total process recovery of 81% for gold and approximately 75% for silver.

The process flowsheet consists of a comminution circuit and flotation plant to recover the sulphide minerals to a concentrate that is fed to the Albion Process^{TM} Plant. The comminution circuit will consist of a crushing plant and SAG and Ball mills grinding to 80 % passing 106 microns. Ground ore will be floated in a rougher/cleaner flotation circuit producing 28 tph of pyrite concentrate. The concentrate will grade 35 % w/w iron, 43 % w/w sulphide and 0.5 % w/w arsenic. The concentrate will contain approximately 15 - 18 g/t gold and 90 g/t silver.

The Albion Process plant will consist of an M10000 IsaMill^{TM}, with 3000 kW of installed power and a series of five 1500 m^{3} Albion Leach Reactors. The ground product from the M10000 IsaMill^{TM} will have an 80 % passing size of 11.5 microns. The Albion Process Plant will be sized to oxidise 180 tonnes per day of sulphide to sulphate.

As the feed to the plant is a refractory pyrite concentrate, the Albion Process^{TM} Plant will operate under near neutral pH conditions, with the pH maintained by the addition of limestone slurry to the reactors off a central ring main. Limestone will be quarried locally. Oxygen will be supplied to the Albion Process^{TM} circuit from a 520 tpd Cryogenic oxygen plant owned and operated by European Goldfields, at a purity of 95 % v/v.

Oxidised residue from the Albion Process^{TM} Plant will be thickened and the thickener underflow cooled in an open void slurry cooling tower. The cooled underflow will then be transferred to a conventional Cyanide Leach plant operating with an INCO cyanide destruct circuit to recover the precious metals. Gold recovery from the Albion Process^{TM} oxidised residue will be 93 % w/w.

The level of pyrite oxidation targeted in the Albion Process^{TM} Plant will be 70 - 75 %, with less than 5 % expected oxidation of the copper and zinc sulphides under the near neutral pH operating conditions. The oxygen consumption will be 270 kg/tonne of Cleaner Concentrate, with a limestone consumption of 450 kg/tonne.

Basic engineering for the Certej project was completed in 2008, and the project is currently undergoing final permitting. Detailed design for the project is expected to begin in 2012.

ENGINEERING CHALLENGES ON THE ROAD TO COMMERCIALISATION

THE ISAMILL^{TM}

The IsaMill^{TM} was developed as a joint project between MIM Holdings and Netzsch GmbH of Germany in the early 1990's, with the first commercial mill installed at Mt Isa in 1994. The IsaMill^{TM} was a revolutionary concept in grinding, extending the economic range for grinding down to 80 % passing 5 – 7 microns. The development of the IsaMill^{TM} presented a host of engineering challenges.

The first engineering challenge in the design of the IsaMill^{TM} was how to impart energy to the fine media required for grinding to fine sizes. Media sizes in the range 2 – 5 mm were required to achieve the 80 % passing 5 – 7 micron grind sizes originally targeted in the IsaMill^{TM}, and traditional tumbling mills will begin to centrifuge the ball charge at media sizes of less than 30 – 35 mm.

Agitated mills were originally developed to overcome the tendency for the media charge to centrifuge, and the Tower Mill and Metprotech mill were early variations. These mills extended the range of grinding media down to 10 – 12 mm, however neither mill could agitated a media bed at finer sizes without significant expansion of the media charge or centrifuging.

To overcome the problems associated with bed expansion and centrifuging, the IsaMill^{TM} was developed with an agitator that consisted of a series of flat discs at right angles to a central shaft. This agitation system can be thought of as a zero pumping head impeller, and so all energy is transferred to the media change by shear, rather than by generation of flow. The shaft sits horizontally within a bed of media retained in a cylindrical shell. The discs were designed with holes toward the midline of the disc to allow slurry circulation around the disc. These holes result in a localised circular slurry eddy between each disc. The combination of 7 to 8 of these slurry flow paths in a typical IsaMill^{TM} results in almost perfect plug flow through the mill. For this reason, the IsaMill^{TM} can be operated in open circuit, without the need for cyclones.

The horizontal configuration chosen by the designers of the IsaMill^{TM} also overcame the significant breakout torque typical in the vertically oriented Tower Mill and Metprotech mill. The breakout torque results from the mass of the media charge sitting vertically on the bottom impellers on the shaft. The horizontal mill layout, and the use of the disc style agitators resulted in virtually no breakout torque requirement for starting the mill.

The horizontal alignment of the mill shaft itself introduced several engineering hurdles, with the mill shaft designed as a cantilevered element, with no steady bearing at the end of the shaft and the shaft weight supported by a single bearing assembly. The shaft is immersed in the slurry and so requires sealing. Several sealing options were trailed, with the mill designers settling on a packed gland, similar in design to a centrifugal slurry pump.

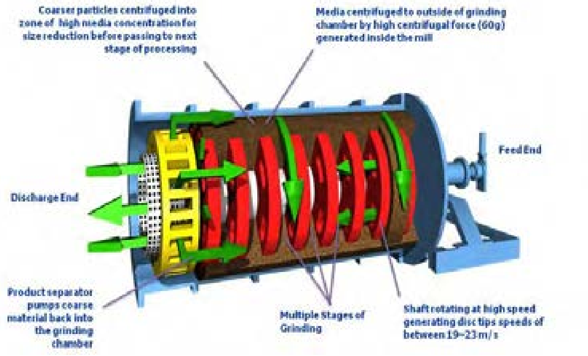


Figure 10
IsaMill^{TM} Operating Mechanism

Media retention within the IsaMill^{TM} was another significant design challenge. Early developments tested screens mounted internally within the mill to retain media and allow the slurry to flow, however these were prone to blockages. The solution to this was elegant, with a change in the final disc geometry and type at the end of the IsaMill^{TM} shaft. The shaft was fitted with an expeller at the end of the shaft, and the distance between the final disc and the expeller was shortened to provide centrifuging of the media bed between the expeller and the final disc. This arrangement is termed the Product Separator. The Product Separator is a very efficient internal classification device, generating up to 20 “g” forces, and provides a very sharp cut size, retaining media within the mill while allowing slurry to flow out of the mill readily. The pressure drop across an IsaMill^{TM} is typically less than 200 kPa. The operating principal for the Product Separator is shown in Figure 10.

Development of the IsaMill^{TM} is ongoing, with Xstrata Technology continually refining aspects of the mill design. Significant work is currently ongoing into new types of wear components to extend disc and liner life within the mill, as well as design of a larger mill, with an installed power of 8 MW.

OXIDATIVE LEACH

The main design challenge of the Albion Leach Reactor is to construct a system capable of achieving the oxygen mass transfer rates required to take advantage of the rapid leach kinetics possible with the finely ground IsaMill^{TM} product. Mass transfer rates in the range 3 – 5 kg.m^{-3}.h^{-1} are required in the oxidative leach. These mass transfer rates correspond to mass transfer coefficients in the range 0.5 – 0.9 s^{-1}.

In conventional atmospheric oxidation systems, such as bacterial leach reactors and fermentation vessels, mass transfer is achieved by shear created by differential acceleration between the gas and slurry phase. The rotational speed of the agitator is used to generate shear. Agitator power is proportional to the agitator tip speed cubed, and so power demand increases as a power law for modest increases in shear and mass transfer requirement.

The power per unit volume required to achieve mass transfer rates in this range is of the order of 2 – 5 kW.m^{-3} of slurry[5]. Agitator tip speeds are in the range 5 – 8 m.s^{-1}, and so the velocity profiles created in the slurry phase are below 10 m.s^{-1} at the tip of the agitator blade, and dissipate as the slurry moves away from the blade. The Albion Leach Reactor is designed to achieve the required oxygen mass transfer rates at a tenth of this power input.

A different approach was taken in the design of the Albion Leach Reactor to lower the power demand. In the Albion Leach Reactor, gas injection at supersonic velocities is used to create the bulk of the shear within the vessel. Gas injection velocities of the order of 500 m.s^{-1} are typical of the HyperSparge oxygen injection lance, compared to the 4 – 8 m.s^{-1} achieved with a typical agitator. Supersonic oxygen injection is a far more efficient method of generating shear than conventional agitation, allowing the total power input into the vessel to be significantly reduced.

The duty requirement for the agitator is reduced, and the agitator tip speed can be decreased, lowering the agitator power draw. Solids suspension in the oxidative leach is not onerous due to the finely ground feed and the slurry is typically homogenous and can be considered non-settling.

Oxygen mass transfer rates are described by the following equation:

$$dO_2/dt = k_L a \ (C_{sat} - C)$$

where:

k_L = liquid film mass transfer coefficient for oxygen into solution ($m.s^{-1}$)
a = the specific gas surface area, which is the ratio of bubble surface area to water volume ($m^2/m^3 = m^{-1}$).
C_{sat} = the solubility of oxygen in the slurry at saturation ($g.m^{-3}$)
C = the steady state oxygen level in the slurry ($g.m^{-3}$)

Typically, the k_L and a terms are combined to represent the mass transfer coefficient for the system – the k_La. It is the k_La term that is most affected by the shear rates in the vessel.

The solubility of oxygen in the slurry at saturation, C_{sat}, is set by the operating temperature of the leach, the hydrostatic pressure in the leach reactor and the leach chemistry. The oxidative leach is designed to maximise the solubility of oxygen in the slurry. The aspect ratio of the leach reactor is increased to provide a higher hydrostatic pressure at the base of the tank to improve the value of the saturated oxygen solubility. The optimum aspect ratio is typically in the range 1.2 – 1.5. Above this range, vessel construction costs tend to increase and outweigh the process benefits achieved with a deeper reactor.

In the Albion Process^{TM} neutral leach, the low dissolved salt background also enhances the saturated oxygen solubility. Total dissolved salt levels are of the order of 25 – 35 $g.l^{-1}$, compared to the 200 – 300 $g.l^{-1}$ typical of acidic leach solutions.

The balance between agitator power and oxygen sparger power is a complex, and is also one of the main testwork outcomes in an Albion Process^{TM} development program. The k_La value for the Albion Leach Reactor is measured during pilot testwork by measuring the rate of increase of oxygen content in the oxidative leach slurry under a specific set of conditions relating to the impeller and HyperSparge. The testwork considers both the final planned size and aspect ratio of the commercial vessel, as well as the HyperSparge supersonic oxygen injection system. Xstrata Technology has dedicated testwork facilities in Brisbane to collect this essential data prior to reactor design. A range of final slurry chemistries is normally evaluated to provide a robust system that can cope with variations in concentrate composition over the life of the project.

During the Albion Process^{TM} development program, a mass transfer correlation is developed for the leach system of the following form:

$$k_L a = A * [P_{total}/V]^{\alpha} * [U_s]^{\beta} . C^T$$

where:

k_La = the mass transfer coefficient for the reactor
V = vessel volume in m^3
U_s = Superficial gas velocity in $m.s^{-1}$
P_{total} = Supersonic sparger input power + impeller power input in $kW.m^{-3}$
C = Temperature constant for the system
T = operating temperature

The balance between the HyperSparge operating conditions and the agitator power and flow requirements is then designed based on this correlation.

The design basis for the Albion Leach Reactor used by Xstrata Technology has been proven to be sound, with above design mass transfer rates achieved with very difficult leach chemistry in both the Spanish and German Albion Process^{TM} plants, at oxygen capture efficiencies of up to 85 %.

Injection of oxygen into a near boiling slurry also presents several design challenges. The time taken to saturate a gas bubble within a near boiling system is less than 1 second[4], and so the injected gas rapidly expands in volume to equilibrate the water vapour pressure at the operating temperature of the reactor. This can lead to ventilated cavities within the tank at high shear zones, such as the impeller and sparger tips. The agitation system needs to be designed for these gas loads, and an impeller must be chosen that can maintain power draw under gassing. Deep downward pumping hollow blade radial flow impellers are preferred. Upward pumping hydrofoils may also be suitable under these gas loads, however this system has not yet been tested commercially. Dual impellers are typically employed due to the aspect ratio of the tank, and so careful consideration must be given to the relative impeller diameters to limit the development of zones of high void fraction within the reactor.

The use of the HyperSparge supersonic injection lance also helps considerably with the injection of oxygen into the near boiling slurry. Oxygen is presented as a coherent jet by the sparger, rather than as individual bubbles, and a significant degree of the oxygen mass transfer occurs prior to the jet resolving into a bubble cloud. This limits the final volume of gas presented to the agitation system.

The heat balance across the Albion Process^{TM} oxidative leach is set by the rate of oxidation of the sulphide minerals. Slurry density is generally allowed to vary down the train to maintain a relatively constant temperature profile across the oxidative leach circuit, and tanks will typically operate in the range 95 – 98 ^{o}C. No cooling is employed within the reactors, with temperature maintained entirely by the evaporation of water from the system through humidification of off gas. No cooling coils or heat exchangers are required.

The vessel temperature will approach boiling, however cannot reach the boiling point in a gas sparged system. Gas bubbles formed in the reactor will consist of a mixture of oxygen, carbon dioxide, nitrogen and steam. The partial pressure of the water vapour will be lower than the pressure at which water would boil due to the un-reacted gases present. A steady state equilibrium will be established in the Albion Leach Reactor at a temperature below the boiling point of water at the plant elevation. This is typically 2 – 3 degrees below boiling for a well insulated tank, or lower for only partially insulated tanks.

The relative number of HyperSparge emplacements in each Albion Leach Reactor is also varied to provide a balanced rate of heat release from each tank in the train. The evaporated water reports to the leach tank exhaust gas, and this is vented from the leach tank via a natural draft chimney. Albion Leach Reactors are covered with a sectioned panel lid to capture the exhaust gas and direct it to the natural draft chimney.

The Albion Process^{TM} oxidative leach circuit can be operated under a range of acidities, and materials of construction are tested during the batch testwork and pilot plant phases. Xstrata Technology has an extensive database of suitable materials for Albion Process^{TM} plant applications.

REFERENCES

[1] Bartsch et al, “Benefits of Using the Albion Process for a North Queensland Project, and a Case Study of Capital and Operating Cost Benefits Versus Bacterial Oxidation and Pressure Oxidation Randol Gold Conference, Perth, 2005
[2] Hourn et al, “Atmospheric mineral leaching”, Australian Patent No. 700850, 1996
[3] Hourn et al, “Method for treating precious metal minerals”, Australian Patent No. 744356, 1999
[4] Middleton et al, “Gas-Liquid Mixing in Turbulent Systems”, Handbook of Industrial Mixing, John Wiley and Sons, 2004.
[5] Smith, J. M et al, “Proc 2nd European Conf Mixing”, BHRA Fluid Engineering, Cranfield, 1977.

ISASMELT™

MOUNT ISA MINES NECESSITY DRIVING INNOVATION

*V. Lawson, H. DeWaal, G. Heferen, N. Aslin, P. Voigt, and M. Hourn

Glencore Technology
Level 10, 160 Ann St
Brisbane, Australia, 4000
*(*Corresponding author: virginia.lawson@glencore.com.au)*

ABSTRACT

Mount Isa Mines (MIM) acquired a reputation for the successful application of R&D to develop break-through technologies for the mining industry starting in the 1978's through until the early 2000's. The ISAPROCESS™ tank-house technology has been licensed to copper refineries throughout the world, and a significant per cent of the world's copper is refined using this technology. Since development in the late 1980's more than 20 ISASMELT™ copper and lead smelting furnaces are now installed in countries around the world. Jameson Cell flotation technology developed jointly by Mount Isa Mines and Professor Graeme Jameson is widely used in the Australian coal mining industry and increasingly in the base-metal and gold industry. The IsaMill™'s developed at Mount Isa and McArthur River made it possible to develop the McArthur River and George Fisher orebodies and has been successfully implemented into base metal fine grinding applications around the world. The most recent commercialised innovation is the atmospheric leach Albion Process™ with its supersonic HyperSparge™ gas sparger, is being adopted as a solution to the increasing complexity of orebodies.

MIM's contribution to the industry was significant given the size and the remote location of its operations with Townsville Copper Refineries more than 1350 km and Mount Isa 1800 km from the nearest state capital of Brisbane. This paper will briefly discuss the development of each of these technologies and why MIM – now owned by Glencore - was so successful innovating and developing such technologies over a period of nearly 40 years.

KEYWORDS

Innovation, Mount Isa Mines, ISAPROCESS™, IsaKidd™, ISASMELT™, IsaMill™, Jameson cell, Albion Process™, HyperSparge™, ZipaTank™

INTRODUCTION

Mount Isa is located in the Gulf Country region of Queensland about 1800 kilometers North West of Brisbane (see Figure 1). It came into existence because of the world class mineral deposits found in the area. In 1923 the orebody containing lead, zinc and silver was discovered by the miner John Campbell Miles. Mount Isa Mines Limited (MIM) was founded in 1924 to develop the minerals discovered by Miles, but production did not begin until May 1931. It paid its first dividend in 1947 after 16 years of troubled production. In 1954 the 1100 copper orebody was discovered and with rapidly rising reserves during the 1950's and 1960's led to the construction of new concentrators to treat lead/zinc/silver ores in 1966 (#2 concentrator) and copper ore's in 1973 (#4 concentrator). The difficult nature of the Mount Isa lead-zinc orebodies has meant that the company had always needed to be at the forefront of mining technology. In the 1970's through to the 1990's, it became a world leader in developing new mining techniques and processing technologies as a response to declining metal prices and rising costs. Mount Isa has been smelting copper since 1953 and lead since the early 1930's. Copper Refining at Mount Isa's fully owned subsidiary of Copper Refineries Proprietary Limited (CRL) had commenced operations in 1959.

Figure 1 – Location of Mount Isa and Townsville relative to Brisbane the nearest Capital City

Technologies to come out of Mount Isa include the ISAPROCESS™ copper refining technology, the ISASMELT™, The Jameson Cell, the IsaMill™, the Albion Process™ and the Hypersparge™. Mount Isa Mines Ltd was acquired by Xstrata in 2003 and Xstrata was then merged with Glencore in 2015. The level of innovation achieved at Mount Isa Mines is unsurpassed and was the result of the difficult nature of the Mount Isa ore bodies and its response to declining metal prices and rising operational costs in the 1970's and 1980's. By the 1990's, Mount Isa had become a world leader in innovative mining techniques and state of the art processing technologies. The processing technologies are discussed below.

INNOVATIONS

Each of the innovations developed at Mount Isa Mines had a driver but the overarching desire was to make technology more efficient and cost effective. Each of these process developments will be discussed separately.

ISAPROCESS^{TM}

The development of the ISAPROCESS^{TM} tank house technology had its beginning in the zinc industry. During the mid-1970s, MIM was considering building a zinc refinery in Townsville to treat the zinc concentrate produced by its Mount Isa operations. As a result, MIM staff visited the zinc smelters using the best-practice technology and found that modern electrolytic zinc smelters had adopted permanent cathode plate and mechanised stripping technology. MIM realised that the copper refineries performance was constrained by the conventional practice of copper starter sheets. The preparation of these copper starter sheets was labour intensive and the overall cycle was several weeks in duration.

MIM initiated a research program aimed at developing similar permanent cathode technology for copper refining. CRL, a subsidiary of MIM, had been operating in Townsville since 1959, using conventional starter-sheet technology and treating blister copper produced in the copper smelter at Mount Isa. Permanent cathode technology was developed and adapted over many years of in-plant experimental work and successfully introduced to the Townsville refinery in 1978. . The fundamental difference between the new ISAPROCESS^{TM} and the conventional starter sheet technology is the use of a permanent reusable cathode blank instead of a non-reuseable copper starter sheet and the introduction of mechanised and automated electrode handling machines replacing labour-intensive manual operations. The vertical edges had plastic strips and the bottom cased in wax to prevent copper cathode from growing around the edges of the cathode plate during, stripping and allowing two separate copper sheets from each cathode plate. This technology led to major advances in the electrode handling systems and automation in copper tank houses. The improved geometry of the cathode plates and the significantly shorter cathode cycle times allowed for increased intensity and efficiency of the refining process. Introduction of permanent cathode technology resulted in higher capacity, better copper cathode quality with less defects, safer operation and a four-fold improvement in productivity. Considerable development work was required to modify the original stripping machines from their zinc cathode origins due to the heavier cathodes. The stripping capacity of the machines has increased from 250 plates per hour to 600 plates per hour in the latest designs. More recent developments include the elimination of wax masking from the cathode plate, robotic electrode handling machines, and the introduction Duplex Stainless Steel cathode plates giving greater durability and corrosion resistance. Through the use of ISAPROCESS^{TM} user forums, to exchange ideas and developments in the technology and to share operational experiences, the technology has enjoyed continued improvement with higher productivity and improved quality at low cost.

Figure 2 – The IsaKidd process

In mid 1981 Falconbridge Limited commissioned a copper smelter near Timmins to treat concentrate from its Kidd Mine. The original copper cathode produced at Kidd suffered from the presence of higher concentrations of lead and selenium and could not meet customer specifications. It was determined that the use of copper starter sheets was preventing the Kidd refinery from meeting its cathode quality targets. Testwork began with the use of permanent stainless steel cathodes after preliminary tests showed a significant reduction in deleterious elements. The Kidd Process cathode used a solid copper header bar welded onto stainless steel resulting in a lower voltage drop than the ISAPROCESS™. Falconbridge began marketing the Kidd Process technology in 1992 providing competition between the two suppliers of permanent cathode technology. Between 1992 and 2006, 25 Kidd technology licenses were sold and 52 ISAPROCESS™ licenses.

The development of the ISAPROCESS™ and Kidd Process set the scene for a run of technology developments that continued until the mid 2000's. Xstrata took over MIM in 2003 and then Falconbridge in 2006. The Kidd Process technology consequently became part of the tank house package and together they have since been marketed as IsaKidd™ representing the dual heritage of the technology. The current robotic stripping machine (Figure 2) is based on over 30 years of copper refining and winning technology. Today over 100 licensees are using IsaKidd™ technology.

ISASMELT™

The sinter plant/blast furnace combination was the dominant technology for lead smelting throughout the 20th century. In the early 1970's companies using this technology came under sustained political and economic pressure as tighter environmental regulations were introduced, and energy costs increased, leading to higher capital and operating costs (Fewings 1988). It was in this environment that Mount Isa Mines sought a process that would improve the performance of the operations at their lead smelter in Mount Isa. After investigating the various processes under development, researchers turned their attention to the Sirosmelt lance. It had recently been developed on a laboratory scale at the Commonwealth Scientific and Industrial Research Organisation (CSIRO) in Melbourne. Following initial investigations Mount Isa Mines recognised the potential of the novel concept for smelting of lead concentrates and embarked on an extensive development program.

In 1978 a joint project was initiated between Mount Isa Mines and CSIRO to investigate the application of the Sirosmelt submerged-combustion technology to the smelting of Mount Isa lead concentrates. The ISASMELT™ process, as it became known, was developed to maturity for smelting copper, nickel, lead and zinc feeds by Mount Isa Mines through the 1980's and 1990's using incremental scale up. Commercialization only occurred once the process had been proven on laboratory, pilot and demonstration scale over many years. Approximately ten years were required for development of the lead and copper ISASMELT™ from crucible to demonstration scale (refer to Figure 4). During this decade the core know-how that was accumulated enabled the development team to reach the point where they were much better equipped to design and construct a full scale commercial plant – the final stage of the scale up process. Key aspects in this process were the selection of the scale up factors and the systematic design, development and re-engineering of several components of the technology. Figure 3 shows a comparison for the scale up stages for the lead and copper ISASMELT™ processes. Pilot scale was defined as unity for scale up comparison.

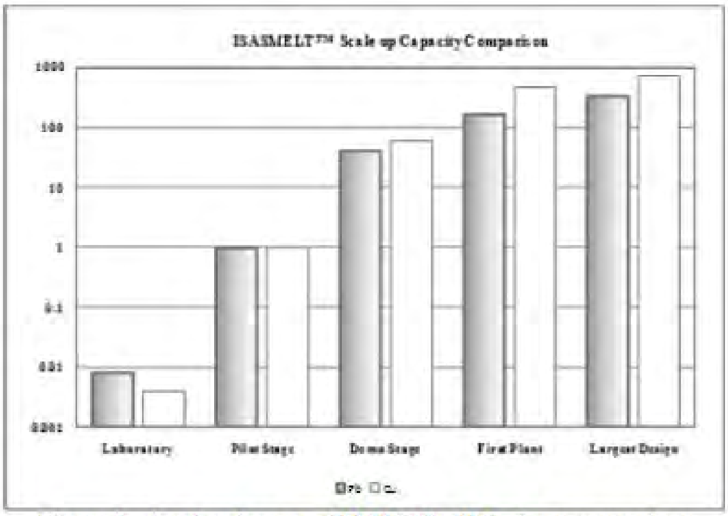


Figure 3 – Lead and copper ISASMELT™ Scale up comparison

During the scale up process, refer to Table 1, several aspects of the technology were developed to a high standard that allowed the ISASMELT™ technology to become a commercial success. As a result, ISASMELT™ technology now operates successfully at numerous plants around the world. The methodical approach to development of the technology has allowed owners to modernise their existing operations or create new businesses with significantly reduced technical risk.

An important parameter in the evolution of the ISASMELT™ technology has been the refractory campaign life. Figure 5 shows the history of the refractory campaigns at the commercial copper ISASMELT™ plant at Mount Isa since commissioning. At the time Mount Isa Mines management considered the installation of water cooling on the furnace refractories undesirable because of the potential for fatal incidents and increased operating costs. As a result the commercial scale furnaces were constructed with minimal water cooling. Although this led to shorter campaign lives initially, a development program was begun that focussed on optimising refractory materials selection and installation methodology. When coupled with process control strategies and continuous on-line monitoring of the bath temperature using systems developed over more than 10 years of operation, it allowed Mount Isa Mines to achieve campaign lives of more than 3 years without using any water cooling of the furnace refractories.

Figure 4 – Tapping matte from the copper ISASMELT at Kazzinc

Table 1 – Key Indicators of ISASMELT™ Plants from pilot to commercial scale

Topic	Unit	Pilot Scale		Demo Scale		First Full Scale		Current Design[1]	
		Pb	Cu	Pb	Cu	Pb[3]	Cu	Pb	Cu
Furnace ID	m	0.4	0.4	1.8	2.3	2.5	3.75	3.6	4.4
Lance Diameter	mm	38	38	150	250	250	350	250	500
Lance Control	-	Manual		Semi Automatic		Semi Automatic		Automatic	
Oxygen Enrichment	%	21	21	21	28	35	45	70	90
Nominal Feed Rate	tph	0.12	0.25	5	15	20	101	40	183
Offgas Treatment	-	Flue System / Baghouse		Gas cooler/ Baghouse		WHB		WHB[2]	

Notes:

ID: Internal Diameter; WHB: Waste Heat Boiler

[1] Refers to maximum throughput

[2] Some of the plants use a combination of radiation section and evaporative cooler for offgas treatment

[3] Refers to the smelting furnace from the two stage lead ISASMELT™ process

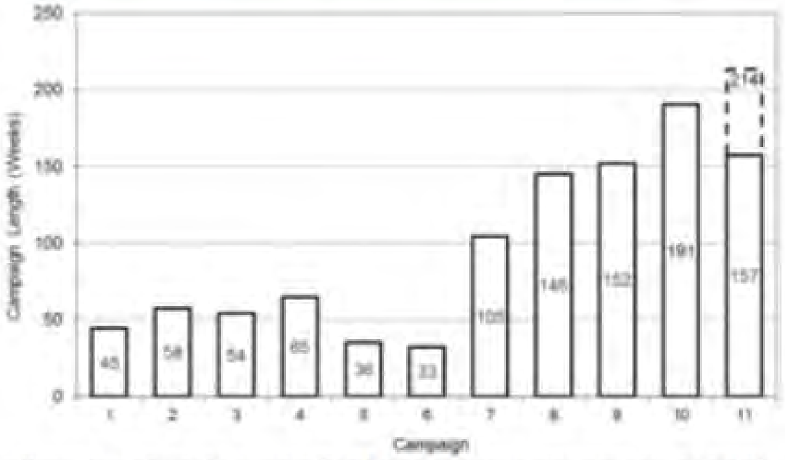


Figure 5 – Mount Isa copper ISASMELT™ plant campaigns (as of 2013)

Jameson Cell

The Jameson Cell (Figure 6) was jointly developed by Mount Isa Mines and Laureate Professor Graeme J Jameson (AO) of the University of Newcastle. Mt Isa had commenced operations with conventional flotation cells but was installing columns in cleaning duties in the mid 1980's. The columns had the benefit of froth washing that was likely to allow significant grade benefits in the very fine lead-zinc circuit. The first observations of the columns was that the collection process was slow necessitating long residence times and large volumes which remains a limitation of columns even today. In 1985 Professor Jameson was commissioned to undertake a project to improve the column sparger design.

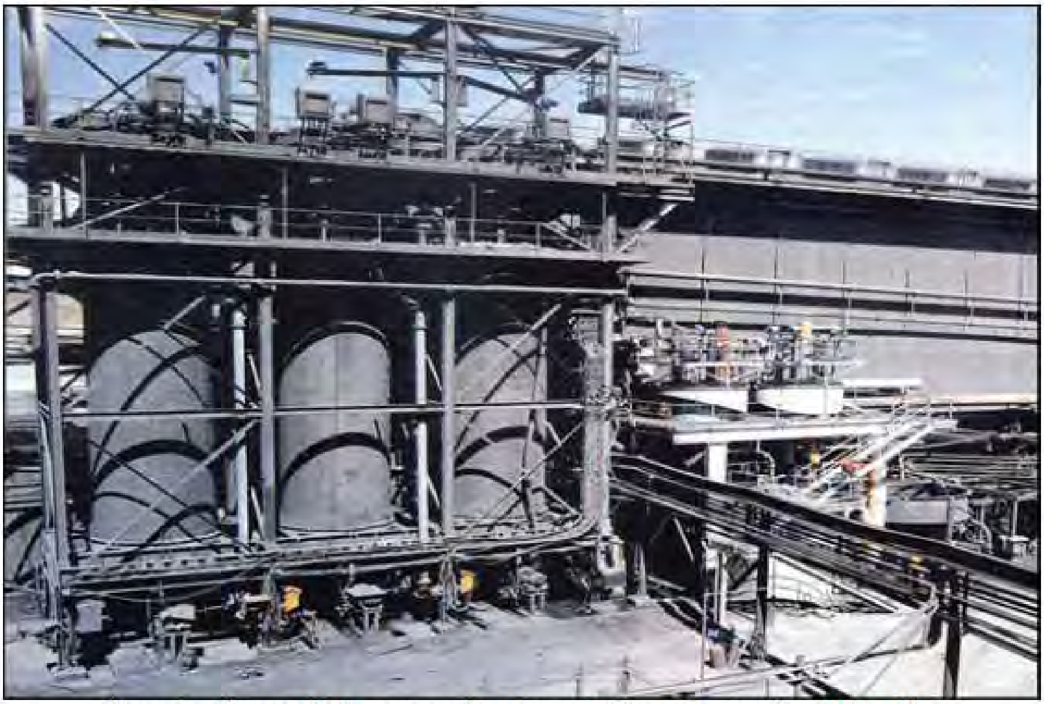


Figure 6 – Jameson Cells compared to columns of the same capacity at Mount Isa

Following initial work to provide an alternate method to bring together bubbles and particles, the downcomer was created. In the downcomer the air and the slurry are co-current with the air being entrained into the plunging jet under vacuum. Investigation showed that all of the bubble particle contact took place in the downcomer and thus the flotation tank could be much smaller. The first application at an industrial scale was in the lead zinc concentrator on the heavy media plant (HMP) lead slimes circuit. The initial improvement in performance were attributed to the very short residence time that allowed the minimisation of oxidation of galena fines. The cells were significantly smaller than the columns and there is no doubt the performance was superior as shown in Figure 3.

The testwork and trials in the early applications showed improved metallurgical performance when operated correctly. The challenge was operating them correctly. The technology hadn't been sufficiently developed to be successfully adopted into plant operations. The cell fell out of favour in base metals and in the 1990's was adopted into the Australia Coal industry and into niche SXEW applications where the main design challenges were resolved. The operability was improved by the introduction of a partial recycle to maintain constant flow and the maintainability of the cell was improved through various design modifications in operating plants. It was a period of continuous improvement. The result was a robust, low maintenance, easy to operate cell with the original features of excellent bubble particle contact.

The final obstacle was overcome when its adaption into the flowsheet was recognised to enable successful installations at the head of cleaner circuits and as low cost brownfield expansions. It is clear that the fast failures have had a significant effect on the success of the cell limiting its adoption into the industry. It is interesting that a significant proportion of sales are to return customers. Once you get over the hurdle of getting a Jameson Cell into your plant then seeing is believing. 2016 was the best year for Jameson cells into base metals and include the first sales back into South America where the cell had been abandoned after the difficulties of operations and maintenance of the Alumbrera installation. The metallurgical performance in Alumbrera was never the issue but the operators and maintainers hated the cells and they failed fast and hard.

The Jameson Cell celebrates its 30th birthday this year and has finally been adopted into mainstream base metals concentrators mainly as cleaner scalper at the head of the cleaner circuit. The cells generally recover up to 80% of the cleaner feed at high grades enabling much lower capital expenditure on the entire circuit. Process performance can be predicted from laboratory and pilot plant testing with demonstrated direct scale-up. It may have taken 30 years but the Jameson Cell is finally a success story. There are many lessons that can be learned from the implementation of innovation into industry from this case study.

IsaMill^{TM}

Unlike the developments of some of the other technologies at Mount Isa where efficiency was the main driver, the IsaMill was developed based on necessity. Figure 7 shows photomicrographs with the same scale of 40 micron demonstrating the increased complexity of Mount Isa ore over Broken Hill ore and the very difficult McArthur River ore. Although McArthur River was discovered in 1955 it was not able to be economically processed until the successful development of ultrafine grinding. McArthur River processing began in 1995 – 40 years after discovery when the IsaMill^{TM} made it technically and economically feasible to grind all of the rougher concentrate to 7 micron to facilitate the rejection of non-sulphide gangue. Even at 7 micron galena liberation is not possible and a bulk zinc-lead concentrate is produced.

Figure 7 – Photomicrograph of a) Broken Hill ore b) Mount Isa ore c) McArthur River ore

Investigations into fine-grinding started at Mount Isa started in the 1970s using conventional grinding technology to increase mineral liberation by grinding to fine sizes. These technologies were not only found to have high power consumption but also proved to be detrimental to flotation performance as a result of pulp chemistry and iron contamination from steel media. These poor results were revisited during pilot plant and tower mill testwork in the 1980s which also showed an inability of tower mills to economically achieve the required sizes. When it became clear that the solution to efficient fine-grinding did not exist in the minerals industry, MIM looked for ideas to "crossover" from other industries that also ground fine particles – pigments, pharmaceuticals, foodstuffs (e.g. chocolate). While these mills operated at a much lower scale and treated high value products they demonstrated the principle that stirring fine media at high speed was highly efficient. The challenge was transferring this concept to continuous, high tonnage and lower-value streams in the minerals industry.

In 1991 the introduction of a Netzsch laboratory stirred mill to the Mount Isa site was a turning point in fine-grinding and ultrafine grinding. The ½ litre bench scale mill resembled a milk shake maker and used fine copper smelter slag as grinding media. Testwork on McArthur River ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the Mount Isa pilot plant. The testwork showed that high speed, inert, horizontal mills could efficiently grind to 7 microns at laboratory scale providing major improvements in metallurgical performance. To make ultrafine grinding applicable to full-scale production a program of development was undertaken between Mount Isa Mines Limited and NETZSCH-Feinmahltechnik GmbH.

After 7 years of development and testing of prototypes in the Mount Isa operations, the IsaMill™ evolved. It was large scale, continuous, and most importantly robust because it was developed by operators. The crucial breakthrough was the perfection of the internal product separator – this allowed the mill to use cheap natural media (sand, smelter slag, ore particles) and to operate in open circuit. These are significant advantages for operating cost and circuit simplicity. Scale-up was tested using trial installations at the Hilton and Mount Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill™ (1.1MW) was installed in the Mount Isa concentrator. Improvements to the technology were continually made by the operators, maintainers and engineers working with the technology.

In 1998 the rights for commercialisation of the IsaMill™ where transferred from Mount Isa Mines Limited to MIM Process Technologies (now Glencore Technology) and under an exclusive agreement with Netzsch. In December 1998, the IsaMill™ technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. The IsaMill™ is now a mainstream fine grinding machine with over 130 installations around the world.

The Albion Process™

In the 1990's, MIM were studying options for the development of the large Frieda River/Nena project in PNG through its subsidiary Highlands Pacific. The Nena ores were not amenable to smelting, due to the elevated arsenic content, and several hydrometallurgical options were examined. Out of this work, MIM developed the Albion Process™, named after the suburb in Brisbane where MIM's development laboratory was located. The Albion Process™ is a combination of ultrafine grinding using Glencore Technology's IsaMill™, followed by oxidative leaching at atmospheric pressure in a series of reactors designed to achieve high oxygen mass transfer efficiency. The HyperSparge™ was also developed to deliver oxygen to the reactors efficiently.

Various small scale continuous pilot plant campaigns were conducted in 1994 and 1995. A larger pilot plant (120kg zinc cathode/day) was constructed in 1997 to conduct testwork as part of a feasibility study on the zinc/gold resources of Pueblo Viejo in the Dominican Republic. Extensive piloting was also conducted on lower grade chalcopyrite concentrates for Cyprus Amax in 1998, and for Mount Isa Mines in 2000. Pre-feasibility and feasibility pilot testing was conducted on the zinc/lead bulk concentrates from McArthur River and Mount Isa in Australia between 2001 and 2005. During this time the Albion Process™ was successfully tested on over 70 different ores and concentrates. The process is designed to recover gold and base metals from refractory ores. The key to the process is the ultrafine grinding stage followed by a hot oxidative leach at atmospheric pressure.

In the period from 1994 until 2004, the Albion Process™ (see Figure 8) was seen as strategic to the MIM/Xstrata group, and was not marketed externally. In 2005, a decision was made to offer the technology to external clients under licence, and a marketing agent – Core Resources, was appointed to market the technology globally. Interest in the technology has been very strong in the subsequent period, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Certej Project. The technology moved into commercial production in 2010 with the commissioning of Glencore's Albion Process™ plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Glencore of a second plant in Germany (16,000 tpa zinc metal). The Las Lagunas refractory gold project commissioned in 2012, and the GPM Gold refractory gold project commissioned in 2013.

Figure 8 – The Albion Process oxidative leach plant in Armenia

The major scale up risk with any oxidative leaching technology is oxygen mass transfer. High agitator power demands are common to achieve the shear rates in the vessel required for effective mass transfer at a commercial scale. A different approach was taken in the design of the Albion Leach Reactor to lower the agitator power demand. Glencore developed the HyperSparge supersonic gas injection lance to provide gas injection velocities of the order of 500 $m.s^{-1}$ within the leaching vessel, compared to the 4 – 8 $m.s^{-1}$ achieved with a typical agitator. Supersonic oxygen injection is a far more efficient method of generating shear than conventional agitation, allowing the total power input into the vessel to be significantly reduced, and greatly reducing the scale up risk for the oxidative leach.

The Albion Process^{TM} was enabled by the fine grinding of the IsaMill^{TM} and the process was designed to deliver a lower cost processing option for treating refractory mineral resources. There are now six operating Albion Process^{TM} plants and the process has now an extensive database of potential applications.

CONCLUSIONS

MIM developed a significant number of processing innovations that are technical and economic successes. The ability to innovate at MIM was enabled by very challenging orebodies and the need to process efficiently to remain economically viable. The success has been attributed to the development of these technologies on an operating site with the R&D group solving the technical issues on small scale. Each subsequent scale up was completed in the operating plants where the operators, maintainers, engineers and metallurgists were required to achieve production goals at each step of the scale up to ensure funding for the next step.

The number of innovations, at MIM, was disproportionate to the scale of operations and may have been enabled by the remoteness of the site and the researchers and operators working collaboratively to solve economic and technical problems. The research group were not capital city based but worked on the same site and were required to assist with installation, commissioning and operation of the various stages. This co-operation led to adoption into the plant and a fast feedback loop for improvements. The ultimate success of the innovations has been their widespread adoption into the mainstream industry where feedback from operating sites based on a user group model has enabled continuous improvement of each of the technologies.

ACKNOWLEDGEMENTS

The authors would like to acknowledge Glencore Copper and Glencore Technology for permission to publish and to all the research and production personnel who enabled and improved the

technologies in their plants. The success of these developments continues with the input from end users in the ongoing development.

REFERENCES

Armstrong, W. "The Isa Process and its contribution to electrolytic copper," paper presented at the Rautomead Conference, Scotland, August 1999.

Fewings J.H., Management of Innovation – The IsaSmelt Process. Presented at AMIRA "Innovation in Metal Production" Technical Meeting at Mount Isa, October 3-4, 1988.

Fountain C. Isasmelt and IsaMills – Models of successful R&D – AusIMM Young Leaders conference, 19-21 March 2002, Kalgoorlie, W.A.

Nihill D.N., Stewart C M. and Bowen P. the McArthur River Mine – The First Years of Operation. The AusIMM '98 Mining Cycle. Mount Isa 19-23 April 1998.

Pease, Joe (August 2005), "Complex leaching becomes much simpler" (PDF), Australian Mining, pp. 26–32, archived from the original (PDF) on 12 September 2009, retrieved 6 January 2010

Pease, Joe September 2016, "Crossing the innovation valley of death" Presented to JKMRC.

GDMB News inside

ERZMETALL
World of Metallurgy

70/2017 No. 1 · January/February
GDMB Verlag GmbH
ISSN 1613 - 2394 · www.GDMB.de

1/2017

Internationale Fachzeitschrift für Metallurgie
International Journal of Metallurgy

70 years

The Technology of Lead Production from Waste

Martina Laubertová, Jana Pirošková, Simona Dociová

In this study the pyrometallurgical and hydrometallurgical processing of lead recovery from waste was investigated. Lead is the softest of the technically important base heavy metals. It is easily rolled and moldable, so it is possible to produce wire rod and tubes as well as sheets and foils. Lead consumption is linked to the manufacture and use of lead-acid batteries and is estimated to account for 85 % of production (Figure 1). Other applications of lead include roofing materials and radiation protection. As regards the environmentally safe production and recycling of base materials, this metal has another positive property: it can be almost fully recycled as often as required without any loss of quality or value. The use of lead in certain applications is limited by several EU regulations. Other uses for lead, which include pigments and compounds, cable sheathing, ammunition and petroleum additives, have been in decline. The use of lead metal in rolled and extruded products such as lead sheets has been fairly stable over the years. The aim of this work is to give a short review of lead-containing waste streams and examine current and future possibilities of obtaining lead from primary and secondary raw materials in EU.

Keywords:

Lead – Recovery – Secondary lead – Technology

Die Technologie der Bleiproduktion aus Abfällen

In dieser Studie werden der pyrometallurgische und der hydrometallurgische Prozess der Rückgewinnung von Blei aus Abfällen untersucht. Blei ist das weicheste der technisch relevanten Schwermetalle. Es ist leicht formbar und kann sowohl als Kabel und Rohre als auch als Platten und Folien Verwendung finden. Die Nachfrage nach Blei ist an die daraus erzeugten Produkte gebunden und die Herstellung von Bleiakkumulatoren wird auf 85 % der gesamten Produktion geschätzt (Abbildung 1). Andere Anwendungen von Blei sind Dächerbau und Strahlungsschutz. Im Hinblick auf umweltfreundliche Produktion und Recyclingfähigkeit hat Blei eine andere positive Eigenschaft: Es kann fast vollständig und beliebig oft ohne Qualitätsverlust recycled werden. Der Gebrauch von Blei in bestimmten Anwendungen ist durch EU-Richtlinien begrenzt. Der Trend anderer Verwendungen von Blei, wie beispielsweise als Pigmente und Verbunde, Kabelummantelungen, Munition oder als Zusatz in Benzin, ist rückläufig. Die Verwendung von Blei in gewalzten und extrudierten Produkten wie Bleiplatten verlief die letzten Jahren stabil. Ziel dieser Arbeit ist es, einen Überblick über Abfallströme mit Bleigehalt zu geben und die aktuellen und zukünftigen Möglichkeiten der Rückgewinnung von Blei aus Primär- und Sekundärmaterialien in der EU zu erarbeiten.

Schlüsselwörter:

Blei – Rückgewinnung – Sekundärblei – Technologie

1 Introduction

Lead is a strategically important metal resource for the industrial development and global economy. Lead is usually found in ore with zinc, silver and copper and is extracted together with these metals. The main lead mineral is galena (PbS). Other common varieties include cerussite ($PbCO_3$) and anglesite ($PbSO_4$). Refined lead is derived from primary material in the form of lead or mixed metal ores and concentrates, and secondary material in the form of scrap and residue. The balance between primary and secondary production has shifted since 1998, and in 2011 secondary sources accounted for more than 77 % of EU production [1].

EU countries produced 343,900 t of lead concentrate, in 2014 the main producers being Poland (1.5 %), Sweden and Turkey. Total world mine production in 2014 was reported as 5.4 mill. t. World refined lead metal production in 2014

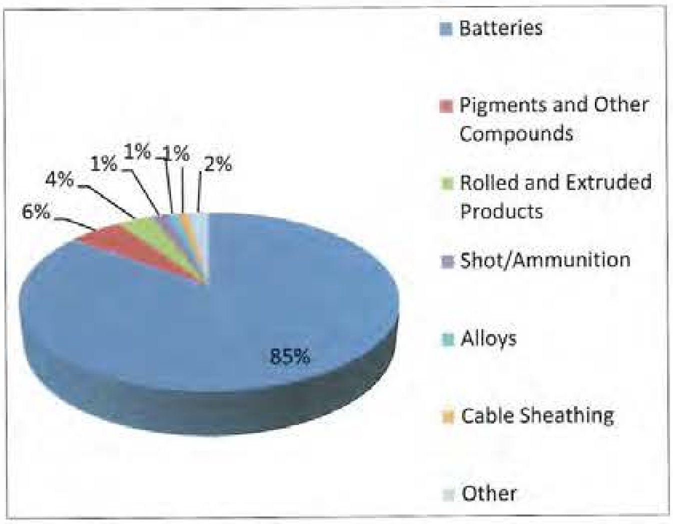


Fig. 1: Principal uses worldwide 2012

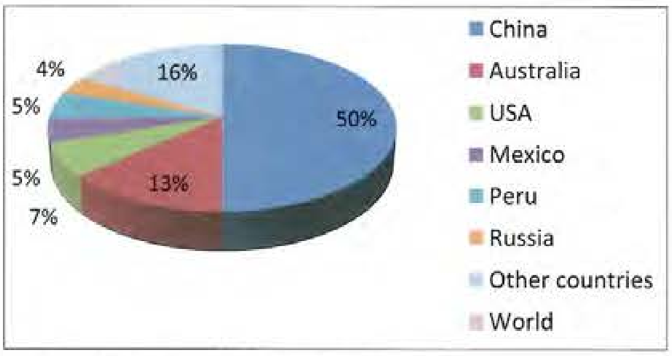


Fig. 2: World mine production of lead 2014

was reported as 10.6 mill. t. China was the world's largest producer of mined lead in 2014, with 50 % of the total world output. The second largest was Australia (13 %), followed by USA (7 %), Peru (4 %) and Mexico (4 %) (Figure 2). China was also the world's largest producer of refined lead in 2014, with 45 % of the total world Output. The second largest were the USA (11 %), followed by India (5 %) and Germany (4 %) (Figure 3) [2].

The secondary refining industry now supplies the majority of lead consumed. Lead-acid batteries are the main source of scrap for secondary refining. The secondary lead industry is characterised by a large number of smaller installations, many of which are independent. There are approximately 30 secondary smelters/refiners in the EU producing from 5000 t/a to 65,000 t/a. They recycle and refine scrap generated in their local area. The number of these refineries is decreasing as the large multinational companies, and the major battery manufacturing groups as well, acquire the smaller secondary facilities or set up their own recycling operations [3].

Umicore, Belgium produces a substantial amount of refined lead. Its capacity is 125,000 t/a. Though different grades were produced in the past, current production contains mostly 99.99 % lead [4].

Boliden, Sweden, produces lead from mines (62,000 t in 2015), lead smelters (26,000 t in 2015) and lead alloys (45,000 t in 2015) from recycled batteries in Bergsöe. Annual total production is around 70,000 t. At the Rönnskär smelter in Skelleftea, lead is extracted from concentrates from Boliden's own mines, as well as from external suppliers. The material contains minimum 99.97 % of lead and is characterized by its low content of silver and bismuth. The metal is cast into ingots weighing 42.5 kg each. At Boliden Bergsöe in Landskrona, lead is extracted and refined from scrapped batteries and other lead-containing recyclables. Bergsöe is one of Europe's largest recyclers of lead batteries. Its main products are lead and lead alloys. Approximately 60 % of the lead produced is sold to the battery industry in Europe, while the remainder is used in the manufacture of e.g. lead sheet and radiation shields [5, 6].

Lead production at Aurubis, Germany, produces refined lead in bars with a lead content of at least 99.985 % Pb. Aurubis produces 30,000 t/a of lead. The metal is employed in the production of rechargeable batteries and sheathing for deep-sea cables [7].

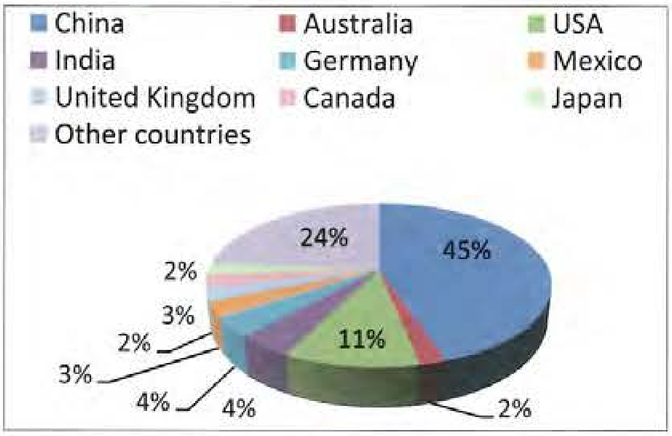


Fig. 3: World production of refined lead 2014

2 The technology of lead production

2.1 Primary lead production

There are two basic pyrometallurgical processes available for the production of lead from lead sulphide or from mixed lead and zinc sulphide concentrates: traditional sintering/smelting in a blast furnace or Imperial Smelting Furnace (ISF); and direct smelting.

Traditional two-stage process: The first stage is to remove the sulphur from the lead ore by roasting the ore in the air. The lead-bearing mineral is converted to lead oxide, and sulphur dioxide gas is released (Equation 1):

$$2PbS + 3O_2 \rightarrow 2PbO + 2SO_2 \qquad 400\text{-}900\ ^\circ C \qquad (1)$$

The second stage is to reduce the lead oxide to metallic lead using carbon (coke) as both the reducing agent and heat source. The lead-oxide rich sinter from the first stage is placed in a blast furnace along with coke and limestone or some other flux (such as silica or iron oxide) (Equation 2).

$$2PbO + C \rightarrow 2Pb + CO_2 \qquad 900\ ^\circ C \qquad (2)$$

A variation of the traditional blast furnace is the Imperial Smelting Process. This operates in a similar way, but allows lead and zinc to be removed simultaneously, the lead in liquid form, and the zinc distilled off as vapour.

Direct smelting: The sintering stage is not carried out separately. Several processes are used for the direct smelting of lead concentrates and some secondary material to produce crude lead and slag: Ausmelt/ISASMELT sometimes in combination with blast furnaces, Kaldo (TBRC), the QSL (bath furnace) integrated processes and electric furnace (used for processing primary lead concentrates together with secondary materials). The Kivcet integrated process is also used and is a flash smelting process.

Hydrometallurgical Processes (which include electrolytic processes) are an alternative approach in obtaining and purifying metallic lead. These offer the advantage that, unlike traditional smelting operations, harmful lead fume and sulphur-containing gases are not evolved (although in modern plants, pollution control systems can reduce emissions to low levels). Hydrometallurgical methods are generally regarded by the industry as a much cleaner approach than pyrometallurgy, and may become important in the future.

The principle for all such methods is that anodes of impure lead are dissolved into an electrolyte and pure lead is deposited on the cathode. At present, this approach is not economical for primary production, except possibly in rare cases where there is a very cheap source of electricity. Electrolytic methods are sometimes used to refine lead which contains relatively small amounts of impurities [8].

2.2 Secondary lead production

Secondary production – the production of lead from scrap, rather than ore is distinct from primary production. There are many plants which are dedicated to production of only secondary lead, while other plants are designed to produce primary lead. However, the processes involved are very similar, and a number of primary production plants now take some scrap as part of their feed material, i.e. produce partly secondary lead, a trend which is likely to increase. Secondary lead can be indistinguishable from primary lead provided it is subjected to sufficient refining steps.

2.2.1 Recovery of lead from lead-acid batteries

Composition of typical lead-acid battery scrap with the content of lead is: lead (alloy) components (grid, poles, etc.) 25 to 30 wt.-%, from this Pb 96 to 98 % and Sb 2 to 4 %, Ca 0.5 %; electrode paste ($PbSO_4$, PbO, PbO_2 and Pb) 35 to 45 wt-%, from this $PbSO_4$ 60 %, PbO (PbO_2) 20 to 30 %, Pb 1 to 2 %. There are two common processes:

Blast furnace recycling process

Only compounds of lead or very crude lead mixtures (such as pastes from batteries, or oxidised lead dust and dross obtained from other operations) need to be smelted. Smelting is not required for clean scrap lead. Smelting was traditionally performed in a blast furnace in a similar way to the extraction of primary lead. The furnace is charged with lead-rich feed, metallurgical coke and possibly hard rubber battery casing [1].

This procedure was developed by the German company VARTA that makes batteries. The procedure is based on the processing of whole batteries after removal of the sul-

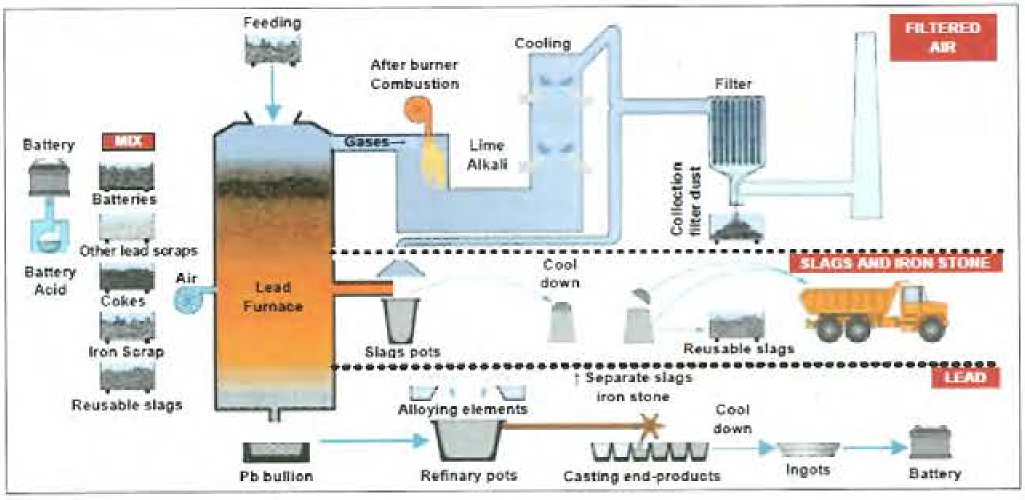


Fig. 4:
Typical shaft furnace battery recovery process [1]

Fig. 5:
A – bullion lead; B – refined lead; C – ingots: slag with matte; D – matte [10]

furic acid in the shaft furnace. This technology is used not only in Germany, Belgium, Russia, Sweden, but also in the Czech Republic. Significant producer of lead from waste in the Czech Republic applying the VARTA process is Kovohuty Příbram, Czech Republic (Figures 4 and 5) [9, 10].

Mechanical battery separation processes followed by smelting
Examples of mechanical separation processes are the MA and CX (Engitec) processes (Figure 6).

The recycling process is divided into the following stages, in which there are strict quality controls for ensuring the regular course of production and achieving maximum product quality: batteries crushing, Materials sorting; high quality polypropylene, Pb < 1500 ppm, materials sorting; heavy plastics (ebonite, PVC, ABS, etc.), materials sorting; high quality materials > 96 % (poles and grids); supply of materials for furnace (lead paste, plates/grids, anthracite, iron, sodium carbonate); casting (furnace) at 1000 °C; molten lead paste for processing/refining; casting (ingots); production of lead (99.99 %) and lead alloys. The desulphurization reaction can be carried out by addition of sodium carbonate, sodium hydroxide or ammonium bicarbonate (Equations 3 and 4).

$$H_2SO_4 + 2NaOH \rightarrow Na_2SO_4 + 2H_2O \quad (3)$$

$$PbSO_4 + 2NaOH \rightarrow Na_2SO_4 + PbO + H_2O \quad (4)$$

2.2.2 Recovery of lead from residues and from flue-dusts from copper smelting

An example for the mentioned pyrometallurgical copper production from secondary raw materials is the company Kovohuty, Krompachy, Slovakia. Copper is smelted in a shaft furnace, converting and fire refining is the next operation of copper anode production. The resulting anode copper is electrolytically refined to cathode copper in the company Montanwerke Brixlegg AG, Austria. As by-products slag and flue dust are generated. The content of individual elements of interest is shown in Table 1. Lead is expelled at a temperature of 900 to 1000 °C in the shaft furnace in the form of PbO and $PbSO_4$ and PbS particular. Only a very small part of Pb is expelled in metallic form.

The following techniques are applied to recover lead and tin in secondary copper smelters:

- Use of a second reduction stage, where lead and also tin are reduced with steel scrap or tin cans in a TBRC process. The tin is refined by crystallisation before the second vacuum distillation stage.
- Use of flue-dust from the secondary copper converter (sometimes also from the blast furnace) as a feed material. Under reducing conditions, zinc is volatilised and

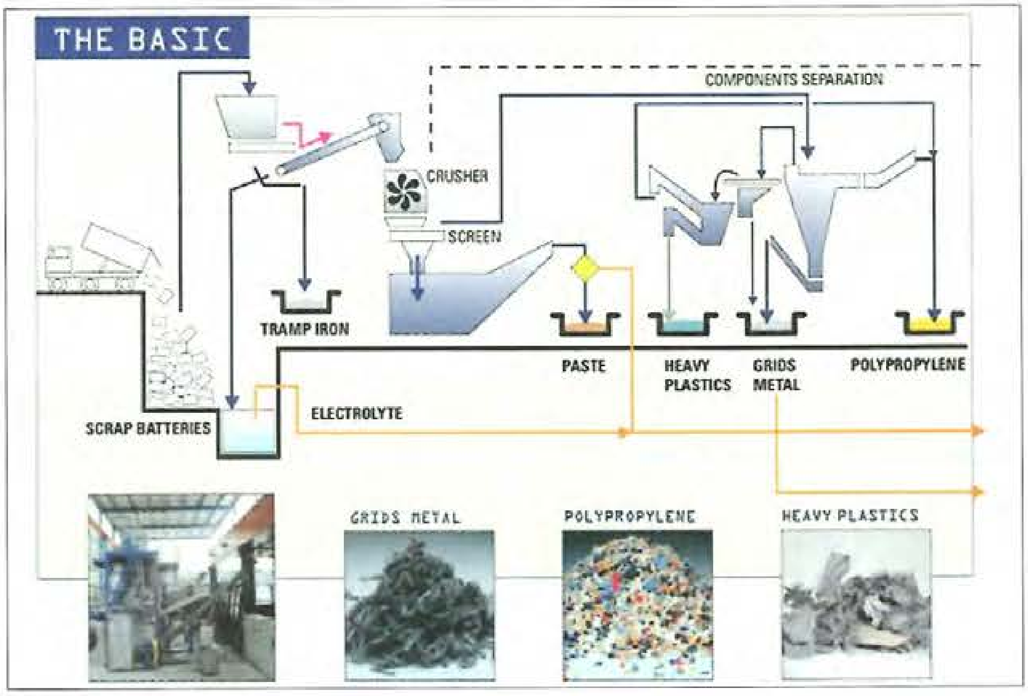


Fig. 6:
Engitec technology [11]

Table 1:
The chemical composition of by-products from secondary copper processing (Kovohuty, a.s., Krompachy, Slovakia) [wt.-%] [12]

Operation		Pb	Sn	Zn	Cu	Use
Smelting (shaft furnace)	slag	0.36-1.2	0.38-0.8	0.5-12 ZnO	0.7-1	Artificial stone for construction works
	dust	5-15	0.4-10.3	45-60	0.7-3	Production of zinc sulphate – closed at the moment
Oxidation process (converter)	slag	8-12	10-12	10-15	30-45	Recovered material to shaft furnace
	dust	4-30	2-20	45-55	0.4-1.5	As pellets to shaft furnace
Copper refining (anode furnace MAERZ)	slag	0.2	1-2	2-4	10.75	Recovered material to shaft furnace
	dust				4-20	As pellets to shaft furnace

recovered as oxide, while lead and tin are produced in the form of an alloy.

- Use of tin-lead alloy furnace technology in a multistage reduction process to produce black copper, tin-lead alloy and iron silicate slag from the KRS or TBRC converter slag.
- Use of a rotary rocking furnace to recover lead from the flue-dust from primary copper smelting.
- Use of electric furnaces for complex secondary materials such as those containing lead/copper, i.e. dusts, drosses, slags, slimes, copper alloy scrap, low-grade matte, lead/copper concentrates and other lead/copper bearing materials [1].

Multi-metal recycling with the Kayser Recycling System (KRS) is done at the Aurubis recycling center in Lünen (Figures 7 and 8). Pyrometallurgical preparation – smelting and refining – takes place in the KRS. The central operation is a submerged lance furnace (ISASMELT) which is almost 13 m high. A special feature is the use of a submerged combustion lance, which is immersed into the furnace from above and supplies the process with heating oil, oxygen and air. The reduction process is very fast in the submerged lance furnace. Charging times are short. The iron silicate sand extracted in that step of the process has a very low residual copper content. Copper, nickel, tin, lead and precious metals contained in the raw materials are enriched in an alloy with a copper content of about 80 %. In a top blown rotary converter (TBRC) the copper content is further enriched to 95 % and tin and lead are separated into a slag. The tin-lead slag is subsequently processed into a tin-lead alloy in the directly connected tin-lead furnace. During the KRS process, zinc is enriched in the KRS oxide, a flue dust.

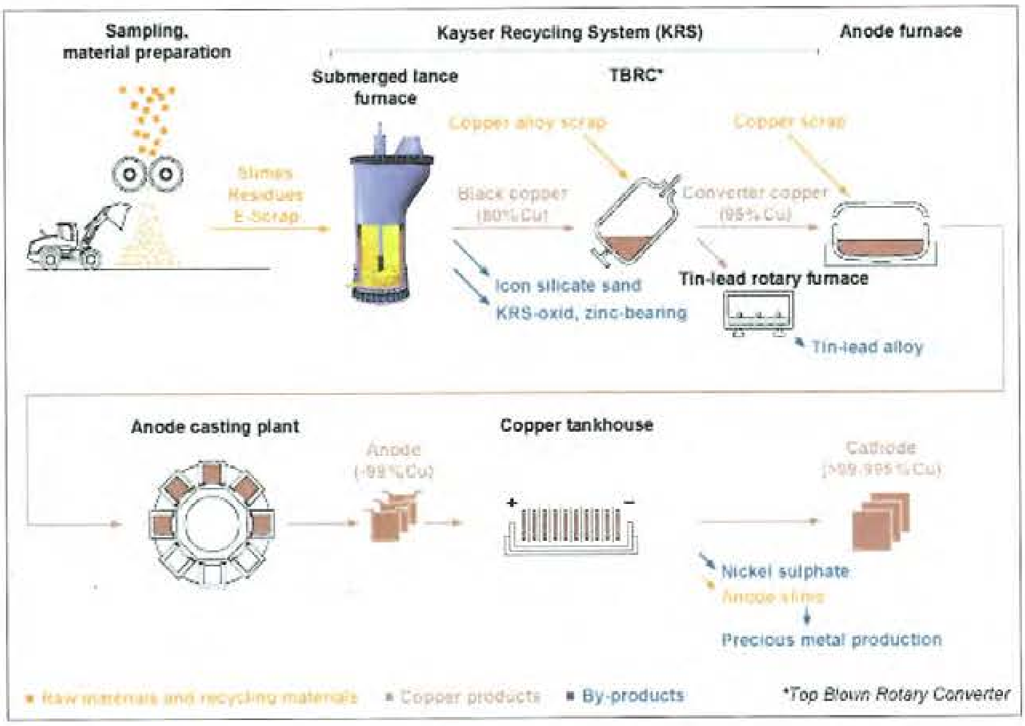


Fig. 7: Copper production at Aurubis, Lünen, Germany [14]

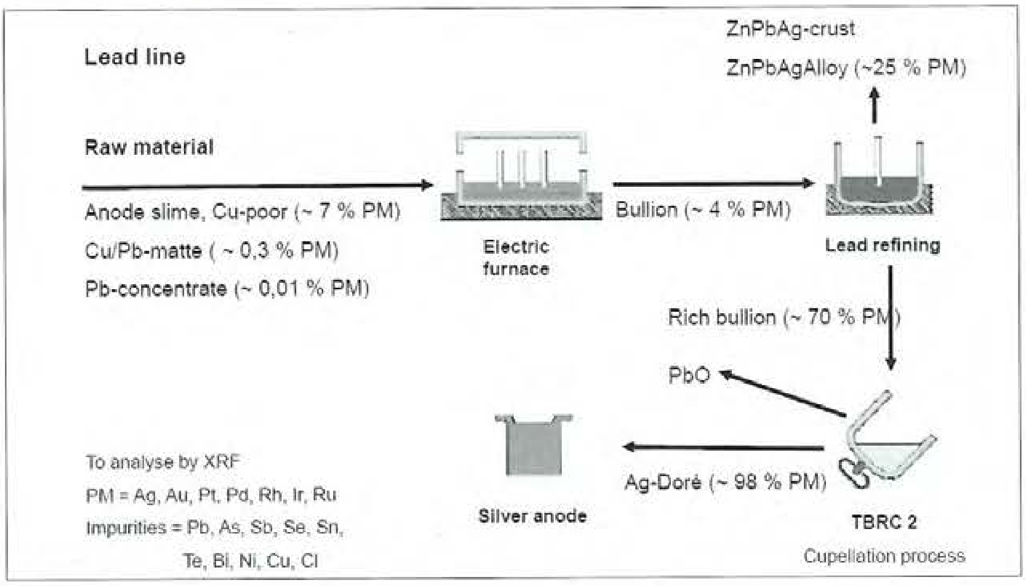


Fig. 8: Lead production at Aurubis, Lünen, Germany [15]

Copper scrap, electronic scrap and residues are used at Aurubis Hamburg. These secondary materials are processed in a modern electric furnace in various melting campaigns. The most important target is the pyrometallurgical separation of lead and copper and the enrichment of precious metals. By-elements still existing during copper production, such as lead, bismuth, antimony and tellurium, are separated in the connected lead refinery and sold as lead

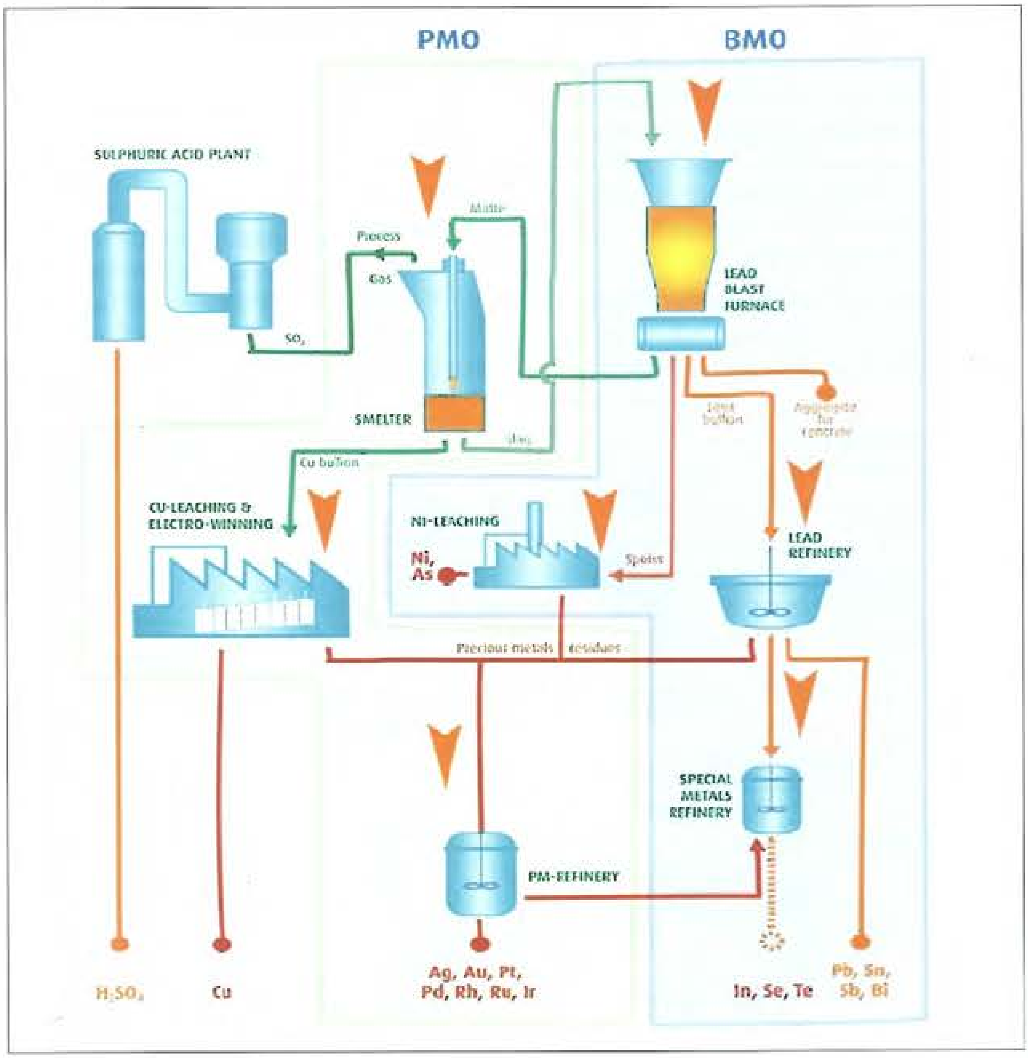


Fig. 9:
Umicore, Hoboken Belgium – Xstrata ISASMELT Technology, Lead Blast Technology [14]; Precious Metals Operations (PMO), Base Metals Operations (BMO)

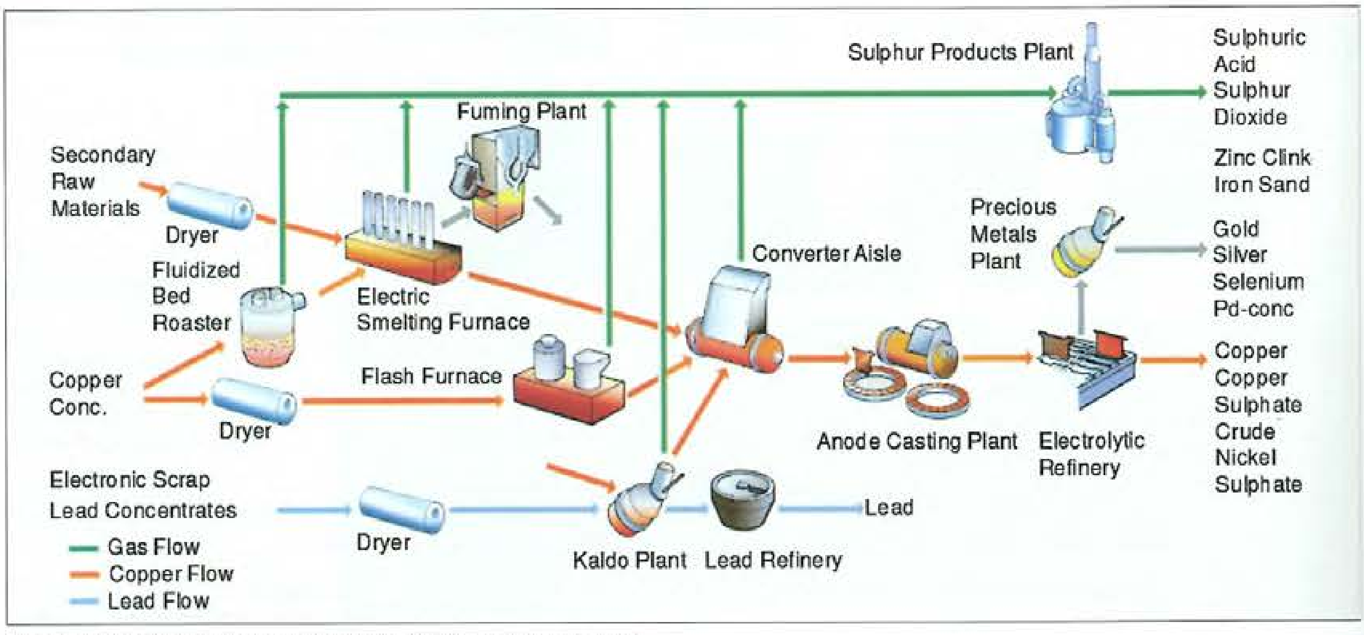


Fig. 10: Outotec Kaldo Technology at Boliden, Rönnskär, Sweden, [6]

bullion, lead-bismuth alloy, antimony concentrates and tellurium concentrates. The precious metals are fortified in a so-called rich lead, which has a precious metal content of about 70 % [13].

At Umicore there are two metals line processes: Precious Metals Operations (PMO) and Base Metals Operations (BMO) – a two-stage batch smelting and converting process. Copper matte remains in the furnace (ISASMELT) for Converting Stage. High lead slag from the furnace is a sub-product for the Base Metals Operation Plant (BMO) Converting Stage main products: converting slag remains in the ISASMELT furnace for the next cycle (Figure 9).

Boliden, Rönnskär, Sweden, – the Kaldo furnace is essentially a slightly leaning cylinder which rotates during the smelting process. The material is fed in and tapped out through the mouth of the furnace. There is no need to put any energy into the furnace: the plastic in the raw material provides sufficient energy for the smelting process (Figure 10).

2.2.3 Recovery of lead from steel scrap

One of the products from the manufacture of iron and steel is dust. The European Union currently produces mainly steel in basic oxygen furnaces (65 %) and in electric arc furnaces (35 %). The difference is mainly in zinc, as well as some heavy metals, if necessary, calcium and lead as well. Due to the high content of Fe in dust (over 30 %), this can be classified as secondary raw material. From the mineralogical study it shows that in the sample of interest most metals are present as franklinite $ZnFe_2O_4$, zincite ZnO, magnetite Fe_3O_4, calcite $CaCO_3$ and quartz SiO_2. Lead is present in phase PbO but also PbSO/$PbSO_4$ form.

The advantage of hydrometallurgical processes is their higher flexibility. Hydrometallurgical processes are also more economical because of lower capital and operating costs. In addition, hydrometallurgy offers the possibility of getting valuable metals from the dust or sludge. There are also environmental benefits in comparison with pyrometallurgy because of no problems associated with off-gases, dust nuisance and noise. However, hydrometallurgy does not offer a solution to the steelmaking waste processing by only one versatile way. The leaching methods are individual and depend not only on the type of the processed waste but on the physical, physico-chemical, chemical and mineralogical properties as well. Research in the field of hydrometallurgical processing is becoming more and more intensive. It is mainly because of the need to process complex raw materials, environmental aspects and also due to the legislative pressure for environmental protection. The content of individual elements of interest is shown in Table 2 [16-18].

Table 2: Chemical composition of dust [wt.-%]

	Fe	Zn	Pb	Ca
BOF (dust from converter)	47.7	2.74	0.18	6.8
EAF (dust from electric furnace)	29.66	18.96	1.94	4.42

For acid leaching a number of acids may be used e.g. acetic acid, sulfuric acid and others. Acid leaching is more effective than alkaline, but Fe, which dissolves causes problems for the subsequent processing of these solutions. Economically, it is appropriate to use sulfuric acid leaching. For practical reasons it is appropriate to infuse normal conditions, the lowest possible temperature and a low acid concentration, which is sufficient to dissolve the metal of interest. Preferred feature of acid leaching is the ability to use more diluted solutions. The disadvantage is that the solution passes through the portion of the iron. In the alkaline digestion are transitioning non-ferrous metals (Zn, Pb, Cd) in the solution. Iron does not pass to the solution and the solid phase remains in the leaching residue. The method for processing dusts requires the use of relatively concentrated solutions and may cause technical problems. In recent years, various industries have begun to successfully implement advanced technologies, which include technology using microwave radiation.

3 Conclusion

Lead production that is part of a complex metallurgical process to recover copper and precious metals remains unaffected. This paper described various waste containing lead and industrial waste generated in the different processes: recovery of lead from lead-acid batteries VARTA and CX (Engitec) technologies, recovery of lead from residues and from flue-dusts from copper smelting and also recovery of lead from steel scrap Xstrata ISASMELT and Outotec Kaldo Technologies. Some of these plants already meet most of the regulations in the EU Industrial Emissions Directive rules for the Best Available Techniques in the non-ferrous metals industry, which will become mandatory as of 2020.

Acknowledgements

This work was supported by the Ministry of Education of the Slovak Republic under grant VEGA 1/0442/17. This work was supported by the Slovak Research and Development Agency under the contract No. APVV-14-0591.

References

[1] Best Available Techniques (BAT) Reference (2014): Document for the Non-Ferrous Metals Industries Final Draft: 1242.

[2] Brown, T.J. et al. (2014): European Mineral Statistics. – British Geological Survey; 2010-2014: 378.

[3] International Lead Association (2012): http://www.ila-lead.org/lead-facts/lead-uses-statistics.

[4] Battery Recycling at Umicore (2016): http://www.umicore.com/en/industries/recycling/umicore-battery-recycling/.

[5] Boliden (2016): Metals for modern life. – http://www.boliden.com/Documents/Press/Publications/Metals%20for%20modern%20life/337-6514%20Metals%20for%20modern%20life%202016_EN.pdf: 32

[6] Borell, M. (2015): Rönnskär Smelter – A versatile and environmentally well adopted Non Ferrous Metal smelter. – http://www.georange.se/upl/files/111436.pdf: 41

[7] Aurubis (2016): Recycling at Aurubis. – https://www.aurubis.com/en/en/corp/products/page-recycling.

[8] Thornton, I., Rautiu, R. & Brush, S. (2001): Lead – the Facts. – http://www.ila-lead.org/UserFiles/File/factbook/chapter4.pdf: 192.

[9] Kunický, Z. & Vurm, K. (2014): 700 let stříbra a olova v Příbramsku, Kovohutě Příbram, 2011, ISBN 978-80-260-0451-6; Best Available Techniques (BAT) Reference Document for the Non-Ferrous Metals Industries, Final Draft: 1242.

[10] Trpčevská, J. & Laubertová, M. (2015): Metal waste and its treatment; ISBN 978-80-553-2365-7: 130.

[11] Lead-acid batteries complete recycling (2016); http://www.engitec.com/it/lead/.

[12] Kovohuty Krompachy, Slovakia (2014): Internal documents.

[13] Wood, J. et al. (2011): Secondary copper processing using Outotec Ausmelt TSL Technology. Conference: MetPlant 2011 Plant Design & Operating Strategies – World's Best Practice: 9.

[14] Modern, Flexible and Clean Technologies for Recovery and Recycling of Valuable Materials (2012). – 2nd International Workshop on Metals Recovery from Mining Wastes – M2R2: 37.

[15] Steffen, M. (2009): From Anode slime to silver-doré. – The LBMA Assaying & Refining, Seminar Aurubis AG, Hamburg, Germany.

[16] Havlík, T. et al. (2012): Acidic leaching of EAF steelmaking dust. – World of Metallurgy – ERZMETALL **65**, 1: 48-56.

[17] Havlík, T. et al. (2012): Leaching of zinc and iron from blast furnace dust in sulphuric acid solutions. – Metall **66**, 6: 267-270.

[18] Trpčevská, J. et al. (2010): Characterization of the bottom dross formed during batch hot-dip galvanizing and its refining. – Acta Metallurgica Slovaca. Roč. 2010, č. 3: 151-156.

Ing. Martina Laubertová, PhD.
Ing. Jana Pirošková, PhD.
Simona Dociová
All:
Technical University of Kosice
Faculty of Metallurgy
Institute of Recycling Technologies
Letna 9
042 00, Kosice
Slovakia
martina.laubertova@tuke.sk

ISASMELT™ - FUNDIR MÁS CON MENOS

A. Burrows[1], G.R.F. Alvear Flores[1], P.J. Mackey[2], Biplop Das[3], E. Herrera[4]

[1]Glencore Technology
Level 10th, 160 Ann Street, Brisbane QLD 4000, Australia
(alistair.burrows@glencore.com.au)

[2]P.J. Mackey Technology Inc., Montreal, QC, Canadá

[3]Sesa Sterlite Ltd., Tuticorin Smelter, Madurai Bypass Road, Tuticorin, Tamil Nadu, India

[4]Southern Copper Corporation
Fundición Ilo, Punta Tablones S/N, Pacocha, Ilo, Moquegua, Perú

RESUMEN

Desde que se construyó la primera planta comercial en Mount Isa, otros veinte hornos ISASMELT™ han entrado en operación, y continúa la construcción de nuevas plantas. La tecnología TLS de fundición de ISASMELT™ sigue atrayendo el interés de posibles usuarios que buscan fundir de manera productiva. Además de los beneficios conocidos en cuanto a productividad y flexibilidad, los usuarios de ISASMELT™ también han logrado mejoras en la eficiencia energética al cambiar sus tecnologías anteriores. Las recientes estadísticas de producción para la operación en Tuticorin, India, muestran que ISASMELT™ ha dado los primeros pasos para superar las 400.000 t / año (base de cátodos de cobre) y esta probablemente será la escala elegida por futuras nuevas fundiciones de cobre que operen exclusivamente como fundiciones a subcontratación. En estas fundiciones se esperará encontrar tecnologías altamente eficientes en cuanto a energía.

PALABRAS CLAVE

Fundición, Cobre, ISASMELT™, TSL, Eficiencia energética

INTRODUCCIÓN

El proceso ISASMELT™, basado en la lanza Sirosmelt inventado por CSIRO y comercializado por Mount Isa Mines (MIM) en las últimas tres décadas, estaba destinado a utilizarse para la fundición de concentrados de plomo. MIM pronto descubrió que esta también era una tecnología adecuada para la fundición de cobre (Arthur & Hunt, 2005; Burford, 2009). Como en el caso de MIM, otras fundiciones de cobre en busca de mejoras en sus procesos, han dado cuenta de las ventajas de sus bajos costos operativos y de su capacidad para producir gases residuales con un alto contenido de SO_2, de modo que el gas proveniente de la fundición pueda tratarse más económicamente en una planta de ácido.

La fundición de cobre Southern Copper Corporation en Ilo, Perú y la fundición de cobre de Sesa Sterlite en Tuticorin, India, tienen cada uno esquemas de fundición y capacidades de hornos de fundición primaria muy similares a la fundición de cobre de MIM. De las tres operaciones, el horno ISASMELT™ en Tuticorin ha producido más cobre por año en los últimos años en forma consistente, y es un ejercicio interesante para comparar el rendimiento específico de las diferentes fundiciones.

DESCRIPCIÓN DE TRES PLANTAS DE COBRE DE ISASMELT™

La tecnología ISASMELT™ es un proceso de fundición por baño en un recipiente vertical revestido con productos refractarios en el que se inserta una lanza de combustión sumergida de diseño especial en un baño de material fundido. El horno se alimenta continuamente con concentrado y fundentes de cobre; se inyecta aire enriquecido con oxígeno en el baño a través de la lanza, creando una agitación muy intensa y una velocidad de reacción rápida. El baño consiste principalmente en escoria fundida en hierro fundido de silicato y mata de cobre. Debido al estado de agitación del baño, la mata y la escoria se sangran juntas periódicamente a través de un único orificio de colada. La separación de fases se lleva a cabo en un horno de aguas abajo.

En Ilo, Perú, Tuticorin, India y Mount Isa, Australia existen fundiciones primarias de cobre que tratan concentrados mayormente de calcopirita, con capacidad de producción de entre 200 a 400 kt / año (en una base de cátodos de cobre). En combinación, estas tres fundiciones representan alrededor de 3,5 Mt / año de tratamiento de concentrado. Cada fundición utiliza un esquema que comprende un horno ISASMELT™ seguido de dos Hornos Rotatorios (RHF por sus siglas en inglés). En estas plantas, la mezcla de mata y escoria producida por el horno ISASMELT™ se lleva alternativamente a uno de los dos RHF por medio de bateas. En estas operaciones, se requiere que los RHF permitan la separación de fase de mata y de escoria, lo que permite que ambos estén limpios para verterlos por separado y que también contenga un depósito de mata que sirva como capacidad de reacción entre la operación continua del horno ISASMELT™ y los ciclos de lote de los convertidores Peirce-Smith aguas abajo. Con pequeñas variaciones en cada fundición, el esquema genérico que describe la sección de fundición de concentrado de las respectivas fundiciones en Ilo, Tuticorin y Mount Isa, está representado adecuadamente por el diagrama esquemático se muestra en la Figura 1.

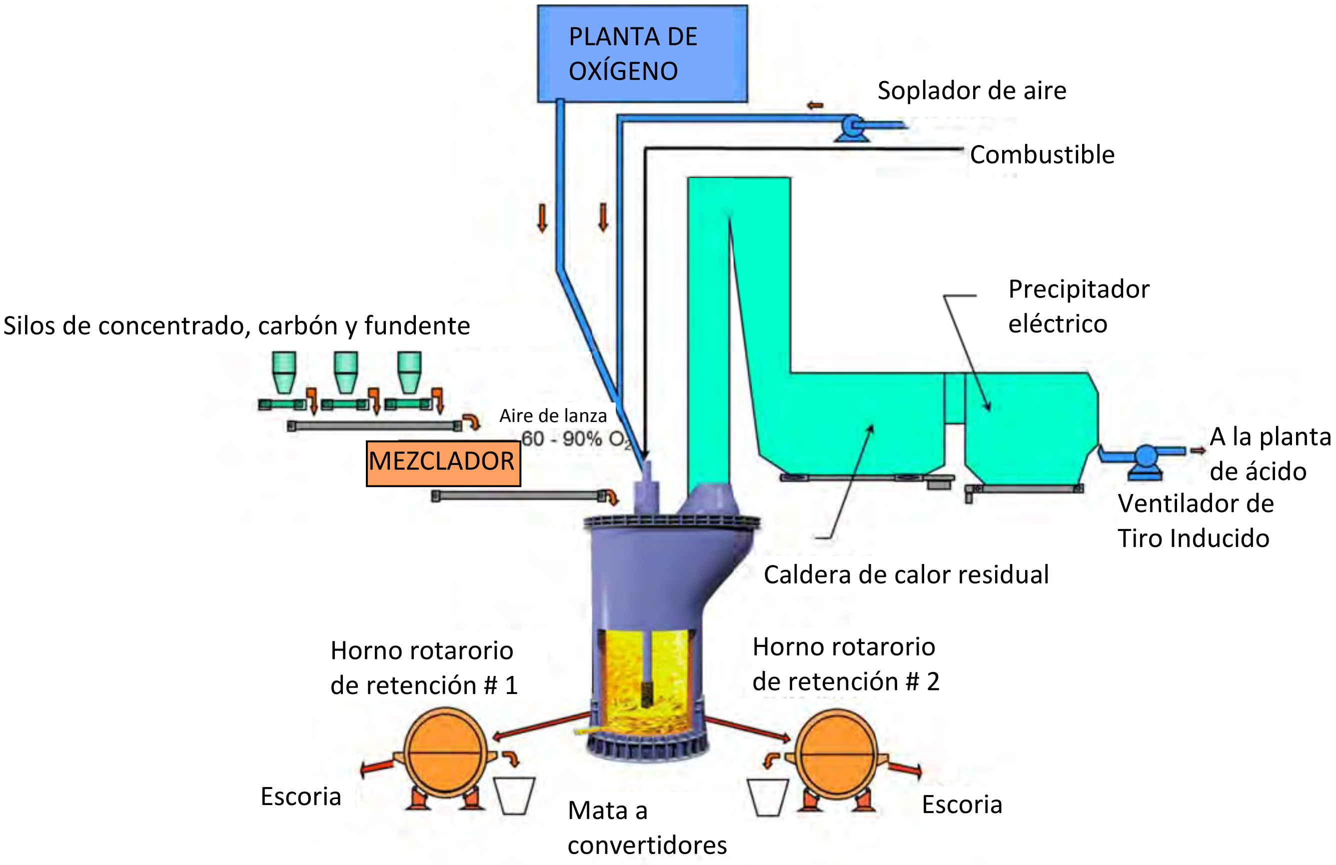


Figura 1 – Esquema genérico de ISASMELT™ como se usa en Mount Isa, Ilo y Tuticorin

Hay variaciones menores en la función y disposición de los elementos auxiliares en cada planta (tales como las calderas y los sistemas de manipulación de alimentación). Las principales diferencias entre cada planta son las siguientes:

- Mount Isa utiliza una caldera de recuperación de flujo Flux, mientras Tuticorin e Ilo utilizan calderas convencionales divididas en secciones de radiación y convección.
- Mount Isa utiliza un granulador de disco para la mezcla de alimentación, mientras que Tuticorin e Ilo utilizan mezcladores de aspa.
- Tuticorin es una fundición a medida, con una variedad de tipos de alimento, mientras que Mount Isa e Ilo son fundiciones internas donde cada uno procesa concentrados de dos minas locales.
- Tuticorin tiende a fundir utilizando un aire enriquecido por lanza, que contiene aproximadamente 85% de O_2, mientras que Mount Isa e Ilo tienden a utilizar alrededor del 65% de O_2 en sus respectivas lanzas.
- Tuticorin enfría su escoria a través de granulación, mientras que Mount Isa e Ilo utilizan Camiones Kress y cucharones para llevar la escoria fundida lejos de la fundición, y dejarla enfriar en represas de tierra.
- Mount Isa muele su escoria para posteriormente recuperar el cobre, lo que no se hace actualmente en Ilo ni en Tuticorin

En otros aspectos, las plantas tienen capacidades similares con estadísticas también similares.

En el plano, el horno ISASMELT™ en cada planta se instaló cerca del pasillo del convertidor, de manera que los RHF puedan descargar la mata en una cuchara a poca distancia de la grúa del pasillo, para su traslado a los convertidores Peirce-Smith. Es una característica de los RHF que la mata para convertidores se pueda mantener caliente por un breve tiempo de transferencia entre el horno y el convertidor; se puede llenar un cucharón con 25 a 40 tonelada en alrededor de un minuto, y no se requieren operadores para la operación de roscado.

En elevación, la disposición genérica del horno ISASMELT™ y de los RHF en Ilo, Mount Isa y Tuticorin se muestra en la figura 2. El horno ISASMELT™ está dispuesto de manera que la transferencia de la batea a los RHF aguas abajo sea fácil y eficiente.

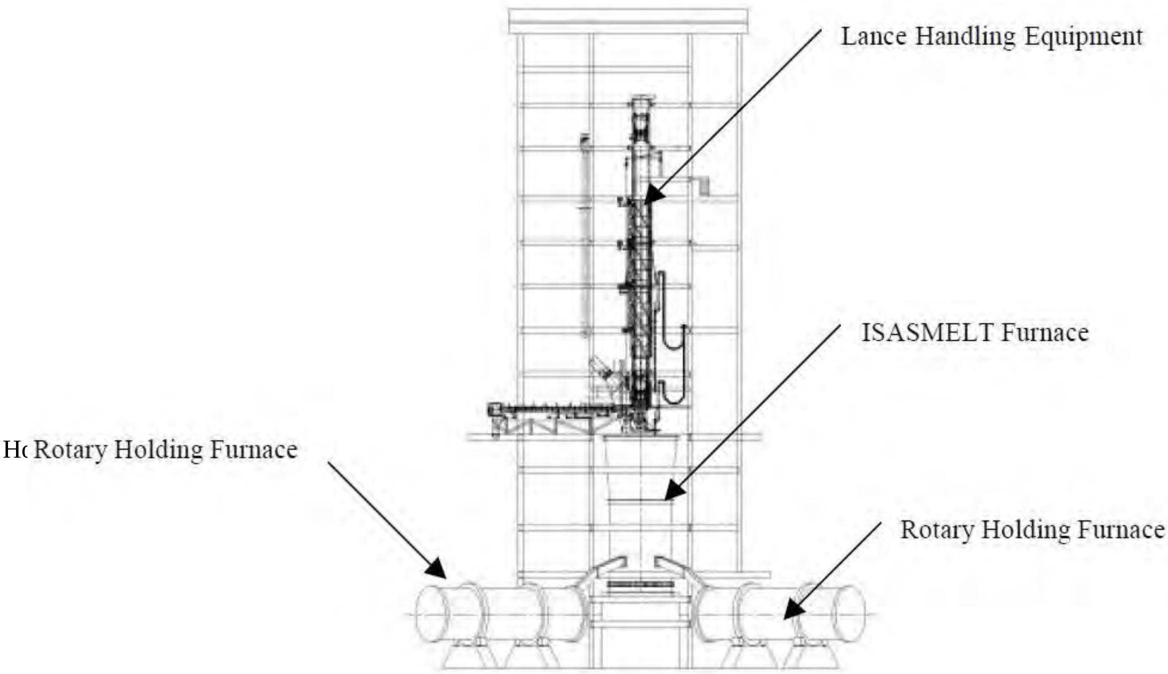


Figura 2 – Elevación genérica de la Planta ISASMELT™, fuente (Arthur & Partington, 2007)

COMPARACIÓN DE LAS TRES OPERACIONES DE COBRE DE ISASMELT™

La diferencia en la escala de producción y las peculiaridades de las materias primas puede complicar la tarea de hacer comparaciones significativas entre las operaciones industriales. En un análisis detallado de este tema, Goonan (2005) utilizó un esquema genérico de fundición de cobre y flujos normalizados de materiales y energía para poder comparar diferentes fundiciones de cobre. Se ha adoptado este enfoque, con la excepción de que los límites considerados en este trabajo sólo se extienden hasta la etapa de fundición primaria y de limpieza de escoria de la fundición y no a las siguientes secciones de conversión y refinamiento de fuego. Este método de comparación ya ha sido presentado previamente (Burrows, Partington y Mascrenhas, 2011), y se utilizará a continuación.

En la Figura 3, se muestra un esquema genérico de fundición primaria. El área de fundición primaria está encerrada dentro de la línea roja segmentada. El material y las corrientes de energía sujetos a comparación están etiquetados individualmente. Todas las mediciones de flujo de masa están en unidades de toneladas por tonelada de cobre producido. Las mediciones de electricidad están en MWh por tonelada de cobre producido. No todas las corrientes son aplicables en cualquier fundición. Por ejemplo, la corriente J se define como cero para un horno de limpieza de escoria, pero sería un valor distinto de cero si la limpieza de escoria se realizara utilizando un proceso de flotación.

Por razones de consistencia, las cifras de electricidad citadas en este trabajo:

- Incluyen la generación de oxígeno industrial;
- Incluyen la compresión del aire por el ventilador de proceso de fundición primaria;
- Incluyen el o los ventiladores de gas de escape de la fundición primaria;
- Incluye a los consumidores menores en el área de fundición primaria; pero,
- Excluye cualquier beneficio de compensación a partir de la electricidad generada utilizando vapor de agua desde el sistema de recuperación de calor residual.

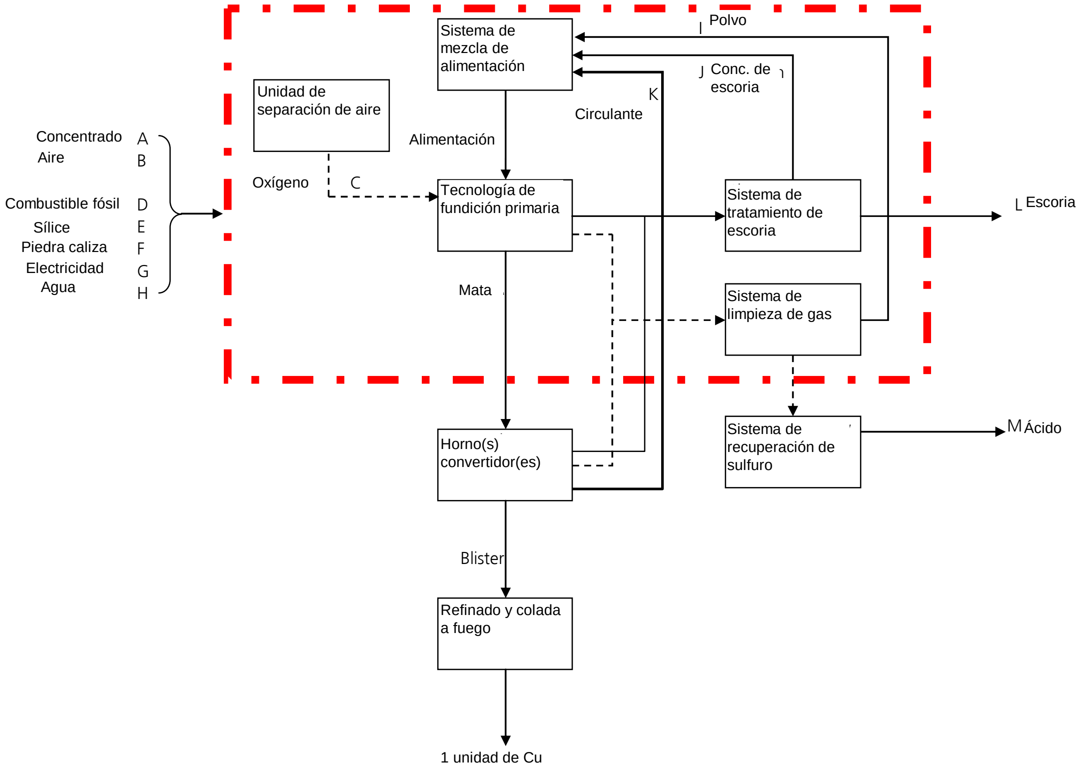


Figura 3 – Diagrama simplificado de la fundición de cobre.

Las historias de las operaciones ISASMELT™ en Mount Isa e Ilo son similares en algunos aspectos. Ambos tenían hornos de fundición con tecnologías obsoletas que fueron reemplazados cuando se instalaron sus plantas ISASMELT™ respectivas. La fundición Tuticorin se diferencia de estas dos operaciones en que la planta ISASMELT™ no fue construida para reemplazar una tecnología obsoleta.

Los datos correspondientes a las tres fundiciones se muestran en la Tabla 1. En el caso de Mount Isa e Ilo, esta tabla incluye los datos de las tecnologías anteriores que se reemplazaron con la instalación de las plantas ISASMELT™. Estos datos permiten una comparación del rendimiento de las diferentes fundiciones y, en el caso de Mount Isa e Ilo, también permiten comparar el desempeño de sus plantas ISASMELT™ y las tecnologías que precedieron a su instalación.

Comparación con las tecnologías anteriores

En comparación con las tecnologías anteriores en Mount Isa e Ilo, está claro que la instalación del esquema ISASMELT™ + RHF ha proporcionado los siguientes beneficios:

- Reducción de la generación de escoria

- Gran reducción del consumo de combustibles fósiles
- Gran reducción del consumo de aire de proceso
- Mucha mayor captura de gases ricos en SO_2 y su explotación en una planta de ácido.
- En contraposición a estos beneficios, también ocurrieron los siguientes cambios:
- Mayor consumo de electricidad por tonelada de cobre, debido a la intensificación del proceso de fundición por la producción y el consumo de oxígeno en el sitio.

Tabla 1 – Parámetros de evaluación comparativa para las fundiciones de cobre que utilizan la tecnología ISASMELT™ + RHF

		Sesa Sterlite Tuticorin India	Southern Copper Corporation Ilo Peru		Glencore Mount Isa Australia	
Tecnologías de fundición primaria		ISASMELT + 2 X RHF	Hornos reverberos + CMT	ISASMELT + 2 X RHF	Hornos reverberos + FSR	ISASMELT + 2 X RHF
Período		FY 2009	CY 2006	2007-2008	FY 1991-92	CY 2008
Concentrado	A	3,39	3,38	3,36	4,52	4,54
Aire	B	0,71Ψ	5,25	0,83	4,35	0,82
Oxígeno	C	0,96	0,26	1,08	0,00	1,00
Combustible fósil	D	0,03	0,33	0,05	0,57	0,09
Sílice	E	0,38	0,29	0,45	0,06	0,18
Caliza	F	0,01	0,07	0,12	0,27	0,01
Electricidad	G	0,70	0,22	0,51	0,13	0,67 *
Agua	H	2,70	0,32	0,09	0,004	0,001
Polvo	I	0,03	0,11	0,05	‡	0,09
Concentrado de escoria	J	0,00	0,00	0,00	‡	0,16
Material Circulate	K	0,15	0,70	0,35	‡	0,20
Escoria	L	2,02	2,62	2,38	3,29	2,59
Ácido Sulfúrico	M	3,03	1,15	3,29	0	3,09

Notas: Corrientes definidas en la Figura 3. Las unidades de potencia son MWh / t de Cu. Las unidades de masa son t / t Cu. * = Incluye la electricidad para la molienda de escoria
Abreviaturas: FSR = Asador de Fluo Sólidos, CMT = Convertidor Teniente, CY - Año Calendario, FY = Año Financiero, ‡ - no disponible, X = incluye aire comprimido para equipos auxiliares

Comparación entre las operaciones de fundición de ISASMELT™ + RHF

Las cuatro características que distinguen a las estadísticas de Mount Isa de las de Ilo y Tuticorin son:

- Menor ley de concentrado (lo que resulta en más toneladas de concentrado requeridas para producir una tonelada de cobre)
- Menor consumo de flujo de sílice por tonelada de cobre
- Menor consumo de agua fresca por tonelada de cobre
- Mayor generación de polvo por tonelada de cobre producido

Algunas de estas características están relacionadas. Por ejemplo, el concentrador adyacente tiene un sistema para añadir relaves ricos en sílice al concentrado de Mount Isa antes de la entrega, lo que resulta en que el concentrado se funde parcialmente y es de baja ley de cobre. El menor consumo de agua se puede explicar por el hecho de que Mount Isa recibe su concentrado desde la mina en forma de lechada y desagua el propio concentrado. Esto se traduce en que se pueda utilizar agua de proceso reciclada y en una reducción concomitante en el consumo de agua fresca. La mayor generación de polvo por tonelada de cobre en Mount Isa se debe a una combinación de:

- El diseño único de la caldera de flujo Flux (que requiere la adición de partículas de piedra caliza, lo que contribuye a la generación de polvo)
- La mayor cantidad de concentrado que se debe procesar para producir una tonelada de cobre

En el caso de Tuticorin, hay dos rasgos distintivos de las estadísticas operativas que difieren de Mount Isa e Ilo. Estos son:

- Mucho mayor consumo de agua
- Menor consumo de combustible fósil por tonelada de cobre

El mayor consumo de agua en Tuticorin se atribuye principalmente a la utilización de un molino de escoria (en comparación con la eliminación de la escoria caliente en Mount Isa e Ilo), y en parte debido a la extracción de vapor de la caldera de recuperación para el consumo de plantas auxiliares. El menor consumo de combustibles fósiles en Tuticorin se debe mucho a la alta disponibilidad de su planta, a menudo superior al 95%, lo que conduce a un menor uso de quemadores que de otro modo serían necesarios para mantener un horno inactivo a temperatura de servicio listo para reanudar la fundición. El enriquecimiento superior del oxígeno utilizado durante la fundición también ayuda a reducir el consumo de combustible.

Perspectivas de futuro para las tres fundiciones de cobre

Mount Isa

Glencore plc está evaluando opciones que pueden resultar en que la fundición de cobre de Mount Isa opere después de la fecha de cierre indicada para diciembre de 2016. Esto podría añadir 4 a 5 años de operación para la fundición que procesa sus propios concentrados, junto con materiales regionales de terceros.

Ilo

La planta ISASMELT™ en Ilo sigue siendo un componente vital de los futuros planes de producción de cobre de Southern Copper Corporation. Su funcionamiento continuado está asegurado para un futuro previsible.

Tuticorin

La planta ISASMELT™ en Tuticorin es parte de la única fundición que suministra cobre a las refinerías electrolíticas, tanto en Silvassa como en Tuticorin. Desde que comenzó a funcionar en mayo de 2005, la planta ISASMELT™ en Tuticorin ha alcanzado o superado su capacidad de diseño original, lo que permite que Sesa Sterlite produzca más de 300.000 t / año de cátodos de cobre de forma consistente. Las cifras de producción se muestran en la Tabla 2.

Tabla 2 – Cifras de producción de la Fundición de Tuticorin (base de cátodos de cobre)

	Producción de Cátodo (kt/año)
FY2006	273
FY2007	313
FY2008	339
FY2009	313
FY2010	334
FY2011	304
FY2012	326
FY2013	353

La progresiva mejora en las cifras de producción se debe a la labor incremental del personal de producción y mantenimiento en Tuticorin que ha realizad mejoras y modificaciones por las cuales las plantas y los servicios auxiliares han dejado de ser un cuello de botella. En el segundo trimestre de 2015, la planta alcanzó una tasa anualizada de 400.000 t / año (cátodo de cobre) por primera vez. El futuro de la planta de Tuticorin ISASMELT™ se ve muy brillante, y sigue estudiando activamente duplicar la fundición. Las futuras plantas ISASMELT™ también tendrán que cumplir y superar los 400 kt / año para mantenerse en la vanguardia tecnológica líder en operaciones de fundición de cobre.

CONCLUSIONES

La tecnología superior ISASMELT™ de fundición de cobre (TSL) sigue atrayendo el interés de posibles usuarios que buscan fundir de manera productiva. Además de los beneficios conocidos en cuanto a productividad y flexibilidad, los operadores ISASMELT™ también han logrado mejoras en la eficiencia energética al cambiar sus tecnologías anteriores. La fundición de cobre de Ilo en Perú y la de Mount Isa en Australia experimentaron reducciones sustanciales en el consumo de combustibles fósiles asociados con la introducción de sus respectivas plantas ISASMELT™

Las recientes estadísticas de producción para la operación en Tuticorin, India, muestran que ISASMELT™ ha dado los primeros pasos en la escala superior a 400.000 t / año (base de cátodos de cobre). Con el beneficio de una mayor capacidad de producción, es probable que mejore aún más su eficiencia energética.

RECONOCIMIENTOS

Los autores desean reconocer a Sesa Sterlite Ltd, Southern Copper Corporation y Glencore plc por facilitar los datos para su publicación.

Los autores desean agradecer a Glencore Technology Pty Ltd., Sesa Sterlite Ltd y Southern Copper Corporation por su autorización para publicar este artículo.

REFERENCIAS

P.S. Arthur and S.P. Hunt (2005), “ISASMELT™ - 25 Years of Continuous Evolution”, Floyd International Symposium on Sustainable Development in Metals Processing, M. Nilmani and W.J. Rankin, Eds., NCS Associates (Australia), 2005, pp 73-94.

P.S. Arthur and P.J. Partington (2007), “Latest Developments with Copper ISASMELT™, Proceedings of Cu 2007, The Carlos Diaz Symposium on Pyrometallurgy, vol. III, pp3-15, Eds. A.E.M. Warner, C.J. Newman, A. Vahed, D.B. George, P.J. Mackey, and A. Warczog, Toronto, TMS, 2007

B. Burford (2009), “The ISASMELT™ Technology Package: Over 30 Years of Innovation” The AusIMM Bulletin, Journal of the Australasian Institute of Mining and Metallurgy, No 1, February 2009, pp.26-30.

A.S. Burrows, P.J. Partington & P. H. Mascrenhas (2011), “ISASMELT™ at Mufulira – Increased Flexibility on the Zambian Copperbelt”, Proceedings of the Fray International Symposium, Ed. F. Kongoli, vol 1, pp217-226, TMS. 2011

T.G. Goonan, (2005), “Flows of Selected Materials Associated with World Copper Smelting”, U.S.G.S. Open-File Report 2004-1395, 2005

ISASMELT™-based Reconstruction of a Lead Plant

Dr.-Ing. Andrej Avrachov, Dipl.-Ing. Pavel Saltykov,
Dipl.-Ing. Jörg J. F. Strecker, Dr.-Ing. Albin Dobersek,
Dipl.-Ing. Turarbek A. Azekenov[1],
Dr.-Ing. Sergey N. Moskalenko[2]

Engineering Dobersek GmbH
Pastorenkamp 31
41169 Mönchengladbach, Germany

[1] Kazzinc Ltd.; [2] Kazzinctech Ltd.
1, Promyshlennaya Street
070002 Ust-Kamenogorsk, Republic of Kazakhstan

Keywords: Slag Casting Machine, Smelter, Refinery, Acid Plant, Lead Plant

Abstract

In 2010-2013 Engineering Dobersek GmbH was involved in a new project to modernize a lead smelting plant for a metallurgical producer in Kazakhstan with an annual production of 100,000 tons of refined lead ingots. The existed lead smelting process was upgraded by the replacement of the traditional sinter plant to a blast furnace operation through the modern TSL technology.

Engineering Dobersek GmbH was responsible for the general design of the smelter plant and the basic and detail engineering as well as the supply of the following auxiliaries:

- feed preparation
- cooling water system
- gas cleaning system
- smelter shop ventilation and heating system
- slag casting machine
- lead bullion discharging system

Also the erection management, commissioning of the delivered equipment and staff training was included in the project scope.

One of the core components was the lead slag casting machine which produces the slag pigs to charge the existing blast furnace for the reduction step. The full automatic two lines casting machine was designed by Engineering Dobersek GmbH with a casting capacity of 45 - 60 t/h. The casting temperature of lead slag is approx. 1,150 °C. The slag is air-cooled by natural convection, then by means of forced convection and at the end of the machine by means of water spray in order to cool the briquettes below 300 °C. The off-gas of the lead smelter is captured very effectively, cooled, cleaned and sent to the sulphuric acid plant. The modernization of the lead plant has accompanied with a significant reduction of harmful emissions into the environment.

Objective: Full-Scale Extraction of own Raw Materials and complete Off-Gas Utilization

Introduction

Kazzinc is a major fully integrated zinc producer with considerable lead, gold and silver credits located in Kazakhstan. Along with increased base of raw materials currently developed by the company the percentage and importance of copper in the entire polymetallic raw material structure increases, too. With the decision to create a copper smelter and a copper refinery plant in order to produce a saleable copper concentrate shipped to the consumers as semi-finished product, the New Metallurgy Project also included the upgrade of the existing lead smelting process by the replacement of the traditional sinter plant to a blast furnace operation through the modern TSL technology.

The New Metallurgy Project

The New Metallurgy Project implemented by Kazzinc was intended to do away with raw material-oriented copper production through construction of two metallurgical plants: a copper smelter and a copper refinery. The 3rd step of this project was the reconstruction of the existing lead plant.

Figure 1: View on Plant in Ust Kamenogorsk

The goal of this project was to reduce the total output of harmful gases and dust into atmosphere with simultaneously increasing the production of valuable products.

Former Lead Production Process

Prior to implementation of the Lead plant in the New Metallurgy Project a standard process including sintering of Lead, gravity and Gold-bearing flotation concentrates, blast smelting of sintered agglomerates and reverts, pyrometallurgical refining of Lead bullion and ingots casting was used at the existing Lead plant. This process was closely linked with the Zinc plant processes, i.e. processing of middlings produced at the Zinc plant.

The following middlings were produced by sinter and reverts smelting: Zinc-bearing slag for fuming, matte for processing in Peirce-Smith converters and Lead bullion for refining.

Zinc oxide after slag fuming was processed at the Zinc plant with Zinc extraction while waste slag was disposed in the waste stockpile.

Minor amounts of Copper blister after matter converting were shipped to the consumers.

Lead bullion was refined, Copper, Tellurium, Arsenic, Antimony, Bismuth, Gold and Silver were removed and the ingots were shipped to the consumers.

The produced quality of Lead metal is 99.985 %.

The precious metals refinery which is an integral part of the Lead plant uses the following process: production of Dore alloy from silvery crust skimmed during Lead refining; hydrometallurgical or electrolytic treatment of Dore with production of Gold, Silver and PGM.

Gold is produced in ingots whereas Silver in ingots and powder. The refined 99.95 % Gold and Silver ingots produced by Kazzinc comply with LBMA Standard “DEER” brand.

Upgraded Lead Production Process and Plant Description

Implementation of the New Metallurgy Project eliminated the following imperfections in the Lead production process:

- large-scale consumption of imported coke
- impossibility to treat Copper-bearing products (Cu cement) from the Zinc plant with their accumulation jeopardizing the local ecology
- generation of lean sulfuric gases and their release to the atmosphere due to inefficiency of their utilization.

ISASMELT™ process was introduced at Kazzinc as a part of the New Metallurgy Project. This process broadened the line of treated Lead-bearing products as opposed to sinter roasting. In addition, the higher-grade slag could be now treated in the blast furnace which is infeasible with high-grade sintered agglomerates due to its melting in the blast furnace. Complete sulfur removal prevents matte generation in the blast furnace.

Process-wise the new plant can be described as follows:

Quite simple feed preparation system designed for processing of concentrates with up to 14 % moisture and consisting of bins, conveyors, feeders and a mixing drum. After the autogenous oxidizing smelting of prepared feed high-grade Lead slag is continuously discharged to the tilt bowl and further to the 2-lines slag casting machine which is shown in figure 2.

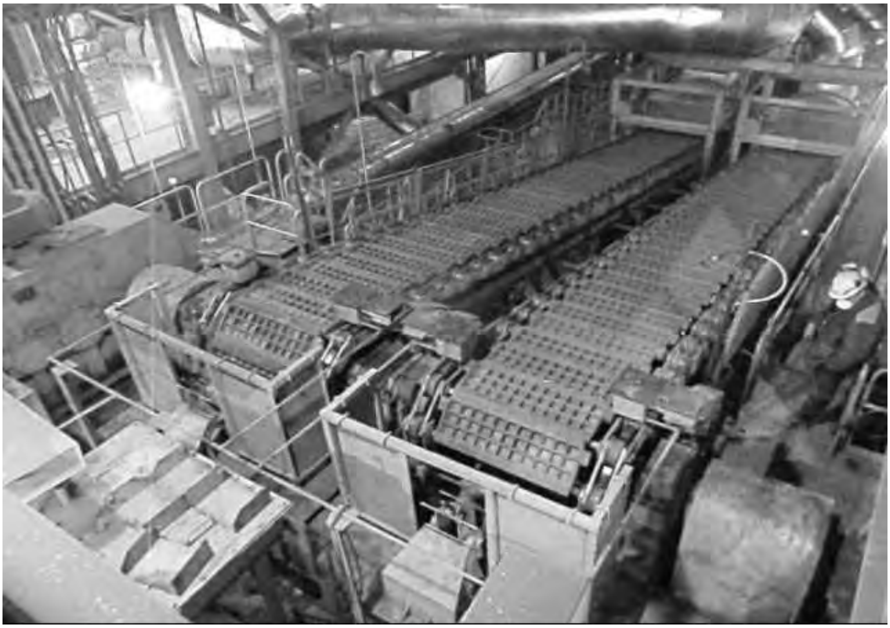


Figure 2: Slag casting machine designed by Engineering Dobersek GmbH

This machine, one of the core components in the new Lead plant, was designed by Engineering Dobersek GmbH and was already the focus of a project for an European noble metal producer. It produces the slag pigs to charge the existing blast furnace for the reduction step. The full automatic two lines casting machine was designed for a casting capacity of 45 - 60 t/h. The casting temperature of Lead slag is approx. 1,150 °C. The slag is air-cooled by natural convection at the beginning of the casting machine, then cooled down by means of forced convection with air blowers to decrease the temperature of about 500 °C and at the end of the machine by means of water sprayed by 20 nozzles each from above as well as from bottom in order to cool the briquettes below 300 °C.

A 3-D view is given in figure 3. Spatial constraints predetermined the design of the machine. The project was implemented in quite space-limited surroundings with production ongoing. After the slag is cooled to the design solidification temperature the pigs are transported to the existing blast furnace for reduction smelting with production of Lead bullion and waste slag.

Rich gases after oxidizing smelting in ISASMELT™ furnace are cooled in the waste heat boiler, purified in the dry electrostatic precipitator (ESP) and supplied to the acid plant. A view into the production hall with the ESP is given in figure 4.

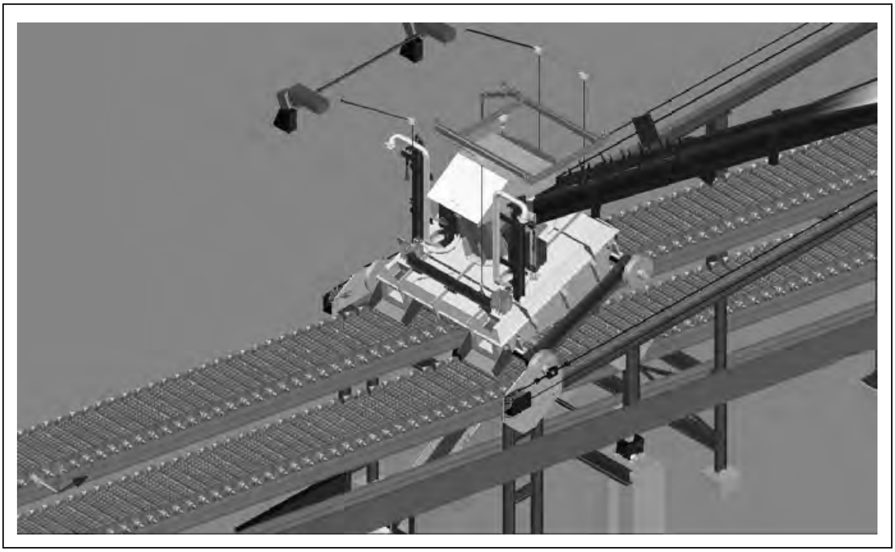


Figure 3: A 3-D view of the slag casting machine

Aspiration gases are captured at source and purified in the existing bag filters.

The modernization of the lead plant has accompanied with a significant reduction of harmful emissions into the environment.

Figure 4: ESP in the Lead plant

Conclusions

Implementation of three segments of the New Metallurgy Project, as the construction of the new copper and acid plants and the reconstruction of the existing lead plant contributed to reduction of air pollution by the metallurgical complex in Ust-Kamenogorsk. Minor emissions from the new copper smelter were more than offset by almost complete utilization of gases from the state-of-the art lead smelter in sulfuric acid production. Previously these diluted gases were released into the atmosphere.

In addition, Kazzinc enjoys a broader product line and higher production output, which along with complex utilization of raw materials and more flexible production processes contributed to higher net profits.

Acknowledgments

The authors would like to thank «Kazzinc Holdings» Ltd for permission to publish this paper.

Lead ISASMELT™ Operations at Ust-Kamenogorsk

A. S. Burrows
Glencore Technology Pty Ltd
Level 10, 160 Ann Street
Brisbane, 4000. Australia

T. Azekenov, R. Zatayev
Kazzinc Ltd
1 Promyshlennaya Street, Ust-Kamenogorsk
070002, Republic of Kazakhstan

Keywords: ISASMELT, TSL, lead smelter, bath smelting, Kazakhstan

Abstract

Ust-Kamenogorsk Metallurgical Complex holds a prestigious role in the world of lead smelting, and has been home to many developments over the past 70 years. Famous among these developments was the design and practical demonstration of the world's first KIVCET furnace thirty years ago. The most recent development was brought to fruition in 2012 with the installation of the world‘s largest Lead ISASMELT™ furnace. The construction and commissioning of the new ISASMELT™ plant was an achievement of delicate 'brownfield' engineering, with multiple interfaces to the existing smelter. Despite the complexity of integrating the new plant into the smelter, no production time was lost on the pre-existing sinter machine and blast furnace flowsheet right up until the first feed was smelted in the Lead ISASMELT™ furnace. This paper includes a description of the engineering challenges, and some results of the subsequent plant operation.

Introduction

Ust-Kamenogorsk Metallurgical Complex (UKMC) is a venerable smelting site, which has maintained a leading position in the development and implementation of non-ferrous pyrometallurgy for over 70 years. Zinc production began in 1943, and lead production began in 1952. UKMC grew to become one of the largest zinc and lead smelting operations in the world. It was for many years the headquarters of zinc and lead metallurgical development within the USSR, and from its technical and operating expertise many processes were developed. Among the most famous is the KIVCET furnace for smelting lead concentrates, which has been adopted by smelters in Italy, Canada and China.

For the past two decades lead production has been exclusively through a sinter plant, of Lurgi design, with most equipment originally installed in 1974. From 2004 onwards the sinter machine gases have been treated by a Haldor-Topsoe wet sulphuric acid plant. Although the Haldor-Topsoe plant is designed to suit low-strength SO_2 gases that would be unsuitable feed for a contact process, the sulphur in gases from the tip end of the sinter machine has only been 50 % utilized. An estimated emission of 13,000 t/y SO_2 has been attributable to these gases in a normal production year.

In 2006 Kazzinc sought to change its lead smelting flowsheet to decrease SO_2 emissions into atmosphere, but with an aim to retain as much as possible of the existing infrastructure, and to continue running the sinter plant up with minimal disruption until the new technology was ready to operate. The new smelter would have myriad interfaces with the existing smelter, and management of these interfaces would require delicate ‘brownfield’ engineering with close cooperation between the smelter operating personnel and the project implementation team.

Lead ISASMELT™ Project

After careful evaluation of the process options Kazzinc chose to build an ISASMELT™ furnace, with the technology, some design services and equipment supplied by Glencore Technology.

Duty of Lead ISASMELT™ Plant

The lead ISASMELT™ plant was designed to treat 291,000 tonnes per year of new lead-bearing material (nett of any fluxes, fuel or dusts that may be generated by the process itself). The Kazzinc feed materials are comprised of three main types:

- Residues from the zinc plant
- Lead concentrates from Kazzinc’s mines
- Lead battery paste from Kazakhstan sources
- Precious metal concentrates, sometimes containing lead, from various sources across Kazakhstan and central Asia.

Depending on the ratio of these materials, which varies significantly across a year, the feed to the lead ISASMELT™ can contain anywhere between 20 - 60 %Pb. The design basis considered that the plant would treat new lead-bearing materials with an average composition as shown in Table 1, at the instantaneous rate of 40 t/h.

Table 1: Design feed blend for Lead ISASMELT™ Plant

Zn (%)	Pb (%)	Cu (%)	S (%)	Fe (%)	SiO_2 (%)	C (%)	As (%)	Sb (%)	H_2O (%)
10.5	42.9	2.4	17.2	9.1	5.6	0.0	0.4	0.20	14.8

During the design stage it was recognised that the ISASMELT™ operating mode would need to afford the operators at Kazzinc a great deal of flexibility in operation, in order to cope with the potential fluctuation in material types and compositions that were foreseen. For most of the range, an oxidative lead smelting operation, producing only lead-rich slag, is the most tolerant and robust processing option available. Consequently the plant design considered that the principle method of operation would be an oxidative lead ISASMELT™ furnace, followed by a reduction furnace. In

Kazzinc's case, the existing blast furnaces were to perform the reduction duty, creating an overall flowsheet reminiscent of the ISASMELT™ operation described for Qujing, China [1].

Integration with Surrounding Plant

The ISASMELT™ plant was designed to:

- Receive pelletised lead concentrate, directly from the existing pelletiser sitting astride the sinter machine;
- Supply cast ingots of lead-rich slag, directly to the existing bins for supply of sinter to the blast furnaces;
- Supply sulfur-rich gas directly to the existing Haldor-Topsoe sulfuric acid plant, via the same duct and through the same exhaust fans used by the sinter machine; and
- Remotely control, via DCS, the existing feed conveyors and flux bin feeders of the sinter machine, which had never been automated previously.

In combination, the engineering required to achieve these objectives was challenging and time-consuming. However, the careful execution of these four objectives was considered essential in order to minimize the shut down time associated with transition from the sinter machine operation to the Lead ISASMELT™ operation, and also to minimize interference with the upstream and downstream operations of a complex 'brownfield' site. A diagram of the Lead ISASMELT™ plant, and the main material transfer points, is shown in Figure 1.

A key feature of the layout in Figure 1, is the relatively small footprint of the plant. The footprint was defined by site infrastructure, and the engineering had to rise to meet the challenge. There was only one suitable location on the site of UKMC where the four integration objectives could reasonably be met, and that was the historic site of three dismantled sinter plants. The site was bounded by the:

- return sinter crushing building to the east;
- sinter skip hoist to the west;
- sinter machine feed conveyor gallery to the north; and the
- return sinter pan conveyor gallery to the south.

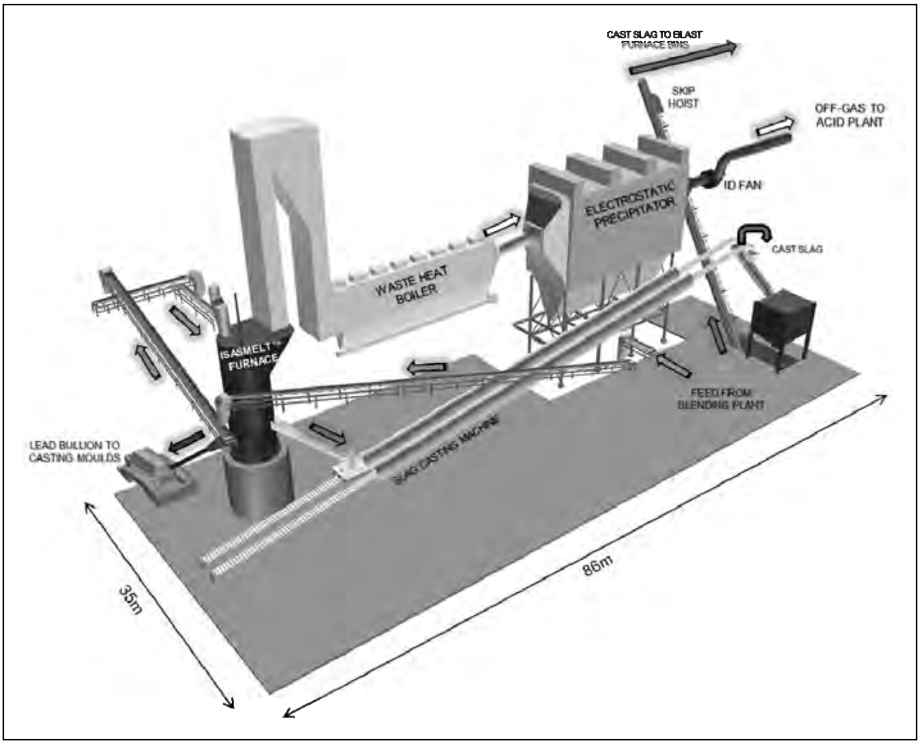


Figure 1: The Lead ISASMELT™ Plant at UKMC

Each of the surrounding unit operations had to continue operating throughout the construction and commissioning periods. Every spare metre of floor space was planned and used to fit the new smelting plant within an 86 m × 35 m footprint. The tightness of fit is illustrated by the artist's impression shown in Figure 2, in which the new (brown) building is contrasted against the old (light green) structures. Design of the compact plant layout, and cross-checking of site features, was performed largely by Kazzinc's in-house engineers with assistance by Engineering Dobersek GmbH.

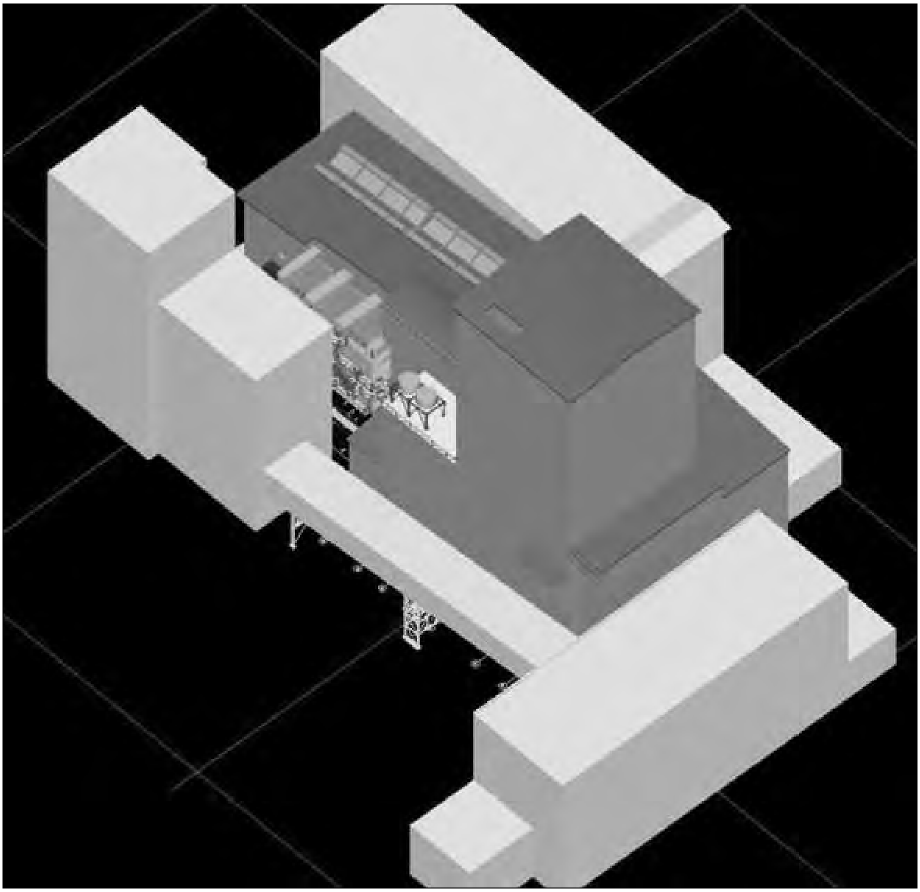


Figure 2: The Lead ISASMELT™ Plant at UKMC

Demolition and Restructure

The plant design took into account the complex task of ‘brownfield’ engineering, including the partial demolition of existing structures. The task would have been simplified if entire structures could have been demolished (see for example Figure 3) but this was not possible. Selective areas on upper floors of multi-storey buildings had to be supported and equipment kept in operation, while demolition of lower floors made way for new equipment. Careful, piece-wise demolition and reconstruction was required.

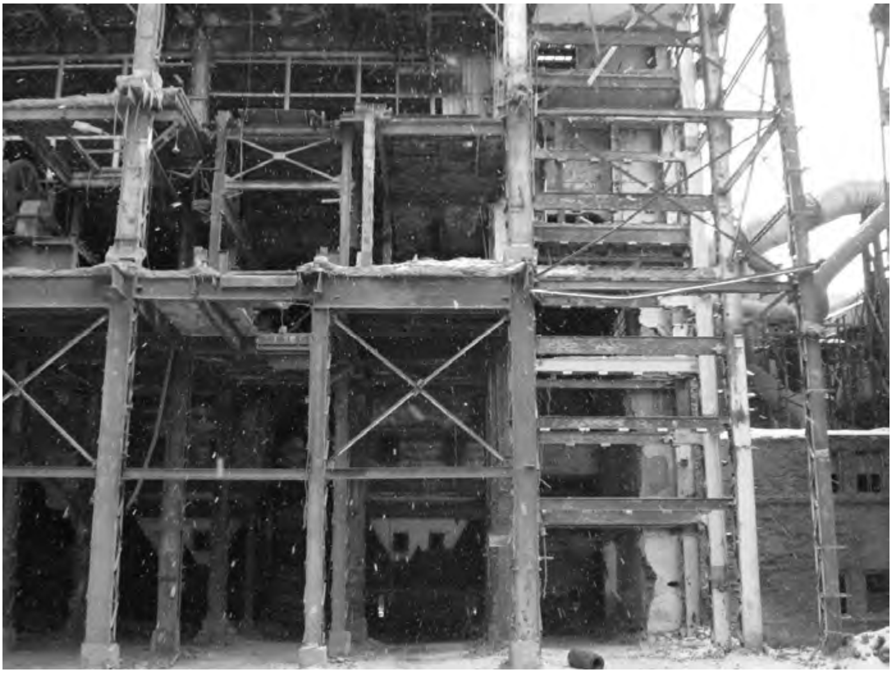


Figure 3: Construction site for the Lead ISASMELT™ Plant at UKMC

Erection

After partial demolition of the surrounding structures had cleared enough room, plant erection began delicately. Excavations to prepare new foundations were done with small machinery, sometimes manually, to enable maximum manoeuvrability around the footings of existing structures. The ISASMELT™ furnace foundation was among the first to be poured, and is shown in Figure 4, in which the partially-exposed footings of several other structures can also be seen at the periphery of the excavation.

Figure 4: Erection of the Lead ISASMELT™ Furnace at UKMC

Cut-in for Services and Materials Transport

The most challenging task in the Lead ISASMELT™ Plant project was to organise the cut-in for services and materials transport. The modernisation of the feed delivery system was a good example of the challenges that had to be met. The feed delivery system to the sinter plant was to be retained and reassigned for duty to the new Lead ISASMELT™ furnace. However, the sinter plant operation was engineered for local control of equipment units and the Lead ISASMELT™ Plant was engineered for automated remote control of equipment units. So each of the operating units in the lead sinter plant feed system had to be re-engineered and automated, ready for the new duty.

Approximately 3 months before the commissioning of the Lead ISASMELT™ Plant, all of the equipment in the lead sinter plant feed system (refer to Figure 5) had been automated, with the final electrical cut-in done during a planned 8-hour plant maintenance shutdown. From that point onwards, the equipment was run under the control of a remotely-situated operator, in what would soon become the Lead ISASMELT™ Plant control room.

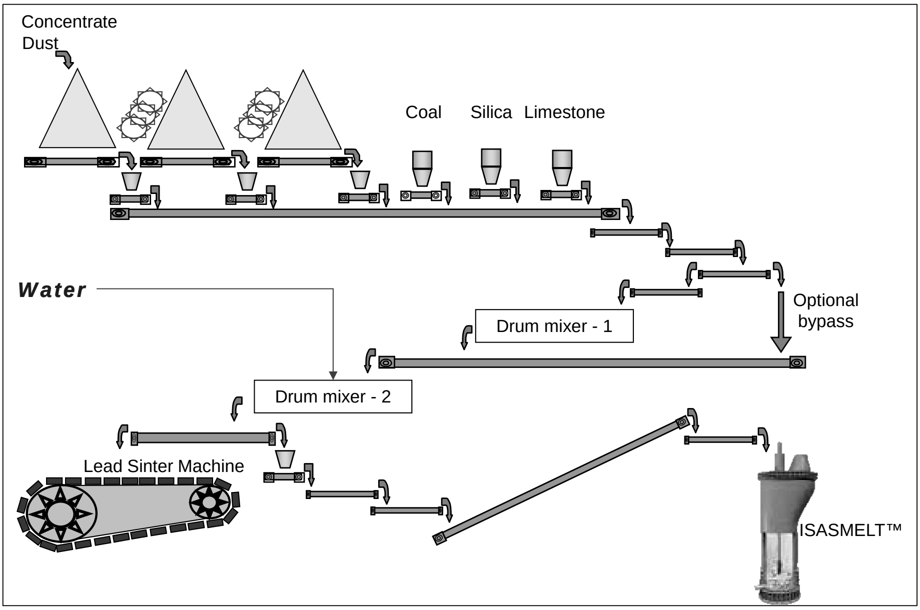


Figure 5: Common feed system of the lead sinter plant and the Lead ISASMELT™ Plant

Operator Training

Operators received both practical and theoretical training in advance of hot commissioning. Off-site training included stints of plant work at operating ISASMELT™ furnaces in Australia and the Peoples' Republic of China. On-site training included rehearsals of operating tasks, workplace safety procedures and plant theory tests.

An unexpected and useful benefit of the automation and early cut-in of the sinter plant feed system was that operators gained many weeks of experience in using the plant automated control system prior to the actual operation of the ISASMELT™ furnace.

Hot Commissioning

The furnace was heated ready for operation by the end of August 2012. Smelting began in September of the same year. There were some challenges to overcome in the following months of operation.

Feed for Start-up

As a result of the complex and time-consuming 'brownfield' engineering of the plant, the time of starting the ISASMELT™ plant was separated by some 6 years from the time of its initial design. Inside that 6 year gap Kazzinc had commissioned new lead mining assets, decommissioned old ones, altered a range of their internal plant recycle streams, and substantially changed the type of feed to be treated in the ISASMELT™ furnace. The combined feed materials (exclusive of fluxes and fuel) had a composition (averaged over the first 10 blended feed beds of 2500 t each) as shown in Table 2 at the time of starting the ISASMELT™ operation in September 2012.

Table 2: Start-up feed blend for Lead ISASMELT™ Plant

Zn (%)	Pb (%)	Cu (%)	S (%)	Fe (%)	SiO_2 (%)	C (%)	As (%)	Sb (%)	H_2O(%)
7.4	30.7	2.66	17.2	12.2	8.5	0.67	0.66	0.22	10.9

The major differences between the design feed composition in Table 1 and the actual feed composition in Table 2 were:

- The lead content was about 12 percentage points lower
- The arsenic content was about 0.3 percentage points higher
- The moisture content was about 4 percentage points lower
- Some organic carbon was present, where none had been expected

These differences contributed to the challenges of the subsequent operation period.

Ramp-up of Production

The first time that the ISASMELT™ plant achieved its design daily cumulative production was 5 weeks after smelting commenced. Maintaining the design production rate for periods longer than a few consecutive days was not possible until many months had passed, and the operation of some surrounding plant items had been optimised such as the slag casting machine, tilt bowl, waste heat boiler, electrostatic precipitator, feeding system, off-gas ducting, etc.

Tapping & Slag Casting Operations

The slag casting is a continuous operation, and requires a continuous feed of molten slag, which is achieved via the specially-designed ISASMELT™ tapping process. Tapped slag is distributed onto each strand of a twin-strand straight-line casting machine designed and supplied by Ormeto Yumz. The slag casting machine performance was found to be sensitive to the slag tapping temperature.

Owing in part to the lower lead grade of the incoming feed, the properties of the tapped slag were different from the vendor's expectations: the slag density was lower and the slag temperature was higher. This in turn meant that the slag casting machine was run at its maximum speed, raising the temperature of the slag briquettes at the discharge end of the casting machine. Some initial difficulty was experienced with reliably disengaging the slag briquettes from their moulds. Remedial actions included:

- Improvements to the design of the anvil that the mould trays impact near the head drum of the slag casting machine.
- Installation of a system for applying mould release agent to the tail end of the slag caster, upstream of where the tapped slag is received.
- Doubling the size of the cells containing the slag briquettes within each mould tray.

After these remedial actions were completed the disengagement of slag from the moulds became reliable.

Dust-handling Operations

The mass of dust collected from the ISASMELT™ furnace off-gas by the waste heat boiler (WHB) and electrostatic precipitator (ESP) was typically equivalent to about 7 % of the mass of the incoming feed materials. The typical dust particle has a diameter of about 10 - 20 µm, and accumulations of dust adhere strongly to steel surfaces. The WHB was sized to accommodate the expected inefficiency of heat transfer, and the sticky dust did not adversely affect the WHB.

However, the ESP operation was impaired by the accumulation of dust near the off-gas inlet, and the ineffectiveness of the rappers to dislodge the dust from this area. The removal of internal steel protrusions eliminated anchor points from which the dust accretions could accumulate. This action, combined with raising the temperature of the incoming off-gas by 50 °C, and improvements to the rapper design, were sufficient remedies for the ESP to attain full functionality.

The WHB was operating with a much reduced off-gas volume compared with design expectations. This is mostly the consequence of the actual feed having a moisture content 4 percentage points lower than design expectations, and therefore a much reduced requirement for combustion of fuel and generation of off-gas. In fact, the typical furnace operation was often autogeneous and the WHB received around 30 % less gas than it was designed for. WHB convection section evaporator bundles were progressively removed so that the optimum heat transfer could be achieved, to keep the ESP below its permissible maximum inlet temperature, but without approaching the acid dew-point.

Blast Furnace Operation

Kazzinc proceeded with the modernisation project assuming that the existing lead blast furnaces would be satisfactory for treating the slag produced in the Lead ISASMELT™ Plant, with little or no capital investment required. This assumption has since been validated. The instantaneous capacity of the ISASMELT™ furnace to generate slag is approximately twice the instantaneous capacity of a Kazzinc blast furnace to smelt slag, so two blast furnaces are in use most of the time.

The ISASMELT™ slag typically contains less CaO than is required to meet the slag chemistry target of the blast furnaces. Consequently, beginning with the introduction of ISASMELT™ furnace slag, the addition of limestone flux to the blast furnace charge has been introduced as a standard operating practice. This has not presented any significant operating difficulty.

One unexpected aspect of the project implementation was that the change-over of blast furnace operation from treating sinter to treating slag was not a one-off step change. The blast furnaces became adept at treating stockpiled material, from either source, and for many months they treated varying compositions of slag and sinter, with the mix varying from 0 - 100 %. This afforded a good opportunity to evaluate the furnace performance when using ISASMELT™ slag as feed, compared to using sinter as feed. The general findings were as follows:

- The blast furnace uses less coke per unit of charge when smelting slag, compared to smelting sinter. The reduction is equivalent to approximately 2 percentage points.
- Addition of a sulphidising agent to the blast furnace charge is necessary to ensure that the excess copper in charge reports to a matte phase, rather than to a less desirable speiss phase.

Post-Commissioning Period

Within a year of the Lead ISASMELT™ plant beginning operation, Kazzinc decided to determine the extent of the operating flexibility of their new furnace. They prepared a special blend of feed containing the approximate composition shown in Table 3. The objective of the trial was to compare the operation of the ISASMELT™ furnace when producing lead bullion and slag, versus production of only slag (which is the typical operation, according to the process described in [2]).

Table 3: Feed Blend for Trial Smelting Period

Zn (%)	Pb (%)	Cu (%)	S (%)	Fe (%)	SiO_2 (%)	As (%)	H_2O (%)
3.9	58.9	0.3	14.5	6.7	3.9	0.1	10.4

The trial was successful, and the lead bullion tapping and casting facilities were commissioned and evaluated against their design targets.

The first replacement of the brick lining occurred 24 months after the furnace was first heated. Good performance of the refractory was noted in most areas of the furnace, and some localised wear areas were identified with remedial actions adopted in advance of the second campaign.

There are ongoing efforts to optimize the material specifications for components of the re-used sinter plant off-gas system, where the introduction of ISASMELT™ off-gas (containing reduced volumes of gas and much stronger concentrations of SO_3) has resulted in accelerated acid corrosion compared with past experience.

Future Prospects

The installation of the ISASMELT™ technology at UKMC is merely the latest in a long line of developments that stretch back over the past 70 years of the venerable plant. Now that it has been successfully incorporated into the flowsheet at UKMC the next challenge is to consider what flowsheet improvements could be made at the blast furnaces, or whether their replacement by modern technology is economically attractive. Evaluation of the options is ongoing. It has been recognised that the presence of the Lead ISASMELT™ Furnace now opens up options for future development that did not exist previously.

Conclusions

A Lead ISASMELT™ Plant has been successfully introduced into the smelting flowsheet of Ust-Kamenogorsk Metallurgical Complex. The new plant was erected and commissioned in a 'brownfield' site with great care taken to optimise the production of the existing smelting assets. Operation began with feed that varied from the original design basis, but the smelting technology was not significantly impaired by this discrepancy. The plant has now treated feeds varying in lead content from 30 - 60 %Pb, at instantaneous operating rates exceeding the original design by 20 %. The introduction of ISASMELT™ technology is the latest in a long list of smelting developments at Kazzinc, and the company goes forward contemplating other technologies for further improvements.

Acknowledgements

The authors would like to acknowledge Kazzinc Ltd for approving data and images for publication. The authors would like to thank Glencore Technology Pty Ltd and Kazzinc Ltd for permission to publish this paper.

References

[1] ERRINGTON, B., ARTHUR P., WANG, J. & DONG, Y. (2005): "The ISA-YMG Lead Smelting Process", Proceedings of the International Symposium on Lead and Zinc Processing, Kyoto, Japan, October 2005, T. Fujisawa et al., Eds., MMIJ, pp. 581-599.

[2] ERRINGTON, B., FEWINGS, J.H., KERAN, V.P., & DENHOLM, W.T. (1987): "The ISASMELT Lead Smelting Process", Trans. Instn. Min. Metall., Section C, vol 96.

Реконструкция свинцового завода с применением технологии ISASMELT™

Д-р. инж. наук Андрей Аврахов, дипл. инж. Павел Салтыков, дипл. инж. Йорг Дж.Ф.Штрекер, д-р инж. наук Альбин Доберсек, дипл. инж. Турарбек А. Азекенов*, д-р инж.наук С.Н.Москаленко[×]

Engineering Dobersek GmbH
Пасторенкамп 31
41169 Мюнхенгладбах, Германия

* ТОО Казцинк
Ул.Промышленная,1
070002 Усть-Каменогорск, Республика Казахстан

[×] ТОО Казцинктех
Ул.Промышленная,1
070002 Усть-Каменогорск, Республика Казахстан

Аннотация

В 2010-2013 компания Инжиниринг Доберсек ГмбХ [Engineering Dobersek GmbH] участвовала в новом проекте модернизации свинцового плавильного завода для казахстанского металлургического производителя с ежегодной производительностью 100 000 тонн чушек рафинированного свинца. Существующий процесс плавки свинца был модернизирован путем замены стандартной агломашины на современную шахтную печь на основе технологии TSL (погружной фурмы).

Инжиниринг Доберсек ГмбХ [Engineering Dobersek GmbH] была ответственна за общий проект плавильного завода, базовое и детальное проектирование, а также поставку следующих вспомогательных систем и оборудования:

- шихтоподготовка
- система водяного охлаждения
- система газоочистки
- система вентиляции и отопления плавильного цеха
- шлакорозливочная машина
- система выпуска чернового свинца

В объем работ по проекту также были включены управление строительством, ввод в эксплуатацию поставленного оборудования и обучение персонала.

Одним из ключевых компонентов являлась машина для розлива свинцового шлака, которая производит шлаковые чушки для загрузки в существующую шахтную печь для восстановления. Компания Инжиниринг Доберсек ГмбХ [Engineering Dobersek GmbH] спроектировала полностью автоматические двухлинейные розливочные машины производительностью 45-60 т/г. Температура розлива свинцового шлака составляет примерно 1 150°C. Охлаждение шлака происходит с помощью воздуха путем естественной конвекции, затем принудительной конвекции и в конце машины путем распыления воды для охлаждения шлаковых брикетов до температуры ниже 300 °C.

Отходящий газ свинцового завода очень эффективно улавливается, затем охлаждается, очищается и направляется на сернокислотную установку. Модернизация свинцового завода позволила значительно сократить вредные выбросы в окружающую среду.

Цель: Комплексное извлечение металлов из собственного сырья и полная утилизация отходящих газов

Введение

Казцинк – крупный интегрированный производитель цинка с большой долей сопутствующего выпуска свинца, золота и серебра, расположенный на территории Казахстана. Наряду с увеличенной сырьевой базой, разработанной компанией в настоящее время, также увеличивается процентное соотношение и важность меди во всей структуре полиметаллического сырьевого материала. Наряду с решением о строительстве медного завода и цеха электролиза меди с целью производства товарного медного концентрата, отгружаемого потребителям в виде полуфабриката, проект Новая Металлургия также включал в себя модернизацию существующего процесса плавки свинца путем замены стандартной агломашины на современную шахтную печь технологии TSL.

Проект Новая Металлургия

Проект «Новая Металлургия», реализованный компанией Казцинк, был направлен на то, чтобы исключить медное производство, ориентированное на сырье, путем строительства двух металлургических объектов: медеплавильного завода и цеха электролиза меди. Третьим этапом данного проекта была реконструкция существующего свинцового завода.

Рисунок 1: Завод в Усть-Каменогорске. Вид сверху.

Целью данного проекта было уменьшение общего выпуска вредных газов и пыли в атмосферу с одновременным увеличением производства ценной продукции.

Прежний процесс производства свинца

До ввода свинцового завода в эксплуатацию по проекту «Новая Металлургия» на существующем свинцовом заводе использовался стандартный процесс, включающий спекание свинцовых, гравитационных и золотосодержащих флотационных концентратов, шахтную плавку агломерата и оборотных материалов, пирометаллургическое рафинирование чернового свинца и отливку чушек. Этот процесс был тесно связан с процессами на цинковом заводе: т.е. переработка произведенных на цинковом заводе промежуточных продуктов.

При плавке агломерата и оборотных материалов получали следующие промежуточные продукты: цинксодержащий шлак для возгонки, штейн для переработки в конвертерах Пирс-Смит и черновой свинец для рафинирования.

Окись цинка после возгонки шлака перерабатывалась на цинковом заводе с извлечением цинка, тогда как отвальный шлак сбрасывался в отвалы.

Незначительное количество черновой меди после конвертирования штейна отгружалось потребителям.

Черновой свинец рафинировался и очищался от меди, теллура, мышьяка, сурьмы, висмута, золота и серебра, а чушки отгружались потребителям.

Качество произведенного свинца составляет 99.985 %.

В аффинажном цехе, который является неотъемлемой частью свинцового завода, используются следующие процессы: производство сплава Доре из серебристой корки, собранной в процессе рафинирования свинца; гидрометаллургическая либо электролитическая переработка сплава Доре с получением золота, серебра и металлов платиновой группы.

Золото выпускается в слитках, тогда как серебро - в слитках и в виде порошка. Выпускаемые компанией Казцинк слитки золота и серебра с чистотой 99.95 % соответствуют бренду “DEER” стандарта LBMA (Лондонской биржи драгоценных металлов).

Модернизированные процессы производства свинца и описание завода

Реализация проекта Новая Металлургия позволила устранить следующие недостатки в процессе производства свинца:

- значительное потребление импортированного кокса
- невозможность переработки медьсодержащих продуктов (цементационная медь) цинкового завода и их накопление, что представляет угрозу местной окружающей среде
- производство газов с низким содержанием серы и их выпуск в атмосферу из-за неэффективности их утилизации.

Процесс ISASMELT™ был внедрен на Казцинке в рамках проекта «Новая Металлургия». Этот процесс помог расширить диапазон перерабатываемых свинец-содержащих продуктов по сравнению с агломерирующим обжигом. Кроме того, шлак с высоким содержанием теперь можно перерабатывать в шахтной печи, что было невозможным с агломератами с высоким содержанием в результате их плавления в шахтной печи. Полное удаление серы препятствует образованию штейна в шахтной печи.

С точки зрения технологического процесса новый завод может быть описан следующим образом:

Для переработки концентратов с содержанием влажности до 14% была разработана достаточно простая система подготовки шихты, состоящая из загрузочных бункеров, конвейеров, питателей и смесительных барабанов. После автогенной окислительной плавки подготовленной шихты, шлак с высоким содержанием свинца непрерывно выпускается в промежуточную мульду и далее на шлакоразливочную машину, состоящую из двух линий, которая показана на рисунке 2.

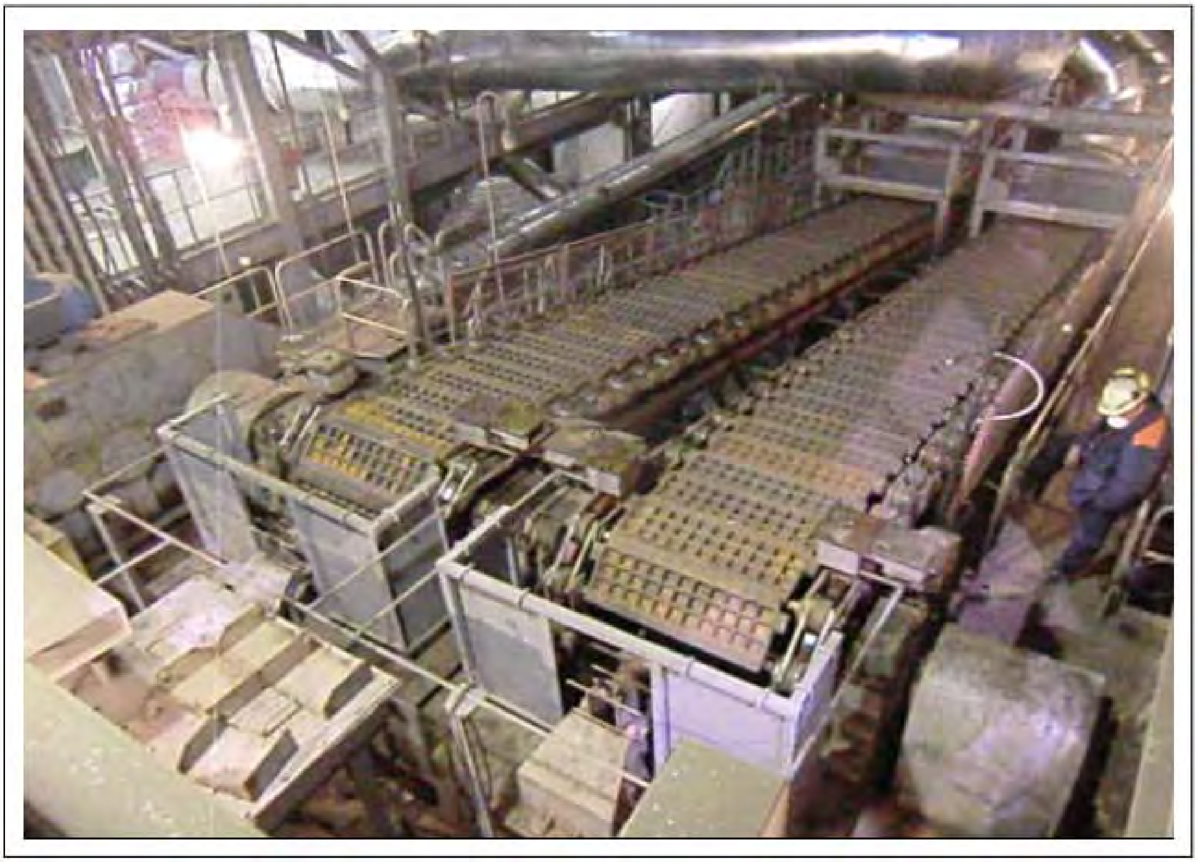


Рисунок 2: Шлакорозливочная машина, разработанная Инжиниринг Доберсек ГмбХ [Engineering Dobersek GmbH]

Данная машина, один из ключевых компонентов нового свинцового завода, была спроектирована компанией Инжиниринг Доберсек ГмбХ [Engineering Dobersek GmbH] и уже была основным элементом проекта для европейского производителя драгоценных металлов. С использованием машины производятся шлаковые чушки для загрузки в существующую шахтную печь для восстановления. Были разработаны полностью автоматические двухлинейные розливочные машины производительностью 45-60 т/ч. Температура розлива свинцового шлака составляет примерно 1 150°C. Охлаждение шлака происходит с помощью воздуха путем естественной конвекции в начале разливочной машины, затем принудительной конвекции с помощью воздуходувок для снижения температуры до 500 °C и в конце машины путем распыления воды через 20 форсунок в верхней части, а также снизу для того, чтобы охладить брикеты ниже температуры 300 °C.

Изображение в 3-D приведено на рисунке 3. Пространственные ограничения определили конструкцию машины. Проект реализовывался в ограниченном пространстве без остановки производства. После охлаждения шлака до проектной температуры затвердевания, чушки транспортируются в существующую шахтную печь на восстановительную плавку для производства чернового свинца и отвального шлака.

Обогащенные газы после окислительной плавки в печи ISASMELT™ охлаждаются в котле-утилизаторе, очищаются в сухом электрофильтре (ESP) и подаются на сернокислотный завод. Вид производственного помещения с электрофильтром ESP см. на рисунке 4.

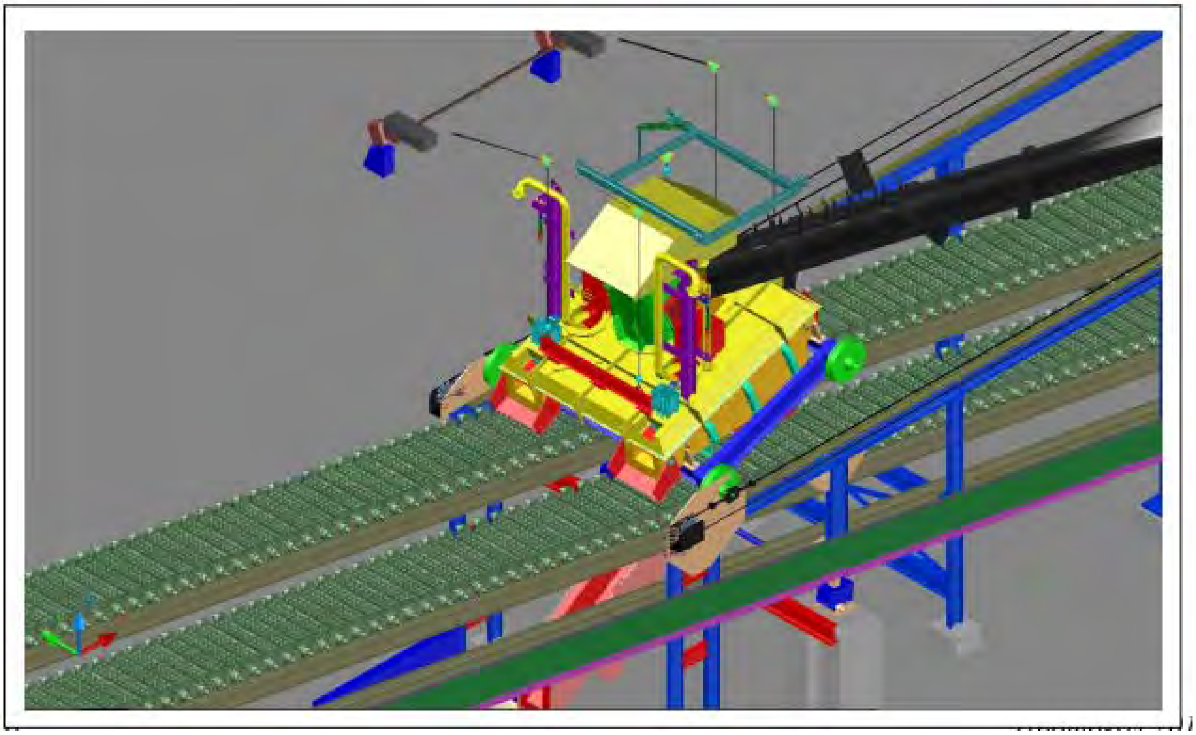


Рисунок 3: 3-D изображение шлакоразливочной машины.

Аспирационные газы улавливаются у источника и очищаются на существующих рукавных фильтрах.

Модернизация свинцового завода позволила значительно сократить вредные выбросы в окружающую среду.

Рисунок 4: Электрофильтр на свинцовом заводе

Выводы

Реализация трех частей в рамках проекта «Новая Металлургия», а именно строительство нового медеплавильного и сернокислотного заводов, а также реконструкция существующего свинцового завода способствовали уменьшению загрязнения воздуха металлургическим комплексом в Усть-Каменогорске. Незначительные выбросы с нового медеплавильного компенсировались практически полной утилизацией газов современного свинцового завода на сернокислотном производстве. Ранее эти разбавленные газы выпускались в атмосферу.

Также, Казцинк расширил ассортимент выпускаемой продукции и увеличил производительность, что наряду с комплексной утилизацией сырья и более гибкими производственными процессами способствовало получению более высокой чистой прибыли.

Подтверждения

Авторы хотели бы поблагодарить ТОО "Казцинк Холдинг" [«Kazzinc Holdings» Ltd] за разрешение опубликовать данный документ.

Свинцовое производство ISASMELT™, Усть-Каменогорск

А. С. Бэрроуз,
Glencore Technology Pty Ltd,
Level 10, 160 Ann Street,
Брисбэн, 4000. Австралия

Т. Азекенов, Р. Затеев
ТОО "Казцинк"
Ул. Промышленная 1, г. Усть-Каменогорск
070002, Республика Казахстан

Аннотация

Усть-Каменогорский металлургический комплекс играет важную роль в свинцовой промышленности и является предприятием, на котором за последние 70 лет было внедрено множество научно-исследовательских разработок. Наиболее известной среди них считается проектирование и демонстрация практической работы первой в мире печи, работающей по технологии КИВЦЭТ, изобретенной 30 лет назад. Новейшая разработка, представляющая собой установку самой большой в мире свинцовой печи ISASMELT™, была внедрена в 2012 г. Строительство и запуск нового завода ISASMELT™ в условиях существующей инфраструктуры является достижением инженерной мысли, учитывая множественные тесные связи с действующим производством. Несмотря на сложности интеграции, агломашина и шахтные печи работали без простоев вплоть до первой загрузки шихты в свинцовую печь ISASMELT™. В данной статье представлено описание инженерных сложностей, а также некоторые результаты последующей работы печи.

Введение

Усть-Каменогорский металлургический комплекс (УКМК) представляет собой достойную металлургическую площадку, сохраняющую за собой ведущее место в разработке и внедрении технологий цветной пирометаллургии в течение вот уже более 70 лет. Производство цинка началось в 1943 г., производство свинца - в 1952 г. В итоге УКМК стал одним из самых крупных цинковых и свинцовых металлургических производств в мире. В течение многих лет он был центром промышленных разработок в области цинковой и свинцовой промышленности на территории бывшего СССР. Технические знания и опыт позволили разработать здесь многие процессы. Печь КИВЦЭТ, предназначенная для плавки свинцовых концентратов, является одной из самых известных разработок, которая затем стала применяться на заводах в Италии, Канаде и Китае.

Последние два десятилетия производство свинца осуществлялось на агломашинах "Лурги", при этом большая часть оборудования была установлена в 1974 г. Начиная с 2004 г., очистка аглогазов осуществляется на сернокислотной установке Haldor-Topsoe. Хотя установка Haldor-Topsoe и рассчитана на газы с низким содержанием SO_2, что является непригодным питанием для контактного процесса, сера, содержащаяся в газах на выходе из агломашины, утилизировалась только на 50%. Расчетные выбросы SO_2, составляющие 13000 т/г, обуславливаются именно этими газами, произведенными в течение обычного производственного года.

В 2006 г. Казцинк начал прорабатывать возможность смены технологии плавки свинца для снижения выбросов SO_2 в атмосферу, при этом сохраняя как можно больше объектов существующей инфраструктуры и продолжая работать на агломашине с минимальными перебоями до того момента, как новая технология будет готова к эксплуатации. Новый завод должен был иметь множество тесных связей с существующим производством, в связи с чем управление этими связями потребовало деликатного проектирования в условиях существующего производства при тесном взаимодействии между персоналом плавильного производства и командой, реализующей проект.

Проект свинцового ISASMELT™ производства

После тщательной оценки технологических вариантов Казцинк остановил свой выбор на строительстве печи ISASMELT™. Технология, некоторые проектные услуги и оборудование поставлялись компанией Glencore Technology.

Производительность свинцового завода ISASMELT™

Свинцовый завод ISASMELT™ проектировался для переработки 291 000 тонн/год нового свинецсодержащего материала (без учета флюсов, топлива или пылей, образующихся в результате самого процесса). Загружаемый материал Казцинка состоит из материалов трех основных типов:

- Обороты цинкового производства
- Свинцовые концентраты с рудников Казцинка
- Свинецсодержащая аккумуляторная паста казахстанских поставщиков
- Концентраты драгоценных металлов, иногда содержащие свинец, из разных источников в Казахстане и в Центральной Азии.

В зависимости от соотношения данных материалов, которое значительно колеблется в течение года, материал, загружаемый в свинцовую печь ISASMELT™, может содержать от 20 до 60% свинца. Техническим заданием на проектирование было предусмотрено, что завод будет перерабатывать новые свинецсодержащие материалы со средним составом, представленным в Таблице 1, при моментной скорости подачи - 40 т/час.

Таблица 1: Проектный состав шихты свинцового ISASMELT™ завода

Zn (%)	Pb (%)	Cu (%)	S (%)	Fe (%)	SiO_2 (%)	C (%)	As (%)	Sb (%)	H_2O (%)
10,5	42,9	2,4	17,2	9,1	5,6	0,0	0,4	0,20	14,8

На стадии проектирования было определено, что режим эксплуатации ISASMELT™ должен быть достаточно гибким, чтобы операторы Казцинка имели возможность справляться с прогнозируемыми колебаниями в типах и составе материалов. Из всего спектра имеющихся вариантов окислительная плавка свинца, в результате которой получается только богатый свинцом шлак, является наиболее приемлемой и надежной опцией. Следовательно, проектом завода был предопределен основной метод эксплуатации - окислительная печь ISASMELT™ для плавки свинца, после которой должна быть восстановительная печь. В случае с Казцинком для восстановления предполагалось использовать существующие шахтные печи, таким образом, общая схема процесса напоминала схему завода в Цюйцзине, Китай [1].

Интеграция с существующим заводом

Завод ISASMELT™ проектировался для:

- Приемки окатанного свинцового концентрата непосредственно с существующего окатывателя, находящегося сверху агломашины;
- Подачи отлитых брикетов богатого свинцом шлака непосредственно в существующую тару, предназначенную для загрузки агломерата в шахтные печи;
- Подачи газа с высоким содержанием серы непосредственно на существующую сернокислотную установку Haldor-Topsoe через тот же самый газоход и через те же дымососы, что и для агломашины;

- Удаленного контроля посредством DCS за существующими загрузочными конвейерами и питателями бункеров флюсов агломашины, которые прежде никогда не были автоматизированы.

Необходимость соответствия всем этим условиям сделала проектирование трудной задачей, требующей больших затрат времени. Однако строгое соответствие этим четырем целям рассматривалось как необходимое условие для сокращения до минимума времени остановки в связи с переходом с эксплуатации агломашины на эксплуатацию свинцовой ISASMELT™ печи, а также для минимизации влияния на процессы вверх и вниз по потоку данного сложного действующего объекта. Схема свинцового ISASMELT™ завода и основные точки перехода материалов представлены на Рисунке 1.

Ключевой характеристикой плана завода на Рисунке 1 является относительно небольшая занимаемая площадь. Размер площадки был обусловлен ее инфраструктурой, и при проектировании приходилось учитывать эту трудность. На территории УКМК было только одно подходящее место, где можно было бы соблюсти все четыре условия – место демонтированных 3х агломашин. Территория площадки была ограничена следующими сооружениями:

- на востоке - участком дробления оборотного агломерата;
- на западе - скиповым подъемником для агломерата;
- на севере - галереей загрузочного конвейера агломашины, и
- на юге - галереей ковшового транспортера оборотного агломерата.

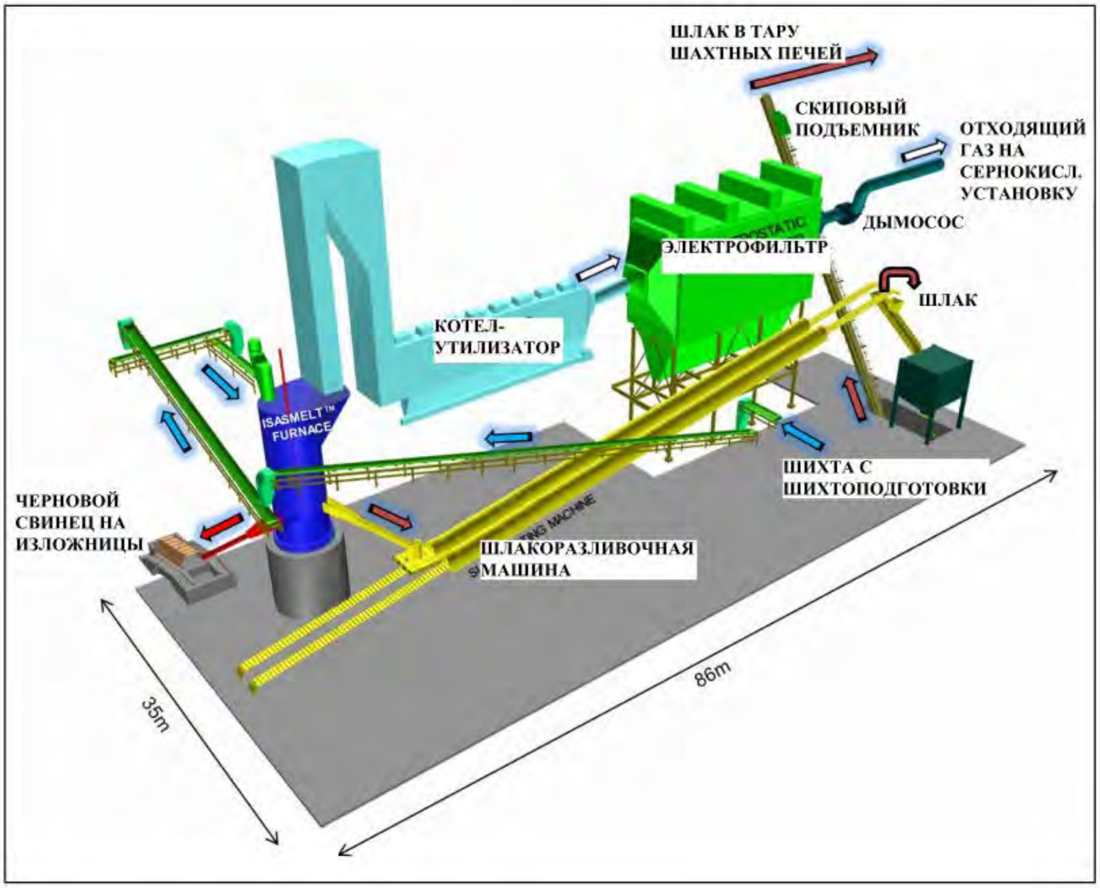


Рисунок 1: Свинцовый завод ISASMELT™ на УКМК

Каждая из установок вокруг площадки должна была продолжать работу на протяжении всего периода строительства и пусконаладочных работ. Каждый свободный квадратный метр был распланирован и использован для нового плавильного цеха площадью 86 x 35 метров. Плотность застройки в схематической форме представлена на Рисунке 2, где новое (коричневое) здание контрастирует со старыми (светло-зелеными) конструкциями. Компактная планировка завода и перекрестная проверка особенностей площадки проводилась в основном собственными инженерными силами Казцинка при содействии компании Engineering Dobersek GmbH.

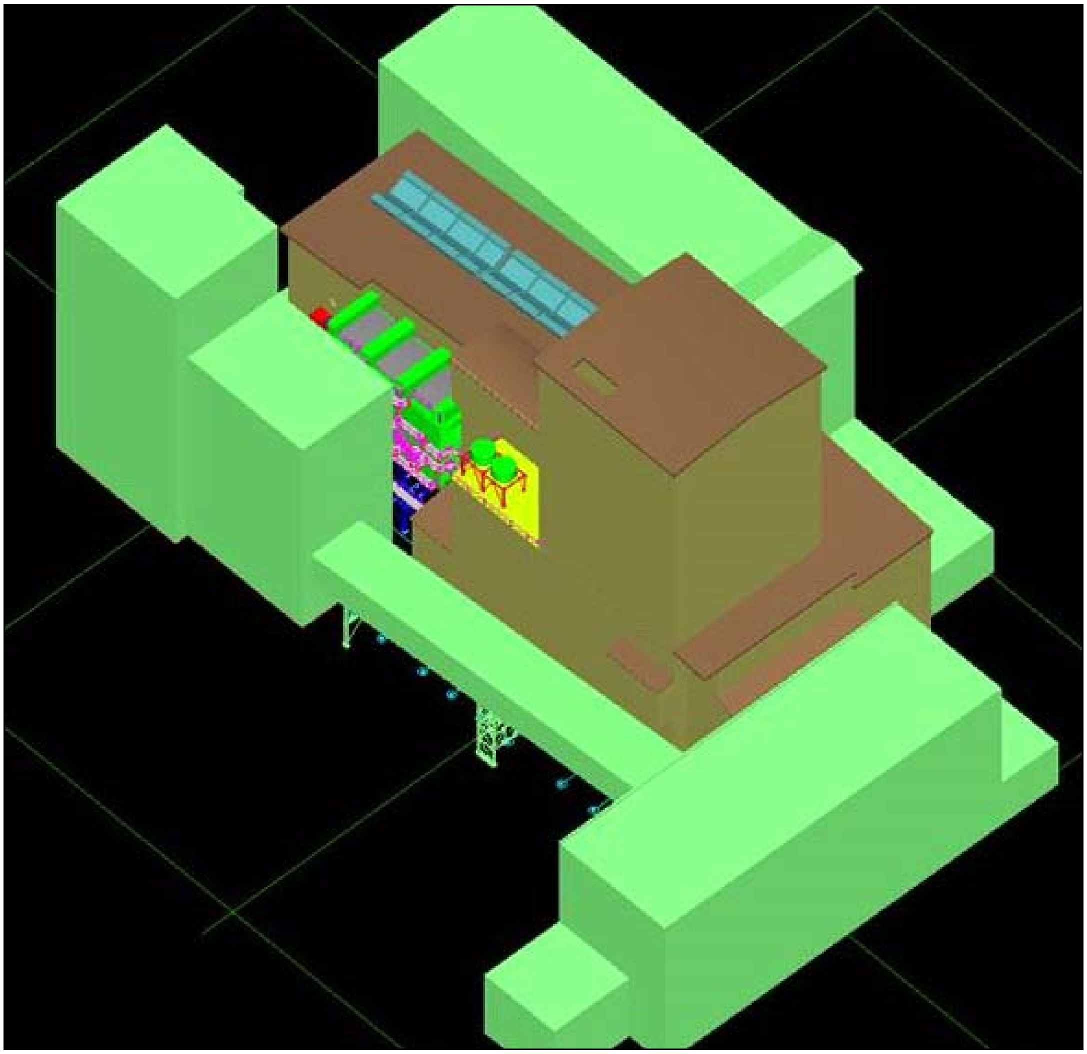


Рисунок 2: Свинцовый завод ISASMELT™ на УКМК

Демонтаж и реконструкция

При проектировании завода учитывалась сложность интеграции с существующим производством, включая частичный демонтаж имеющихся конструкций. Задача была бы упрощена, если бы полнстью демонтировались все конструкции (см. Рисунок 3), но это было невозможно. Выборочные участки на верхних этажах многоэтажных зданий должны были иметь опору, а оборудование - продолжать работу, в то время как производился демонтаж нижних этажей под новое оборудование. Демонтаж необходимо было проводить аккуратно, по частям.

Рисунок 3: Строительная площадка Свинцового завода ISASMELT™ на УКМК

Строительные работы

После частичного демонтажа конструкций вокруг и расчистки достаточного пространства началось осторожное сооружение завода. Земляные работы по подготовке новых фундаментов производились небольшими механизмами, иногда вручную, чтобы обеспечить максимальную маневренность вокруг опор существующих конструкций. Фундамент печи ISASMELT™ отливался одним из первых, он представлен на Рисунке 4, на котором также частично видны опоры нескольких других конструкций по внешнему краю земляных работ.

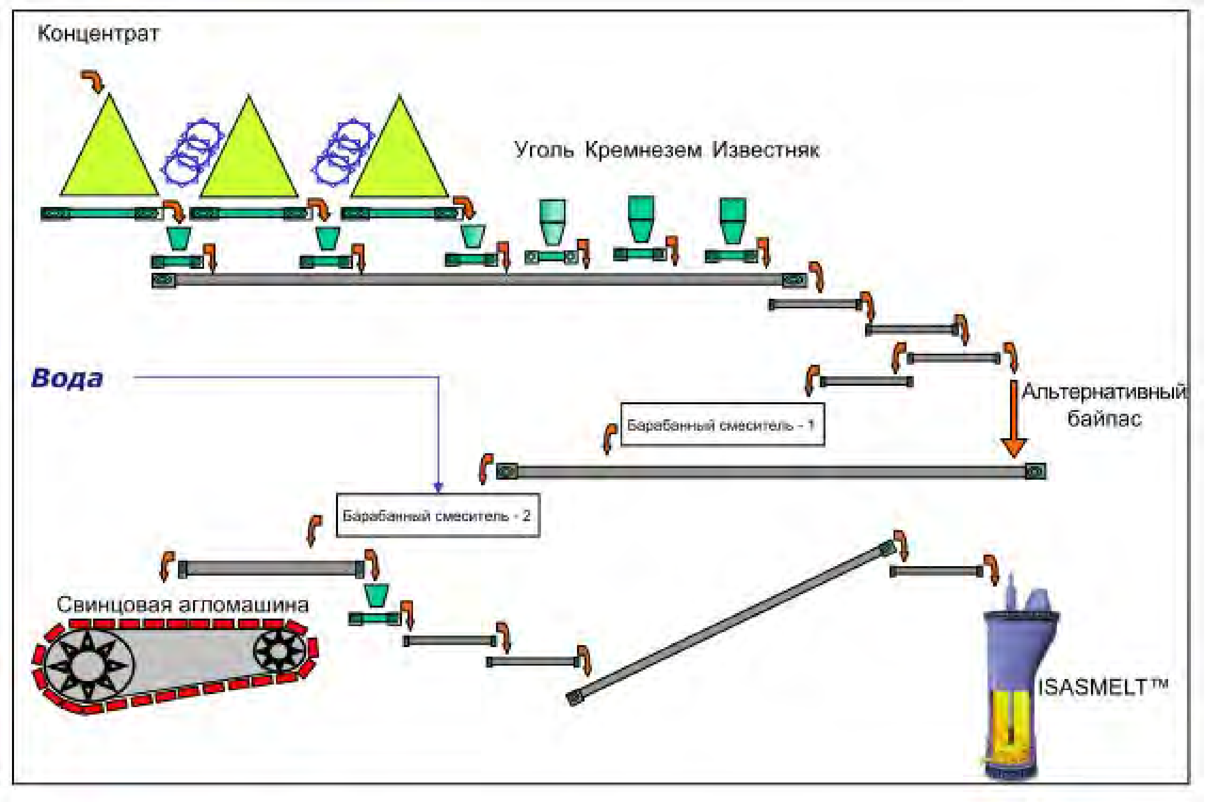
Рисунок 4: Монтаж свинцовой печи ISASMELT™ на УКМК

Подключение инженерных сетей и транспортировка материалов

Наиболее трудной задачей проекта свинцового завода ISASMELT™ была организация подключения инженерных сетей и транспортировки материалов. Модернизация системы подачи шихты стала отличным примером трудностей, которые пришлось решать. Необходимо было сохранить систему подачи шихты на агломерационную установку и перепрофилировать ее для подачи в свинцовую печь ISASMELT™. Однако оборудование агломерационной установки проектировалось для непосредственного управления, в то время как на свинцовом заводе ISASMELT™ планировалось внедрить автоматизированное дистанционное управление единицами оборудования. Таким образом, каждую из единиц оборудования системы подачи шихты на агломерационную установку нужно было перепроектировать и автоматизировать, прежде чем использовать для новых целей.

Примерно за 3 месяца до пусконаладочных работ на свинцовом заводе ISASMELT™ все оборудование системы загрузки свинцового агломерата (см. Рисунок 5) уже было автоматизировано, а электрические подключения производились во время плановой 8-часовой остановки на техническое обслуживание. С этого момента оборудованием дистанционно управлял оператор из помещения, которое вскоре стало операторной свинцового завода ISASMELT™.

Рисунок 5: Общая система подачи шихты на свинцовую агломашину и на свинцовый завод ISASMELT™.



Обучение операторов

Операторы прошли как практическое, так и теоретическое обучение до начала горячих пусков. Обучение на предприятиях Австралии и Китайской народной республики включало основы эксплуатации ISASMELT™ печей. Обучение на собственной площадке состояло из отработки эксплуатационных задач, процедур техники безопасности на рабочем месте и тестирования на знание теоретических аспектов эксплуатации завода.

Неожиданным и полезным преимуществом от автоматизации и подключения к электросети на раннем этапе стало то, что операторы получили возможность в течение многих недель нарабатывать опыт использования автоматизированной системы управления заводом еще до ввода ISASMELT™ печи в эксплуатацию.

Горячие пуски

Печь была разогрета и готова к эксплуатации к концу августа 2012 года. Плавка началась в сентябре того же года. В течение следующих нескольких месяцев предстояло решить еще несколько проблем.

Загрузка шихты для запуска

По причине сложных и продолжительных работ по проектированию завода на территории действующего предприятия с начала проектирования до запуска ISASMELT™ завода прошло около 6 лет. За эти 6 лет Казцинк ввел в эксплуатацию новые активы, разрабатывающие свинцовые месторождения, вывел из эксплуатации старые, внес изменения во внутренние рециркуляционные потоки завода и существенно изменил тип шихты, которую предполагалось перерабатывать в ISASMELT™ печи. Комбинированная шихта (без учета флюсов и топлива) на момент ввода в эксплуатацию завода ISASMELT™ в сентябре 2012 года имела состав, представленный в Таблице 2 (среднее значение первых 10 штабелей смешанной шихты по 2500 тонн каждый).

Таблица 2: Состав шихты свинцового ISASMELT™ завода при запуске

Zn (%)	Pb (%)	Cu (%)	S (%)	Fe (%)	SiO_2 (%)	C (%)	As (%)	Sb (%)	H_2O (%)
7,4	30,7	2,66	17,2	12,2	8,5	0,67	0,66	0,22	10,9

Основными отличиями проектного состава шихты, представленного в Таблице 1, и реального состава, представленного в Таблице 2 были:

- Содержание свинца ниже примерно на 12 процентных точек
- Содержание мышьяка выше примерно на 0,3 процентных точки
- Содержание влаги ниже примерно на 4 процентных точки
- Наличие органического углерода в некотором количестве, хотя его присутствие изначально вообще не предполагалось

Данные отличия создавали трудности в последующий период эксплуатации.

Наращивание объемов производства

Впервые завод ISASMELT™ достиг своей проектной суточной производительности через 5 недель после начала плавки. Однако поддержание проектной производительности в течение периодов более длительных, чем несколько последовательных дней, не представлялось возможным в течение нескольких месяцев, пока не были оптимизированы вспомогательные единицы оборудования – шлакоразливочная машина, промежуточная мульда, КУ ЭФ система подачи шихты, газоходный тракт и тд.

Операции выпуска и розлива шлака

Розлив шлака - это непрерывная операция, требующая непрерывной подачи расплавленного шлака, которая достигается благодаря специально разработанному ISASMELT™ процессу выпуска. Выпущенный шлак распределяется на каждую из двух лент проточной машины розлива шлака, спроектированной и поставленной Ormeto Yumz. Обнаружилось, что эксплуатационные свойства шлакоразливочной машины зависят от температуры выпускаемого шлака.

Частично по причине более низкого содержания свинца в поступающей шихте свойства выпускаемого шлака отличались от ожиданий поставщика: плотность шлака была меньше, а его температура - выше. Это в свою очередь означало, что шлакоразливочная машина работала на максимальной скорости, повышая температуру брикетов шлака на разгрузочном конце разливочной машины. Изначально возникли некоторые трудности с извлечением шлаковых брикетов из изложниц. Действия по устранению данной проблемы включали:

- Изменения в конструкции наковальни, о которую ударяются изложницы рядом с головным барабаном шлакоразливочной машины.
- Установку системы нанесения средства, облегчающего выемку из изложниц, в хвостовой части шлакоразливочной машины, на выходе из которой принимается разлитый шлак.
- Увеличение в два раза размера ячеек с шлаковыми брикетами на каждом лотке для изложниц.

После завершения этих действий выемка шлака из изложниц стала более надежной.

Переработка пыли

Масса пыли, собираемой из отходящих газов печи ISASMELT™ котлом-утилизатором (КУ) и электрофильтром (ЭФ), как правило, составляла 7% от общей массы поступающей шихты. Типичная частица пыли имела диаметр около 10-20 μм, а скопления пыли крепко прилипали к стальным конструкциям. Размер КУ был рассчитан таким образом, чтобы соответствовать ожидаемым потерям при теплопередаче, а липкая пыль не влияла отрицательно на сам КУ.

Однако работа ЭФ нарушалась скоплением пыли рядом с входным отверстием отходящих газов и неэффективностью работы молотков по выбиванию пыли на данном участке. Удаление внутренних выступов позволило сократить точки, на которых накапливался пылевой настыль. Этой меры вместе с повышением температуры поступающего отходящего газа на 50 °C и улучшением конструкции молотков было достаточно для того, чтобы ЭФ заработал полноценно.

КУ работал на существенно сокращенном объеме отходящего газа по сравнению с проектными значениями. В основном, это было следствием того, что реальная шихта имела содержание влаги на 4 процентные точки ниже, чем проектные значения, и таким образом необходимость в сжигании топлива и генерировании отходящего газа была существенно сокращена. Фактически обычная работа печи зачастую была автогенной, а КУ получал примерно на 30% меньше газа, чем допускает его проектная мощность. Испарители конвекционной секции КУ постепенно удалялись, пока не была достигнута оптимальная передача тепла, благодаря чему температура ЭФ поддерживалась ниже допустимой максимальной температуры на входе, но при этом она не приближалась к точке кислотной росы.

Эксплуатация шахтных печей

Казцинк продолжал работу в рамках проекта модернизации, исходя из предположения, что существующих свинцовых шахтных печей будет достаточно для переработки шлака, получаемого на свинцовом заводе ISASMELT™, при этом капитальные затраты будут минимальны. Это предположение впоследствии подтвердилось. Мгновенная мощность печи ISASMELT™ по получению шлака примерно в два раза больше мгновенной мощности шахтной печи Казцинка по плавке шлака, поэтому большую часть времени в эксплуатации находится две шахтные печи.

Шлак ISASMELT™ обычно содержит меньше CaO, чем требуется для соответствия целевым значениям химии шлака шахтных печей. Следовательно, начиная с момента загрузки шлака из печи ISASMELT™, добавление известнякового флюса в загрузку шахтных печей стало стандартной эксплуатационной практикой. Это не вызвало каких-либо затруднений.

Одним из неожиданных аспектов реализации проекта стало то, что перепрофилирование работы шахтных печей с переработки агломерата на переработку шлака осуществлялось не одномоментно. В шахтных печах успешно переплавлялся штабелированный материал либо из одного, либо из другого аппарата, а также в течение многих месяцев перерабатывались шлак и агломерат различного состава в смеси, варьирующейся от 0-100%. Это дало хорошую возможность для оценки производительности печи при использовании шлака ISASMELT™ в качестве загружаемого материала по сравнению с использованием агломерата. Общие результаты следующие:

- В шахтной печи используется меньше кокса на каждую единицу загрузки при плавке шлака по сравнению с плавкой агломерата. Снижение произошло примерно на 2 процентных точки.
- Добавление сульфидирующего вещества в загрузку шахтной печи необходимо для обеспечения перехода избыточной меди в штейновую фазу, а не в нежелательную шпейзовую.

Период после пусконаладочных работ

На первом году после начала эксплуатации свинцового завода ISASMELT™ Казцинк принял решение об определении технологической гибкости новой печи. Специалисты подготовили специальную шихтовую смесь, примерный состав которой представлен в Таблице 3. Целью испытания было сравнение эксплуатации печи ISASMELT™ при производстве чернового свинца и шлака с эксплуатацией при производстве только шлака (что является типичным способом эксплуатации в соответствии с процессом, описанным в [2]).

Таблица 3: Состав шихты во время испытательного периода

Zn (%)	Pb (%)	Cu (%)	S (%)	Fe (%)	SiO_2 (%)	As (%)	H_2O (%)
3,9	58,9	0,3	14,5	6,7	3,9	0,1	10,4

Испытания прошли успешно, а устройства выпуска и розлива чернового свинца были введены в эксплуатацию и оценены по сравнению с проектными целями.

Первая замена футеровки производилась через 24 месяца после первого разогрева печи. Было отмечено хорошее состояние футеровки большей части зон печи, также были локализованы зоны износа и предприняты предупредительные меры для проведения второй кампании.

Компания прикладывает постоянные усилия для оптимизации спецификации материалов для изготовления частей вновь используемой системы отходящих газов агломерационной установки, так как подача отходящих газов ISASMELT™ (содержащих сокращенный объем газа и более сильную концентрацию SO_3) привела к ускоренной кислотной коррозии по сравнению с предыдущим опытом.

Перспективы на будущее

Внедрение технологии ISASMELT™ на УКМК стало самым последним из длинной череды усовершенствований за последние 70 лет эксплуатации на данном почетном предприятии. Теперь, когда технология успешно внедрена в схему УКМК, следующей задачей становится рассмотрение возможности модернизации шахтных печей либо экономическая привлекательность их замены на современную технологию. В данный момент продолжается оценка возможных вариантов. Было признано, что наличие свинцовой печи ISASMELT™ в данный момент открывает новые возможности для развития, которых ранее не существовало.

Выводы

Свинцовый завод ISASMELT™ был успешно внедрен в плавильную технологическую схему Усть-Каменогорского металлургического комплекса. Новый завод строился и вводился в эксплуатацию на территории действующего производства, при этом большое внимание уделялось оптимизации существующих плавильных активов. Эксплуатация началась с подачи шихты, которая отличалась от первоначального технического задания, но на технологию плавки это отличие существенно не повлияло. В данный момент завод перерабатывает разнообразную шихту, содержание свинца в которой варьируется в диапазоне 30 – 60 % при мгновенной производительности, превышающей на 20% исходные проектные параметры. Внедрение технологии ISASMELT™ стало самым последним из длинной череды усовершенствований металлургического производства на КАЗЦИНКе – но компания на этом не останавливается и уже сейчас рассматривает уже другие технологии для дальнейшей модернизации.

Благодарности

Авторы желают выразить благодарность ТОО «Казцинк» за разрешение использовать данные и фотографии для публикации. Авторы хотели бы выразить благодарность компании Glencore Technology Pty Ltd и ТОО «Кацинк» за разрешение публикации данной статьи.

Использованная литература

[1] ERRINGTON, B., ARTHUR P., WANG, J. & DONG, Y. (2005): “The ISA-YMG Lead Smelting Process”, Proceedings of the International Symposium on Lead and Zinc Processing, Kyoto, Japan, October 2005, T. Fujisawa et al., Eds., MMIJ, pp. 581-599.

[2] ERRINGTON, B., FEWINGS, J.H., KERAN, V.P., & DENHOLM, W.T. (1987): “The ISASMELT Lead Smelting Process”, Trans. Instn. Min. Metall., Section C, vol 96.

ПЛАВКА КАЗАХСТАНСКИХ КОНЦЕНТРАТОВ В МЕДНОЙ ПЕЧИ ISASMELT™, г. УСТЬ-КАМЕНОГОРСК

A.S. Burrows[1] и G.R.F. Alvear F.[2]
Xstrata Technology
[1]*Level 10, 160 Ann St. Брисбен, Квинслэнд 4000, Австралия*
[2]*Level 10, 700 West Pender St. Ванкувер, Британская Колумбия, Канада V6C1G8*
aburrows@xstratatech.com

А.Т. Тыныбаев
ТОО «Казцинк»
ул. Промышленная, 1
070002 Усть-Каменогорск
Республика Казахстан

АННОТАЦИЯ

В июле 2011 г. на территории Усть-Каменогорского металлургического комплекса ТОО «Казцинк», Усть-Каменогорск, Казахстан, была запущена новая медеплавильная печь ISASMELT™, в рамках проекта строительства плавильного завода на территории действующего металлургического комплекса. До этого момента, Казцинк производил преимущественно цинк и свинец, в связи с чем вся технологическая цепочка производства меди была создана заново, начиная с шихтоподготовки и заканчивая обвязкой катодов. На проектную производственную мощность вышли за 14 месяцев.

Для успешной реализации проекта, пришлось решить множеством проблем: ввоз новой технологии в Республику Казахстан, наработка опыта местным производственным персоналом при работе с ней; плавка сложных полиметаллических концентратов Восточно-Казахстанской области без ущерба для производственной мощности или качества продукции; и внедрение технологической схемы производства меди в общую схему металлургического комплекса, так, чтобы извлечь наибольшую экономическую выгоду из полиметаллического характера сырья.

В данной работе описывается, как медеплавильная печь ISASMELT™ была встроена в эту стратегию, и как согласованные действия команд компаний Казцинк и Экстрата Текнолоджи привели к успешной реализации данного проекта.

ВВЕДЕНИЕ

Ранее Казахстан входил в состав Советского Союза, Республика получила независимость более 20 лет назад. Это крупнейшая внутриматериковая страна в мире, по площади больше территории Западной Европы. Страна богата полезными ископаемыми, которые распределены по всей территории. ТОО «Казцинк» является крупным полностью интегрированным производителем цинка. Компания была основана в 1997 г. в результате слияния трех основных компаний, производящих цветные металлы, в Восточном Казахстане. Помимо цинка концентраты, получаемые на рудниках Казцинка, имеют значительное содержание свинца, меди и драгоценных металлов. Центром Восточно-Казахстанской области является Усть-Каменогорск, город с население около 350 000 человек. Усть-Каменогорск - это центр транспортировки и переработки концентратов основных металлов уже более века. С 1943 года в Усть-Каменогорске располагается металлургический комплекс, специализирующийся на производстве цинка, свинца, серебра, золота, сурьмы, висмута и различных побочных продуктов. Усть-Каменогорский Металлургический Комплекс (УК МК) является одним из активов Казцинка. В 2005 г., когда Казцинк решил построить медеплавильный завод и цех электролиза на УК МК, компания обратилась к Экстрата Теклолоджи для предоставления технологий ISASMELT™ и ISA PROCESS™ для нового медного завода.

Несмотря на богатый опыт УК МК в плавке металлов, ранее медь не являлась основным продуктом производства. Около 5000 т/г, черновой меди производилось в качестве побочного продукта на свинцовом заводе. Для медеплавильного цеха и цеха электролиза понадобился абсолютно новый завод. По масштабам новый завод схож с проектом нового строительства, если не считать того, что его расположение в середине крупного металлургического комплекса имело все трудности, связанные с расширением существующего проекта. Концепция нового медного завода заключалась в том, что номинальная производительность должна составлять 70 000 т/г катодной меди, завод должен перерабатывать полиметаллические медные концентраты и ряд побочных продуктов от рафинации цинка и свинца, допускать колебание концентраций малых элементов и иметь возможность к расширению в будущем. ISASMELT™ была логичным выбором для первичной медеплавильной печи, которая могла соответствовать всем этим требованиям. В течение последующих 6 лет многое изменилось в процессе оценки, расположения, проектирования, строительства и ввода в эксплуатацию медного завода, но первоначальные цели проекта остались неизменными.

При производительности всего 70 000 т/г меди медеплавильный завод УК МК является небольшим по мировым стандартам. Он имеет высокое отношение постоянных затрат к переменным. Для того, чтобы проект приносил прибыль, основными рабочими характеристиками завода должно быть следующее: нормальная работа при колебаниях загрузки полиметаллических концентратов, которые поставляются с местных рудников Казцинка, извлечение драгоценные металлы, содержащиеся в загрузке, отмена оплаты транспортных расходов и штрафов, связанных с продажей концентратов на сторону.

ОБСУЖДЕНИЕ

Описание завода

Медеплавильный завод УК МК имеет традиционную технологическую схему. Плавка концентрата осуществляется в медеплавильной печи ISASMELT™, штейн конвертируется в одном из двух P-S конверторов, и огневое рафинирование проводится в одной из двух анодных печей с последующим розливом анодов. Сернокислотный завод может принимать газ от печи ISASMELT™ и от одного из P-S конверторов, находящегося на продувке. Технологическая схема представлена на Рисунке 1.

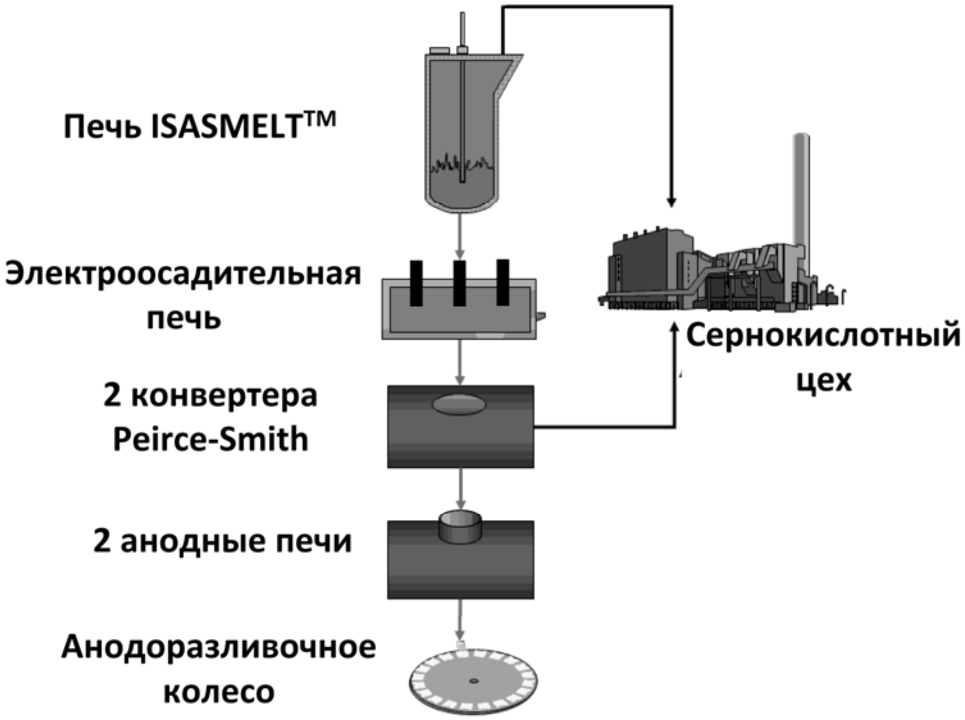


Рисунок 1: Принципиальная технологическая схема Медеплавильного завода

По данной технологической схеме удаление примесей осуществляется при помощи специального выпуска шлака из P-S конвертора во время продувки на медь, и через пыли, образующиеся от плавильных печей. После проведения расчетов было определено, что прямая рециркуляция пылевых потоков является нецелесообразной из-за накопления летучих примесей. Вместо этого была проведена модификация существующего металлургического объекта для проведения на нем выщелачивания пыли, и, таким образом, удаления целевых элементов и производства чистой цементационной меди, которая затем может перерабатываться на плавильном заводе.

Шихта плавильного завода

Среди стран бывшего Советского Союза, марка МООК представляет самое высокое качество реализуемой катодной меди, технические характеристики которой соответствуют марке А Лондонской биржи металлов. Исходными данными для проектирования медного завода УК МК и цеха электролиза меди было производство катодов марки МООК из концентрата, содержащего значительное количество свинца, мышьяка, цинка, висмута и сурьмы. В литературных источниках было отмечено [1], что возможность печи ISASMELT™ в плане допуска и удаления летучих примесных элементов будет очень целесообразной для медеплавильного завода Казцинка. На этапе проектирования был рассмотрен переход примесей в промежуточные продукты и рециркуляционные потоки. Анализ основных элементов «проектной» шихты завода приведен в Таблице 1.

Таблица 1 – Проектный состав шихты медеплавильного завода Казцинка

	% по весу							
	Cu	Pb	Zn	Fe	S	SiO2	Sb	As
Проект	25,0	2,68	3,14	24,85	31,74	5,11	0,16	0,55

Медеплавильный завод ISASMELT™

Медеплавильный завод ISASMELT™ на УК МК включает подачу шихты в медеплавильную печь ISASMELT™ и транспортировку отходящих газов и расплавленных продуктов, получаемых в печи. Технологическая схема, схематично изображенная на Рисунке 2, схожа с заводами, описание которых было выполнено ранее [2, 3]. Различные концентраты смешиваются в цехе шихтоподготовки, и смешанный концентрат загружается в один из двух бункеров для концентрата с помощью мостового грейферного крана. Из питательных бункеров концентрат загружается в печь ISASMELT™. Питательные бункеры также используются для точной подачи взвешенного количества угля, оборотов, известнякового флюса и кремнеземистого флюса на сборный конвейер, а затем в печь ISASMELT™. Влажная шихта загружается с конечного питательного конвейера через вентиляционное отверстие в своде печи, а затем и в перемешиваемую ванну расплава. Свод печи является составной частью котла-утилизатора.

Подача технологического воздуха и технического кислорода (чистота 98%) на фурму ISASMELT™ контролируется автоматически с помощью Распределительной системы управления (DCS), которая используется по всему плавильному заводу. Погружение фурмы в ванну с жидким шлаком также контролируется автоматически системой DCS, так что наконечник фурмы всегда находится в оптимальном положении в расплавленной ванне независимо от ее уровня.

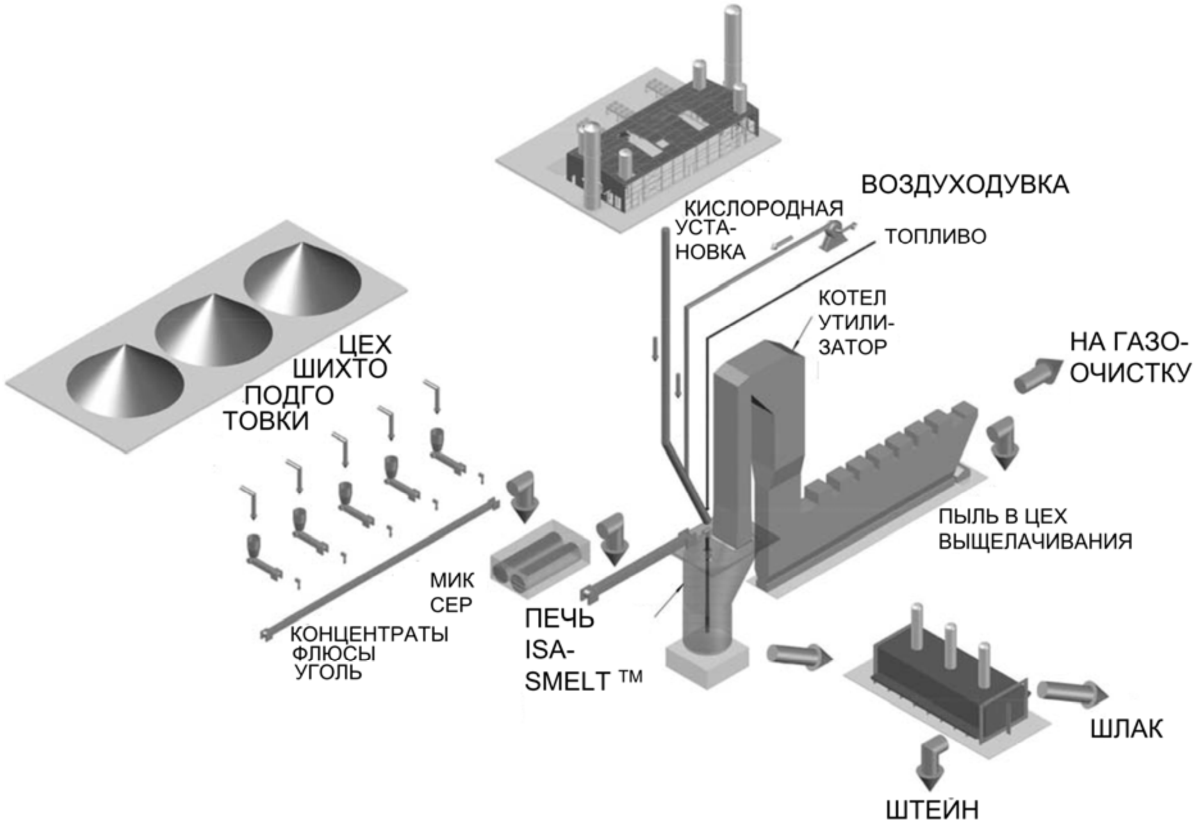


Рисунок 1: Схема участка плавильной печи ISASMELT™.

Штейн и шлак, получаемые в печи ISASMELT™ периодически выпускаются из одиночного водоохлаждаемого выпускного блока, располагающегося в основании печи. Эти материалы перетекают в электроосадительную печь, находящуюся возле печи ISASMELT™. Проектное содержание штейна составляло 60%, но Казцинком было определено, что оптимальный баланс производства может быть достигнут приблизительно при 55-58 %Cu в штейне.

Отходящие газы процесса покидают печь через отверстие в своде печи ISASMELT™ и переходят в котел-утилизатор. Охлажденный газ выходит из котла-утилизатора при температуре приблизительно 350°C, а затем очищается от пыли в электрофильтре.

Переход к новой технологии

В пирометаллургии изменения, вносимые в работу заводов, технологические схемы и принятые практики, обычно протекают медленно. Переход к новой технологии, инновации и внедрение новых технологий всегда сопряжено с техническими трудностями. Важным преимуществом технологии ISASMELT™ является то, что она без труда может использоваться в новых инновационных областях применения заказчиками по всему миру.

Управленческая команда Медного завода Казцинка подготовилась к выводу плавильного завода на проектную мощность и эффективному внедрению незнакомой технологии в эксплуатационную и управленческую практику УК МК. В случае с медным заводом ISASMELT™ процессу перехода к новой технологии способствовали значительные вложения времени и человеческих ресурсов в обучение эксплуатационного и ремонтного персонала, аналогично другим лицензиатам ISASMELT™[4]. Для персонала Казцинка проводилось теоретическое обучение в учебных классах, практическое обучение на других лицензированных заводах ISASMELT™, а также практическое обучение на площадке.

Практическое обучение в течение трех месяцев было проведено на медеплавильном заводе Xstrata Copper в Маунт-Айза, Австралия. Там персонал Казцинка прошел полноценное практическое обучение под руководством опытного технологического и ремонтного персонала. Это позволило специалистам Казцинка, находящимся на обучении, стать компетентными операторами и понять основы строения печи ISASMELT™ и приемы ее эксплуатации. В разгар программы обучения на Маунт Айза делегация от Казцинка, в состав которой входили обучающиеся, руководители и переводчики, насчитывала в общей сложности 45 человек. Потребовалась тщательная организация для управления их графиком работ и для гарантии того, что обучение является как эффективным, так и безопасным.

Дополнительно к обучению на Маунт Айза персонал Казцинка посетил медеплавильный завод Муфулира компаниии Mopani, использующий технологию ISASMELT™. Данный завод находится на Медном поясе в Замбии. Привлекательной особенностью данного обучения являлось то, что один и тот же поставщик (SMS Siemag AG) был выбран для поставки электроосадительной печи как на УК МК, так и на плавильном заводе Муфулира. Таким образом, обучаемые специалисты Казцинка имели возможность получить опыт интегрированной эксплуатации печи ISASMELT™ с электроосадительной печью с тремя электродами.

Помимо полномасштабного обучения на заводе инженеры систем управления Казцинка провели несколько недель со специалистами Xstrata Technology в Брисбене, Австралия, работая совместно над конфигурацией и тестированием Распределенной системы управления (DCS), которая лежит в основе работы ISASMELT™. Обучение, пройденное ими за это время, дало гарантию, что местная группа технического обслуживания сможет в полной мере участвовать в процессе запуска завода.

В последние три месяца, предшествующие запуску завода, когда медеплавильный завод ISASMELT™ был готов к работе, но незначительные задержки оказывали влияние на оборудование вниз по потоку и препятствовали запуску, персонал Казцинка и Xstrata Technology организовывали многочисленные обучения на площадке для операторов медного завода ISASMELT™. По сложности данные обучения отличались, начиная от отдельных единиц оборудования до всего завода, включая подвижный транспорт и контроль всех входящих потоков материалов.

В течение всего периода проектирования, закупок, строительства, установки, запуска и выхода на рабочие режимы для персонала Казцинка было полезно посещать лицензионные семинары ISASMELT™, которые проводились в Индии, Перу, Бельгии и Замбии, а также обмениваться информацией с предприятиями-аналогами и такими же обладателями лицензий в других странах и на других предприятиях. Запуск нового плавильного завода всегда представляет собой комплексную и трудную задачу. Спокойствие и уверенность можно почерпнуть, обсуждая с другими заводами их прошлый опыт и получая от них ценные советы.

Запуск завода

Для абсолютно нового медеплавильного завода и цеха электролиза, особенно если они находятся в холодном климате, запуск завода – это не единичное событие, которое происходит в определенный день. По необходимости работы занимают много недель, так как в эксплуатацию вводятся различные части технологической схемы. Сезонные установки и оборудование вводятся в эксплуатацию по отдельному графику, а именно, навесы для оттаивания концентрата, нагревательные приборы и вентиляторы зданий, установки парового обогрева и оборудование подачи топлива. Концентрат доставляется, хранится и смешивается за недели до его использования. Запас анодов должен быть накоплен в Цехе электролиза до того, как на все секции ванн будет подано напряжение.

Эксплуатационная группа УК МК внедрила дальновидный метод запуска всего сернокислотного завода (проект SNC Lavalin) до начала запуска медеплавильного завода. Это стало возможным при помощи подсоединения газохода к существующим заводам плавки цветных металлов и подачи газов с высоким содержанием SO_2 в течение нескольких месяцев, предшествующих запуску медеплавильного завода. Подобным образом воздухоразделительная установка (спроектированная Криогенмаш) была запущена и работала на полную мощность задолго до того, как она понадобилась для Медного завода.

Отчасти благодаря стараниям и планированию во время пуско-наладочных работ, УК МК достиг 100% проектного проплава на медеплавильном заводе ISASMELT™, который измерялся каждую 8-часовую смену, в течение нескольких дней после первой загрузки шихты в печь ISASMELT™. Однако работа по выходу завода на проектную мощность на этом не заканчивается. Следующий год работы был затрачен на достижение согласованной и надежной работы всех систем медеплавильного завода. Это то, что необходимо для стабильной работы всего завода. Персонал Экстрата Текнолоджи оказывал определенную помощь в данной работе, когда бы к нему не обратились. В течение нескольких первых недель после принятия решения о разогреве печи ISASMELT™ Экстрата Текнолоджи имела возможность оказывать помощь всем технологическим сменам, имея на площадке по два человека на смену в течение 24 часов в день. Необходимость в оказании помощи значительно сократилась, когда эксплуатационная группа приобрела опыт и компетентность при выполнении своих задач. Через несколько месяцев команда специалистов УК МК работала самостоятельно на профессиональном уровне и продолжала оптимизировать и улучшать работу медного завода.

Эксплуатация

Эксплуатационная статистика стандартного месяца работы на медном заводе ISASMELT™, взятая через 7 месяцев после запуска, приведена в Таблице 2. При средней месячной производительности, превышающей 80% проектной мощности после нескольких месяцев работы, очевидно, что подбор большей части методов эксплуатации, выбор оборудования, установка и запуск были выполнены верно и основательно.

Таблица 2 – Эксплуатационная статистика медеплавильного завода ISASMELT™, Казцинк

	Ед.	Февраль 2012 г.
Готовность	%	89,4
Концентрат	т/ч	32,0
Кремнезем	т/ч	7,7
Обороты	т/ч	2,6
Уголь	т/ч	0,4
Известняк	т/ч	1,3
Влагосодержание шихты	т/ч	9
Срок службы фурмы	дни	6
Cu в шихте	%	21,7
Cu в штейне	%	57,0
Поток на фурму ISASMELT™	нм^3/ч	16500
O_2 в воздухе фурмы	%	58

Отклонения в шихте плавильного завода

Одним из главных преимуществ, выявленных Казцинком в применении технологии ISASMELT™, была возможность перерабатывать медный концентрат с широким диапазоном составов. На самом деле, состав щихты, имевшейся в наличии при запуске плавильного завода (смотрите Таблицу 3), существенно отличается от проектных данных по некоторым элементам. Отличие включает в себя пропорциональное увеличение содержания свинца и сурьмы в фактическом концентрате более, чем на 50%, и пропорциональное снижение кремнезема в фактическом концентрате более, чем на 50%, по сравнению с проектным концентратом.

Таблица 3 – Сравнение фактически загружаемой шихты медного завода Казцинка с проектными данными

	% по весу							
	Cu	Pb	Zn	Fe	S	SiO2	Sb	As
Проект	25,0	2,68	3,14	24,85	31,74	5,11	0,16	0,55
Факт	25,68	4,35	3,53	26,65	32,92	2,39	0,25	0,49
Абсолютное отклонение	+0,68	+1,57	+0,39	+1,80	+1,18	-2,72	+0,09	-0,06
Относительное отклонение	+3%	+59%	+12%	+7%	+4%	-53%	+56%	-10%

Проблема более низкого уровня кремнезема в медном концентрате была решена посредством простого увеличения кремнезема с добавляемым флюсующим материалом. Повышенный уровень сурьмы в концентрате усложнил процесс производства катодной меди марки МООК.

Повышенный уровень свинца в концентрате, который был достаточно высоким с самого начала, потребовал внесения модификаций в технологическую схему и разработки стратегий управления после завершения процесса запуска. Основной стратегией для устранения последствий более высокой загрузки свинца является организация дополнительного отвода шлака в конце каждой медной продувки P-S конвертера и его измельчение и флотация. Данный подход позволил избежать накопления возвращающегося в цикл свинца, что происходило при подаче шлака в электроосадительную печь для электротермического восстановления.

Сочетания этих мер было достаточно для предотвращения проблем, таких как самопроизвольное образование чернового свинца после химического восстановления шлака в электроосадительной печи, и на плавильном заводе был освоен процесс переработки медных концентратов с высоким содержанием свинца без больших затруднений. В особенности в работе котла-утилизатора медеплавильной печи ISASMELT™ (проект компании Foster Wheeler) не возникало никаких проблем, связанных с составом концентрата. Опасения о том, что внутренние детали котла-утилизатора будут покрываться свинецсодержащей настылью, что уменьшит эффективность теплопередачи, существовавшие на этапе планирования проекта, оказались необоснованными. Пристальное внимание, уделенное детальному проектированию размеров, формы и профиля потоков котла-утилизатора позволило Казцинку получить надежную систему утилизации отходящих газов для медеплавильной печи ISASMELT™.

ВЫВОДЫ

Усть-Каменогорский металлургический комплекс запустил новый медеплавильный завод и цех электролиза в основном для переработки сырья с их собственных рудников и металлургических объектов. При проектировании завода планировалось, что полиметаллические концентраты будут являться основным компонентом загрузочного материала.

Персонал ТОО «Казцинк» и Экстрата Текнолоджи работали совместно над интенсивной программой обучения и пуско-наладочным работам для подготовки персонала и оборудования завода к соответствующим функциям при работе завода. Завод начал работать нормально, и продолжает свою работу. В будущем планируется расширить производительную мощность завода, и проект завода рассчитан на реализацию подобных расширений при необходимости.

БЛАГОДАРНОСТЬ

Авторы выражают благодарность ТОО «Казцинк» и Экстрата Текнолоджи за разрешение опубликовать данную статью.

СПРАВОЧНАЯ ЛИТЕРАТУРА

1. G. R. F. Alvear F, P. Arthur, P. Partington, “Feasibility to Profitability with Copper ISASMELT™”, Copper 2010, Volume 2– Pyrometallurgy I, J. Harre, Ed., GDMB, Hamburg, 2010
2. Y. Li, P. Arthur, “Yunnan Copper Corporation’s new smelter – China’s first ISASMELT”, Yazawa International Symposium on Metallurgical and Materials

Processing, Volume II – High Temperature Metal Production, F. Kongoli, K. Itagaki, C. Yamauchi, H.Y. Sohn, Eds., TMS, Warrendale, 2003

3. J. Ross and D. de Vries, “Mufulira smelter upgrade project – ‘Industry’ Smelting on the Zambian Copperbelt”, Pyrometallurgy 05, Capetown, Minerals Engineering International, 2005
4. H. Walqui, C. Noriega, P. Partington, &G.R.F. Alvear F, “SPCC’s 1,200,000 TPA Copper ISASMELT™”, Sohn International Symposium on Advanced Processing of Metals and Materials, Volume 8 – International Symposium on Sulfide Smelting 2006, F. Kongoli & R.G. Reddy, Eds., TMS, 2006

SMELTING OF KAZAKHSTAN CONCENTRATES AT UST-KAMENOGORSK USING A COPPER ISASMELT™ FURNACE

A.S. Burrows[1]& G.R.F. Alvear F.[2]
Xstrata Technology
[1]Level 10, 160 Ann St. Brisbane, QLD 4000, Australia
[2]Level 10, 700 West Pender St. Vancouver, BC, Canada V6C1G8
aburrows@xstratatech.com

A.T. Tynybayev
Kazzinc Ltd
1 Promyshlennaya St.
070002 Ust-Kamenogorsk
Republic of Kazakhstan

ABSTRACT

In July 2011 a new copper ISASMELT™ furnace was commissioned at Kazzinc Ltd's Metallurgical Complex in Ust-Kamenogorsk, Kazakhstan. It was part of a brownfield smelter project, situated inside an existing metallurgical complex. Kazzinc had up to that time been primarily a zinc and lead producer so the entire copper circuit was new, from feed blending through to cathode strapping. Within 14 months it was achieving its design production capacity.

For the project to be brought to successful fruition, numerous challenges had to be met: new technology had to be brought into the Republic of Kazakhstan and the local operating team had to become competent to operate it; the complex polymetallic concentrates of the East Kazakhstan region had to be smelted without compromising the production rate or product quality; and the copper production circuit had to be integrated into the overall metallurgical complex to derive greatest economic benefit from the polymetallic nature of the raw materials.

This paper describes how the copper ISASMELT™ furnace fitted into this strategy, and how the coordinated actions of Kazzinc and Xstrata Technology personnel brought this project to fruition.

INTRODUCTION

Formerly part of the Soviet Union, the Republic of Kazakhstan has been an independent nation for more than 20 years. It is the world's largest landlocked country by land area, being larger than Western Europe. It has a wealth of mineral resources spread throughout the country. Kazzinc Ltd is a major, fully integrated, zinc producer established in 1997 through the merger of Eastern Kazakhstan's three main non-ferrous metals companies. In addition to zinc, the concentrates from Kazzinc’s mines contain significant amounts of lead, copper and precious metals. The capital of Eastern Kazakhstan province is Ust-Kamenogorsk, a city of about 350,000 people, which has been a hub for transporting and processing base metals concentrates for more than a century. Since 1943, Ust-Kamenogorsk has also been home to a multi-smelter metallurgical complex specialising in the production of zinc, lead, silver, gold, antimony, bismuth and various other by-products. The Ust-Kamenogorsk Metallurgical Complex (UKMC) is one of Kazzinc’s assets, and in 2005 when Kazzinc wished to add a copper smelter and refinery to UKMC it approached Xstrata Technology to provide the ISASMELT™ and ISA PROCESS™ technologies for the new copper plant.

Despite UKMC’s rich smelting history, copper had not previously been a major product. About 5000 tonnes per annum (tpa) of blister copper was produced as a by-product of lead smelting. A completely new plant was required for the copper smelter and refinery. The new plant would be akin to a greenfield project in scope, except that its location amid a major metallurgical complex had all of the complications associated with a brownfield expansion project. The concept of the new copper plant was that it should have a nominal production capacity of 70,000 tpa of cathode copper, be able to treat polymetallic copper concentrates and a range of by-products from zinc and lead refining, be tolerant of minor element fluctuations, and be readily expandable in the future. For the primary copper smelting furnace, an ISASMELT™ was a logical choice to meet these objectives. During the course of the next 6 years, many things changed in the process of evaluating, locating, designing, constructing and commissioning the copper plant, but the original project objectives remained unchanged.

With a capacity of just 70,000 tpa of copper, the UKMC copper smelter was destined to be small by world standards, and subject to a high ratio of fixed:variable costs. For the project to make sense, the key performance characteristic would be that the smelter could run well on a fluctuating supply of polymetallic concentrates sourced locally from Kazzinc mines, recover the precious metals contained therein, and eliminate the transportation and penalty costs associated with sales of concentrate to third parties.

DISCUSSION

Plant Description

The copper smelter at UKMC has a conventional flowsheet. Concentrate smelting is performed in the copper ISASMELT™ furnace, matte is converted in one of two P-S Converters, and fire refining occurs in two anode furnaces, prior to anode casting. The sulphuric acid plant is able to accept gas from the ISASMELT™ furnace and one blowing P-S Converter. The flowsheet is shown schematically in Figure 1.

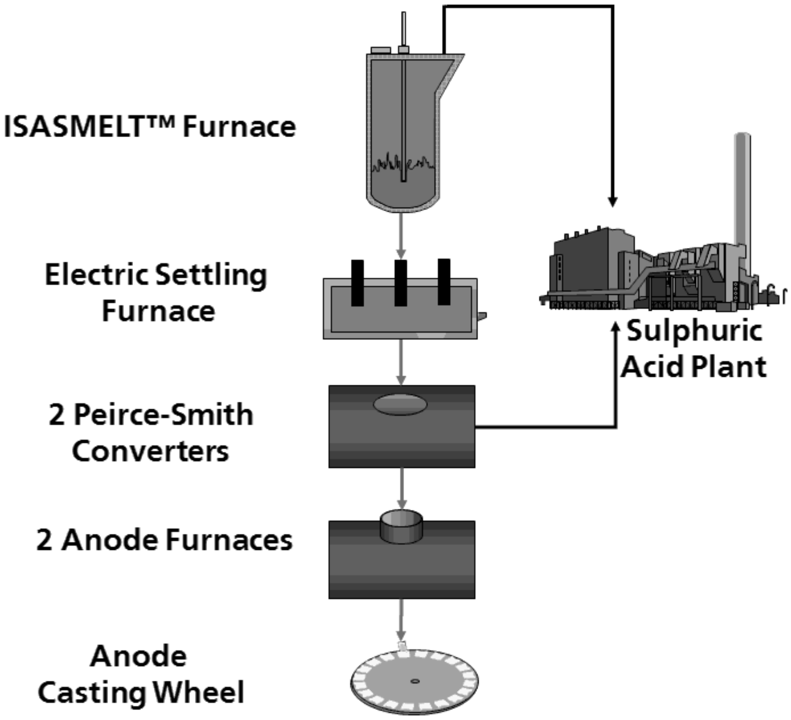


Figure 1: Schematic Diagram of Copper Smelter Flowsheet

Within this flowsheet, impurity removal was to occur via a dedicated bleed of slag from the P-S Converter copper blow, and via dusts produced from the smelting furnaces. Direct recycle of dust streams was calculated to be impractical owing to the build-up of volatile impurity elements. Instead, an existing metallurgical facility was modified to perform dust leaching duties, thereby removing target elements and producing a clean copper cement that could be recycled to the smelter.

Smelter Feed

Among countries of the former Soviet Union, the highest purity designation of traded cathode copper is referred to as MOOK grade, with technical specifications comparable to LME grade A copper. The design basis of the UKMC copper smelter and refinery was to produce MOOK grade cathode from a concentrate feed containing significant quantities of lead, arsenic, zinc, bismuth, and antimony. It has been noted elsewhere [1] that the ability of the ISASMELT™ furnace to tolerate and eliminate volatile impurity elements was of benefit for the Kazzinc copper smelter. The deportment of impurities to intermediate products and recycle streams was considered during the design phase. Assays for the major elements of the "design" smelter feed are shown in Table 1.

Table 1 – Kazzinc Copper Smelter Design Feed Composition

	Wt, %							
	Cu	Pb	Zn	Fe	S	SiO2	Sb	As
Design	25.0	2.68	3.14	24.85	31.74	5.11	0.16	0.55

Copper ISASMELT™ Plant

The copper ISASMELT™ plant at UKMC encompasses the supply of feed materials to the copper ISASMELT™ furnace and the handling of the off-gases and molten products produced in the furnace. The flowsheet, shown schematically in Figure 2, is similar to plants that have been described previously [2,3]. A variety of concentrates are blended in a blending plant, and the blended concentrate is loaded into one of two concentrate bins using an overhead grab crane. From the feed bins the concentrate is accurately dispensed to the ISASMELT™ furnace. Feed bins are also used for dispensing accurately weighed amounts of coal, reverts, limestone flux and silica flux onto a collector conveyor and then to the ISASMELT™ furnace. The moist feed mix is charged from the final feed conveyor through a ventilated port in the furnace roof and into the agitated molten bath below. The roof of the furnace is an integral part of the waste heat boiler.

Delivery of process air and tonnage oxygen (98% purity) to the ISASMELT™ lance is controlled automatically be the Distributed Control System (DCS) that is used throughout the smelter operation. Lance immersion in the molten slag bath is also controlled automatically by the DCS, so that the lance tip is always at the optimum position in the molten bath regardless of the bath level.

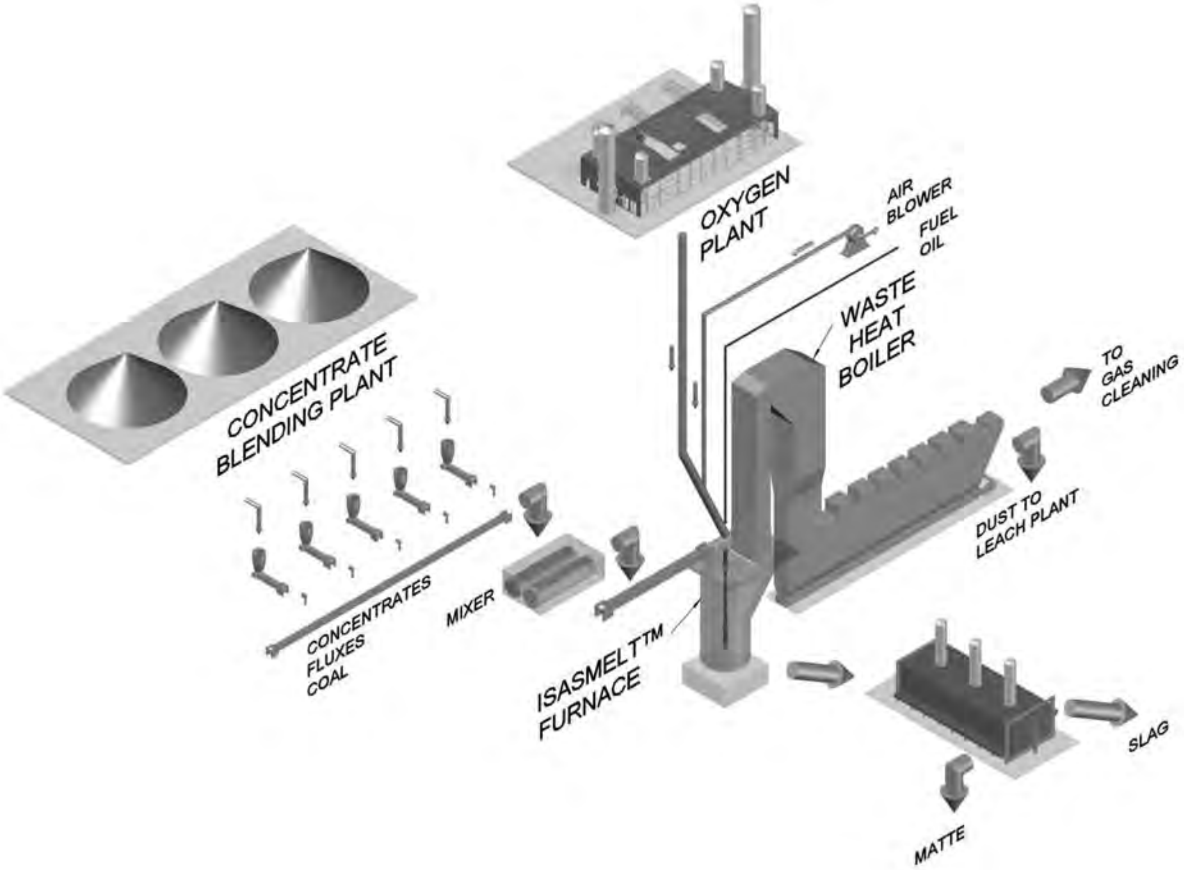


Figure 1: Schematic Diagram of Copper ISASMELT™ Plant

Matte and slag produced in the ISASMELT™ furnace are tapped intermittently from a single water-cooled tapping block at the base of the furnace. These flow into an electric settling furnace located next to the ISASMELT™ furnace. The design matte grade was 60%, but Kazzinc has found that an optimum balance of production can be achieved using approximately 55-58 %Cu in matte.

Off-gases from the process exit the furnace through a hole in the ISASMELT™ furnace roof and pass into a waste heat boiler. Cooled gas leaves the waste heat boiler at approximately 350°C and is then de-dusted in an electrostatic precipitator.

Technology Transfer

In the world of pyrometallurgy, changes to plants, flowsheets and accepted practices are usually slow. Technology transfer, innovation and implementation of new technologies always face technical challenges. An important strength of the ISASMELT™ technology has been its capacity and versatility to be incorporated into new innovative applications, by customers throughout the world.

The Kazzinc copper smelter management team made preparations to improve the ramp-up of the copper smelter, and to effectively integrate foreign technology into the operational and management practices of UKMC. In the case of the copper ISASMELT™ plant, the technology transfer process was aided by a significant investment of both time and people in the training of operations and maintenance personnel, as has been practised by other ISASMELT™ licensees [4]. Theoretical training in the classroom, practical training at other ISASMELT™ licensee plants, and in situ practical training were all employed to the benefit of Kazzinc personnel.

About three months of practical training occurred at Xstrata Copper's copper smelter in Mount Isa, Australia. There, the Kazzinc personnel received valuable hands-on training from experienced operations and maintenance staff. This enabled Kazzinc trainees to become competent operators and understand the basics of the ISASMELT™ furnace hardware and operating technique. At the peak of the Mount Isa training programme, the delegation of Kazzinc trainees, supervisors and interpreters numbered 45 in total, requiring careful organisation to manage their activity schedule and ensure that the training was both effective and safe.

Supplementary to the training at Mount Isa, Kazzinc personnel visited the copper ISASMELT™ plant at Mopani's Mufulira smelter in the Copper belt of Zambia. An attractive feature of this training exercise was that a common vendor (SMS Siemag AG) was chosen for the electric settling furnace at both UKMC and the Mufulira smelter, so Kazzinc trainees were able to develop experience in the integrated operation of an ISASMELT™ furnace with a 3-in-line electric settling furnace.

In addition to the full scale plant training, Kazzinc control system engineers spent several weeks with Xstrata Technology specialists in Brisbane, Australia to collaborate in the configuration and testing of the DCS, which is at the heart of ISASMELT™ operation. The training that they received during this time helped to ensure that the local maintenance team was able to participate fully in the plant commissioning.

In the final three months preceding smelter start-up, when the copper ISASMELT™ plant was ready to run, but minor delays affected downstream equipment and prevented start-up, Kazzinc and Xstrata Technology personnel organised numerous plant training drills for the ISASMELT™ operators. These varied in sophistication from single equipment items up to the entire plant, including live transport and control of all input material flows.

Throughout the entire design, procurement, construction, installation, commissioning and ramp-up period, Kazzinc personnel benefited from attending ISASMELT™ Licensee Workshops held in India, Peru, Belgium and Zambia, and from the exchange of information that was possible with their peers and fellow licensees from other countries and other companies. Commissioning a new smelter is always a complex and challenging task and there is comfort and reassurance to be gained from discussing with others their past experiences and tips for success.

Plant Start-up

For a completely new copper smelter and refinery complex, especially one in a cold climate, a plant start-up is not a single event that takes place on a particular day. By necessity the activities span many weeks as various parts of the flowsheet are brought into production. Seasonal plant and equipment are brought into operation on schedules of their own, such as concentrate thawing sheds, building heaters or ventilators, steam heating plants and fuel supply equipment. Concentrate is delivered, stored and blended weeks in advance of its consumption. An inventory of anodes must accumulate in an electrolytic refinery before entire cell sections can be energised together.

At UKMC the operations team implemented an astute practice of commissioning an entire sulphuric acid plant (designed by SNC Lavalin) ahead of the copper smelter. This was made possible by interconnecting ducts with existing non-ferrous smelting plants and the supply of SO_2-rich gases for several months in advance of commissioning the copper smelter. Similarly the air separation unit (designed by Kryogenmash) was also commissioned and operating at full capacity well in advance of its required duty for the copper smelter.

In part because of the care and planning that occurred during pre-commissioning, UKMC achieved 100% of the copper ISASMELT™ plant’s design smelting capacity, measured on an 8-hour shift basis, within a few days of first adding feed to the ISASMELT™ furnace. However, the job of ramping up a smelter to full production does not end there. The following year of operation was spent achieving both consistency and reliability across the whole copper smelter, qualities that are necessary for stable plant operation. Xstrata Technology personnel provided some assistance in this endeavour, whenever asked to do so. For the first few weeks after deciding to heat the ISASMELT™ furnace Xstrata Technology was able to assist all operations shift crews, with two specialists per shift, on a 24 hour basis. The required assistance diminished significantly as the operations team gained experience and competence in their tasks. Within a couple of months the UKMC team were “flying solo” with proficiency, and have continued to optimise and improve the smelter operation ever since.

Operation

Operating statistics of a typical month in the copper ISASMELT™ plant, taken 7 months after start-up, are shown in Table 2. For the average monthly production to be greater than 80% of design capacity within a few months of operation, it is clear that the greater proportion of operating practice, equipment selection, installation and commissioning must have been performed with care and thoroughness.

Table 2 – Operating Statistics from Kazzinc Copper ISASMELT™ Plant

	Unit	February 2012
Availability	%	89.4
Concentrate	t/h	32.0
Silica	t/h	7.7
Reverts	t/h	2.6
Coal	t/h	0.4
Limestone	t/h	1.3
Feed Moisture	%	9
Lance life	days	6
Cu in feed	%	21.7
Cu in matte	%	57.0
ISASMELT™ Lance Flow	Nm^3/h	16500
O_2 in Lance Air	%	58

Variations in Smelter Feed

One of the main advantages identified by Kazzinc in employing the ISASMELT™ process was its ability to treat a wide range of copper concentrate compositions. In fact, the composition of the feed available at the start of smelter operation (refer to Table 3) deviated significantly from the design basis for certain elements. The differences included a proportional increase of lead and antimony in the actual concentrate by more than 50 %, and a proportional decrease of silica in the actual concentrate by more than 50 %, compared with the design concentrate.

Table 3 – Comparison of Actual Kazzinc Copper Smelter Feed at Start-up with Design

	Wt, %							
	Cu	Pb	Zn	Fe	S	SiO2	Sb	As
Design	25.0	2.68	3.14	24.85	31.74	5.11	0.16	0.55
Actual	25.68	4.35	3.53	26.65	32.92	2.39	0.25	0.49
Absolute Deviation	+0.68	+1.57	+0.39	+1.80	+1.18	-2.72	+0.09	-0.06
Relative Deviation	+3%	+59%	+12%	+7%	+4%	-53%	+56%	-10%

The lower level of silica in the copper concentrate required the simple action of greater addition of silica flux material. The increased level of antimony in the concentrate made the task of producing MOOK grade cathode more challenging.

The increase in the level of lead in the concentrate, which was reasonably high to begin with, required modification to the flowsheet and management strategies after the commissioning process was complete. The main strategy for coping with the higher loads of lead was to install a bleed by milling and flotation of the slag from the end of each P-S Converter copper blow. This approach avoided the buildup of recirculating lead that occurred when the slag was charged to the electric settling furnace for electrothermal reduction.

In combination, these measures have been sufficient to prevent problems, such as inadvertent generation of lead bullion by chemical reduction of slag in the electric settling furnace, and the smelter has settled into its role of treating lead-laden copper concentrate without great difficulty. In particular, the waste heat boiler for the copper ISASMELT™ furnace (designed by Foster Wheeler) has experienced no problems attributable to the concentrate composition. Concerns raised during project planning that the internal components of the waste heat boiler would be coated with lead-bearing accretions, thereby reducing its heat transfer efficiency, proved unfounded. Careful attention to the detail design of the waste heat boiler size, shape and flow profile have allowed Kazzinc to have a robust off-gas system for their copper ISASMELT™ furnace.

CONCLUSIONS

The Ust-Kamenogorsk Metallurgical Complex has commissioned a new copper smelter and refinery, treating mostly feeds sourced from its own mines and metallurgical facilities. The new plant was conceived with a future vision of polymetallic concentrates being a major component of the feed material.

Personnel from Kazzinc and Xstrata Technology worked together on an extensive programme of training and pre-commissioning activities to prepare the people and plant equipment for their respective roles in plant operation. The plant started well, and continues to run well. Plans exist for expanding the plant production capacity in the future and the plant is designed to accommodate such expansion should it be required.

ACKNOWLEDGMENTS

The authors would like to acknowledge Kazzinc Ltd and Xstrata Technology Pty Ltd for permission to publish this paper.

REFERENCES

1. G. R. F. Alvear F, P. Arthur, P. Partington, “Feasibility to Profitability with Copper ISASMELT™”, Copper 2010, Volume 2– Pyrometallurgy I, J. Harre, Ed., GDMB, Hamburg, 2010
2. Y. Li, P. Arthur, “Yunnan Copper Corporation’s new smelter – China’s first ISASMELT”, Yazawa International Symposium on Metallurgical and Materials Processing, Volume II – High Temperature Metal Production, F. Kongoli, K. Itagaki, C. Yamauchi, H.Y. Sohn, Eds., TMS, Warrendale, 2003
3. J. Ross and D. de Vries, “Mufulira smelter upgrade project – ‘Industry’ Smelting on the Zambian Copperbelt”, Pyrometallurgy 05, Capetown, Minerals Engineering International, 2005
4. H. Walqui, C. Noriega, P. Partington, &G.R.F. Alvear F, “SPCC’s 1,200,000 TPA Copper ISASMELT™”, Sohn International Symposium on Advanced Processing of Metals and Materials, Volume 8 – International Symposium on Sulfide Smelting 2006, F. Kongoli & R.G. Reddy, Eds., TMS, 2006

Кинетика восстановления шлака Isasmelt с высоким содержанием свинца при температуре от 1073К до 1473К

Xinmei HOU[1),2)] *and Baojun ZHAO*[2)] *and Kuo-Chih CHOU*[1)]

1) Государственная лаборатория Прогрессивной Металлургии Пекинского Научно-Технологического Университета, Пекин 100083, Китай

2) Пирометаллургический Научно-Исследовательский Центр, Университет Квинсленда, Сент-Люсия, Брисбен, Квинсленд 4072, Австралия

Реферат: Использование печи Isasmelt в значительной степени упрощает технологический процесс и позволяет более эффективно улавливать SO_2 и неорганизованные выбросы свинца. В дополнение, шлак с высоким содержанием свинца содержит малое количество серы (<0,5%) по сравнению с агломератом, что в свою очередь сокращает ряд экологических проблем, связанных с работой шахтной печи. Следовательно, печь Isasmelt имеет многообещающие перспективы применения в области производства свинца и может заменить стандартные агломерационные установки в современном производстве свинца. Богатый свинцом шлак из печи Isasmelt используется в качестве сырья свинцовой шахтной печи для производства металлического свинца. Для оптимизации работы свинцовой шахтной печи, важно очень хорошо понимать все реакции и кинетику процессов, происходящих во время восстановления свинцового шлака в шахтной печи. В данном исследовании также проводились комплексные эксперименты по восстановлению свинцового агломерата в углеродных тиглях в температурном диапазоне от 1073 до 1473К. Для сравнения кинетики восстановления проводилось комплексное исследование на промышленных агломератах и синтетических шлаках. Для каждой пробы степень завершенности реакции измерялась с помощью определения объема полученного газа при данной температуре и времени. Для всех шлаков, анализируемых при фиксированной температуре, скорость реакции между шлаком и углеродом, сначала была очень быстрой, а затем снижалась по мере повышения степени завершенности реакции. Между шлаком и углеродом при температурах ниже 1173К происходит лишь ограниченное взаимодействие. При температурах выше 1173К скорость реакции значительно повышается с ростом температуры. В случае с агломератом и шлаком было установлено, что при температуре 1473К между пробой агломерата и углеродом формируется слой металлического свинца. Слой металлического свинца сокращает контактную поверхность между агломератом и углеродом, тем самым снижая общую скорость реакции. Жидкая фаза в составе исследуемого шлака начинает формироваться примерно с 1073К. Доля жидкой фазы увеличивается с ростом температуры. Реакция восстановления происходит в процессе реакции жидкое-твердое. Кинетика реакции контролируется химически на начальном этапе; а диффузией при более длительном интервале времени. Рассчитали условную энергию активации 83,8 кДж/моль (контроль химической реакцией) и 224,9 кДж/моль (контроль диффузией) для реакции восстановления агломерата. Для синтетического шлака энергия активации реакции составила 102,9 кДж/моль на начальной стадии и 259,4 кДж/моль на стадии контроля диффузией.

Ключевые слова: Восстановительная реакция, шлак с высоким с содержанием свинца, кинетика

1. ВВЕДЕНИЕ

По сравнению со свинцовыми шахтными печами, которые в большей степени применяются в технологии получения первичного металлического свинца [1], использование печи Isasmelt в значительной степени упрощает технологический процесс и позволяет более эффективно улавливать SO_2 и неорганизованные выбросы свинца. В дополнение шлак с высоким содержанием свинца содержит малое количество серы (<0,5%) по сравнению с агломератом и, следовательно, уменьшает экологические проблемы, связанные с работой шахтной печи [2]. Следовательно, печь Isasmelt имеет многообещающие перспективы использования в области производства свинца и может заменить стандартные агломерационные установки в современном производстве свинца. Например, на заводе YMG в Цюйцзин, провинция Юньнань, Китай, агломерационная установка была заменена на свинцовую печь Isasmelt, что в свою очередь дало свои положительные результаты в определенной степени [3, 4].

В технологическом процессе, где используется печь Isasmelt, богатый свинцом шлак из печи Isasmelt используется в качестве сырья для свинцовой шахтной печи для производства металлического свинца. Для оптимизации работы свинцовой шахтной печи, важно очень хорошо понимать все реакции и кинетику процессов, происходящих во время восстановления свинцового шлака в шахтной печи. Твердый углерод широко используется в качестве реагента для восстановления свинца из шлака с высоким содержанием свинца. В литературе имеется несколько статей о кинетике восстановления оксида свинца или свинцового шлака. Upashya K.

[5] исследовал кинетику восстановления оксида свинца углеродом, растворенным в железе при 1673K, в жидком шлаке. Он показал, что скорость реакции на начальной стадии регулируется химической реакцией, а затем - переносом масс. H. H. Kinaev др. [6] измерили скорость восстановления шлака PbO-FeO-Fe_2O_3-CaO-SiO_2 с использованием углерода с концентрацией PbO от 3 до 100% при температуре 1423-1573K. Результаты показали, что восстановление свинцовых шлаков углеродом - это контролируемая химическая реакция. Дополнительно B. J. Zhao и др. [7] проводили эксперименты на промышленном свинцовом агломерате и синтетическом свинцовом шлаке с условными микроструктурами и характеристиками восстановления в газовых смесях CO/CO_2. В настоящее время в литературе очень мало информации о поведении шлака богатого свинцом в процессе восстановления с использованием твердого углерода в качестве восстановителя при низкой температуре, а именно от 1073-1473K, что имеет важное значение для промышленности.

Цель настоящего исследования изучить поведение шлака Isasmelt в процессе восстановления при низкой температуре. В данном исследовании будут измеряться и сравниваться характеристики восстановления промышленного свинцового агломерата и синтетических свинцовых шлаков. Исходя из этих данных, будет анализироваться кинетика реакции восстановления.

2. ЭКСПЕРИМЕНТАЛЬНАЯ ЧАСТЬ

2.1 Подготовка проб свинцового агломерата и шлака

Проба агломерата, которая используется в данном исследовании, представляет собой стандартный свинцовый агломерат с завода YMG в Цюйцзин, провинции Юньнань, Китай. Состав исходной пробы, которая использовалась в эксперименте по восстановлению, представлен в Таблице 1. Железо может присутствовать в пробе как двухвалентное, так и трехвалентное, но специальных анализов не проводилось. Все железо было рассчитано по Fe_2O_3 в целях описания на основе предположения, что пробы были получены путем окисления в воздухе.

Синтетические свинцовые шлаки подготавливались в лаборатории из порошков PbO, ZnO, Fe_2O_3, SiO_2 и $CaCO_3$ высокой степени чистоты. Содержание PbO и Fe_2O_3 практически одинаковое в пробе, с меньшим соотношением CaO/SiO_2 (как показано в Таблице 1).

Гранулированную смесь весом примерно 1000 г помещают в алюмосиликатный тигель и нагревают в печи до 1373-1423K в течение 60 минут. После этого шлак медленно охлаждают в тигле. Для перемешивания шлака во время эксперимента использовали аргоновый газ, который подавали со скоростью 5000 мл/мин.

Таблица 1 Состав исходных проб, используемых в экспериментах по восстановлению

Проба			Состав (вес%)			
	Fe_2O_3	ZnO	CaO	SiO_2	PbO	CaO/SiO_2
Агломерат	16,1	9,1	12,0	9,3	53,6	1,30
Синтетический шлак	16,9	11,4	6,7	11,9	51,8	0,56

2.2 Эксперимент в восстановительной среде

Аппаратная схема эксперимента по восстановлению показана на Рис. 1. Эксперименты проводились в алюминиевой реакционной трубке с внутренним диаметром 19 мм. Процедура теста заключается в ведении тигля в реакционную трубку из нижней части печи. Тигель (внешний диаметр 18мм, внутренний диаметр 14 мм и 40мм высота) из чистого графита, удерживали на платформе, состоящей из перевернутого алюмооксидного тигля, удерживаемого на корпусе термопары из оксида алюминия. Для контроля температуры шлака использовали термопару Pt/Pt-13% Rh. Воздух продували из реакционной трубки с помощью аргона высокой чистоты. Платформу и графитовый тигель медленно поднимали на алюминиевой трубке в горячую зону предварительно нагретой печи. В соответствии с требованием температуру печи предварительно измеряли. Когда тигель оказывался в горячей зоне, промывку газом останавливали и реакционную трубку герметично закрывали.

Затем в эксперименте в тигель добавляли 5г пробу шлака (диаметром 1 мм) из верхней части печи. Реакция начиналась сразу после того, как проба шлака соприкасалась с графитовым тиглем. Газ CO/CO_2, полученный в результате реакции между графитом и шлаком, улавливали водосодержащим манометром, в результате чего вода перемещалась в контейнер на весах (как показано на Рис.2). Объем газа CO/CO_2, получаемый в результате реакции восстановления, постоянно измеряли по весу вытесненной воды. Реакцию останавливали путем снижения температуры пробы и тигля до комнатной температуры под воздействием потока Ar газа. Также проводились предварительные эксперименты по восстановлению свинцового агломерата и синтетических шлаков в углеродных тиглях в диапазоне температур от 1073 до 1473K.

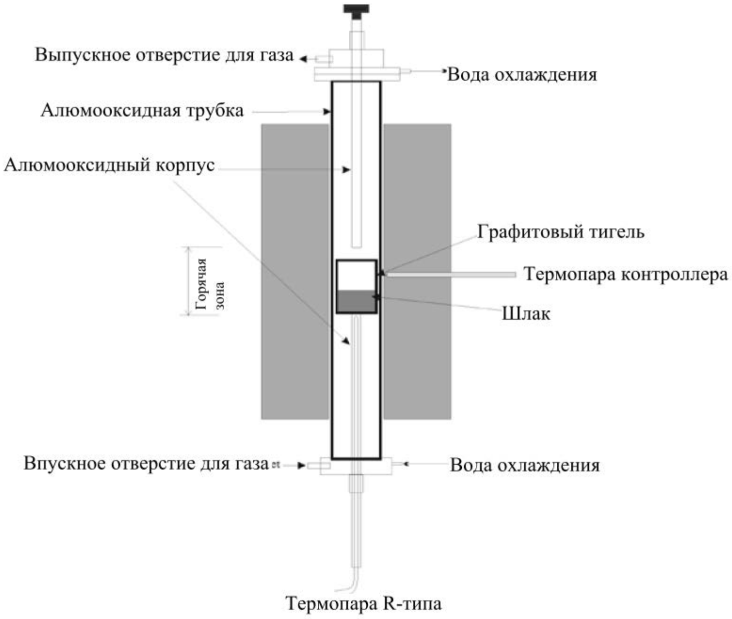


Рис 1. Аппаратная схема эксперимента по восстановлению в вертикальной печи.

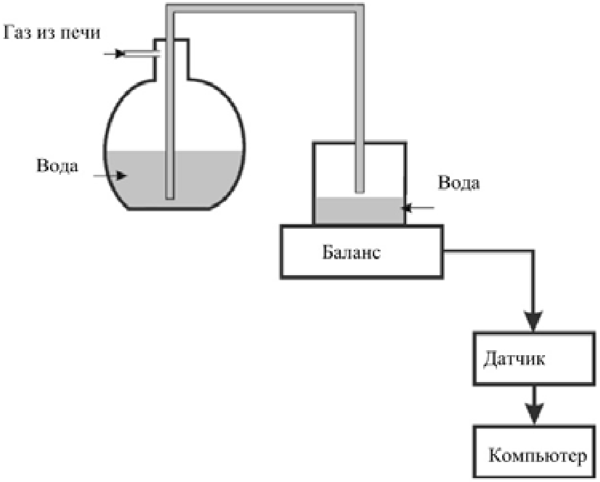


Рис 2. Аппаратная схема сбора газа и система регистрации данных

2.3 Анализ микроструктуры

Исходные и восстановленные пробы закрепляли в эпоксидной смоле, полировали и изучали методом оптической и сканирующей электронной микроскопии, чтобы определить имеющиеся фазы. Состав стеклянной и твердой фаз, содержащихся в шлаках, определяли методом электронного зондового рентгеновского микроанализа (EPMA JEOL 8200L, Japan Electron Optics Ltd., Toyo, Япония). EPMA проводили в условиях ускоряющего напряжения

15кВ и 15нА зондового тока на основе измерений Duncumb-Philibert. Также применяли процедуру коррекции ZAF. Средняя точность измерений методом EPMA в пределах ±1 вес%.

3. РЕЗУЛЬТАТЫ И ОБСУЖДЕНИЕ

3.1 Ликвидус системы шлака богатого свинцом - углерода

Согласно литературе богатые свинцом шлаки переходят в жидкую фазу только при температуре выше 1473К [5-7]. В исследованном диапазоне температур шлаки богатые свинцом были в твердой и жидкой фазах. Реакция между углеродом и шлаками с высоким содержанием свинца представляет собой сложный процесс, который включает в себя реакции между твердым-жидким и твердым-твердым. Для сравнения отметим, что для химической реакции между твердым и жидким более предпочтительны жидкий шлак и углерод. Количество жидкой фазы можно спрогнозировать с помощью компьютерной программы термодинамического моделирования FactSage на основе оптимизированной термодинамического базы данных, разработанной авторами специально для этой системы [8]. Соотношение основных фаз, содержащихся в системе шлака богатого свинцом - углерод, при различных температурах представлено в Таблице 2. Из таблицы видно, что переход в жидкую фазу начинается с температуры 1073К и ее соотношение увеличивается с ростом температуры, в результате реакция восстановления проходит легче при более высокой температуре.

На Рис.2 а и б показаны стандартные микроструктуры агломерата (а) и синтетического шлака (б), восстановленные при температуре 1073К в течение 60 минут в графитовом тигле соответственно. В обоих видах шлака присутствует некоторая часть жидкой фазы помимо кристаллических фаз, что соответствует результатам FactSage. По жидкой фазе проводились комплексные измерения. Результаты измерений EPMA по жидкой фазе приведены в Таблице 3. Из таблицы видно, что содержание жидких фаз, рассмотренных в эксперименте, практически соответствуют прогнозам FactSage, что еще раз доказывает целесообразность программного обеспечения.

Необходимо отметить, что в шлаке присутствует оксид железа в виде Fe_2O_3. Так как эксперименты проводили в графитовом тигле, Fe_2O_3 восстанавливается до FeO. Поэтому оксид феррита по программе FactSage прогнозировали в виде FeO.

Таблица 2 Состав ликвидуса шлака с высоким содержанием свинца при различных температурах спрогнозирован FactSage

Проба	*Состав (вес%)*						
		Содержание жидкой фазы	FeO	PbO	CaO	SiO_2	ZnO
Агломерат	1073K	39,2	4,16	86,922	0,438	6,13	2,35
	1173K	43,9	7,36	80,9	1,18	8,05	2,51
	1273K	51,7	8,92	76,4	2,12	9,42	3,14
	1373K	69,3	9,72	68,5	4,76	10,6	6,42
	1473K	100	12,4	55,1	8,44	13,0	11,06
Синтетический шлак	1073K	10,5	5,82	89,9	0,998	1,792	1,49
	1173K	57,4	7,63	85,3	1,67	2,79	2,61
	1273K	65,7	9,54	79,5	2,88	3,62	4,46
	1373K	74,6	12,4	67,3	6,36	6,31	7,63
	1473K	91,8	14,3	55,6	11,9	9,01	9,19

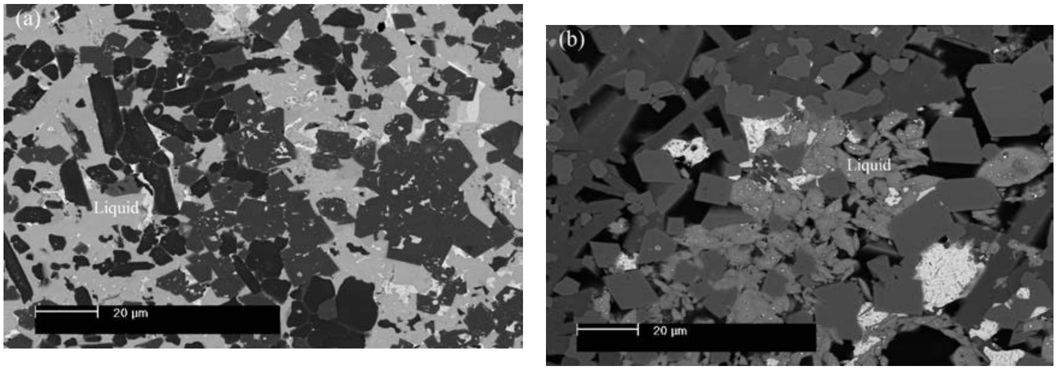


Рис.3 Обратнорассеянное микроизображение шлака с высоким содержанием свинца, полученное с помощью растрового микроскопа, после реакции при температуре 1073K в течение 60 минут в графитовом тигле (а) агломерат; (б) синтетический шлак

Таблица 3 Экспериментально определенный состав жидкой фазы в шлаке при 1073K

Проба	*Состав (вес%)*					
		FeO	PbO	CaO	SiO_2	ZnO
Агломерат	1073K	5,2	89,0	0,1	4,3	1,1
	1073K	5,0	90,3	0,1	3,1	1,3
	Среднее	5,1	89,7	0,1	3,7	1,2
Синтетический шлак	1073K	2,1	77,2	0,3	18,0	2,6
	1073K	2,7	72,9	1,2	21	2,3
	Среднее	2,5	75,1	0,75	19,5	2,45

3.2 Эксперименты по восстановлению в изотермических условиях

Эксперименты по восстановлению в изотермических условиях проводили в температурном диапазоне от 1073К до 173К с интервалом 100К. Следует отметить, что газ, полученный в результате реакции между агломератом/шлаком и углеродом фактически представляет собой смесь CO и CO_2. В настоящем исследовании соотношение CO к CO_2 не определялось. Однако общий объем CO и CO_2, полученный в результате реакции, определяет относительную степень восстановительных реакций.

На Рис. 4 и 5 показаны результаты экспериментов по восстановлению агломерата и синтетического шлака соответственно. Для каждой пробы результаты показаны в виде объемов полученного газа как функции от времени реакции. На Рис. 4 и 5 можно увидеть, что при фиксированной температуре объем получаемого газа сначала резко увеличивается, а затем по ходу реакции медленно уменьшается. Это свидетельствует о том, что кинетика реакции восстановления контролируется химически на начальном этапе; а диффузией при более длительном интервале времени. В жидкой фазе был обнаружен металлический свинец (как показано на Рис.3), что говорит о химической реакции на границе жидкого шлака и углерода.

Кроме того, можно отметить лишь ограниченное взаимодействие между агломератом/шлаком и углеродом при температурах ниже 1173К. При температурах выше 1173К объем полученного газа значительно увеличивается с ростом температуры. Во всем диапазоне исследуемых температур агломерат выделяет больший объем газа (Рис.4) по сравнению с синтетическим шлаком (Рис.5), но разница в объемах зависит от температуры. Основная причина, вероятно, заключается в том, что соотношение жидкой фазы в пробе агломерата выше, чем в синтетическом шлаке (как показано в Таблице 2), что и обеспечивает большую реакционную поверхность в реакции жидкое/твердое.

В процессе реакции восстановления агломерата (Рис.4) можно заметить, что в течение первых 20 минут, объем образующегося газа увеличивается с повышением температуры. Однако после 20 минут реакции объем газа, образовавшийся при температуре 1473К, стал меньше, чем при температуре 1373К. При осмотре пробы при температуре 1473К было обнаружено, что слой металлического свинца образовался на границе между пробой агломерата и углерода. Скорее всего, в этом случае слой металлического свинца сокращает площадь контакта агломерата и углерода, тем самым сокращая общую скорость реакции.

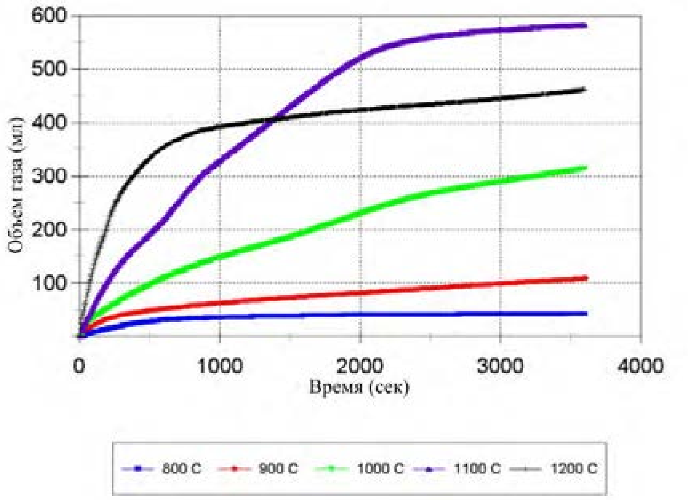


Рис.4 Образование газа из свинцового агломерата в графитовом тигле определяется как функция от времени реакции

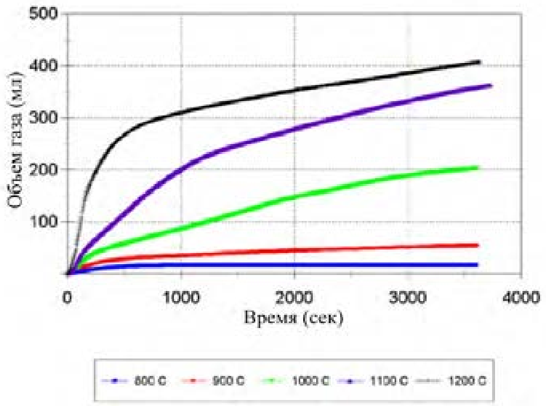
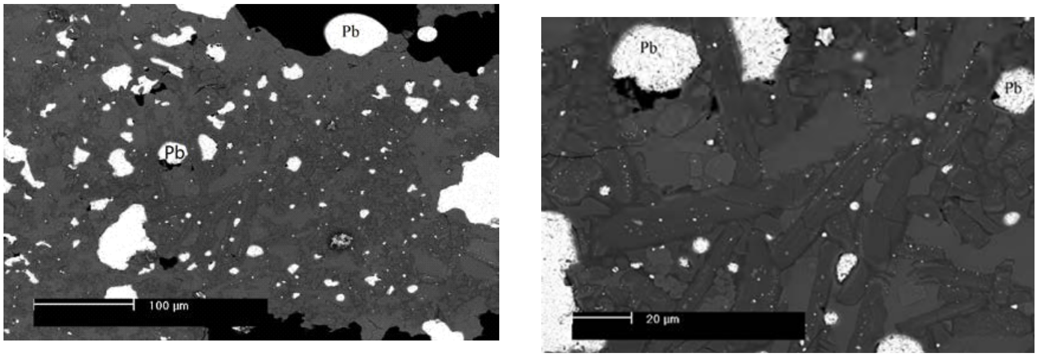


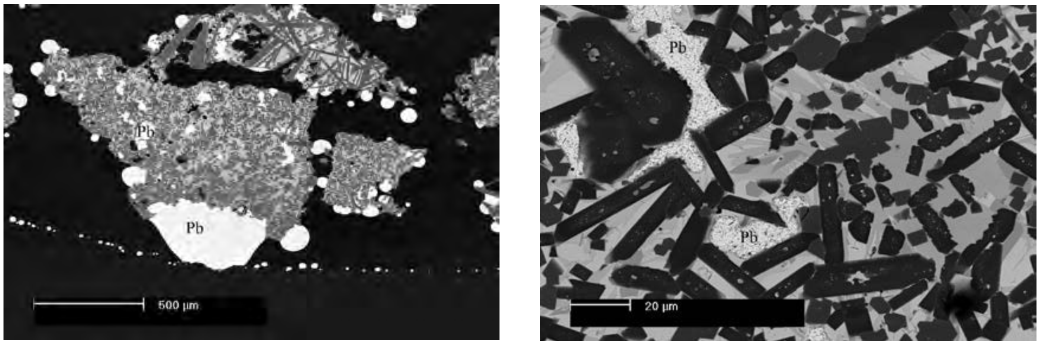
Рис. 5 Образование газа из синтетического шлака в графитовом тигле определяется как функция от времени реакции

3.3 Микроструктурный анализ

На Рис.6 показаны стандартные микроструктуры агломерата (а) и синтетического шлака (б), восстановленные при температуре 1173К в течение 60 минут в графитовом тигле. На обратно рассеянной электронной микрофотографии можно увидеть, что металлический свинец присутствует в обеих системах, что свидетельствует о том, что при этой температуре прошла реакция восстановления. Можно также заметить некоторую пористость (показано в виде черных участков на Рис.6), которая создается за счет затвердевания металлического свинца в процессе перехода из жидкой фазы в твердую фазу. Присутствие металлического свинца в жидком шлаке (как показано на Рис.6) подтверждает тот факт, что реакция восстановления в основном проходила на границе жидкое-твердое, а именно на линии контакта жидкий шлак - углерод.



(а)



(b)

Рис.6 Обратно рассеянное микроизображение шлака с высоким содержанием свинца, полученное с помощью растрового микроскопа, после реакции при температуре 1173К в течение 60 минут в графитовом тигле (а) агломерат; (б) синтетический шлак

3.4 Кинетика восстановления

Потенциальные реакции между отдельными компонентами жидкой фазы шлака и углерода:

PbO(шлак)+C(тв)=Pb(ж)+CO(г) (1)

Fe_2O_3(шлак)+C(тв)=$2FeO$(шлак)+CO(г) (2)

ZnO(шлак)+C(тв)=Zn(г)+CO(г) (3)

FeO(шлак)+C(тв)=Fe(тв)+CO(г) (4)

Поскольку содержание ZnO и FeO в исходном шлаке очень мало, их реакцией с углеродом можно пренебречь. Только уравнения Реакции (1) и (2) - полное восстановление. Реакция восстановления в основном происходит, когда PbO и Fe_2O_3 находятся в жидком состоянии. Соотношение рассчитывалось по FactSage. Следовательно, объем образующегося газа можно очень просто рассчитать по уравнению (1) и (2).

Из приведенного выше анализа видно, что реакция восстановления между свинцовым шлаком и углеродом может быть разделена на два этапа: на первом этапе реакция контролируется химической реакцией на границе контакта жидкое/твердое, а на втором этапе диффузией при более длительном интервале времени. В следующем разделе описывается, каким образом можно проанализировать кинетику реакции с помощью модели сжимающегося ядра [9-11].

Что касается агломерата, на начальном этапе кинетика его реакции контролировалась с помощью химической реакции и может быть описана следующим уравнением:

$$(\Delta m) = k_r t \tag{5}$$

где Δm - изменение массы. k_r - константа реакции, а t - время реакции. Поскольку существует линейная зависимость между изменением массы и объемом газа, получаемого в процессе реакции Соотношение между объемом газа и временем реакции должно также удовлетворять уравнению (5). Из приведенного выше уравнения можно рассчитать константу реакции при разных температурах. Так как соотношение между константой реакции и энергией активации описывается уравнением Аррениуса:

$$k=Ae^{-E/RT} \tag{6}$$

где k - коэффициент частоты (м.с $^{-1}$), E - энергия активации (Дж.моль$^{-1}$), R - константа газа (Дж.моль $^{-1}$.K^{-1}) и T-температура (K). Условную энергию активации можно рассчитать по наклону lnk и 1/T (Рис.7), она должна равняться 83,8 кДж/моль.

При более длительном интервале времени реакция контролируется диффузией и может описываться следующим образом [12]:

$$(\Delta m)^2=k_d t \qquad (7)$$

где k_d - константа реакции. Поскольку реакция при 1473K проходила необычно, о чем говорилось в предыдущем разделе. Константа реакции при этой температуре не учитывается при расчете условной энергией активации. Аналогично условную энергию активации можно рассчитать по наклону lnk и 1/T (Рис.8), она должна равняться 224,9 кДж/моль.

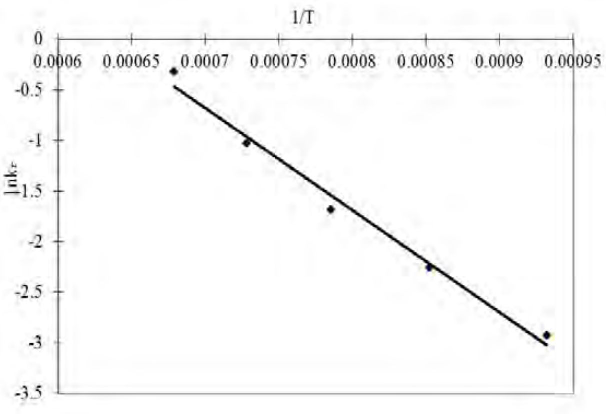


Рис.7 Соотношение lnk и 1/T для реакции восстановления агломерата на начальном этапе

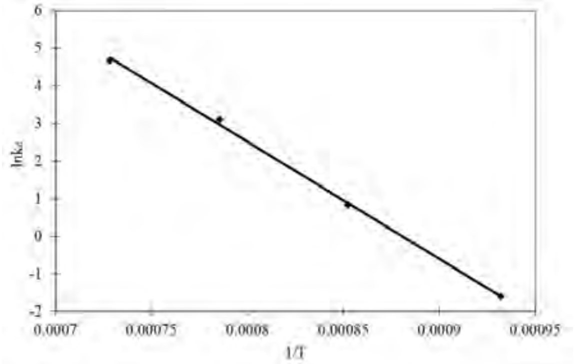


Рис.8 Соотношение lnk и 1/T для реакции восстановления агломерата на начальном этапе

Аналогичным способом рассчитывалась условная энергия активации для реакции синтетического шлака и углерода, равная 102,9 кДж/моль на начальной стадии и 259,4 кДж/моль на стадии контролированию диффузией. Более высокую энергию активации для восстановления синтетических шлаков углеродом можно объяснить его микроструктурой (Рис.8б). Условная энергия активации сопоставима с энергией, о которой сообщается в литературе [8].

4. ЗАКЛЮЧЕНИЕ

Реакция восстановления проб свинцового агломерата и синтетического шлака была выполнена с помощью графита в температурном диапазоне 1073-1473K. Скорость реакции между шлаком богатым свинцом и графитом была определена экспериментальным образом с помощью измерения объема газа, полученного в ходе реакции. Термодинамические расчеты были выполнены с помощью компьютерной программы FactSage с целью прогнозирования фазовых соединений для выбранного состава шлака с углеродом и в равновесии с металлическим свинцом. Результаты показали следующее:

(1) Жидкая фаза в составе исследуемого шлака начинает формироваться примерно с 1073K. Доля жидкой фазы увеличивается с ростом температуры.

(2) Для всех шлаков, анализируемых при фиксированной температуре, скорость реакции между шлаком и углеродом, сначала была очень быстрой, а затем скорость снижалась по мере повышения степени завершенности реакции.

(3) Между шлаком и углеродом при температурах ниже 1173K происходит лишь ограниченное взаимодействие. При температурах выше 1173K скорость реакции значительно повышается с ростом температуры.

(4) Реакция восстановления происходит в процессе реакции жидкое-твердое. Кинетика реакции контролируется химически на начальном этапе; а диффузией при более длительном интервале времени. Рассчитали условную энергию активации 83,8 кДж/моль (контроль химической реакцией) и 224,9 кДж/моль (контроль диффузией) для реакции восстановления агломерата. Для синтетического шлака энергия активации реакции составила 102,9 кДж/моль на начальной стадии и 259,4 кДж/моль на стадии контроля диффузией.

Благодарность

Авторы выражают признательность Национальному Научному Фонду Китая No.51104012 за финансовую поддержку.

Справочная литература

[1] T. Imai, M. Sakata, K. Morita, Y. Kondo, A. Horiuchi and M. Kusano. Lead smelters survey, Lead & Zinc’05, Kyoto, Japan, 2005, 419-478.

[2] H. T. He, The QSL Lead Slag Fuming Process Using an Ausmelt Furnace Translated selectively, Nonferrous smelting, 2003(1): 50-53.

[3] B. Errington, J. Wang, P. Arthur and Y. Dong. The ISA-YMG lead smelting process, Lead & Zinc’05, Kyoto, Japan, 2005, 587-599.

[4] J. K. Wang, B. J. Zhao, G. Yang, P. Hayes. Properties and reduction mechanism of lead-rich slags, Nonferrous metals, 2004,6:5-8.

[5] Upashya K. Kinetics of reduction of lead oxide in liquid slag by carbon in iron. Metall. Trans. B, 1986, 17B: 271-279.

[6] N. N. Kinaev, E. Jak, P. C. Hayes. Kinetics of reduction of lead smelting slags with solid carbon. Scand. J. Met., 2005, 34: 150-157.

[7] B. J. Zhao, B. Errington, E. Jak, P. Hayes. Gaseous reduction of isasmelt lead and lead blast furnace sinters. Can.

Metall. Quart., 2010, 49(3):241-248.

[8] E. Jak, S. Degterov, P. C. Hayes, A. D. Pelton, Thermodynamic modeling of the system PbO-ZnO-FeO-Fe2O3-CaO-SiO2 in zinc lead smelting, 5th Intl. Conf. on Slags and Fluxes, Sydney, ISS, Warrendale, 1997, 621-628.

[9] M.D.Pritzker. Model for parallel surface and pore diffusion of an adsorbate in a spherical adsorbent particle, Chem.

Eng. Sci., 2003, 58:473 – 478.

[10] F. Veglio, M.Trifoni, F. Pagnanelli, L.Toro, Shrinking core model with variable

activation energy: a kinetic model of manganiferous ore leaching with sulphuric acid and lactose. Hydrometallurgy, 2001, 60, 167–179.

[11] A.Velardo, M. Giona, A. Adrover, F. Pagnanelli, L. Toro,Two-layer shrinkingcore model: parameter estimation for the reaction order in leaching processes, Chem. Eng. J., 2002, 90, 231–240.

[12] W. B. Jepson. On the parabolic rate law. J. Electrochem. Soc., 1960, 107(1), 53-55.

ESCALAMIENTO GRADUAL DE ISASMELT™ - LA CLAVE PARA EL ÉXITO

P. S. Arthur and G. R. F. Alvear F.

Xstrata Technology
Level 4, 307 Queen Street
Brisbane, Qld. 4000, Australia
isasmelt@xstratatech.com

Palabras clave: fundición de cobre, fundición de plomo, reciclaje TSL, desarrollo del proceso, plata piloto

RESUMEN

El desarrollo de un nuevo proceso de fundición es arriesgado. Muchos procesos nuevos en la industria de metales no cumplieron las expectativas iniciales de sus inventores. Grandes sumas de capital se han invertido en tecnologías que, o bien no proporcionaron un rendimiento satisfactorio de la inversión, o no funcionaron en absoluto y se tuvieron que ser reemplazadas. Una de las claves del éxito en el desarrollo de procesos es el incremento gradual en su desarrolo, permitiendo identificar y rectificar problemas lo antes posible. De esta manera, se puede minimizar el costo de cualquier rectificación que se requiera. El proceso de ISASMELT™ ha sido desarrollado totalmente por Mount Isa Mines entre los años 1980 y 1990, a través de un proceso de incremento gradual. La comercialización sólo se produjo una vez que el proceso fue probado en escalas de laboratorio, piloto y de demostración durante varios años. Como resultado, ahora la tecnología ISASMELT™ opera con éxito en numerosas plantas de todo el mundo. El enfoque metódico para el desarrollo de la tecnología ha permitido a los usuarios modernizar sus operaciones o crear nuevas empresas con reducción significativa del riesgo técnico.

INTRODUCCIÓN

En el mundo de la pirometalurgia, los cambios tecnológicos ocurren lentamente. La adopción de nuevas tecnologías ocurre con poca frecuencia, ya que los requisitos de inversión de capital para nuevas fundiciones son generalmente muy altos y la implementación de nuevas tecnologías resulta en desafíos técnicos importantes. En consecuencia, a pesar de que los avances innovadores pueden crear oportunidades para que las empresas exploren nuevas fronteras, tales revoluciones son poco frecuentes. En la mayoría de los casos las empresas prefieren seleccionar tecnologías probadas para minimizar el riesgo y garantizar la rápida mejora de sus instalaciones. En algunos casos, sin embargo, las tecnologías probadas no son capaces de entregar los beneficios económicos y ambientales necesarios. En tales situaciones se requieren avances y la innovación técnica juega un papel esencial. Para que los procesos innovadores se implementen con éxito, se deben superar muchos obstáculos. Es aquí donde se requiere un enfoque sistemático en el proceso de innovación para mitigar el riesgo. Se requiere de equipos multidisciplinarios que trabajen en colaboración sobre una base científica sólida, para llegar a la etapa en que un prototipo se pueda probar. Una vez que se haya desarrollado y probado dicho prototipo, todavía existe la pregunta de si el nuevo proceso funcionará y será económicamente viable a escala comercial. Un proceso de crecimiento incremental minimizará los riesgos técnicos y financieros, lo que aumenta la probabilidad de un resultado exitoso.

Este artículo pretende registrar la secuencia de los pasos dados en el desarrollo de la tecnología TLS ISASMELT™, con énfasis en las etapas piloto y de demostración como pasos fundamentales para la comercialización exitosa.

DESARROLLO DE LA TECNOLOGÍA LÍDER DE ISASMELT™

Origen

La combinación planta de sinterización / alto horno fue la tecnología dominante en la fundición de plomo durante todo el siglo 20. A principios de los años 1970, las empresas que empleaban esta tecnología se vieron sometidas a una sostenida presión política y económica según se fueron introduciendo regulaciones ambientales más estrictas (en gran parte debido a un reconocimiento de la naturaleza tóxica de plomo), y los costos de energía se incrementaron, lo que llevó a mayores costos operativos y de capital [1]. La industria buscó soluciones en la innovación tecnológica, tratando de:

- Investigar las innovaciones y mejoras del proceso
- Desarrollar tecnologías más limpias y eficientes
- Reducir los costos a través de un mayor rendimiento y una mayor eficiencia

Se hicieron esfuerzos en todo el mundo para desarrollar procesos directos de fundición de plomo. Se introdujeron procesos como Kivcet, QSL y Kaldo para reemplazar la tecnología convencional. Los desarrolladores tuvieron que hacer frente a muchos desafíos tecnológicos tales como minimizar la generación de humos, optimizar la química de la escoria, mejorar el rendimiento refractario y la eficacia en la reducción de la escoria.

Fue en este ambiente que Mount Isa Mines buscó un proceso que mejoraría el rendimiento de las operaciones en su fundición de plomo en Mount Isa, Australia. Después de investigar los distintos procesos en fase de desarrollo, los investigadores centraron su atención en la lanza Sirosmelt. Recientemente, esta última se había desarrollado a escala de laboratorio en la Organización de Investigación Científica e Industrial (CSIRO por sus siglas en inglés) en Melbourne. A raíz de las investigaciones iniciales, la administración de Mount Isa Mines reconoció el potencial del nuevo concepto para la fundición de concentrados de plomo y se embarcó en un extenso programa de desarrollo.

En 1978 se inició un proyecto conjunto entre Mount Isa Mines y CSIRO para investigar la aplicación de la tecnología de combustión sumergida Sirosmelt a la fundición de concentrados de plomo de Mount Isa. En el momento de esta decisión, la tecnología Sirosmelt ya se había utilizado para demostrar la reducción de escoria del convertidor de cobre en Mount Isa [2,] en una escala de planta piloto.

La administración de Mount Isa Mines decidió que la clave para la implementación exitosa de esta tecnología era optar por un enfoque metodológico. Al contar con etapas claramente definidas, sería posible resolver los desafíos técnicos desde los fundamentos básicos del proceso de fundición de plomo hasta la viabilidad comercial.

Modelado termodinámico

El punto de partida fue el modelado termodinámico. El paquete Chemix de CSIRO permitió determinar las condiciones más adecuadas para la operación del proceso. Por ejemplo, se concluyó que la estrategia óptima de fundición dependía de la ley de concentrado, como sigue [3]:

- Para los concentrados de plomo de alta ley (65 a 75%) era factible la producción directa de algunos lingotes. Esto pudo lograrse operando bajo condiciones sub-estequiométricas. Se requirió una segunda etapa de reducción de escoria para producir los lingotes de plomo restantes.
- Para concentrado de plomo de menor ley (45-50%) la generación de humos pudo reducirse mediante la realización del proceso de fundición bajo condiciones altamente oxidantes a la temperatura más baja posible. Este proceso generaba una escoria rica en óxido de plomo que posteriormente se pudo reducir para producir lingotes de plomo.

Pruebas a escala crisol

Se inició un programa de prueba de laboratorio a escala crisol para obtener datos fundamentales en la química del proceso de fundición de plomo. El programa se llevó a cabo en el laboratorio de CSIRO con la participación directa del personal de Mount Isa Mines. Un ingeniero metalúrgico de Mount Isa Mines fue adscrito a CSIRO para trabajar en las pruebas a escala crisol. Esta participación directa fue reconocida en su momento como un paso crucial para establecer la confianza necesaria en la nueva tecnología, y desarrollar a un profesional que más tarde estaría involucrado en trabajo de pruebas a escala piloto. Algunos de los aspectos investigados fueron:

- Generación de volátiles de plomo versus contenido de plomo en la alimentación
- Efecto del potencial de oxígeno en la generación de volátiles de plomo
- Cinética de la reducción de la escoria

Los resultados obtenidos en el trabajo de pruebas a escala crisol se consideraron prometedores como para invertir en la siguiente etapa del proceso de desarrollo mediante la ampliación de una unidad de planta piloto. Con la conclusión del trabajo de pruebas a escala crisol en CSIRO y la decisión de avanzar a escala piloto, la responsabilidad del desarrollo del proceso pasó al Departamento de Investigación y Desarrollo en Mount Isa.

Primera etapa en el proceso de ampliación: Plan piloto

La primera planta piloto de ISASMELT™ fue comisionada en Mount Isa, en septiembre de 1980. Esta planta fue diseñada para una velocidad de alimentación nominal de 120 kg / h de concentrado. Se llevó a cabo un diseño experimental para confirmar la viabilidad de un proceso de dos etapas para la fundición de concentrados de plomo y reducción de la escoria, y para obtener estimaciones razonables de las tasas de volatilización de plomo, desgaste del refractario y las tasas de combustible. La planta piloto funcionó durante un período prolongado, entregando una cantidad importante de datos operacionales que se pudieron utilizar como la base para la siguiente etapa del desarrollo en la escala de demostración. La Tabla 1 resume las principales conclusiones del trabajo de la planta piloto [4].

Mount Isa Mines y CSIRO posteriormente obtuvieron una patente para el "proceso de fundición de plomo de alta intensidad" [5]. Este proceso se denomina Proceso ISASMELT™. El término "ISASMELT" se utilizó posteriormente para describir la combinación del diseño del horno y la lanza modificada que incorpora los desarrollos y el know-how de las operaciones de Mount Isa Mines.

Tabla 1 – Hallazgos principales en la planta piloto de plomo ISASMELT™

Tópico	Principales hallazgos
Volatilización del Plomo	Controlado a lo largo de una amplia gama de composiciones y temperaturas de escoria
Aumento del contenido de oxígeno del aire de lanza	Disminución de las necesidades de combustible Disminución de la producción de compuestos volátiles de plomo No hay un aumento significativo en el desgaste de la lanza
Desgaste del refractario	Menos de lo esperado Mayor durante la reducción que durante la oxidación
Reducción de escoria	Demostrado en la operación por lotes Procesado rápidamente añadiendo carbón global
Humo del Zinc	Muy dependiente de la temperatura

Segunda etapa en el proceso de ampliación: Planta de Demostración

Los resultados en la planta piloto de 120 kg / h motivó a la administración de Mount Isa Mines a construir una planta de demostración con el fin de evaluar el potencial de la operación comercial del proceso. Se decidió por un caudal de aproximadamente 5 t / h, porque era aproximadamente de un cuarto a un décimo del tamaño de una planta comercial, lo que permitiría una futura ampliación dentro de niveles de riesgo aceptables. La intención era que la planta de demostración se utilizara para obtener información adicional sobre el desgaste refractario y de lanza, el consumo de combustible, la generación de compuestos volátiles de plomo y la higiene. Para mejorar la transferencia de la tecnología al personal de operaciones, la planta de demostración se instaló dentro de la fundición de plomo existente. El objetivo era hacer que la nueva tecnología fuera parte de la operación de fundición de plomo y que el personal de operaciones la considerara como propia. Se decidió construir la planta en dos etapas. La primera, encargada en septiembre de 1983, fue un solo horno programado para 5.9 t / h de material de alimentación. Se utilizó exclusivamente para la fundición de oxidación del concentrado. La gran cantidad de escoria producida se agregó a la planta de sinterización como reemplazo parcial para la sinterización de retorno. La incorporación de escoria de plomo en la mezcla de alimentación de sinterización, incrementó la capacidad efectiva de la planta de sinterización, y por tanto, la de la fundición de plomo en general en aproximadamente un 10%.

Una vez que la primera etapa demostró el funcionamiento de la etapa de fundición durante un período de más de un año, se construyó la segunda etapa de la planta y fue comisionada en agosto de 1985. El segundo horno fue construido adyacente al primero para la transferencia de escoria fundida oxidada entre ellos. Inicialmente se llevaron a cabo más de 150 ensayos de reducción de lotes. Posteriormente, en 1987, los dos hornos de demostración fueron operados de forma simultánea, con procesos continuos de:

- Fundición del concentrado de plomo en el horno de fundición
- Colada y transferencia de escoria fundida alta en plomo al horno de reducción
- Reducción de la escoria con carbón en trozos a 1170-1200 ºC
- Colada de plomo crudo y desecho de la escoria juntos a través de una única piquera

La operación a escala de demostración permitió que el equipo de desarrollo de procesos reconociera y adaptara varios aspectos del diseño para mejorar la operación de la planta a escala comercial. La Tabla 2 resume algunas de las principales conclusiones de la planta de demostración [6-7].

Tabla 2 – Principales hallazgos en la planta de carbón ISASMELT™ de demostración

Tópico	Principales hallazgos	Efecto
Diseño de lanza	Rediseño de lanzas para permitir el funcionamiento a baja presión de aire	Reducción en el costo operativo y de capital
Adición de combustible	No se requiere un sistema de inyección neumática para combustible sólido. Se añadió combustible sólido en el horno de oxidación (principal contribuyente para el balance de calor) con el concentrado. Correcto ajuste de temperatura logrado con inyección de aceite a través de la lanza.	Reducción de costos de operación y de capital
Tasa de Volatilizacion	Menos de la mitad de la planta piloto	Menor costo operativo debido a la reducción en la recirculación de polvo
Desgaste del refractario	Menos de una décima parte de la planta piloto	Menor costo operativo

Un factor clave para el éxito económico y para la continuación del proceso de mejora del diseño fue que la planta de demostración era parte de la cadena de producción global de la fundición. Hacia abril de 1989, la planta había tratado más de 125.000 toneladas de concentrado de plomo, agregando valor a la operación de Mount Isa Mines. Este hecho indica claramente que el proceso no sólo tuvo mérito técnico, sino que también podría ser un éxito financiero, y la administración se motivó a invertir más en su desarrollo.

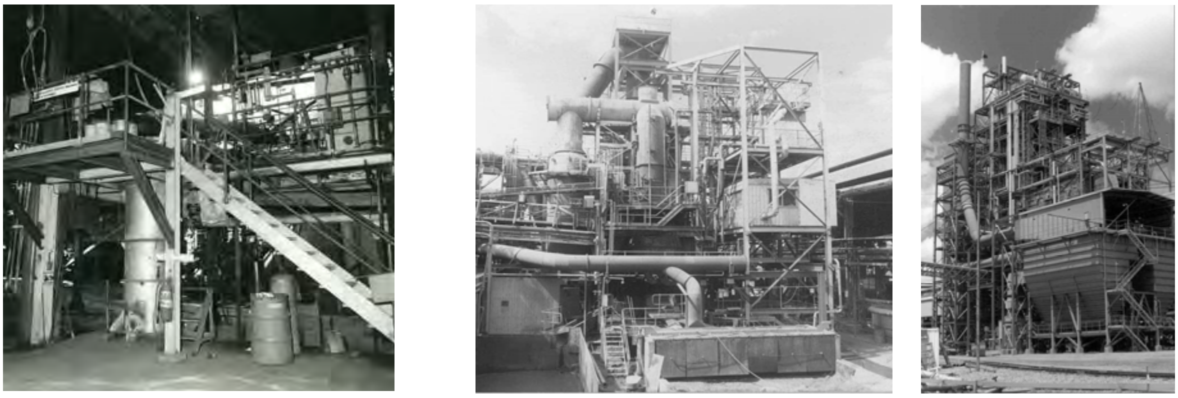
Etapa final en el proceso de ampliación: Implementación comercial

Tras el éxito de la fundición de plomo como demostración, la administración de Mount Isa Mines aprobó la construcción de una planta de plomo ISASMELT™ de dos etapas. La primera etapa se diseñó para oxidar completamente el concentrado de plomo y así producir una escoria alta en plomo, mientras que la segunda se diseñó para reducir esta escoria alta en plomo a fin de producir lingotes de plomo y escoria de descartes. Esta planta comenzó su actividad en febrero de 1991. Algunas diferencias entre el diseño de la planta de demostración y la planta comercial fueron el uso de un vertedero de plomo para permitir una colada continua del producto de lingotes de plomo, la instalación de calderas de recuperación, y la inyección de carbón a través de la lanza en el horno de reducción.

La etapa de oxidación demostró rápidamente la capacidad para fundir el concentrado de Mount Isa a tasas de 20 t / h, produciendo una escoria alta en plomo [8]. La operación de fundición era estable y se lograron largas vidas útiles de la lanza del refractario. Las condiciones de oxidación y un muy agitado baño en el horno de fundición lograron suprimir la formación de humos de sulfuro de plomo.

Sin embargo, la etapa de reducción inicialmente resultó ser más problemática, con vidas cortas de lanza y bloqueo y desgaste del sistema neumático de inyección de carbón, lo que derivó en contenidos de plomo erráticos en la escoria final. Estos problemas fueron superados por el rediseño de los equipos de inyección de carbón y la modificación del diseño de la punta de la lanza. Un mejor rendimiento se produjo durante 1993, cuando se dispuso de oxígeno adicional y se fundieron hasta 36 t / h de concentrados de plomo con aire de lanza con enriquecimiento de oxígeno de 33 a 35%.

La Figura 1 muestra una secuencia de fotos de las diferentes etapas de la ampliación del proceso de plomo de ISASMELT™ en Mount Isa.



(a) (b) (c)

Figura 1 – Plantas de plomo Lead ISASMELT™: (a) escala de planta piloto, (b) escala de demostración, (c) escala comercial

DESARROLLO DE LA TECNOLOGÍA DE COBRE ISASMELT™

El desarrollo del proceso de cobre ISASMELT™ se produjo en paralelo al del proceso de plomo y siguió el mismo enfoque sistemático para su ampliación. La fundición de cobre de Mount Isa Mines operaba dos hornos de reverbero en ese momento y la administración reconoció que había una necesidad de sustituir el proceso con una nueva tecnología que permitiría un consumo energético significativamente menor y una captura económica de dióxido de azufre en el gas de escape para su conversión a ácido sulfúrico. Después de evaluar las distintas opciones del proceso en cuanto a la oferta de los proveedores de tecnologías y siguiendo el exitoso trabajo en la planta piloto con el proceso de fundición de plomo, decidieron adaptar dicho proceso para la fundición de concentrados de cobre. La investigación fundamental además de extensas pruebas de la planta piloto llevó a la instalación de una planta de demostración a escala dentro de la fundición de cobre en Mount Isa en 1987. La Tabla 3 muestra algunas de las mejoras en los procesos, logradas durante la operación de la planta de demostración ISASMELT™ para cobre.

El funcionamiento de la planta a escala de demostración permitió resolver temas claves del proceso, independientemente de la corriente principal del proceso. Mientras tanto, esta planta recuperó su inversión inicial durante los primeros 14 meses de operación. Durante este tiempo, se fundió concentrado de escoria de convertidor acumulada por 10 años [9]. Se realizó una serie de ensayos a gran escala para demostrar el proceso a altas tasas de alimentación, lo que reduce la ampliación desde la planta de demostración a gran escala a un factor de 2. El enfoque sistemático para la ampliación durante el desarrollo de la tecnología nuevamente permitió una cuidadosa evaluación de los aspectos claves de diseño en la escala apropiada, minimizando el impacto financiero en las operaciones de la compañía. Como resultado, se obtuvo un diseño de proceso mucho más robusto para su aplicación comercial. El buen funcionamiento de la planta llevó a la decisión de construir la primera planta comercial de cobre ISASMELT™. Esta planta comenzó su actividad en 1992. Varias publicaciones han informado acerca de las principales conclusiones del proceso de desarrollo ISASMELT™ en la fundición de cobre de Mount Isa Mines desde que la planta comercial comenzó su actividad [10-14]. La capacidad de detectar y resolver problemas y aspectos críticos del proceso previo a la construcción de la primera planta comercial, sin duda contribuyó a su éxito.

Tras cinco años de funcionamiento de la planta comercial, se llevó a cabo en 1997 una prueba a gran escala para demostrar que la tasa de alimentación de concentrado podría elevarse a 150 t / h a partir de la tasa de diseño de 100 t / h. Esta demostración exitosa permitió que la administración de Mt Isa Mines aprobara más gastos de capital para ampliar la capacidad de producción de la planta de 180.000 a 240.000 tpa de cobre contenido. Este fue el último paso en la ampliación del proceso de cobre para las operaciones de Mount Isa.

Otro resultado de la exitosa operación de la planta de demostración fue la decisión de comercializar la tecnología ISASMELT™ para empresas externas. Cyprus Miami en Arizona (ahora Freeport McMoRan Miami) fue autorizada para diseñar y construir su planta ISASMELT™ con base en la planta de demostración. La planta de Miami se construyó al mismo tiempo que la primera planta comercial en Mount Isa. Desde principios de los años 1990, Xstrata ha proporcionado licencias de sus diseños ISASMELT™ a empresas de todo el mundo. La tabla 4 muestra una lista de las plantas de cobre ISASMELT™ diseñadas desde que se construyó la planta de demostración.

Tabla 3 – Principales resultados de la planta de demostración de ISASMELT^{TM} para cobre

Tema	Etapa inicial del proceso de desarrollo	Etapa final del proceso de desarrollo
Vida útil de la lanza	Muy baja: pocos minutos a horas	Alta: Se lograron más de 20 días de operación después de la modificación del diseño y el desarrollo de técnicas para posicionar con precisión la lanza en el baño
Tasa de alimentación	15 t/h de concentrado con la operación basada en aire	Se lograron 48 t/h durante las pruebas utilizando aire enriquecido con oxígeno
Enriquecimiento del oxígeno	Aire – sin adición de oxígeno	El enriquecimiento del oxígeno del aire permitió un aumento de la capacidad del fundidor de cobre en un 20%
Disponibilidad de la planta	Primer año bajo 70% (Diseño 75%)	Se logró la disponibilidad de diseño luego del primer año. El logro máximo superó el 95%
Vida útil del refractario	El primer revestimiento duró 14 semanas	El quinto revestimiento duró 90 semanas

DESARROLLO DE ISASMELT™ PARA RECICLAJE

El proceso ISASMELT™ también ha sido desarrollado para el reciclaje de materiales de desecho como cobre chatarra, residuos electrónicos y metales preciosos. Umicore Precious Metals y Aurubis obtuvieron licencia para operar la tecnología ISASMELT^{TM} que se basa en los diseños originales desarrollados en Mount Isa Mines.

Umicore necesitaba una tecnología de fundición flexible que formaría el núcleo de su nuevo complejo de reciclaje en Hoboken, Bélgica. Ellos colaboraron con Mount Isa Mines para desarrollar la nueva aplicación de la tecnología ISASMELT^{TM}. Siguiendo el ejemplo del proceso de ampliación utilizado con éxito en Mount Isa, Umicore inicialmente contrató a Xstrata para llevar a cabo un programa detallado de planta piloto a fin de poner a prueba los límites de su nuevo proceso. Tras el éxito de las pruebas de la planta piloto, se construyó una planta de demostración y fue operada en la fundición de Hoboken. Se identificó una serie de parámetros claves del proceso durante la operación de la planta de demostración, de manera que el riesgo técnico se redujo considerablemente cuando se diseñó el horno comercial ISASMELT™ a gran escala. La fundición comercial se puso en marcha a finales de 1997 y ahora está procesando hasta 360.000 toneladas al año de materias primas secundarias.

Tabla 4 – Plantas de Cobre ISASMELT™

N°	Fecha de puesta en marcha	Dueño de la Planta	Ubicación de la planta	Tipo	Capacidad de diseño
1	1987	Mount Isa Mines Limited	Mount Isa, Australia	Demostración	120.000 t/año concentrado
2	1991	AGIP Australia Pty Ltd.	Radio Hill, Australia	Níquel/Cobre Comercial	60.000 t/año concentrado
3	1992	Freeport McMoRan Miami	Arizona, EE.UU.	Comercial	700.000 t/año concentrado
4	1992	Mount Isa Mines Limited	Mount Isa, Australia	Comercial	1.000.000 t/año concentrado
5	1996	Vedanta	Tuticorin, India	Comercial	200.000 t/año concentrado[1]
6	1997	Umicore	Hoboken, Bélgica	Cobre Secundario Comercial	200.000 t/año material mezclado
7	2002	Aurubis	Lünen, Alemania	Cobre Secundario Comercial	150.000 t/año material mezclado
8	2002	Yunnan Copper	Kunming, China	Comercial	800.000 t/año concentrado
9	2005	Vedanta	Tuticorin, India	Comercial	1.200.000 t/año concentrado
10	2006	Mopani Copper Mines	Mufulira, Zambia	Comercial	650.000 t/año concentrado
11	2007	Southern Peru Copper Corporation	Ilo, Perú	Comercial	1.200.000 t/año concentrado
12	2009	Yunnan Copper	Chuxiong, China	Commercial	500.000 t/año concentrado
13	2009	Yunnan Copper	Chambishi, Zambia	Comercial	350.000 t/año concentrado
14	2010	Yunnan Copper	Liangshan, China	Comercial	500.000 t/año concentrado
15	2011	Kazzinc Ltd	Ust-Kamenogorsk, Kazakhstan	Comercial	290.000 t/año concentrado
16	2012	Doe Run Peru	La Oroya, Perú	Comercial	280.000 t/año concentrado
17	2012	Vedanta	Tuticorin, India	Comercial	1.360.000 t/año concentrado

Nota 1: Capacidad inicial de diseño 60.000 tpa de cobre en mata, pero ampliada a 150.000 tpa de cobre en mata con un mayor enriquecimiento de oxígeno.

Aurubis también necesitaba adaptar la tecnología ISASMELT™ para su planta de reciclaje de cobre en Lünen, Alemania. Una vez más se llevó a cabo una serie de ensayos detallados en la planta piloto Xstrata para identificar los parámetros clave del proceso antes del diseño y construcción de la planta comercial. Esta última se puso en marcha en 2002. Los ensayos adicionales de la planta piloto se utilizaron para probar una variedad más amplia de material de alimentación. Como resultado, la planta ahora procesa más de 220.000 tpa de materiales de cobre secundarios.

FACTOR DE ESCALAMIENTO UTILIZADO EN EL DESARROLLO DE LA TECNOLOGÍA ISASMELT™

El desarrollo de una nueva tecnología de fundición lleva tiempo. Requiere de una disciplina rigurosa para evitar la tentación de saltarse etapas a fin de acelerar el desarrollo. Se necesitaron aproximadamente diez años para el desarrollo de la tecnología ISASMELT™ para plomo y cobre desde el crisol hasta la escala de demostración. Durante esta década, el conocimiento que se fue acumulando permitió que el equipo de desarrollo alcanzara un punto en el que estaban mucho mejor equipados para diseñar y construir una planta comercial a gran escala, etapa final del proceso de ampliación. Los aspectos claves en este proceso fueron la selección de los factores de la ampliación y del diseño sistemático, el desarrollo y la reingeniería de varios componentes de la tecnología. La figura 2 muestra una comparación de las etapas de escalamiento de los procesos ISASMELT™ para plomo y cobre. La escala piloto se define como la unidad de comparación.

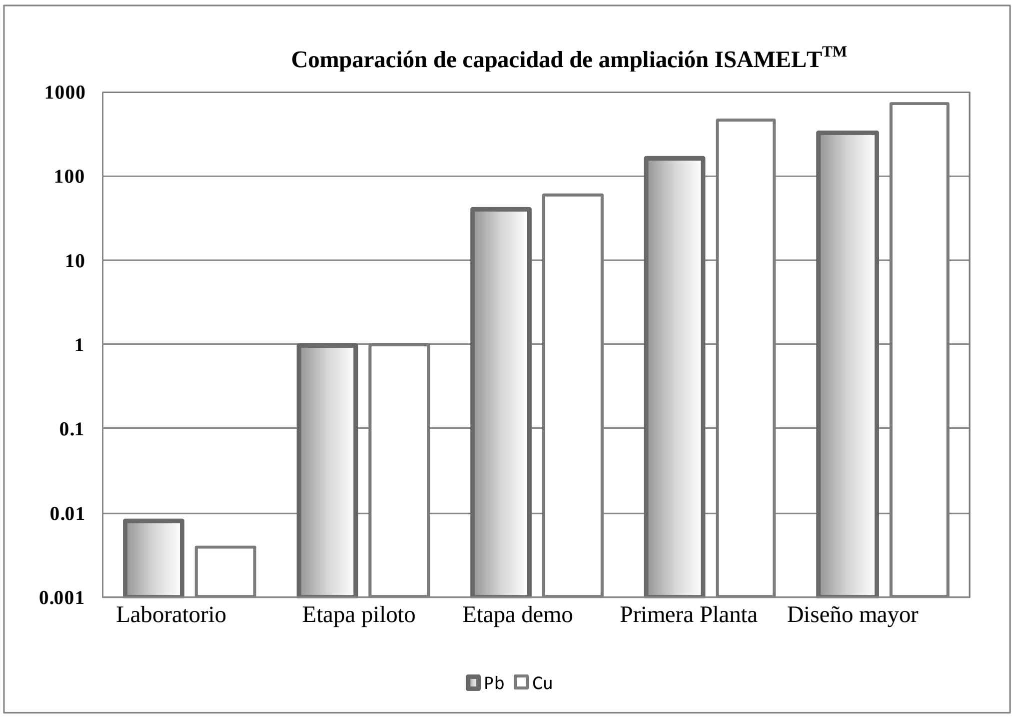


Figura 2 Comparación de ampliación de ISASMELT™ para plomo y cobre

Durante el proceso de ampliación se desarrollaron varios aspectos con un alto nivel, lo que permitió que la tecnología ISASMELT™ se convirtiera en un éxito comercial. La Tabla 5 compara algunos de los componentes e indicadores claves de las tecnologías ISASMELT™ para plomo y cobre, en las diferentes etapas de sus procesos de desarrollo.

Un parámetro importante en la evolución de la tecnología ISASMELT™ ha sido la vida útil del refractario. La Figura 3 muestra la historia de las campañas de refractarios en la planta comercial ISASMELT™ para cobre en Mount Isa desde la puesta en marcha. Hubo un momento en que la administración de Mount Isa Mines consideró inadecuada la instalación de refrigeración por agua en los refractarios del horno debido a la posibilidad de incidentes fatales y el aumento en los costos de operación. Como resultado, se construyeron hornos a escala comercial con enfriamiento mínimo de agua. Si bien en un principio esto acortó la vida de la campaña, se inició un programa de desarrollo que se centró en optimizar la selección de materiales refractarios y la metodología de instalación. Al combinar esto con estrategias de control de procesos y con monitoreo continuo en línea de la temperatura del baño utilizando sistemas desarrollados por más de 10 años de operación, Mount Isa Mines pudo lograr alargar la vida de la campaña por más de 3 años sin usar ninguna refrigeración por agua de los refractarios del horno.

Tabla 5 – Indicadores claves de las plantas ISASMELT™ desde la escala piloto hasta la comercial

Tema	Unidad	Escala Piloto		Escala Demo		Primera Escala Completa		Diseño Actual[1]	
		Pb	Cu	Pb	Cu	Pb[3]	Cu	Pb	Cu
Id del horno	m	0,4	0,4	1,8	2,3	2,5	3,75	3,6	4,4
Diámetro de lanza	mm	38	38	150	250	250	350	250	500
Control de lanza	-	Manual		Semi Automático		Semi Automático		Automático	
Enriquecimiento de oxígeno	%	21	21	21	28	35	45	70	90
Tasa nominal de alimentación	tph	0,12	0,25	5	15	20	101	40	183
Tratamiento de emanación	-	Gases de escape/ Mangas		Enfriador de Gas / Manga		WHB		WHB[2]	

Notas:
ID: Diámetro interior; WHB: Caldera de calor residual
[1] Se refiere al rendimiento máximo
[2] Algunas plantas utilizan una combinación de sección de radiación y enfriador evaporativo para el tratamiento de gases de escape
[3] Se refiere al horno de fundición del proceso ISASMELT™ de dos etapas para plomo

CONCLUSIONES

Un proceso de escalamiento gradual puede llevar al éxito en el desarrollo de procesos de fundición innovadores. Se ha seguido esta metodología desde que en Mount Isa Mines se reconoció el potencial de la lanza Sirosmelt en la década de 1970. Esto ha sido un aspecto clave en el éxito del desarrollo de la tecnología ISASMELT™. Siguiendo paso a paso las diferentes etapas del proceso desde la escala de laboratorio pasando por la experimentación y demostración hasta la aplicación comercial, se pudo tratar e identificar problemas críticos. De esta manera, el riesgo técnico se redujo en cada etapa y también disminuyó al mínimo el costo total de la inversión. La participación del personal de operaciones en todas las etapas del desarrollo permitió crear un sentido de propiedad y entusiasmo por el proceso, lo que hizo que los problemas se abordaran y se resolvieran mientras todavía se podía. Con las dificultades propias en la escala de demostración, los procesos de cobre plomo demostraron en gran medida su viabilidad económica antes de que se requirieran grandes inversiones de capital para su implementación comercial.

Como las innovaciones evolucionan desde su conceptualización inicial, se van adaptando. Quienes la van adoptando, llevan la innovación a nuevas fronteras. Este ha sido el caso con la tecnología ISASMELT™. Desde su exitosa comercialización a principios de los 1990, ISASMELT™ se ha incorporado en las plantas de todo el mundo y ahora se utiliza para una variedad de aplicaciones. Se espera que crezca el uso de ISASMELT™ para fundición secundaria. Precious Metals con su aplicación secundaria de cobre-plomo-pgm en Hoboken, Bélgica y Aurubis con su planta de cobre secundario en Lünen, Alemania, constituyen ejemplos exitosos de la tecnología ISASMELT™ aplicada a nuevas fronteras. Ambas plantas alcanzan altos estándares ambientales con el uso de la tecnología ISASMELT™ además de operar con una viabilidad económica muy superior en comparación con la tecnología convencional. Ellos, junto con los procesos de plomo y cobre primarios comercializados en Mount Isa, representan el resultado final de un programa bien planificado y con éxito consumado de la ampliación gradual.

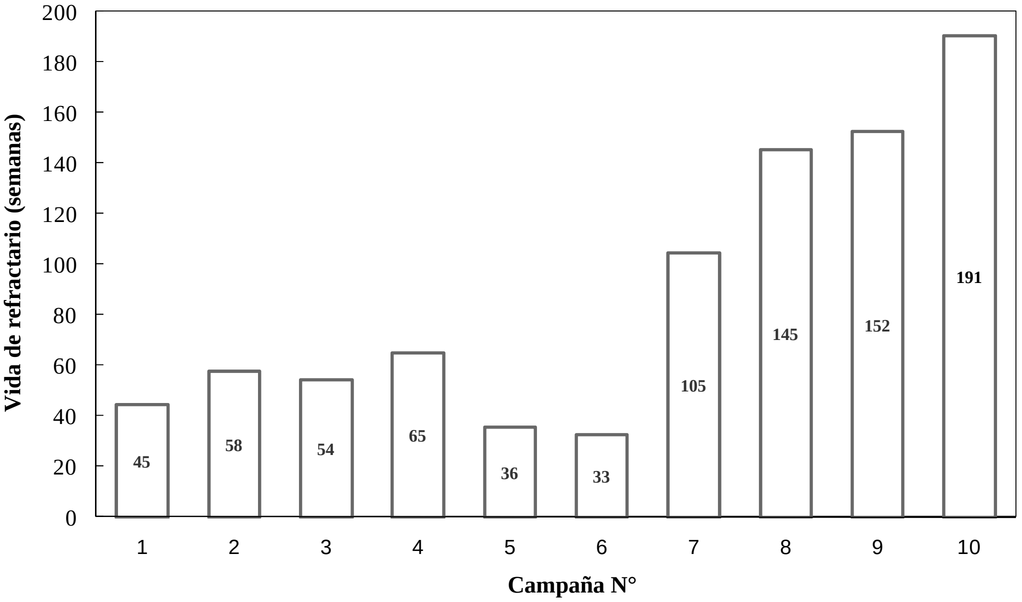


Figura 3 – Campañas de Planta de Cobre ISASMELT™ en Mount Isa (2010)

Un resultado de este largo y deliberado programa de desarrollo y su posterior comercialización, es que Xstrata ha probado diseños flexibles y eficientes para este proceso de fundición que ha estado en operación comercial desde hace más de 20 años. Durante ese tiempo, Xstrata Technology ha transferido la tecnología ISASMELT™ con éxito a numerosas fundiciones de todo el mundo. Los propietarios pueden utilizar la tecnología para modernizar las operaciones existentes o crear nuevas empresas con un mínimo riesgo técnico.

REFERENCIAS

1. J.H. Fewings, “Management of Innovation – The ISASMELT Process”, Innovation in Metal Production Technical Meeting, AMIRA, Mount Isa, 3-4 October 1988.

2. W.J. Errington, J.H. Fewings, V.P. Keran and W.T. Denholm, “The ISASMELT Lead Smelting Process”, Transaction. Institution of Mining and Metallurgy, Section C, 96, March 1987, C1-C6.

3. Ibid 2.

4. P.S. Arthur and S.P. Hunt, “Isasmelt – 25 Years of Continuous Evolution”, Sustainable Developments in Metal Processing, Floyd Symposium, M. Nilmani and W. J. Rankin, Eds., Melbourne, Australia, 2005, 73-94.

5. W.T. Denholm, J.M. Floyd, W.J. Errington and A.N. Parry, “High Intensity Lead Smelting Process”, U.S. Patent 4514222, 18 November 1982.

6. Ibid 2.

7. Ibid 4.

8. W.J. Errington, J.S. Edwards and P. Hawkins, “ISASMELT Technology – Current Status and Future Development” Colloquium: Trends in Base Metals Smelting and Refining, South African Institute of Mining and Metallurgy Eds., South Africa, 1997.

9. C.R. Fountain, J.M.I. Tuppurainen, N.R. Whitworth and J.K. Wright, “ New Developments for the Copper ISASMELT Process”, Extractive Metallurgy of Copper, Nickel and Cobalt, Volume II, C. A. Landolt, Eds., TMS, Warrendale, 1993, 1461-1473.

10. R. Player, “Copper ISASMELT – Process Investigations”, The Howard Worner International Symposium on Injection in Pyrometallurgy, M. Nilmani and T. Lehner, Eds., TMS, Warrendale, 1996, 439-446.

11. J. S. Edwards, “ISASMELT – a 250,000 tpa Copper Smelting Furnace”, AusIMM ’98 – AusIMM ’98 - The Mining Cycle, Mount Isa, Australia 1998, 395-400.

12. P. Arthur, B. Butler, J. Edwards, C. Fountain, S. Hunt and J. Tuppurainen, “The ISASMELT Process – An Example of Successful Industrial R&D”’ Yazawa International Symposium on Metallurgical and Materials Processing, Volume II – High Temperature Metal Production, F. Kongoli, K. Itgakai, C. Yamauchi, H. Y. Sohn, Eds., TMS, Warrendale, 2003, 281-291.

13. G.R.F. Alvear F., P. Arthur and P. Partington, “Feasibility to Profitability with Copper ISASMELT™”, Copper 2010, Volume 2 Pyrometallurgy I, GDMB Eds., Dusseldorf, Germany, 615-630.

14. Ibid 9.

ISASMELT™ EN MUFULIRA – AUMENTO EN LA FLEXIBILIDAD DEL CINTURÓN DE COBRE EN ZAMBIA

Alistair Burrows[1], Phil Partington[1], John Sakala [2], Paul Hermit Mascrenhas[2]

[1]Xstrata Technology, Lv 4, 307 Queen St., Brisbane, QLD, 4000, Australia;
[2]Mopani Copper Mines plc, Mufulira Smelter, PO Box 22000, Kitwe, Zambia;

Palabras clave: ISASMELT™, Eficiencia, Cinturón de cobre

RESUMEN

Desde que se construyó la primera planta en Mount Isa, veinte plantas ISASMELT™ hab sido construidas o están en construcción. Recientemente se ha incorporado a la familia de plantas ISASMELT™ la fundición de Mufulira en el cinturón de cobre de Zambia, donde la planta ISASMELT™ reemplazó a un secador rotatorio y a un horno de fusión eléctrico. El cambio ha sido beneficioso para la fundición de Mufulira.

La tecnología de fusión de cobre primaria ISASMELT™ de lanza sumergida (TSL) sigue atrayendo el interés de posibles usuarios que buscan fundir de manera productiva. Además de los beneficios conocidos en cuanto a productividad y flexibilidad, los operadores ISASMELT™ también han logrado mejoras en la eficiencia energética. Las estadísticas de producción de la fundición Mufulira muestran que el consumo de energía por unidad de producción se ha reducido en forma importante con la introducción de la tecnología ISASMELT™.

INTRODUCCIÓN

El proceso ISASMELT™, basado en la lanza Sirosmelt inventado por CSIRO y comercializado por Mount Isa Mines (MIM) en las últimas tres décadas, estaba destinado a utilizarse para la fundición de concentrados de plomo. MIM pronto descubrió que esta también era una tecnología adecuada para la fundición de cobre [1,2]. Como en el caso de MIM, otras fundiciones de cobre en busca de mejoras en sus procesos han dado cuenta de las ventajas de sus bajos costos operativos y de su capacidad para producir gases residuales con un alto contenido de SO_2, de modo que el gas proveniente de la fundición pueda tratarse más económicamente en una planta de ácido.

Mopani Copper Mines Plc (MCM) decidió, a finales de 2003, reemplazar su horno de fundición eléctrica y modernizar su planta de fundición en Mufulira con la instalación de un horno ISASMELT™ y un horno eléctrico de sedimentación de mata (MSEF). La nueva instalación fue diseñada para el tratamiento de concentrados externos, además de los producidos en las minas MCM [3]. El horno ISASMELT™ fue diseñado, construido y puesto en marcha en 30 meses, y desde entonces la producción ha aumentado constantemente en los años posteriores, con opciones de continuar aumentándola en el futuro.

Las características del horno ISASMELT™ permiten que se adapte fácilmente a fundir concentrados de cobre de un amplio rango de composición. Esta capacidad de adaptación ha permitido a la fundición Mufulira aceptar una mezcla de alimentación que habría hecho ineficiente e improductivo el esquema de fundición anterior.

ESQUEMA DE LA FUNDICIÓN MUFULIRA ANTES DE 2006

La fundición de Mufulira comenzó su actividad en 1937 [4]. De 1991 a 2006, la capacidad de fundición fue de alrededor de 400.000 toneladas de concentrados de cobre por año, en su mayoría procedente de las minas de MCM (es decir, las minas Mufulira y Nkana) a través de un esquema que incluía un secador rotatorio que utilizaba aceite combustible pesado (HFO por sus siglas en inglés) de diseño FFE y un horno eléctrico de arco sumergido de 36MVA de diseño Elkem, con seis electrodos en línea. No había instalaciones para captura de dióxido de azufre [3].

El horno eléctrico propiamente tal producía poco SO_2 directamente, por su rol de "fundir" una mezcla de concentrado rico en cobre y relativamente bajo en hierro, produciendo mata con una ley natural superior al 50% en peso de Cu. Las composiciones típicas de concentrados de las minas de MCM se muestran en la Tabla 1.

Tabla 1 – Composición normal de concentrados MCM

	%Cu	%Fe	%S	$\%SiO_2$	%CaO	%MgO	$\%Al_2O_3$
Mufulira	38-41	15	23	16	0.8	2,1	2,5
Nkana	30-32	23	30	12	1.5	1,1	1,9

El horno eléctrico tenía una capacidad nominal de tratamiento de concentrados de 420.000 toneladas por año (tpa). Para el año 2003, se reconoció que el horno eléctrico requeriría la reconstrucción o sustitución por tecnología alternativa. MCM eligió la segunda opción, en parte debido a que con una nueva tecnología de fundición se podría producir ácido sulfúrico a bajo costo. También fue ventajoso construir la nueva planta adyacente a la antigua, conectada a la nave de Convertidores Peirce Smith (PS), con el objetivo de lograr una transición perfecta desde la configuración antigua a la nueva al final de la operación del horno eléctrico.

ESQUEMA DE LA FUDICIÓN MUFULIRA DESPUES DEL AÑO 2006

En la Figura 1 se muestra el nuevo esquema de fundición primaria de la fundición Mufulira. El horno ISASMELT™ es la característica central de la planta.

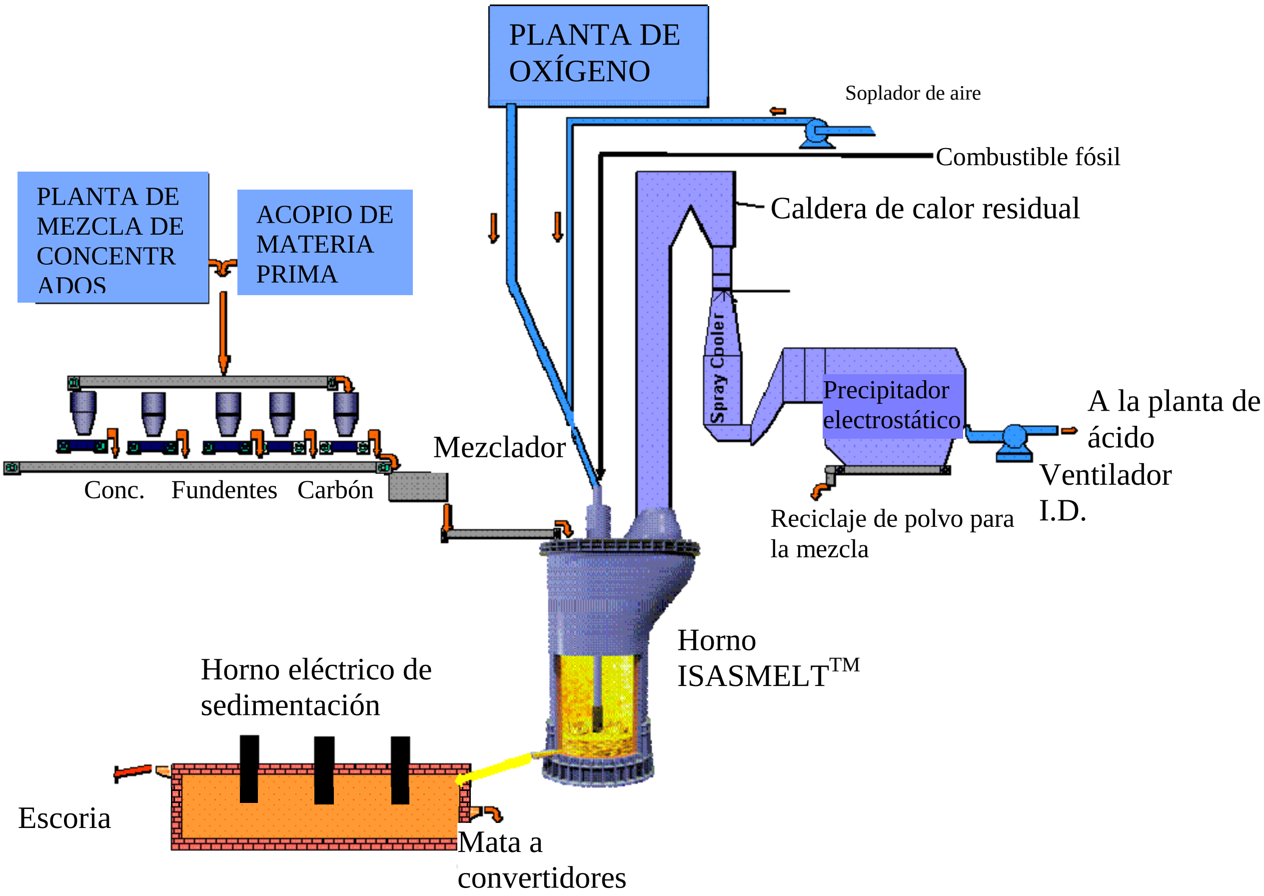


Figura 1- Esquema Planta ISASMELT™ en Mufulira

En la planta se incluye lo siguiente:

- Un área mejorada de manipulación y de almacenamiento de concentrado.
- Una planta de oxígeno criogénico de 650 tpd.
- Un horno eléctrico para sedimentación de mata (MSEF por sus siglas en inglés), capaz de procesar los productos generados en el horno ISASMELT™ y la escoria de los convertidores Peirce-Smith; generando una escoria para ser acopiada en el vertedero de escorias.
- Una planta de ácido sulfúrico de contacto simple para procesar los gases de generados en el horno ISASMELT™.

En conjunto, estos elementos son responsables de la producción actual de cobre de la fundición de Mufulira y sus perspectivas de futuros acuerdos económicos para la elaboración de concentrado desde dentro del cinturón de cobre de Zambia y la República Democrática del Congo (RDC).

Rol de Xstrata Technology en el desarrollo de la fundición de Mufulira

Xstrata Technology (XT), licenciante del proceso ISASMELT™, participó en el desarrollo e implementación del nuevo esquema de fundición en Mufulira, lo que incluyó actividades claves en el diseño, suministros, capacitación de los operadores y puesta en marcha.

El alcance de trabajo de XT se dividió en las siguientes etapas:

Ingeniería de diseño:

En marzo de 2004, MCM firmó un Acuerdo de Servicios Profesionales con XT y un Acuerdo de Ingeniería con un contratista de ingeniería, adquisición y gestión de construcción (EPCM). La ingeniería de diseño para el horno ISASMELT™ se llevó a cabo en la oficina XT en Australia con la participación activa del equipo del propietario y de la oficina EPCM en Sudáfrica. El uso regular de herramientas avanzadas de gestión de proyectos permitió a las tres partes compartir y revisar documentos en ambos lados del Océano Índico. XT revisó la ingeniería mecánica y estructural que entregó el contratista EPCM durante esta etapa. Se hicieron estudios HAZ-Op tanto en el equipo diseñado por XT como en todo el equipo de interfaz.

Suministro del equipo XT:

Como parte del acuerdo de servicios con MCM, el suministro del equipo básico para la planta ISASMELT™ se adjudicó a XT. El paquete básico incluye equipos de elevación automática de lanza, lanzas, quemadores especializados, sistemas de control de combustible, caldera de calor residual (WHB) con enfriador de gas evaporativo, ventiladores de escape para combustión e higiene, sopladores de aire para proceso de combustión y lanza e instrumentación especializada.

Diseño y suministro del sistema de control de procesos:

Una parte importante del éxito de la tecnología ISASMELT™ es su sistema de control de procesos. XT en asociación con MIPAC entregó un paquete completo de control de procesos para la nueva planta de fundición. Esto incluye el hardware distribuido de control del sistema y el software y la configuración de un amplio sistema de control de planta, que abarca el control para la preparación planta de alimentación, operación ISASMELT™, WHB, sistema de gases de escape del proceso de fundición, planta de ácido y planta de oxígeno.

Programa de capacitación:

Se necesita una capacitación exhaustiva para transmitir la tecnología de manera exitosa. El proceso de capacitación de MCM se llevó a cabo en dos fases:

- Capacitación a operaciones mantenimiento por tres meses a escala completa en la fundición MIM en Australia.
- Capacitación in situ durante la puesta en marcha de la planta donde se explicó partida de los hornos en frío y en caliente, partidas de prueba en frío, puesta en marcha, paradas, y casos de mantenimiento y solución de problemas.

Asistencia técnica:

Parte importante del proyecto fue la asistencia técnica por parte de XT a MCM durante la ingeniería, antes y después de la puesta en marcha del proyecto. Los servicios técnicos incluyen la adscripción de personal de XT a MCM y a la fundición Mufulira para llevar a cabo las siguientes funciones:

- Ayudar con la instalación de los elementos claves de equipo.
- Ayudar a supervisar la instalación de todo el sistema de control de la planta.
- Ayudar a supervisar la puesta en marcha y el inicio de la planta.

La planta fue puesta en servicio en septiembre de 2006, 30 meses después de la firma del Acuerdo de Servicios profesionales, y se muestra en la Figura 2.

Figura 2 – Planta ISASMELT™ en la Fundición de Mufulira

CAMBIOS IMPORTANTES EN EL SUMINISTRO DE CONCENTRADO

En el momento de iniciar el proyecto de actualización de la fundición Mufulira, en MCM se anticipaba una producción relativamente constante de aproximadamente 400.000 tpa de concentrado desde las minas y concentradoras de MCM (Nkana y Mufulira) [3], los que iban a entregar gran parte de la alimentación al horno ISASMELT™. Normalmente, Nkana producía el 55-60% de la producción extraída de Mopani y Mufulira producía el resto [4]. Durante los últimos años, los concentrados acordados comercialmente, sobre todo de la mina First Quantum Minerals "Kansanshi", se han vuelto cada vez más frecuentes debido al aumento del rendimiento de la fundición. En la Tabla 2, se muestra una composición normal del nuevo material con contenido de cobre que actualmente recibe la fundición Mufulira

Tabla 2 - Composición normal del nuevo material de alimentación a la fundición Mufulira

	%	%Cu	%Fe	%S	%SiO_2	%CaO	%MgO	%Al_2O_3
Mufulira	15	38-41	15	23	16	0,8	2,1	2,5
Nkana	35	30-32	23	30	12	1,5	1,1	1,9
Kansanshi#	50	26	27	33	8	0,3	0,7	1,4

Otro concentrados de la mina también están dentro de los acuerdos comerciales, pero este es el componente más destacable.

Como se indica en la Tabla 2, también se ha producido un efecto secundario al recibir más concentrado de una tercera parte: una disminución en la ley promedio de cobre que alimenta la fundición. Considerando que el concentrado de Mufulira contiene una cantidad importante de bornita y calcosina, los concentrados de tratados comercialmente en su mayoría sólo contienen calcopirita. Aunque esto hubiera sido perjudicial para la producción de la fundición utilizando el esquema original de fundición primaria, lo que incluía un secador rotatorio y un horno eléctrico, MCM ha podido fundir en forma productiva mediante el esquema actual que incluye un horno ISASMELT™ y un MSEF. De hecho, ha sido un ajuste excelente. La fundición Mufulira produce ahora, por unidad de producción de cobre, mucho más ácido sulfúrico de lo que hubiera sido el caso. El ácido se puede utilizar para operaciones de lixiviación o se puede vender a otras operaciones en la región.

COMPARACIÓN DEL ESQUEMA ANTERIOR VERSUS EL ACTUAL

Se ha tratado de hacer una comparación estadística de los méritos relativos a los esquemas actuales y anteriores. No existe un método estándar para hacer esto. La diferencia en la escala de producción y las peculiaridades de las materias primas pueden complicar la tarea de hacer comparaciones importantes entre las operaciones industriales. En un análisis detallado de este tema, Goonan [5] utiliza un esquema genérico sobre fundición de cobre y flujos normalizados de materiales y energía, para efectos de comparar diferentes fundiciones de cobre. Se adoptó el enfoque de Goonan, con la modificación necesaria de que los límites considerados en este trabajo sólo se extienden alrededor de la etapa de fundición primaria y de limpieza de escoria y no hacia las secciones de conversión y refinación a fuego.

En la Figura 3 se muestra un esquema genérico de fundición primaria. El área de fundición primaria está encerrada dentro de la línea roja segmentada. El material y las corrientes de energía sujetos a comparación están etiquetados individualmente. Todas las mediciones de flujo de masa están en unidades de toneladas por tonelada de cobre producido. Las mediciones de electricidad están en MWh por tonelada de cobre producido. No todas las corrientes son aplicables a todas las fundiciones. Por ejemplo, la corriente J se define como cero para un horno eléctrico de limpieza de escoria, pero este sería un valor distinto de cero si la limpieza de escoria se realizara utilizando un proceso de flotación.

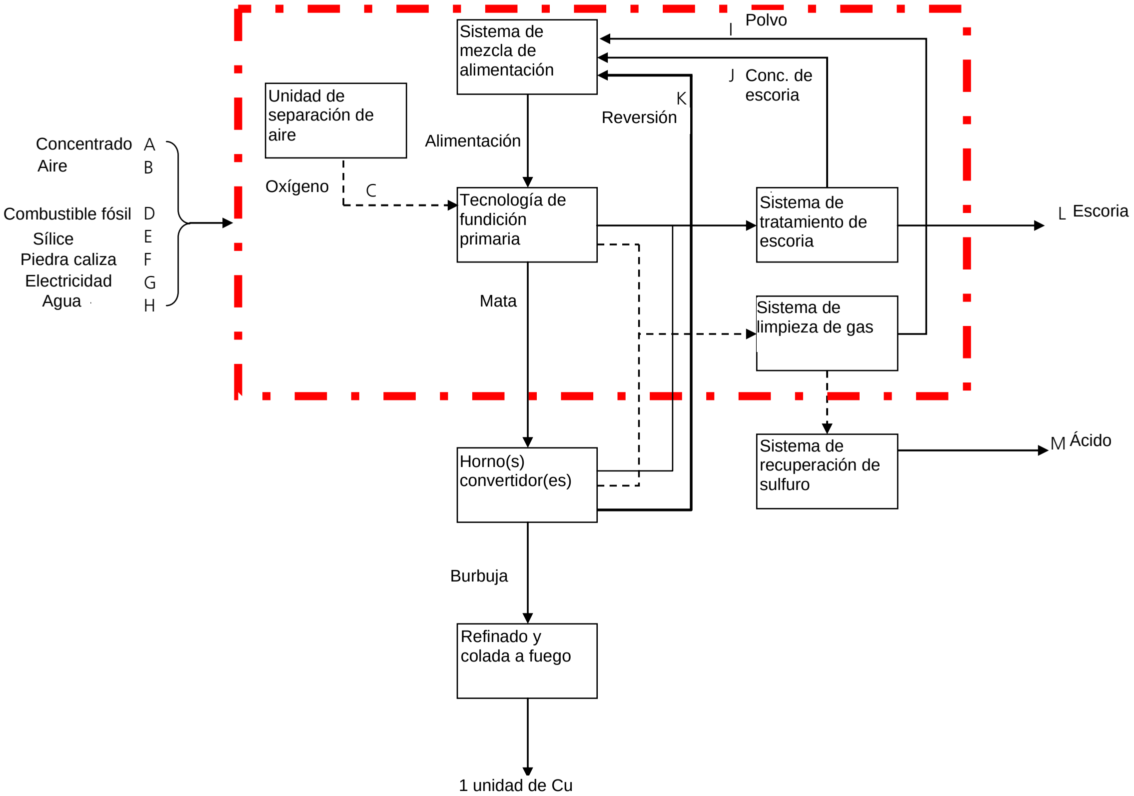


Figura 3 – Diagrama simplificado de una fundición de cobre.

Para efectos de consistencia, las cifras de electricidad citadas en este trabajo:

- Incluyen la generación de oxígeno industrial;
- Incluyen la compresión de aire por el ventilador del proceso de fundición primaria;
- Incluye la electricidad para el MSEF;
- Incluye el gas de escape del o de los ventiladores de fundición primaria;
- Incluye a los consumidores menores de toda el área de fundición primaria; pero,
- Excluye cualquier beneficio de compensación a partir de la electricidad generada utilizando vapor de agua desde el sistema de recuperación de calor residual (esto, en caso de la fundición de Mufulira es igual a cero).

Los principales resultados de la comparación del antes y el después en la fundición de Mufulira son que se notó un ahorro sustancial en el consumo de electricidad, y un aumento (desde cero) en la generación de ácido sulfúrico cuando se expresa por unidad de producción de cobre. El resultado del cambio del horno eléctrico de fundición por uno ISASMELT™ y MSEF, fue que el consumo específico de energía eléctrica se redujo considerablemente (en un 39%), incluso teniendo en cuenta la introducción de una planta de oxígeno de 650 tpd. La caída del consumo específico de energía eléctrica fue acompañada por un pequeño aumento en el consumo específico de combustible fósil. Estos datos se muestran en la Tabla 3.

Tabla 3 – Comparación de antes y después en la fundición de Mufulira

		Mopani Copper Mines Mufulira Zambia	
Tecnologías de Fundición Primaria		EF	ISASMELT + MSEF
Período		CY 2004	CY 2009
Concentrado	A	1,87	2,80
Aire	B	0	0,27
Oxígeno	C	0	0,67
Combustible fósil	D	0,02	0,15
Sílice	E	0	0,06
Caliza	F	0,24	0
Electricidad	G	1,57	0,96
Agua	H	0	0,26
Polvo	I	0,07	‡
Concentrado de escoria	J	0	0
Reversiones	K	0,35	0,52
Escoria	L	1,36	1,61
Ácido sulfúrico	M	0	1,14

Notas: Las corrientes se definen en la Figura 3. Las unidades de corriente son en MWh / t Cu. Las unidades de masa son en t / t Cu.
Abreviaturas: EF = Horno eléctrico, CY – Año calendario, ‡ – no disponible

COMPARACIÓN CON USUARIOS SIMILARES DE ISASMELT™

La fundición Mufulira es la tercera fundición de cobre que reemplaza un horno eléctrico de fundición por un horno ISASMELT™, y lo combina con un MSEF. Las dos fundiciones anteriores que lo hicieron son Freeport Miami [6], en Arizona y Yunnan Copper en Kunming, en China [7], respectivamente. Es instructivo ver cómo Mufulira se compara con estas dos operaciones de fundición.

Las historias de las operaciones ISASMELT™ en Miami y Kunming son similares en algunos aspectos. Ambas tenían hornos eléctricos de fundición de seis en línea que fueron modificados para actuar solamente como hornos de sedimentación. La fundición Mufulira se diferencia de estas dos operaciones en que tiene un MSEF especialmente diseñado para acompañar al horno ISASMELT™. Las tres fundiciones producen mata con un contenido de cobre similar (52-59%), aunque Miami y Kunming comienzan con concentrados de cobre de menor ley.

Los datos comparativos de los tres hornos de fundición se muestran en la Tabla 4. Los datos indican que la fundición de Mufulira utiliza menos aire y menos oxígeno en el horno ISASMELT™ y produce menos escoria desde el MSEF que las de Miami y Kunming, cuando se expresa por tonelada de cobre producido. Esto se espera de una planta de tratamiento de concentrado de ley superior.

La fundición de Mufulira consume más agua en la planta ISASMELT™ que la de Miami. Esto se puede atribuir a la utilización parcial de enfriamiento por evaporación para el gas de escape en Mufulira, mientras que Miami y Kunming enfrían los gases de escape de ISASMELT™ a una temperatura similar utilizando la sección de convección de sus respectivas calderas de calor residual.

En otros aspectos, cada planta tiene capacidad similar y ha logrado estadísticas similares.

Tabla 4 - Parámetros de evaluación comparativa para las fundiciones de cobre que utilizan la tecnología ISASMELT™ + MSEF

		Freeport Miami EE.UU.	Yunnan Copper Kunming China	Mopani Copper Mines Mufulira Zambia	
Tecnologías de Fundición Primaria		ISASMELT + MSEF	ISASMELT + MSEF	EF	ISASMELT + MSEF
Período		CY 2007	CY 2006	CY 2004	CY 2009
Concentrado	A	3,45	4.25	1,87	2,80
Aire	B	1,29	0.88	0	0,27
Oxígeno	C	1,02	0.82	0	0,67
Combustible fósil	D	0,18	0,14	0,02	0,15
Sílice	E	0,16	0,06	0	0,06
Caliza	F	0,05	0,00	0,24	0
Electricidad	G	0,94	0,86	1,57	0,96
Agua	H	0,014	0,50	0	0,26
Polvo	I	0,02	0,007	0,07	‡
Concentrado de escoria	J	0	0	0	0
Reversiones	K	0,33	0	0,35	0.52
Escoria	L	2,41	2,62	1,36	1,61
Ácido sulfúrico	M	3,35	3,33	0	1,14

Notas: Las corrientes se definen en la Figura 3. Las unidades de corriente son en MWh / t Cu. Las unidades de masa son en t / t Cu.
Abreviaturas: EF = Horno eléctrico, CY = Año calendario, ‡ = no disponible

El hecho de que la fundición de Mufulira sea la única que tiene un MSEF construido especialmente, mientras que Miami y Kunming modificaron sus hornos eléctricos de fundición anteriores, no parece haber influido en el consumo específico de energía del esquema general de fundición primaria. El MSEF de Mufulira [3] es un horno eléctrico de tres en línea, con capacidad de 12 MVA que usa electrodos Søderberg. Las dimensiones del horno son 18,1 m x 7,4 m x 5,7 m (Largo x Ancho x Alto) y su capacidad de retención de mata es de aproximadamente 400 tons. Una pequeña cantidad de coque alimenta al MSEF, principalmente para reducir la escoria más oxidada del convertidor PS. El MSEF ha demostrado ser una eficaz unidad de limpieza de sedimentación y escoria para la fundición de Mufulira. Produce una escoria de descarte con un contenido de cobre de menos de 0,7% en peso de Cu, y también juega un rol importante en retener suficiente mata para permitir una capacidad de reacción entre el horno ISASMELT™ continuo y los Convertidores PS por lotes. El MSEF complementa muy bien el horno ISASMELT™.

PERSPECTIVA FUTURA PARA LA FUNDICIÓN DE MUFULIRA

Los parámetros claves del proceso para la planta ISASMELT™ de la fundición Mufulira para mayo de 2010 se presentan en la Tabla 5. Al momento en que se escribió este artículo, la planta estaba funcionando en alta disponibilidad con una velocidad de alimentación de hasta 105 t / h (de concentrado de cobre, con exclusión de corrientes de recirculación internos, en base seca).

La producción de fundición en Mufulira todavía está en expansión a medida que se van produciendo mejoras continuas a la operación. Se prevé una modernización y expansión en la operación del Convertidor PS, lo que también dará lugar a la instalación de otra planta de ácido sulfúrico para capturar los gases de azufre restantes que se emiten actualmente a laatmósfera.

Tabla 5 – Parámetros normales de proceso para la Planta ISASMELT™ en Mufulira a mayo de 2010

Parámetro	Valor	Unidad
Tasa normal de alimentación de concentrado	90-100	seco t/h
Contenido promedio de cobre en el concentrado	29	%
Humedad promedio del material	9	%
Tasa promedio de alimentación de flujo de sílice	1,5	seco t/h
Tasa promedio de alimentación de carbón	1,5	seco t/h
Tasa promedio de alimentación de reversiones	15	seco t/h
Ley promedio de cobre de la mata	54-55	%
SiO_2/Fe promedio en escoria	0,95	--
Tasa promedio de flujo de aire de lanza ISASMELT™	6.0	Nm^3/s
Contenido promedio de oxígeno en el aire de lanza ISASMELT™	76-78	%
Rango de temperatura del baño	1175 - 1185	°C

Se han estado abriendo nuevas minas en el cinturón de cobre de Zambia, y se prevén más para los próximos años. Mufulira puede aprovechar el aumento del suministro de concentrado en esta región y su futuro se ve seguro. También es ventajosa la producción de ácido adicional, ya que se cuenta con consumidores locales que ya poseen operaciones hidrometalúrgicas y que pretenden abrir otras.

CONCLUSIONES

La fundición de Mufulira ha modernizado su esquema de fundición incluyendo un horno ISASMELT™ y un MSEF. El funcionamiento de la planta actual partió cuando la tecnología de fundición anterior, un secador rotatorio y un horno eléctrico de fundición, llegaron al final de su vida útil. La producción de la fundición ha aumentado de manera constante utilizando la nueva tecnología para tratar comercialmente una creciente cantidad de concentrado de cobre a fin de complementar la producción de las minas Nkana y Mufulira de MCM.

El consumo específico de energía eléctrica (es decir, el consumo por unidad de cobre producido) se redujo en un 39% después de introducir el nuevo esquema de fundición en Mufulira. La fundición Mufulira está operando según estadísticas similares a otras fundiciones bien establecidas (Miami y Kunming), que también están usando un esquema ISASMELT™ y MSEF

RECONOCIMIENTOS

Los autores desean reconocer a Mopani Copper Mines, Yunnan Copper Corporation y Freeport McMoran Miami por facilitar los datos para su publicación.

Los autores igualmente desean agradecer a Xstrata Technology y Mopani Copper Mines por su autorización para publicar este artículo.

REFERENCIAS

[1] P.S. Arthur and S.P. Hunt, “ISASMELT™ - 25 Years of Continuous Evolution”, *Floyd International Symposium on Sustainable Development in Metals Processing*, M. Nilmani and W.J. Rankin, Eds., NCS Associates (Australia), 2005, pp 73-94.

[2] B.Burford, “The ISASMELT™ Technology Package: Over 30 Years of Innovation” *The AusIMM Bulletin, Journal of the Australasian Institute of Mining and Metallurgy,* No 1, February 2009, pp.26-30.

[3] Ross, J. & de Vries, D., “Mufulira Smelter Upgrade Project – ‘Industry’ Smelting on the Zambian Copperbelt”, *Pyrometallurgy 2005,* Capetown, Minerals Engineering International, 2005.

[4] “Mopani to Invest US$0.25Billion on Copperbelt in 2006”, *Mining Review Africa*, Issue 1, pp20-23, 2006

[5] Goonan, T.G., “Flows of Selected Materials Associated with World Copper Smelting”, *U.S. G.S. Open-File Report 2004-1395,* 2005

[6] Bhappu, R.R, Larson, K.H. & Tunis, R.D., “Cyprus Miami Mining Corporation, Smelter Modernisation Project Summary and Status”, *EPD Congress 1994,* pp555-570, TMS, 1993

[7] Yi Feng Shi, “Yunnan Copper’s ISASMELT – Successful Smelter Modernization in China”, *Proc. of the Sohn Symposium*, TMS, 2006

艾萨炉直接炼铅的生产实践与讨论

云南驰宏锌锗股份有限公司曲靖分公司铅厂　代龙果

摘　要：本文简要介绍了驰宏公司艾萨炉炼铅的先进生产工艺，讨论了富氧顶吹炼铅中的物理模型及物理现象，有针对性地研究了烤铅渣及煤粒在高温熔池中的物理化学行为，分析了物料耗氧量的影响因素及温度对艾萨炉直接炼铅的影响。最后，针对目前的艾萨炉炼铅生产工艺提出几点建议。

关键词：艾萨炉；直接炼铅；富氧顶吹；熔池熔炼

1　前言

在传统炼铅法的烧结—鼓风炉流程中，存在许多难以克服的弊端。随着能源、环境污染控制以及生产效率和生产成本对冶炼过程的要求越来越严格，传统炼铅法受到多方面的严峻挑战。为充分利用硫化铅精矿粒子的化学活性和氧化热，减少对环境的污染，提高资源利用率，降低生产成本。云南冶金集团引进开发具有自主知识产权的富氧顶吹熔炼技术，该技术具有环保节能效果好、原料适应性强、产能提升潜力大等优点。

云南驰宏公司富氧顶吹熔炼炉于2005年6月点火投料开始试生产，到2005年12月基本打通了流程，生产趋于稳定。目前，富氧顶吹炼铅技术已经基本成熟，具有处理硫酸铅渣、硫化铅精矿、炼铅烟尘及氧化矿等多种物料的生产工艺。该技术体现了驰宏公司强大的技术研发运用能力以及对国家环保节能的积极响应与支持。

2　艾萨炉炼铅生产流程简述

艾萨炉生产粗铅是一个高度自动化的控制过程，从炉料的配料、上料、熔炼过程气氛和温度的控制以及设备运行状况的监控等作业主要都由艾萨炉主控室通过DCS系统完成控制。其生产控制与备料系统、余热利用系统、收尘系统、硫酸厂、制氧站高度关联，任何一方出现问题都会影响正常投料。艾萨炉炼铅工艺流程如图1所示。

该流程较传统炼铅工艺有以下优点：

（1）对原料适应性强，不仅可以处理硫化铅精矿，还可以处理硫酸铅渣、烟尘返料、氧化铅矿。

（2）采用富氧熔池熔炼，强化了冶金过程，充分利用硫化精矿放出大量热量，熔炼效率与热利用率均较高。

（3）炉体密封好，漏风少，烟气量大为减少，提高烟气SO_2利用率，为硫酸生产提供条件。

（4）余热发电，热源回收再利用充分；收尘系统收尘效果好，尾气达标排放，降低了对环境的污染。

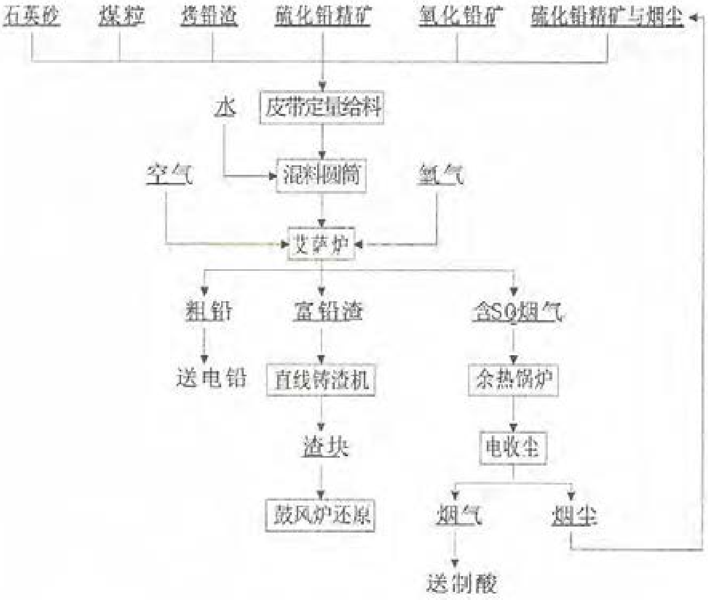


图 1 艾萨炉炼铅工艺流程图

3 艾萨炉熔炼模型及物理现象

3.1 艾萨炉熔炼物理模型

根据熔池熔炼相关理论，熔池熔炼是在气-液-固三相形成的卷流运动[1]中进行物理化学反应的过程。进入熔池的生料，在高温熔池内发生氧化脱硫与造渣反应。喷枪喷入的空气和化学反应生成的气体对熔池起到强力搅拌的作用，强化了冶金过程。

目前，国内外公开报道研究艾萨炉炉内熔体行为的论文较少，张波[2]等人通过冷态模拟实验研究，提出引起体系内熔体流动的动力主要是气泡浮力，而不是湍流的粘性力。并将熔池反应区域划分为四个区：主要反应区，燃料在此区域迅速燃烧，硫化铅精矿发生剧烈氧化反应，并放出大量的热；次要反应区，加入的炉料随熔体向炉壁方向流动，炉料预热融化，并同一起扩散的氧发生少量氧化反应；循环流区，炉料以及喷入气体中的少量气泡从上向下流动，形成回流循环，气泡中残留的氧与精矿和煤充分发生反应；相对静止区，富集粗铅，有利于渣铅分离。

3.2 泡沫渣的形成与影响

（1）泡沫渣的形成

在硫化铅精矿熔炼生产粗铅过程中，熔池中的硫化物、煤粒与氧气反应生成大量的 SO_2、CO 和 CO_2 气体，在炉内形成泡沫状熔体。如果熔渣粘度大，生成的气泡在熔体中无法正常逸出，泡沫渣便会逐渐积累长大，很容易形成较为稳定的泡沫熔渣。

（2）泡沫渣对艾萨炉熔炼的影响

泡沫渣的存在，对艾萨炉炼铅产生了巨大的影响。泡沫渣不仅会造成生产事故，而且严重影响了渣铅分离，导致熔渣排放困难。

4　物料熔炼分析

艾萨炉炼铅原料有硫酸铅渣、氧化铅精矿、硫化铅精矿和烟尘，以及石英石及煤。各种物料的性质不同，所发生的化学反应及作用也不同，这一点在熔炼过程中充分得到体现。由于硫酸铅渣与煤在艾萨炉熔炼过程中的特殊性，下面对硫酸铅渣与煤在熔炼过程中的行为进行分析。

4.1　硫酸铅渣冶金过程

烟化炉处理炼铅炉渣与锌浸出渣，产出氧化锌粉，此类氧化锌粉含有害杂质 F、Cl 较多，成分复杂，一般要做单独浸出处理，所得浸出渣便是硫酸铅渣。

硫酸铅渣主要由硫酸盐组成，因此必须分析 $PbSO_4$ 等硫酸盐的物理化学性质）。

硫酸盐受热分解为金属氧化物和 SO_3。$CaSO_4$ 在 800℃到 1000℃的情况下，很难分解完全。$ZnSO_4$ 的稳定性比硫酸铅小得多，故在此只对 $PbSO_4$ 化学性质进行分析。

在艾萨炉熔炼过程中，硫酸盐（$PbSO_4$）在炉内发生的化学反应主要有：

$2PbSO_4 == PbO \cdot PbSO_4 + SO_3$　　$\Delta H^0 = 404kJ$

$PbSO_4 + PbS = 2Pb + 2SO_2$　　$\Delta H^0 = 426kJ$

$3PbSO_4 + PbS = 4PbO + 4SO_2$　　$\Delta H^0 = 792kJ$

以上三个化学反应均为吸热反应。同时，由于硫酸铅渣品位较低，带入杂质较多，在熔炼过程中必须吸收大量热量造渣。对艾萨炉正常熔炼造成不利影响。

硫酸铅渣中硫酸盐（$PbSO_4$）很难分解，为此研究硫酸铅分解反应的动力学。当温度高于于 860℃时，$PbSO_4$ 的热分解开始反应为

$2PbSO_4 == PbO \cdot PbSO_4 + SO_3$

当温度高于 960℃时，$PbSO_4$ 的发生分解反应，并伴随熔融现象，进一步升温 $PbSO_4$ 便在液态下分解。

通过以上分析，要加速硫酸铅渣熔炼反应速度，可通过提高炉内熔池温度，使其提高到 1000℃以上；也可通过加大熔池搅拌力度，增快硫酸铅渣生成气体的扩散速度。

4.2　煤的燃烧反应

在艾萨炉炼铅时，加入煤粒主要是为了补充热量，保证炉内化学反应顺利进行，同时提高渣的流动性，有利于渣铅分离及富铅渣顺利放出。

通过生产实践，认为添加了煤粒的物料进入炉内，主要存在与 O_2 的燃烧放热、还原 PbO 生成铅的化学反应。由于煤粒的燃烧、还原反应是非均相反应，反应完全需要很长时间，颗粒较大的煤粒在熔池内循环发生上述反应。根据微区化学环境的不同，表现出的化学现象也不相同。

煤的燃烧反应发生在主要反应区，该区域内氧气充足，可以直接燃烧放热，加热熔池。主要反应为：

$C+O_2==CO_2+408kJ$

$C+CO_2==2CO-162kJ$

$2C+O_2==2CO+246kJ$

在主要反应区未燃烧完的煤粒随着炉渣依次进入次要反应区、循环流区、相对静止区。在这几个区域内，由于氧气较少，在此主要发生炭还原 PbO 的反应。炭还原 PbO 生成 Pb，Pb 又被 O_2 氧化生成 PbO，这一循环等于炭燃烧反应放热。。

煤粒在高温熔池中的反应为非均相反应，反应需要时间长，很容易随炉渣进入 3 区、4 区，并且在整个过程中，生成大量的气泡。气泡的存在，带动了煤粒在熔池中做循环运动，强化了传热传质。根据图 4，气泡的浮游现象带动硫化铅与粗铅共熔体，进入 1 区，促进粗铅液中 PbS 的氧化反应，同时，提高粗铅温度与质量。

煤粒在熔渣中穿透能力特强，很容易进入熔池下部，并反应产生大量气泡。一方面，对提高熔池下部温度，提高粗铅质量，加速熔池搅拌都是非常有利的。另一方面，在熔池下部产生大量的气泡，严重影响到渣铅分离与熔渣的顺利放出，所以在放铅放渣前，必须停止煤粒加入，这对艾萨炉炼铅是不利的。

5 空气系数

在艾萨炉熔炼控制过程中，空气系数的确定非常重要。它的大小，不仅关系到产出富铅渣品位的高低，而且关系到粗铅质量的好坏。同时，空气系数的确定也是艾萨炉操作控制中的难点。下面将对影响空气系数的因素进行分析。

5.1 硫化铅精矿

硫化铅精矿是艾萨炉生产粗铅的主要原料。其品位通常并不很稳定，品位高时，空气系数应当较低；而品位较低时，应相应调高空气系数。在生产当中，应当严格把好物料关，稳定物料品位，及时分析，并根据分析结果作出正确调整。

5.2 硫酸铅渣

上面已经全面分析了硫酸铅渣的性质。在高温状态下，它的化学反应并不消耗氧气，相反，由于 PbO 的生成，起到提供氧的作用。但是，硫酸铅渣品位较低，杂质较多，故应采用较高的

空气系数，以保证所产出的富铅渣成分稳定。

5.3 烟尘

艾萨炉生产中，通过在自产铅精矿矿中配入烟尘的方法处理鼓风炉烟尘、艾萨炉烟尘。在烟尘中，铅主要以 $Pb_2(SO_4)O$ 及 $PbSO_4$ 的形式存在，且烟尘品位在55%左右，高于富铅渣品位期望值。所以在大量加入烟尘时，应当相应地调低空气系数，达到降低富铅渣品位的目的。

5.4 氧化铅矿

氧化铅矿主要以白铅矿（$PbCO_3$）的形态存在。白铅矿在700℃时，分解压已达到 $2.9*10^6$Pa。在高温状态下，快速发生分解，化学反应为：

$PbCO_3==PbO+CO_2$

在反应过程中，吸收大量的热。通过化学反应可知，氧化铅矿在艾萨炉熔炼过程中，不消耗氧气，相反，由于其品位较高，生成大量的 PbO，为熔炼过程提供大量的氧。所以在加入大量氧化铅矿的情况下，应当适当降低空气系数。

5.5 煤

煤燃烧需要大量的氧。在计算风量、氧量的过程中，为了保证进入炉内的煤粒能完全燃烧放热，煤燃烧时，部份多余的氧未被消耗。在这种情况下，每小时加入煤量越大，氧气富余量也越大。所以，在熔炼时应参考用煤量对空气系数作出相应调整。

5.6 油

柴油在艾萨炉熔炼中起着重要作用。在使用柴油时，由于喷枪的特殊结构，油嘴喷出的雾化油滴在喷枪口快速燃烧。在喷枪口附近，空气量大，温度高，柴油得到完全燃烧，消耗大量的氧气。同时，由于柴油燃烧激烈，造成气体快速膨胀破裂，熔渣局部温度过高，致使大量氧气未能进入熔池深部，氧气利用率降低。

6 温度对艾萨炉的影响

在火法冶金中，可以说生产的不是金属，而是炉渣。特别是艾萨炉生产粗铅过程中，高温炉渣的性质直接关系到粗铅的质量、数量。更重要的是它的性质直接关系到冶炼过程能否顺利完成，生产能否继续进行下去。炉渣的性质主要由渣型、渣温控制。原料成分相对难于控制，加上在操作控制中不确定因素太多。所以，在冶炼过程中，改变炉渣物理化学性质的方法主要是通过提高炉渣温度实现。

7 结语

艾萨炉熔炼生产粗铅已经成功地用于驰宏公司。建成投产至今，艾萨炉工作人员克服了各种各样的难题，付出了勤劳的汗水。它的成功运行，显示了它强大的生命力。可以预言，它将

取代传统炼铅法成为直接炼铅的主流工艺之一。

然而，任何生产艺都有其不足之处，都需要不断地改进，才能在众多直接炼铅法中占有一席之地。作者对目前的艾萨法炼铅工艺有以下几点改进建议：

（1）改进“正常熔炼—调整排放”的操作控制模式。该操作模式造成了各项操作控制参数的波动，极大地影响了工艺总结与改进，严重影响设备使用寿命。

（2）加强对影响空气系数的因素研究，降低艾萨炉主控操作难度，稳定富铅渣品位、炉温等众多参数。

（3）操作控制应以稳定富铅渣品位为目标。目前，富铅渣品位波动较大，严重影响鼓风炉还原工艺控制，同时影响“艾萨炉氧化—鼓风炉还原—烟化炉吹炼”的经济性。

总之，艾萨炉炼铅属新工艺，在投产至今取得的成果是巨大的。在这个过程中，取得了很多新知识，新经验。在加强经验应用的过程中，我们不应该满足于眼前取得的成果，更应该积极努力，完善该生产工艺。

参考文献

[1] 蒋荣生.提高铅艾萨炉连续熔炼水平的探索和实践 [J].云南冶金,2007,2:36

[2] 张波，洪新，陈朝轶，施哲.艾萨炉水模拟研究 [J].中国稀土学报,2008,8:26

ISACONVERT™ – Непрерывное конвертирование никелевого / МПГ штейна с использованием шлака на основе феррита кальция

М.Л. Беккер, С. Николич, Дж.Р.Ф. Альвеар

Процесс ISASMELT™ представляет собой технологию плавки в жидкой ванне с продувкой методом погружения сверху (TSL), которая развивается и оптимизируется на протяжении последних 25 лет. К концу 2011 года общая установленная мощность технологии ISASMELT превысит 9 млн. тонн в год питания, перерабатываемого на меде- и свинцовоплавильных заводах по всему миру. На промышленных установках в Бельгии и Германии также применяется периодическое конвертирование меди в печах ISASMELT. Данная технология TSL является в равной степени эффективной при непрерывном конвертировании и поэтому получила название ISACONVERT™. Компания Xstrata Technology (XT) недавно запатентовала новый процесс ISACONVERT для непрерывного конвертирования никелевого/МПГ (металлы платиновой группы) штейна с использованием шлаковой системы на основе феррита кальция. В данной статье описывается разработка этого нового процесса и представлена концептуальная технологическая схема внедрения процесса на действующем металлургическом производстве меди/МПГ.

ВВЕДЕНИЕ

Технология продувки методом погружения сверху (TSL) ISASMELT™ хорошо известна как одна из стандартных технологий плавки первичной меди.[1] С момента начала промышленного производства на первой установке на комбинате Маунт Айза в 1991 году построено более 20 установок ISASMELT. Процесс был быстро принят в плавке первичной меди и первичного и вторичного свинца, и к концу 2011 года общая установленная мощность технологии ISASMELT превысила 9 млн. тонн в год. История развития технологии подробно описана в литературе.[2-4] Технология TSL в равной степени эффективна при плавке сульфидных концентратов никеля.[5,6] Первая печь ISASMELT для плавки меди и никеля была построена в 1991 году по заказу Agip Australia.[5]

Процесс ISASMELT также хорошо подходит как для периодического, так и непрерывного конвертирования медного штейна в черновую медь[7,8] и низкосортного никелевого штейна в файнштейн[9] – процесс ISACONVERT™. В Европе периодическое конвертирование в печи ISASMELT применяется на двух заводах, а именно на заводе Umicore Precious Metals, Хобокен, Бельгия,[10] и на заводе Aurubis AG, Люнен, Германия,[11] с 1997 и 2002 годов, соответственно.

ПРОЦЕСС КОНВЕРТИРОВАНИЯ НИКЕЛЯ / МПГ ISACONVERT

Технология ISACONVERT имеет много общего с печью ISASMELT.[8] Она может быть без каких-либо сложностей реализована в закрытом исполнении с целью свести к минимуму выбросы в окружающую среду. Процесс использует технологию продувки TSL, обеспечивающую высокоэффективное смешивание и взаимодействие твердого штейна и флюса, которые могут подаваться через свод печи. Использование совершенных систем управления технологическим процессом позволяет в значительной степени автоматизировать работу печи. Поскольку печь является вертикальной, установка занимает очень малую площадь и может быть с легкостью включена в технологическую схему существующего металлургического завода, дополнив или заменив применяющуюся технологию. Значительно меньший объем отходящих газов процесса ISACONVERT, по сравнению с конвертером Пирса-Смита, позволяет сократить капитальные и эксплуатационные затраты на системы сбора и очистки отходящих газов.[8]

Последние прикладные исследования в совокупности с новаторскими испытаниями пилотной установки сделали возможным внедрение процесса ISACONVERT в промышленных масштабах для непрерывного конвертирования меди.[8] Установка ISACONVERT в разрезе показана на рисунке 1. Эта технология была доработана компанией Xstrata Technology (XT) для решения задач непрерывного конвертирования низкосортных никелевых/МПГ штейнов в высокосортные файнштейны. Результатом работ стал запатентованный процесс конвертирования никеля ISACONVERT. Как и в процессе конвертирования меди ISACONVERT, в процессе конвертирования никеля/МПГ используется шлаковая система на основе феррита кальция.

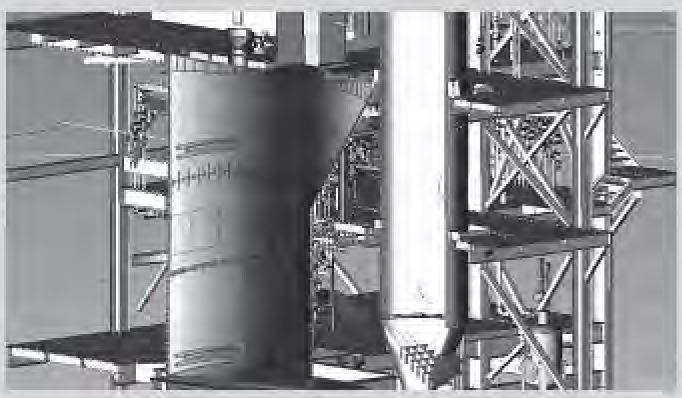


Рисунок 1. Конструкция печи ISACONVERT™ в разрезе.

ПРИНЦИП ДЕЙСТВИЯ ПРОЦЕССА КОНВЕРТИРОВАНИЯ НИКЕЛЯ / МПГ ISACONVERT

Питанием металлургического завода, перерабатывающего сульфид никеля, как правило, является никель-медный концентрат, который также может содержать небольшое количество кобальта и металлов платиновой группы.[12] Технологические схемы никелевой печи ISASMELT проработаны и представлены в имеющихся публикациях.[9]Как правило, питание металлургического завода по производству МПГ имеет более низкое содержание никель-медных сульфидов и более высокое содержание огнеупорных оксидных материалов по сравнению с питанием традиционных металлургических заводов по производству никеля.[12] Продуктом плавки никель-медного концентрата или питания цикла МПГ, как правило, является штейн с высоким содержанием железа, который подвергают дальнейшей обработке, почти исключительно в нескольких конвертерах Пирса-Смита, чтобы получить конечный штейн с низким содержанием железа, так называемый файнштейн. Исключение составляют металлургический завод Ватервал компании в ЮАР, на котором применяется процесс конвертирования Anglo Platinum (ACP),[13] и металлургический завод компании Stillwater Mining Company (SMC) в штате Монтана, США, на котором установлены вращающиеся конвертеры верхней продувки (TBRC).[14] В процессах ACP и SMC гранулированный штейн с высоким содержанием железа конвертируются в файнштейн, однако питание поступает непрерывно только в процесс ACP.

Непрерывное конвертирование никеля/МПГ не является новой технологией и исследовано с точки зрения повышения производительности и снижения уровня выбросов по сравнению с традиционными периодическими конвертерами Пирса-Смита. Как отмечено выше, основные принципы процесса уже реализованы в промышленном масштабе в установке ACP. Кроме того, компания Vale Inco предприняла масштабную программу исследований и разработок, занявшую 10 лет, с целью промышленного внедрения процесса непрерывного конвертирования никеля.[15]

Компания XT провела исследования конвертирования никеля/МПГ с использованием технологии ISACONVERT[5,9,16] и успешно получила на пилотной установке файнштейн, содержащий менее 4% масс. железа. Следует отметить, что процесс ISACONVERT для никеля/МПГ является процессом непрерывного конвертирования с непрерывной подачей в ванну штейна и воздуха/кислорода. Ванна состоит из штейна и шлака в соответствии с заданным составом продукта. Процесс эффективно работает в условиях, аналогичных тем, которые наблюдаются в конечной точке процесса конвертирования, в настоящее время применяющегося при периодическом конвертировании никеля/МПГ.

На рисунке 2 показано, как процесс ISACONVERT может быть интегрирован в технологическую схему действующего металлургического завода первичного продукта (EPSF). Гранулированный штейн EPSF, известняковый флюс, закупленное питание, вынос печи, топливо, воздух и кислород непрерывно подаются в никелевую/МПГ печь ISACONVERT. Продукт, жидкий файнштейн, периодически выгружают через летку и, в зависимости от требований последующего рафинировочного или аффинажного завода, либо гранулируют для гидрометаллургической переработки[17], либо медленно охлаждают для отделения никель-сульфидных, медных сульфидных и легированных фаз путем флотации перед рафинированием или аффинажем.[18]

Шлак выгружают через отдельную летку и возвращают в EPSF для извлечения ценных металлов. Отходящие газы печи ISACONVERT направляют в котел-утилизатор для утилизации тепла и обеспыливания с использованием электростатического осадителя, а затем на установку серной кислоты для извлечения серы. Вся пыль, собранная системами обработки газов, возвращается в печь ISACONVERT.

Если никелевый/МПГ конвертер ISACONVERT устанавливается вместо существующих конвертеров Пирса-Смита, может потребоваться переоборудование установки серной кислоты EPSF. Это позволит утилизировать поток высококонцентрированного диоксида серы (SO_2), производимый печью ISACONVERT. Технологическую схему ISACONVERT легко модифицировать таким образом, чтобы очистка шлака происходила отдельно от EPSF; в этот и другие варианты могут быть внесены изменения, позволяющие оптимизировать извлечение и свести к минимуму единичные операции и перемещение расплава.

Процесс конвертирования никеля/МПГ ISACONVERT, представленный на рисунке 2, обладает двумя основными преимуществами по сравнению с традиционным периодическим конвертированием в конвертере Пирса-Смита: Во-первых, процесс ISACONVERT отличается низким и постоянным объемным расходом отходящих газов с высоким содержанием SO_2, которые могут перерабатываться в обычной установке серной кислоты. Это важное преимущество, учитывая строгие экологические нормы, регулирующие современные и будущие промышленные выбросы и промышленную гигиену. Хотя одним из возможных вариантов сбора отходящих газов конвертера Пирса-Смита является установка камина, при этом также требуется второй камин для сбора выбросов, миновавших первый камин, что, как правило, влечет значительные дополнительные расходы. В процессе ISACONVERT конвертирование происходит в один этап и в одной печи, что позволяет использовать максимальное обогащение кислородом с минимальным разбавлением воздухом. Во-вторых, процесс ISACONVERT позволяет использовать в качестве питания твердый штейн, тем самым устраняя необходимость в подаче расплавленного штейна в ковше, а также уменьшая риск неконтролируемых выбросов, что улучшает состояние промышленной гигиены на заводе. Использование твердого питания также позволяет развязать стадии плавки и конвертирования, что повышает гибкость процесса и упрощает обслуживание и эксплуатацию завода.

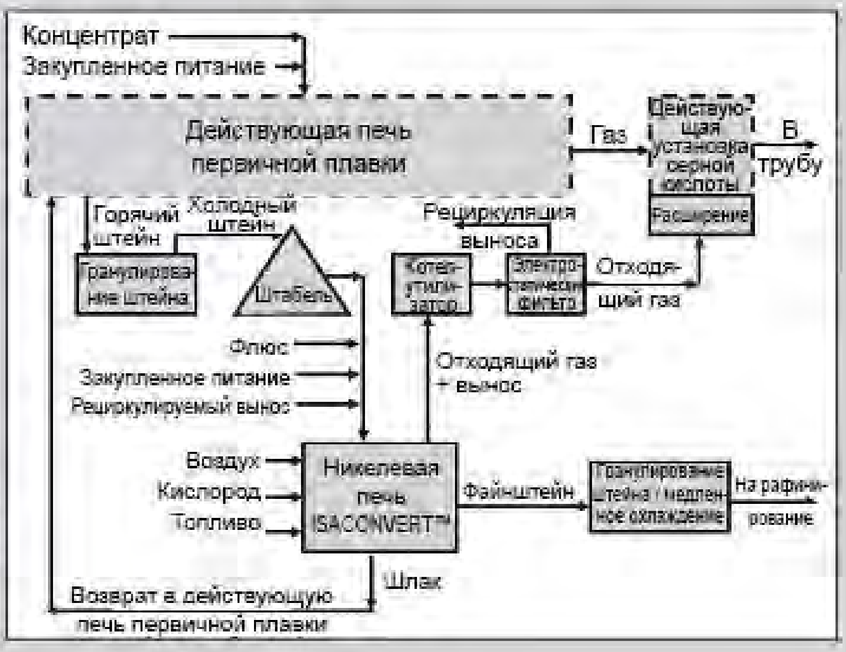


Рисунок 2. Технологическая схема процесса конвертирования никеля/МПГ ISACONVERT™.

ХИМИЧЕСКИЙ СОСТАВ ШЛАКА ПРИ НЕПРЕРЫВНОМ КОНВЕРТИРОВАНИИ

Как в периодическом конвертере Пирса-Смита, так и в непрерывном процессе конвертирования никеля/МПГ АСР используется шлаковая система на основе силиката железа (фаялит). В конвертерных печах Пирса-Смита из расплавленного штейна плавильной печи, как правило, получают конечный штейн, содержащий 2-3% масс. железа. Быстрое осаждение магнетита (преимущественно никель-феррита) в шлаке обуславливает содержание железа в конечном штейне конвертера Пирса-Смита не ниже примерно 2% масс. Некоторые операторы (Vale-Inco[19] и Lonmin [20]) позволяют оставшемуся шлаку отвердеть внутри конвертера Пирса-Смита, прежде чем продолжить цикл продувки, чтобы снизить содержание железа в штейне. Практика отверждения в цикле окончательной продувки приводит к образованию шлака, насыщенного диоксидом кремния и магнетитом и заключающего в себе файнштейн, который может быть извлечен только в новом цикле конвертирования,[20] что обуславливает неэффективность процесса. Если процесс продолжается при содержании железа в штейне ниже 2% масс. без затвердевания шлака, происходит либо чрезмерный унос магнетита/шлака в конечный штейн, либо чрезмерное наростообразование внутри печи.[15],[20]

Первоначальная технологическая схема ACP предполагала двухстадийный периодический процесс получения файнштейна: на первой стадии содержание железа в штейне снижалось до ~13% масс., а на второй стадии до ~3% масс.[13] Из-за трудностей, связанных с определением момента начала второй стадии конвертирования, происходили выбросы шлака из-за чрезмерного окисления ванны.[13] Периодический характер процесса приводил к неудовлетворительному или неполному перемешиванию и неравновесной стратификации расплавов внутри печи. Последующее быстрое перемешивание слоев расплава из-за возмущений в ванне приводило к взрывному вспениванию содержания ванны при низком содержании железа в штейне.[13] По этим причинам и для обеспечения постоянного потока SO_2 высокой концентрации, поступающего на установку по переработке отходящих газов, процесс ACP был переведен в непрерывный режим с непрерывной подачей в печь гранулированного штейна и конвертированием с конечным содержанием железа в штейне 3% масс.[13]

В программе исследований, предпринятой Vale Inco,[15] рассматривалось три подхода к непрерывному конвертированию: собственная технология взвешенной плавки,[21] технология конвертирования в ванне с верхней продувкой кислородом и донным перемешиванием азотом и технология конвертирования в ванне типа Noranda/El Teniente.[22] Необходимо отметить, что основная цель программы заключалась в разработке технологии непрерывного конвертирования, применимой на никелевом заводе Коппер Клифф, где требования рафинировочного завода обуславливают содержание железа в штейне не выше 0,5% масс. Хотя это технически осуществимо, испытания показали, что окисленный железо-силикатный шлак, получаемый при столь низком содержании железа, нестабилен и в процессе эксплуатации происходит чрезмерное наростообразование в реакторах.[22] Поэтому предпочтение было отдано "двухстадийному" подходу с непрерывным конвертированием до содержания железа в штейне на уровне 2-3% с последующей периодической доводкой для получения конечного штейна желаемого сорта.

Предварительные пилотные испытания процесса ISACONVERT были проведены с использованием шлаков на основе силиката железа при конвертировании двух сортов первичного штейна, как показано в таблице I. Был получен конечный штейн, содержащий 0.7-13.5% масс. железа (первичный штейн 1)[16] и 2.2-10 % масс. железа (первичный штейн 2).[9] Результаты предварительных испытаний конвертирования железо-силикатного шлака показали, что для получения файнштейнов, содержащих менее 2% масс. железа, необходимо значительно повысить температуру процесса, чтобы обеспечить текучесть железо-силикатного шлака. Поэтому были рассмотрены возможности применения альтернативной шлаковой системы.

Таблица I. Первичный никелевый/МПГ штейн, использованный для конвертирования при испытаниях ISACONVERT™

Элемент	Среднее содержание в штейне 1 (% масс.)	Среднее содержание в штейне 2 (% масс.)
Ni	16.2	44.5
Cu	10.6	9.7
Co	0.53	2.9
Fe	40.1	25.3
S	26.9	17.2
SiO_2	—	0.3

Конвертирование никелевого штейна с использованием шлака на основе феррита кальция успешно применяется в промышленном масштабе компанией SMC с 1991 года в конвертерах TBRC, производящих файнштейн, содержащий около 2% масс. железа.[12] Процесс TBRC представляет собой периодический процесс, аналогичный конвертерам Пирса-Смита. Первоначально для конвертирования в процессе TBRC компания SMC использовала шлак на основе силиката железа, но при низком содержании железа в штейне в результате чрезмерного окисления происходило внезапное вспенивание шлака, приводившее к потере шихты и создававшее потенциальную угрозу для безопасности оборудования и операторов.[14] Поэтому SMC внесла изменения в химический состав шлака, заменив кремнеземистый флюс известняковым, чтобы предотвратить нестабильное состояние ванны. Учитывая успешное применение шлаков на основе феррита кальция в периодическом конвертировании никелевого штейна на заводе SMC и доказанную способность процесса ISACONVERT и других процессов использовать шлаки на основе феррита кальция для производства меди,[7,8,23] компания XR изучила возможность применения данной шлаковой системы в непрерывном процессе TSL при производстве никелевого/МПГ штейна.

Шлаковая система на основе феррита кальция успешно применяется в технологиях непрерывного конвертирования меди с середины 1970-х годов.[23] Полезные свойства шлаков на основе феррита кальция при конвертировании меди были описаны в 1980-х годах в работах Язава[24,25] и Такэда.[26] Эти свойства включают в себя способность жидкой фазы содержат более высокие концентрации трехвалентного железа при высоких кислородных потенциалах, меньший объем шлака, меньшие потери ценных металлов и большую текучесть.

Таблица II. Сравнение шлаков на основе феррита кальция и силиката железа

Условия при Fe<4% В масс. в штейне	Шлак на основе феррита кальция	Шлак на основе силиката железа
Доля ценных металлов, попадающих в штейн (Ni, Cu и Co)	Выше	Ниже
Доля примесей, попадающих в шлак (As и Sb)	Выше	Ниже
Минимальная температура ликвидуса шлака (FactSage)	~1250°C	>1400°C
Текучесть	Выше	Ниже

Фонт[27] и Энао[28] исследовали в лабораторных условиях применение шлаков на основе феррита кальция при конвертировании никелевого штейна. Эти исследователи приводили к равновесию расплав штейна и шлака при 1500°C или 1600°C в течение определенных периодов времени при заданном парциальном давлении кислорода (P_{O_2}) и диоксида серы (P_{SO_2}) (задается соотношениями S_2/SO_2 и CO/CO_2). В конце каждого равновесного теста для анализа отбирались охлажденные образцы штейна и шлака. Результаты показали, что, по сравнению со шлаками на основе силиката железа, шлаки на основе феррита кальция обеспечивали более высокое извлечение ценных металлов (Ni, Cu и Co) при большей доле примесных элементов (As и Sb) в шлаковой фазе.[27,28]

XT использовала пакет термодинамического моделирования FactSage[29], чтобы убедиться, что полезные свойства системы на основе феррита кальция с точки зрения текучести и концентрации трехвалентного железа также применимы к процессу конвертирования никелевого/МПГ штейна. Эти результаты моделирования, а также данные фундаментальных исследований[27,28] приведены в таблице 11.

НЕПРЕРЫВНОЕ КОНВЕРТИРОВАНИЕ НА ПИЛОТНОЙ УСТАНОВКЕ С ИСПОЛЬЗОВАНИЕМ ШЛАКОВ НА ОСНОВЕ ФЕРРИТА КАЛЬЦИЯ

Цель пилотных испытаний конвертирования заключалась в том, чтобы определить химию технологического процесса получения никеля/МПГ ISACONVERT при конвертировании штейна первичной плавки с высоким содержанием железа в файнштейн с низким содержанием железа с использованием шлака на основе феррита кальция. Результаты эксплуатации пилотной установки ISACONVERT описаны в литературе.[9] Типичный состав штейна первичной плавки, использованного во всех испытаниях ISACONVERT с шлаками на основе феррита кальция, близок к составу штейна 1 в таблице I. Штейн первичной плавки был успешно конвертирован в штейн с содержанием железа в пределах от 2,6 до 8,0% масс.

В ходе испытаний в пилотную печь подавали твердый штейн и известняковый флюс с расходом 100-150 кг/ч твердой штейна без дополнительной обработки. Воздух и кислород подавались через отдельные ротаметры при соотношении с выходом 25-35% об. от общего обогащения кислородом. Природный газ подавался через копье для поддержания температуры ванны в пределах 1300-1380°C.

Пилотные испытания показали, что жидкие шлаки были получены при всех условиях проведения испытаний. Состав шлаков ISACONVERT, полученных в ходе испытаний, показан в таблице III.

Коэффициенты распределения никеля, меди и кобальта, определенные уравнением 1, показаны в зависимости от сорта штейна (сумма массового процента никеля, меди и кобальта) на рисунках 3-5, соответственно. На графиках сравниваются коэффициенты распределения, полученные при испытаниях процесса ISACONVERT с использованием шлака на основе феррита кальция, с показателями конвертеров Пирса-Смита и процесса TSL с использованием шлака на основе силиката железа. Были использованы следующие источники данных по конвертированию в конвертере Пирса-Смита и процессе TSL с использованием железо-силикатного шлака: Результаты проведенных XT пилотных испытаний конвертирования штейна в процессе TSL с использованием шлака на основе силиката железа;[9] и результаты отбора проб из конвертеров Пирса-Смита на металлургическом заводе Фэлконбридж компании Xstrata Nickel (XNi)[30] и на завод

Томпсон компании Vale Inco.[31]

$L_x^{s/m}$=(% масс. X в шлаке)/(% масс. X в штейне) (1)

где L - коэффициент распределения, s - шлак, X - элемент, распределение которого необходимо установить, m - штейн.

Сравнение данных как пилотной установки, так и металлургического завода показывает, что доля никеля (рисунок 3) и кобальта (рисунок 5), приходящаяся на шлак, ниже в процессе конвертирования никеля/МПГ ISACONVERT с использованием шлака на основе феррита кальция, чем в процессах с использованием шлака на основе силиката железа. Рисунок 5 показывает, что процесс конвертирования никеля/МПГ ISACONVERT с использованием шлака на основе феррита кальция значительно превосходит другие процессы с точки зрения распределения кобальта: доля кобальта в шлаке более чем на 40% ниже при содержании в штейне выше 76% масс. Доля меди (рисунок 4) в шлаке процесса ISACONVERT с использованием шлака на основе феррита кальция ниже, но только при высоких сортах штейна, соответствующих файнштейну. Для расчета коэффициента распределения меди на заводе XNi доступная информация является неполной.[30] Желаемый сорт файнштейна никеля/МПГ, как правило, соответствует требованиям последующего рафинировочного / аффинажного производства, устанавливающим допустимый уровень железа и серы в штейне. Сравнение процесса конвертирования никеля/МПГ ISACONVERT с использованием шлака на основе феррита кальция и процессов с использованием шлака на основе силиката железа с точки зрения сорта штейна по содержанию железа[9,30,31] приведено на рисунке 6. При фиксированном содержании железа в штейне процесс конвертирования никеля/МПГ ISACONVERT с использованием шлака на основе феррита кальция позволяет получать штейн с массовой долей никеля, меди и кобальта, более чем на 5% превышающей аналогичный показатель процессов конвертирования с использованием шлака на основе силиката железа.

Таблица III. Состав шлака ISACONVERT™

Элемент	Диапазон (% масс.)
Ni	4.1-12.8
Cu	0.9-2.3
Co	1.6-2.9
Fe	46.2-51.0
CaO	11.9-15.6
SiO_2	1.7-2.2

Более высокий сорт файнштейна, получаемого при использовании шлака на основе феррита кальция, является результатом более низкого содержания серы в фазе штейна по сравнению с процессами, использующими шлак на основе силиката железа (см. рисунок 7). Исходный штейн, поступающий в процесс конвертирования никеля/МПГ, как правило, имеет низкое содержание серы (см. таблицу I), и конвертирование с использованием шлака на основе феррита кальция делает возможной дальнейшую металлизацию штейна до содержания серы менее 15% масс. Штейн, получаемый в процессах конвертирования никелевого штейна с использованием шлака на основе силиката железа, содержит больше серы, как минимум 20% масс., что обуславливает более низкий сорт штейна по сравнению с процессом конвертирования никеля/МПГ ISACONVERT. Минералогический анализ штейна конвертера Пирса-Смита показал, что металлизированные фазы файнштейна отличаются доминированием никеля и выступают в качестве собирателей МПГ.[32] Увеличение металлизации процесса ISACONVERT (см. рисунок 7), следовательно, должно также привести к увеличению концентрации и переноса МПГ в конечный файнштейн по сравнению с конвертером Пирса-Смита или процессом TSL с использованием шлака на основе силиката железа.

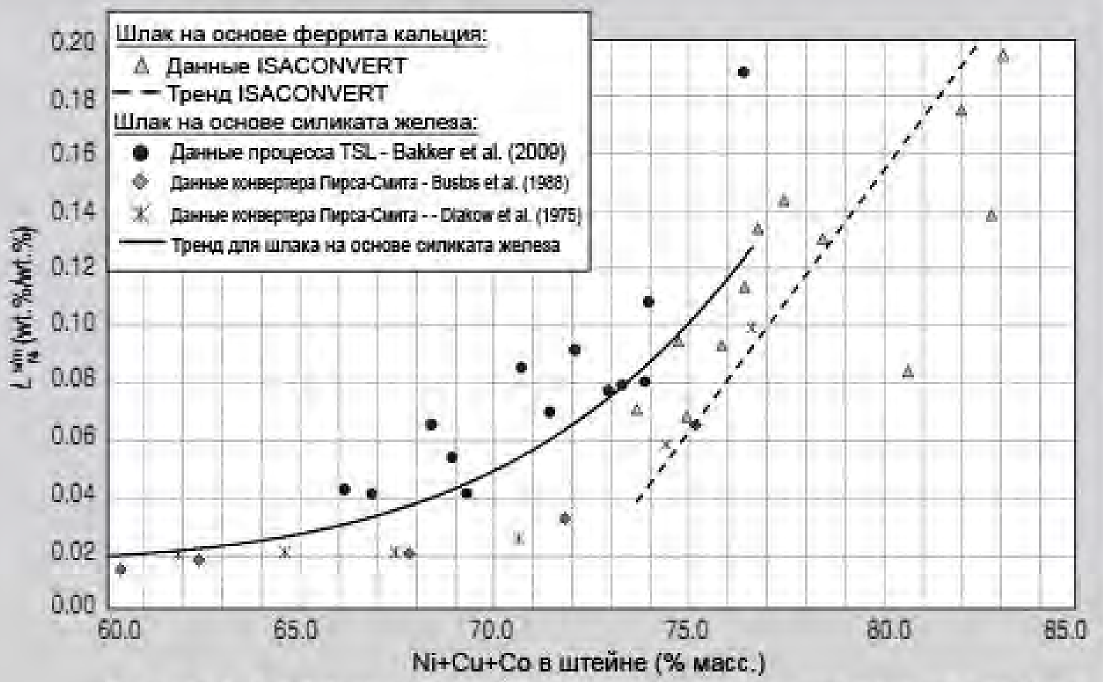


Рисунок 3. Коэффициент распределения никеля между шлаком и штейном в зависимости от сорта файнштейна (% масс. Ni+Cu+Co в штейне).

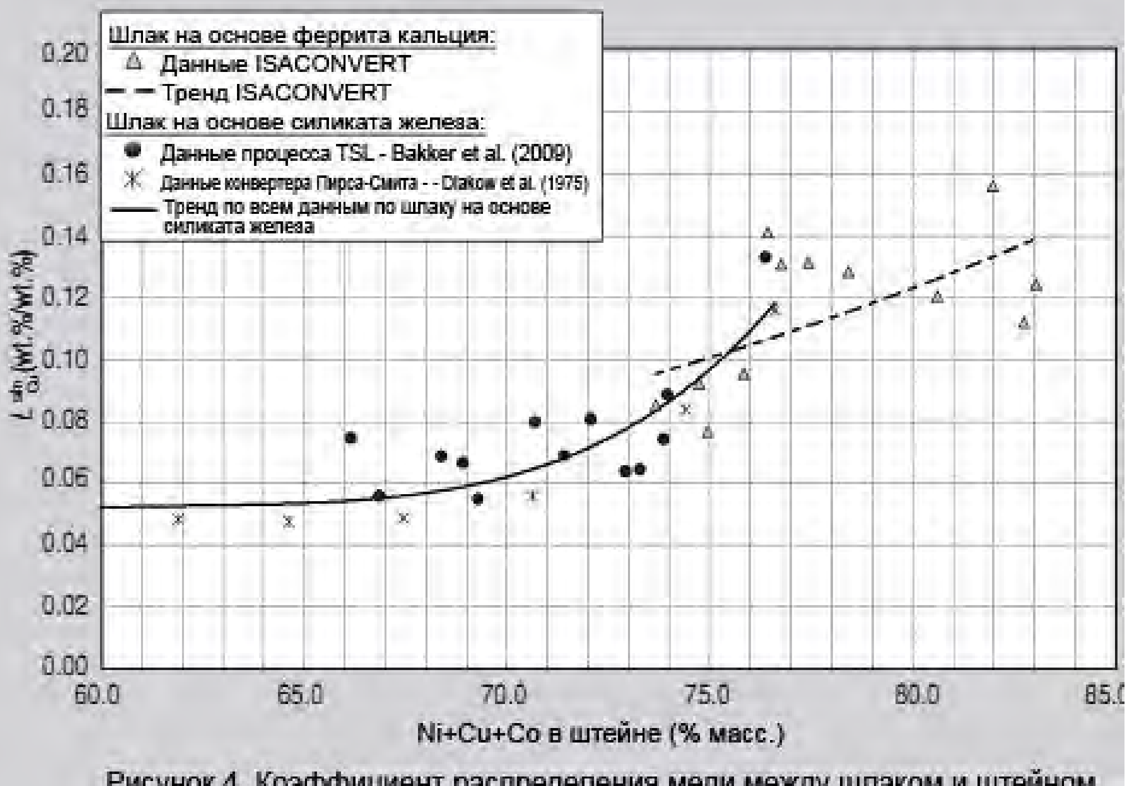


Рисунок 4. Коэффициент распределения меди между шлаком и штейном в зависимости от сорта файнштейна (% масс. Ni+Cu+Co в штейне).

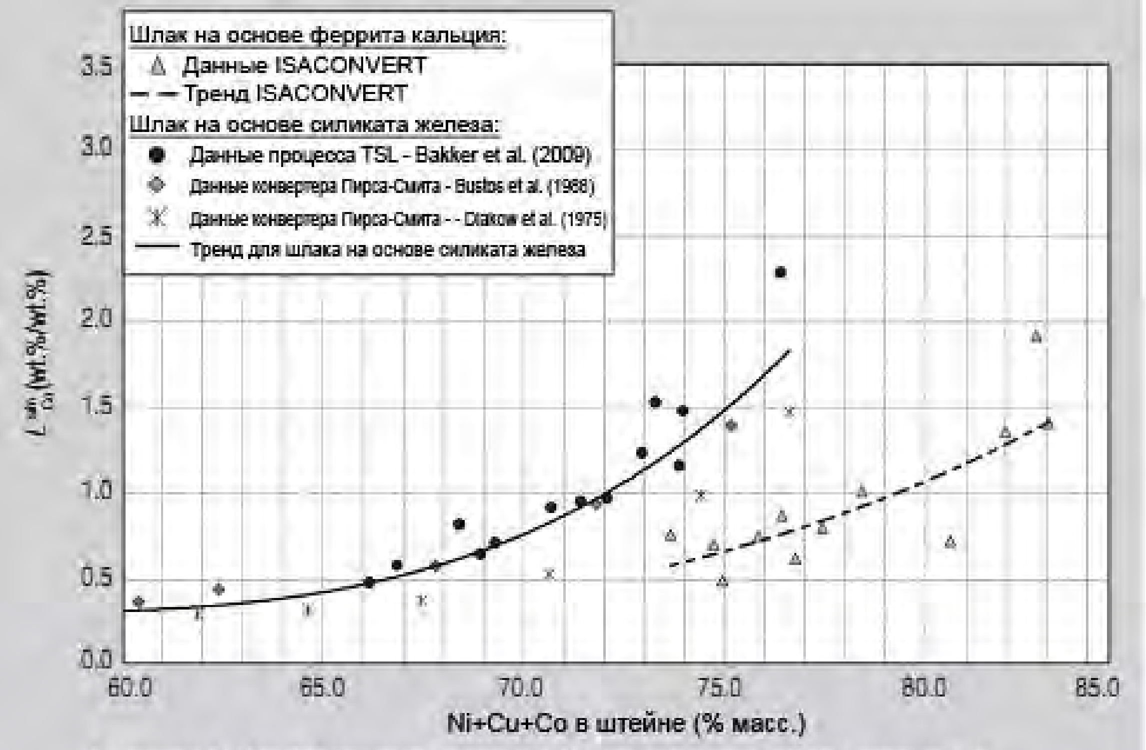


Рисунок 5. Коэффициент распределения кобальта между шлаком и штейном в зависимости от сорта файнштейна (% масс. Ni+Cu+Co в штейне).

ВЫВОДЫ

Технология ISASMELT TSL широко применяется в плавке первичных и вторичных меди и свинца. Периодическая плавка и конвертирование с помощью технологии ISASMELT также хорошо известны. Технология является в равной степени эффективной при непрерывном конвертировании и поэтому получила название ISACONVERT™.

Особенности, которые обуславливают привлекательность технологии ISACONVERT при конвертировании меди и в равной степени применимы к конвертированию никеля/МПГ, следующие: (1) Постоянный объемный расход отходящих газов с высоким содержанием SO_2, которые могут перерабатываться в обычной установке серной кислоты. (2) Конвертирование происходит в один этап и в одной печи, что позволяет использовать максимальное обогащение кислородом с минимальным разбавлением воздухом. (3) В качестве питания может быть использован твердый штейн, что устраняет необходимость подачи расплавленного штейн в ковше, снижает неконтролируемые выбросы и позволяет развязать плавку и конвертирование, увеличивая гибкость и упрощая техническое обслуживание и эксплуатацию металлургического завода.

Процесс ISACONVERT успешно опробован в полупромышленном масштабе в качестве технологии конвертирования медного/никелевого штейна. Результаты показали, что, по сравнению с процессами с использованием шлака на основе силиката железа, процесс конвертирования никеля/МПГ ISACONVERT с использованием шлака на основе феррита кальция: (1) обеспечивает более высокий сорт файнштейна (сумма массового процента никеля, меди и кобальта в штейне) благодаря более низкому конечному содержанию серы в штейне; (2) обеспечивает более высокое извлечение ценных металлов в файнштейн. Последнее уменьшает долю никеля и меди в шлаке, в свою очередь, более чем на 40% уменьшая долю кобальта в шлаке.

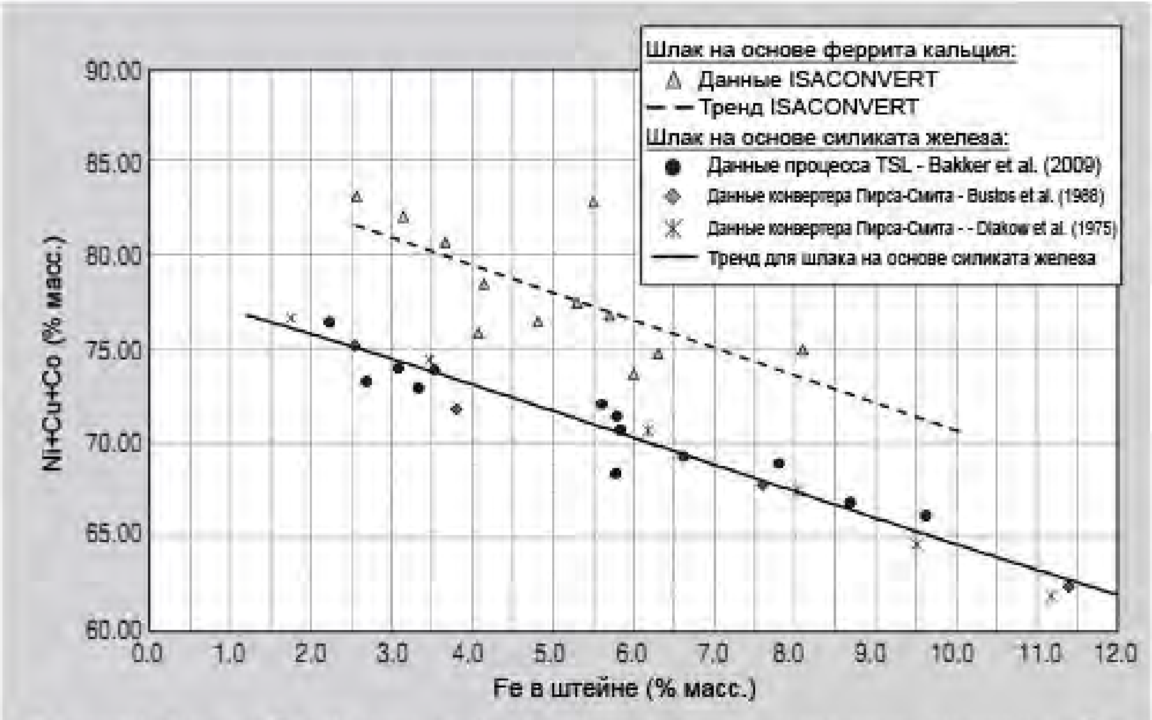


Рисунок 6. Сорт штейна (% масс. Ni+Cu+Co) в зависимости от содержания железа в штейне.

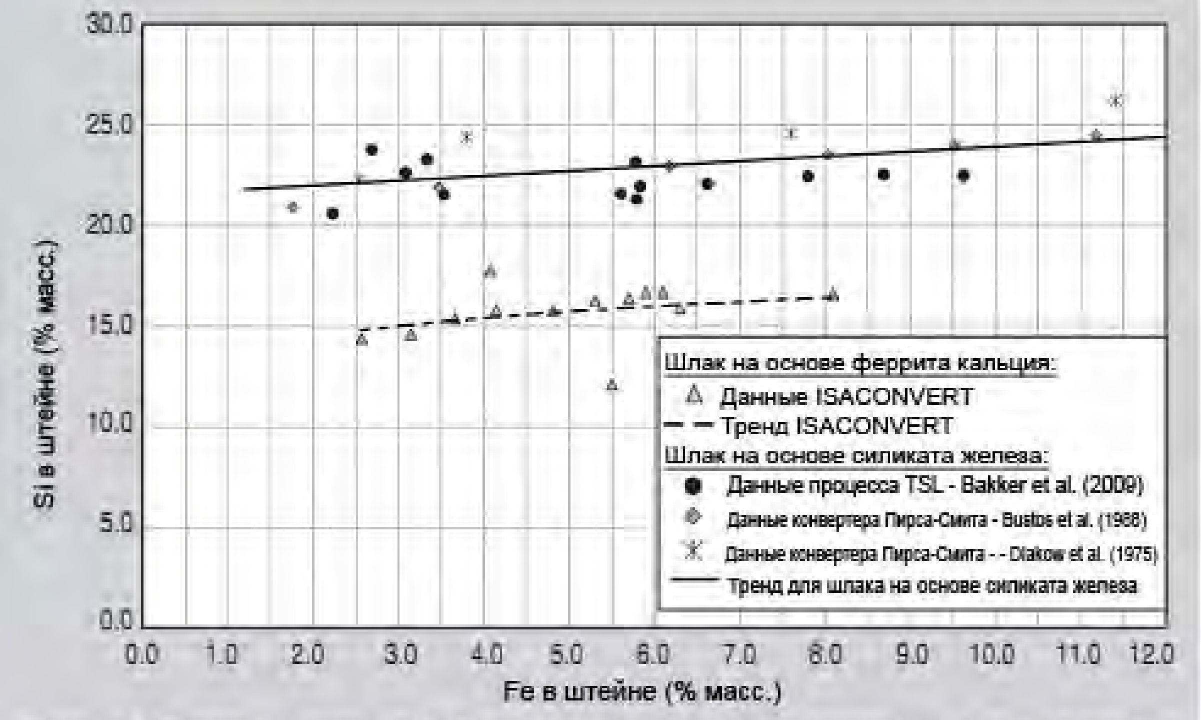


Рисунок 7. Содержание серы в файнштейне в зависимости от содержания железа в штейне.

БЛАГОДАРНОСТИ

Авторы выражают признательность компании Xstrata Technology за разрешение опубликовать настоящую статью.

Список использованной литературы

1. C. Diaz, Proc. Copper 2010, vol. 7 (Hamburg, Germany: GDMB, 2010), pp. 2543-2562.

2. G. Alvear, P. Arthur, and P. Partington, Proc. Copper 2010, vol. 7 (Hamburg, Germany: GDMB, 2010), pp. 615-630.

3. B. Burford, AusIMM Bulletin, No. 1 (February 2009), pp. 26-30.

4. B. Errington, P. Arthur, J. Wang, and Y. Dong, Proc. Inf I. Symp. Lead and Zinc Processing, ed. T. Fujisawa et al. (Tokyo: MMIJ, 2005), pp. 581-599.

5. P. Bartsch, B. Anselmi, and C.R. Fountain, “The Radio Hill Project,” ed. W.A. Pyrosem, E.J. Grimsey, and N.D. Stockton (Perth, Australia: Murdoch University Press, 1990).

6. M. Zhou, A. Wan, G Li, R. Baldock, and H. Li, Supplemental Proceedings Volume 1: TMS2010-Materi- als Processing and Properties (Warrendale, PA: TMS, 2010), pp. 525-534.

7. J.S. Edwards and S. Jahanshahi, “Copper Converting,” U.S. patent 5,888,270 (30 March 1999).

8. S. Nikolic, J.S. Edwards, A.S. Burrows, and G.R.F. Alvear, International Peirce-Smith Converting Centennial, ed. J. Kapusta and T. Warner (Warrendale, PA: TMS, 2009), pp. 407-414.

9. M.L. Bakker, S. Nikolic, and P.J. Mackey, “ISA- SMELT™ TSL Applications for Nickel” (Presented at Nickel Processing 2010, Minerals Engineering International, Falmouth, U.K., 2010).

10. F. Vanbellen and M. Chintinne, Advanced Processing of Metals and Materials, ed. F. Kongoli and R.G. Reddy (Warrendale, PA: TMS, 2006), pp. 43-52.

11. S. Schmidt, Aurubis AG-Huttenwerke Kayser, unpublished (March 2006).

12. A.E.M. Warner, C.M. Diaz, A.D. Dalvi, P.J. Mackey, A.V. Tarasov, and R.T. Jones, JOM, 59 (4) (2007), pp. 58-72.

13. P. Viviers and K. Hines, First Extractive Metallurgy Operators’ Conference (Brisbane, Australia: AusIMM, 2005), pp. 101-108.

14. G.K. Roset, J.W. Matousek, and P.J. Marcantonio, JOM, 44 (4) (1992), pp. 39-42.

15. A.E.M. Warner and C.M. Diaz, An Overview of the Metallurgy of Nickel-Copper

Matte Converting, Yazawa International Symposium Metallurgical and Materials Processing: Principles and Technologies, ed. F. Kongoli, K. Itagaki, C.Yamauchi, and H.Y. Sohn (Warrendale, PA: TMS, 2003), pp. 113-129.

16. PGM Matte Converting, XT Internal Report (1996).

17. E.O. Stensholt, O.M. Dotterud, E.E. Henriksen, P.O. Ramsdal, F. Stalsen, and E. Thure, CIM Bulletin, 94 (1051) (2001), pp. 101-104.

18. A.E.M. Warner, J. Liu, F. Javor, R. Lawson, W. Shellshear, T. Hoang, and R. Falcioni, Converter and Fire Refining Practice, ed. A. Ross, T. Warner, and K. Scholey (Warrendale, PA: TMS, 2005), pp. 27-43.

19. S. Marcuson, International Peirce-Smith Converting Centennial, ed. J. Kapusta and T. Warner (Warrendale, PA: TMS, 2009), pp. 103-126.

20. G.A. Bezuidenhout, J.J. Eksteen, W. Wendt, and W. Persson, “Implementation of Semtech Optical Spectrometry System at Lonmin for Converter Fe End-point Control” (Presented at Nickel Processing 2010, Minerals Engineering International, Falmouth, U.K., 2010).

21. G.S. Victorovich, Proceedings of the Paul E. Que- neau International Symposium, Volume 1, ed. R.G. Reddy and R.N. Weizenbach (Warrendale, PA: TMS, 1993), pp. 501-529.

22. J.R. Donald, A.E.M. Warner, A.D. Dalvi, K. Scholey, A. Ross, and C. Harris, Converter and Fire Refining Practices, ed. A. Ross, T. Warner, and K. Scholey (Warrendale, PA: TMS, 2005), pp. 295-307.

23. W.G.L. Davenport et al., Extractive Metallurgy of Copper, 4th ed. (Oxford, U.K.: Pergamon, 2002).

24. A. Yazawa, Y. Takeda, and Y. Waseda, Can. Metall. Q., 20 (1981), pp. 129-134.

25. A. Yazawa and Y. Takeda, Trans. Jpn. Inst. Met., 23 (1982), pp. 328-333.

26. Y. Takeda, S. Nakazawa, and A. Yazawa, Can. Metall. Q., 19 (1980), pp. 297-305.

27. J.M. Font, “Phase Equilibrium and Minor Elements Distribution between Slag and Matte Phases in Nickel Smelting” (Ph.D. thesis, Tohoku University, 1999), pp. 63,66.

28. H.M.Z. Henao, “Phase Equilibrium between Ni-S or Ni-Fe Melt and Slag in Nickel Smelting” (Ph.D. thesis, Tohoku University, 2003), pp. 54,60.

29. C.W. Bale, A.D. Pelton, and W.T. Thompson, “Facility for the Analysis of Chemical Thermodynamics” (FactSage version 6.1) (Ecole Polytechnique de Montreal, 2002).

30. A.A. Bustos, S.W. Ip, G. O’Connell, G.H. Kaiura, and J.M. Toguri, Extractive Metallurgy of Nickel and Cobalt, ed. G.P.Tyroler and C.A. Landolt (Warrendale, PA: TMS, 1988), pp. 335-354.

31. J.S. Diakow, Y.F. Mak, and R.G. Orr, “Metallurgy of the Converting Process in the Thompson Smelter" (Paper presented to the 14th Annual Conference of Metallurgists, Edmonton, Alberta, The CIM, Montreal, Quebec, Canada, August 1975).

32. E. Thyse, G. Akdogan, and J.J. Eksteen, “The Effect of Changes in Iron-endpoint during Peirce-Smith Converting on PGE-containing Nickel Converter Matte Mineralization” (Presented at Nickel Processing 2010, Minerals Engineering International, Falmouth, U.K., 2010).

Главные металлурги М.Л. Беккер и Дж.Р.Ф. Альвеар и инженер-металлург С. Николич являются работниками компании Xstrata Technology, (Level 4, 307 Queen Street, Brisbane, 4000, Australia). С доктором Беккером можно связаться по электронной почте mbakker@xstratatech.com.

Minerals Engineering 24 (2011) 610–619

Contents lists available at ScienceDirect

Minerals Engineering

journal homepage: www.elsevier.com/locate/mineng

ISASMELT™ TSL – Applications for nickel

M.L. Bakker [a,*], S. Nikolic [a], P.J. Mackey [b]

[a] Xstrata Technology, Level 4, 307 Queen Street, Brisbane 4000, Australia
[b] PJ Mackey Technology Inc., Kirkland, Quebec, Canada

ARTICLE INFO

Article history:
Available online 25 October 2010

Keywords:
Extractive metallurgy
Pyrometallurgy
Sulfide ores
Oxide ores

ABSTRACT

The ISASMELT™ process is a top submerged lance (TSL) bath smelting technology which has been developed and optimised over the last 25 years. By the end of 2011, the total installed capacity of the ISASMELT™ technology will exceed 9,000,000 tonnes per year of feed materials in copper and lead smelters around the world. The technology is equally effective for smelting nickel sulfide concentrates, converting nickel mattes, and producing ferronickel from lateritic ores. This paper demonstrates how the features that make ISASMELT™ attractive for copper and lead smelting may be applied equally to nickel smelting and converting operations. Conceptual flowsheets are presented, supported by results from recent pilot plant and bench-scale testwork.

1. Introduction

The introduction of lance technology has enabled the use of stationary furnaces of simple design and very high reaction rates. The prior art involved the injection of gases into liquid slags or mattes predominantly through tuyeres, with inherent design complications and refractory problems. The top submerged lance (TSL) bath smelting technology was developed to commercial success at the Mount Isa smelting complex in Australia in the early 1990s, and subsequently called "ISASMELT™". Development of the process has focussed on smelting of lead and copper concentrates or secondaries over the last 30 years and commercial ISASMELT™ furnaces operated by Xstrata and external licensees currently have a combined annual smelting capacity exceeding 9,000,000 tonnes of feed.

2. The ISASMELT™ process

ISASMELT™ technology is based on a furnace design which is readily enclosed to eliminate emissions to the surrounding environment. It uses submerged lance injection technology to provide highly efficient mixing and reaction of feed materials in a molten slag bath. The use of advanced process control systems results in the furnace operation being largely automated. Being a vertical furnace, a very small footprint of floor space is required to accommodate the plant and so it can generally be easily retro-fitted into existing smelters to either augment or replace existing technology. The process concept is shown in Fig. 1.

During 30 years developing and operating submerged lance technology on large scale plants, significant technical improvements have occurred in areas such as furnace design, feed preparation systems, off-gas handling, operating and process control strategies, refractory management and operator training. The combined experience led to what is well recognised by the marketplace today as the "ISASMELT™ technology package", a technology package licensed to external clients – Burford (2009). Many of the improvements implemented by plant operators have been passed onto, and adopted by, other licensees. Exchange of ideas and technical improvements occurs through ad hoc visits to fellow licensee sites and through regular licensee workshops arranged by Xstrata Technology. Fig. 2 shows the location of the commercial plants that have been licensed to date.

3. Agip nickel ISASMELT™ plant

It is less well known that the ISASMELT™ process was also adapted for the treatment of nickel bearing feeds at a very early stage of its development. A large amount of pilot-scale testwork was completed during the 1980s on nickel deposits owned by Mount Isa Mines Ltd. Testwork was also performed for AGIP Australia Pty Ltd., who owned the Radio Hill deposit in Western Australia – Bakker et al. (2009). In 1991, AGIP decided to construct a semi-commercial ISASMELT™ plant to produce nickel–copper matte from concentrate feed. This was the first ISASMELT™ plant built and commissioned for an external client. Earlier in 1991, the first commercial-scale ISASMELT™ plant had been commissioned to produce lead bullion from concentrates in Mount Isa. A year later, two commercial-scale ISASMELT™ plants were commissioned to produce copper matte from concentrates; one in Mount Isa and the other in Miami, Arizona.

---

* Corresponding author.
E-mail address: mbakker@xstratatech.com.au (M.L. Bakker).

0892-6875/$ - see front matter © 2010 Published by Elsevier Ltd.
doi:10.1016/j.mineng.2010.09.016

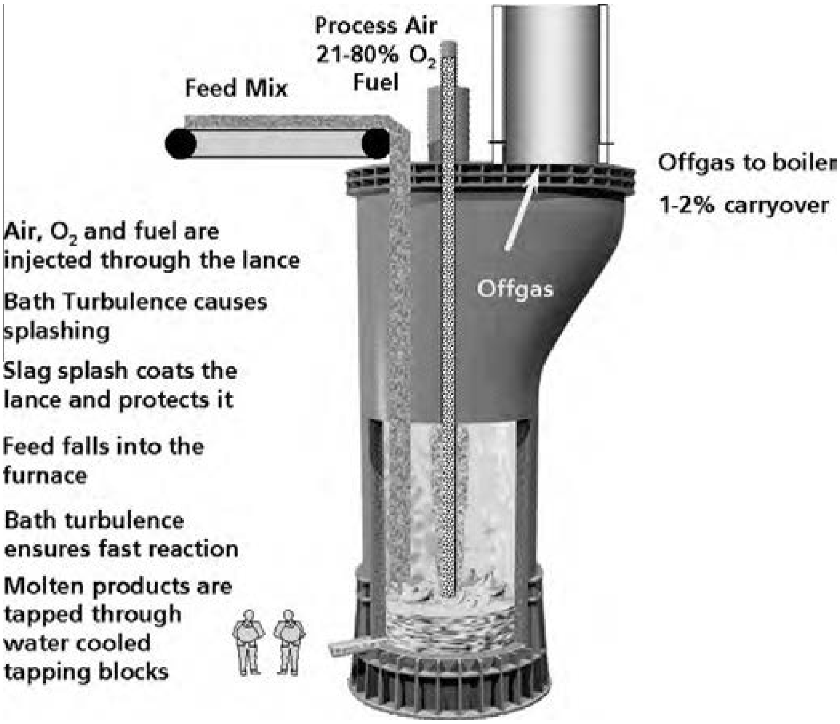


Fig. 1. The ISASMELT™ concept.

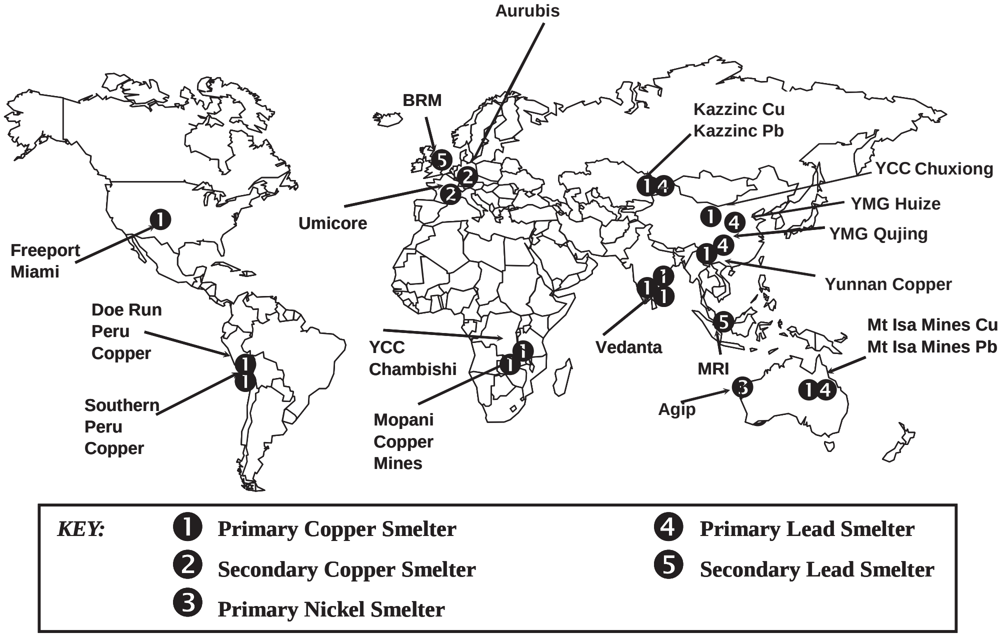


Fig. 2. ISASMELT™ plant locations.

The AGIP nickel ISASMELT™ commercial-scale plant was commissioned in September 1991 and within 3 months was running at design capacity of 7.5 t/h concentrate – Arthur and Hunt (2005). It produced 45 wt% nickel/copper matte from a concentrate containing approximately 7 wt% nickel and 3.5 wt% copper. Photographs of the plant are shown in Fig. 3 – Bakker et al. (2009). At the time, the plant was deemed a technical and operational success, however due to a large drop in nickel price, the facility was forced to close near the end of 1991; the plant has since been demolished.

4. The ISACONVERT™ process

A combination of pilot plant copper and nickel–copper continuous converting trials and the commercial batch copper converting operations show the ISASMELT™ furnace to be well suited to the duty of continuous converting – the ISACONVERT™ furnace – Edwards and Alvear (2007).

The ISACONVERT™ technology is based on the same design as the ISASMELT™ furnace which allows it to be readily enclosed to eliminate emissions to the surrounding environment. It uses the

Fig. 3. Photographs of AGIP nickel ISASMELT™ plant.

submerged lance injection technology to provide highly efficient mixing and reaction of solid matte and flux, which is dropped in through the roof of the furnace. As with the ISASMELT™ furnace the ISACONVERT™ has a small footprint of floor space and can generally be easily retro-fitted into existing smelters.

The union between recent applied research and pioneering pilot plant work has made possible the industrial-scale implementation of the ISACONVERT™ process – Nikolic et al. (2009). A cutaway image of an ISACONVERT™ plant is shown in Fig. 4.

Application of both the ISASMELT™ and ISACONVERT™ furnace technology to copper production allows for a very compact site that requires significantly less area than a conventional copper smelter. The fact that the smelting and converting processes are very similar and easy to control via a proven robust process control system also simplifies operations and logistics within the smelter, allowing all operations to be carried out from one central control room. These factors lower the operating costs of a new smelter compared with the conventional Peirce–Smith technology. The significantly reduced off-gas volume from the ISACONVERT™ process will result in lower capital and operating costs for off-gas collection and cleaning systems.

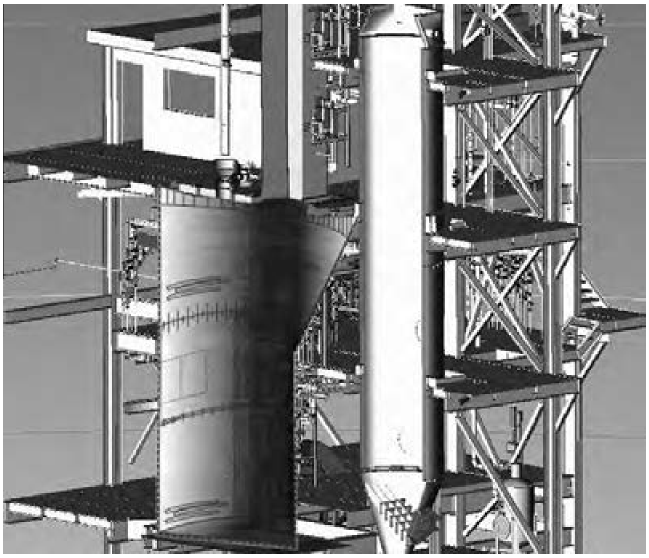


Fig. 4. Cutaway of a design for the ISACONVERT™ furnace.

5. Nickel ISASMELT™ sulfide smelting process concept

The feed to a nickel sulfide smelter typically consists of a nickel–copper concentrate, which may also contain minor amounts of cobalt and platinum group metals. The smelting product from such feeds is generally a primary, high iron smelting matte which is further processed, typically in Peirce–Smith converters, to produce finished, high grade matte, often referred to as ‘‘Bessemer matte”.

The 2007 TMS survey of worldwide nickel sulfide smelters reported that flash smelting accounted for approximately 70% of nickel smelter output (based on 2004–2005 reporting data), with electric furnace smelting representing the balance – Warner et al. (2007). However, the same features that make the ISASMELT™ process an attractive alternative for copper and lead smelting can be equally applied to nickel smelting – Bakker et al. (2009).

The basic process block diagram for a primary smelting nickel ISASMELT™ plant is shown in Fig. 4. The nickel ISASMELT™ furnace continuously processes concentrate feed, fluxes, and recycled dust. The product liquid matte and slag is tapped periodically from the ISASMELT™ vessel to a separate slag cleaning furnace via a single taphole. Off-gases from the ISASMELT™ furnace are directed to a waste heat boiler for heat recovery and de-dusted using an electro-static precipitator before being sent to a sulfuric acid plant for sulfur capture. The primary smelting matte is transferred by ladles from the settling furnace to the Peirce–Smith converters for production of Bessemer matte; alternatively continuous converting could be used, refer below. Discard slag is also intermittently tapped from the settling furnace for disposal. All dusts collected from the gas handling systems are recycled to the ISASMELT™ furnace.

The possibility of using the nickel ISASMELT™ process to produce Bessemer matte with low iron content directly from concentrate feed is also shown in Fig. 4, as a dotted line. This would obviate the need for Peirce–Smith converting altogether. The direct nickel ISASMELT™ process is similar to the Direct Outokumpu Nickel (DON) process where the metal values in the smelting slag are recovered in the slag cleaning furnace. The DON process was first applied to the Harjavalta plant in Finland in 1995 and then at a much smaller scale at the Fortaleza plant in Brazil in 1998 – Mäkinen and Taskinen (2006). The nickel ISASMELT™ thus has the flexibility to produce either primary smelting matte (Fe in matte greater than 15 wt%) or Bessemer matte (Fe in matte less than 4 wt%) depending on the customer's requirements.

6. Nickel ISACONVERT™ process concept

Matte generated from the primary smelting of nickel concentrate is almost exclusively processed to Bessemer matte using multiple units of Peirce–Smith converters. The exceptions are the Anglo Platinum Watervale smelter in South Africa, where the Anglo Platinum Converting Process (ACP) is employed and Stillwater Mining Company smelter in Montana, USA, where top blown rotary converters are used. In both cases, the respective converting processes treat granulated primary smelting matte, however, of the two, only the ACP is fed continuously.

Continuous nickel converting is not a new concept and has been investigated previously for improved productivity and emission control compared to the traditional Peirce–Smith batch converters. As noted above the ACP plant has already commercialised the basic process concept. In addition, Vale Inco has for over a decade carried out a substantial research and development program investigating continuous nickel converting – Warner and Diaz (2003). Over about this period, Vale Inco investigated three approaches for continuous converting; their own flash converting (Victorovich (1993)), oxygen top blowing-nitrogen bottom stirring bath converting technology, and using Noranda/El Teniente type bath converting technology (Donald et al. (2005)). It is noted that the primary goal in this work was to develop a continuous converting technology applicable for the Copper Cliff, Ontario nickel plant, where downstream refining requires a very low, 0.5 wt%, iron in matte. While technically feasible, in the tests (Donald et al. 2005) the more oxidized slag produced at this low iron content proved somewhat problematic; instead preference was given to a "two-stage" approach involving continuous converting to about 2–3% Fe, followed by batch finishing for final matte grade adjustment.

It is in the production of Bessemer mattes, down to less than 4 wt% iron, where the ISACONVERT™ is seen to have a niche. The basic process block diagram for a nickel ISACONVERT™ plant is shown in Fig. 5. It should be noted that the nickel ISACONVERT™ process is a truly continuous converting process with matte and air/oxygen continuously fed to the bath. The bath consists of the converting products at all times, as the operating conditions effectively fix the process at what is, for Peirce–Smith converters, the end point of the converting reactions.

The nickel ISACONVERT™ furnace continuously processes primary smelting matte feed, fluxes along with recycled slag cleaner matte and dusts. The product liquid Bessemer matte is tapped periodically from the matte taphole to: (1) A granulation system or, (2) Slow cooling system (depending on refinery). Slag is tapped from a separate taphole from the ISACONVERT™ vessel to a separate slag cleaning furnace where metal values are recovered by adding reductant (coke and/or concentrate) to produce a separate matte and a cleaned slag phase. The matte from the slag cleaning furnace is granulated and recycled to the ISACONVERT™ furnace, whereas the slag is discarded. Off-gases from the ISACONVERT™ furnace are directed to a waste heat boiler for heat recovery and de-dusted using an electro-static precipitator before being sent to a sulfuric acid plant for sulfur capture. All dusts collected from the gas handling systems are recycled to the ISACONVERT™ furnace.

The presented nickel ISACONVERT™ process offers two important advantages compared with the traditional batch Peirce–Smith Converting.

The first advantage of the ISACONVERT™ process is the generation of a constant volume of high strength SO_2-containing gases that can be treated in a conventional sulfuric acid plant. This is an important benefit considering current and future stringent environmental regulations affecting both plant emissions and in-plant hygiene. While fitting tight converter hoods remains the option to maintain adequate SO_2 levels in converter off-gas, this approach coupled with the additional need for secondary hooding to control fugitives emissions all add to the overall cost of Peirce–Smith converters. The ISACONVERT™ offers a one-step converting process that can utilize high levels of oxygen enrichment coupled with minimal air dilution, thus producing a lower volume off-gas readily amenable to treatment in a sulfuric acid plant.

A second advantage is the flexibility offered by the use of solid matte as the feed material, thus eliminating ladle transfers, and

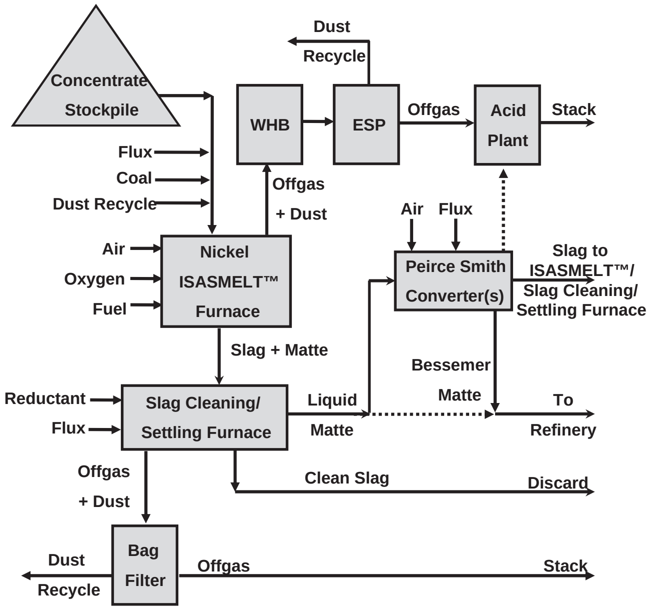


Fig. 5. Process block diagram for a nickel ISASMELT™ plant.

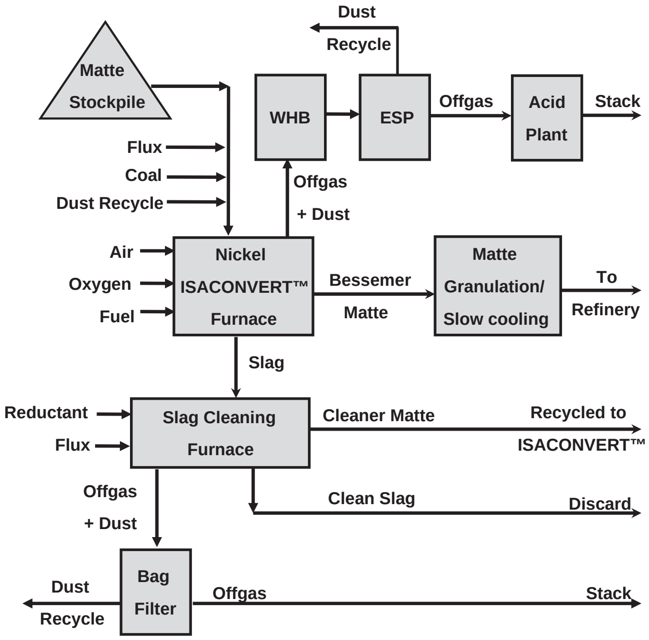


Fig. 6. Process block diagram for a nickel ISACONVERT™ plant.

further lowering opportunities for fugitive emissions with the resulting improvement in-plant hygiene. Further, the use of solid/granulated feed allows decoupling of the smelting and converting steps, giving added flexibility and simplifies the maintenance and operational aspects of the smelter.

7. Nickel ISASMELT™ laterite smelting process concept

The 2007 TMS survey of worldwide nickel laterite smelters reported that the Rotary Kiln-Electric Furnace technology (RKEF) is almost exclusively used for the smelting/reduction of nickel-bearing laterites – Warner et al. (2006). The RKEF process typically treats a low-iron magnesium–aluminium–silicate saprolite material (or sometimes a “transitional” material); the electric furnace (EF) products are a ferro-nickel alloy for the market (after refining) and a discard slag. In some cases, sulfur is added at the reduction kiln, thus producing a low sulfur matte at the EF which is finished to a Bessemer matte by conventional converting. The ISASMELT™ technology has recently been investigated for the smelting of these type of nickel laterite ores.

The basic process block diagram for a nickel laterite smelting ISASMELT™ plant is shown in Fig. 7. The ISASMELT™ furnace continuously processes dried laterite feed and recycled dust, with the smelting heat provided by coal and oxygen-enriched air; flux addition may be considered depending on the ore used. The product liquid alloy and slag are periodically tapped via a single taphole from the ISASMELT™ vessel to a separate slag settling/cleaning furnace. Alternatively the ISASMELT™ furnace can be used for smelting/pre-reduction, maximising fuel energy usage, with the majority of the reduction, and therefore nickel alloy generation, completed in the slag cleaning furnace. Another variation would be to include a calcination step, for example, using a flash calciner before the ISASMELT™ unit. ISASMELT™ off-gases can be post-combusted in the upper section of the furnace (with the addition of some oxygen-enriched air as required), imparting heat to the incoming feed. Off-gases from the top of the ISASMELT™ furnace are then directed to a waste heat boiler for heat recovery and de-dusted using an electro-static precipitator before being sent to the site stack. The product liquid ferro-nickel alloy is periodically tapped from the electric furnace to: (i) A granulation system, or (ii) A casting system (depending on product specifications), or (iii) A hot metal ladle for further refining as needed. Discard slag is also intermittently tapped from the settling furnace for disposal; and advantageously if required, heat recovery. All dusts collected from the gas handling processes are recycled to the ISASMELT™ furnace.

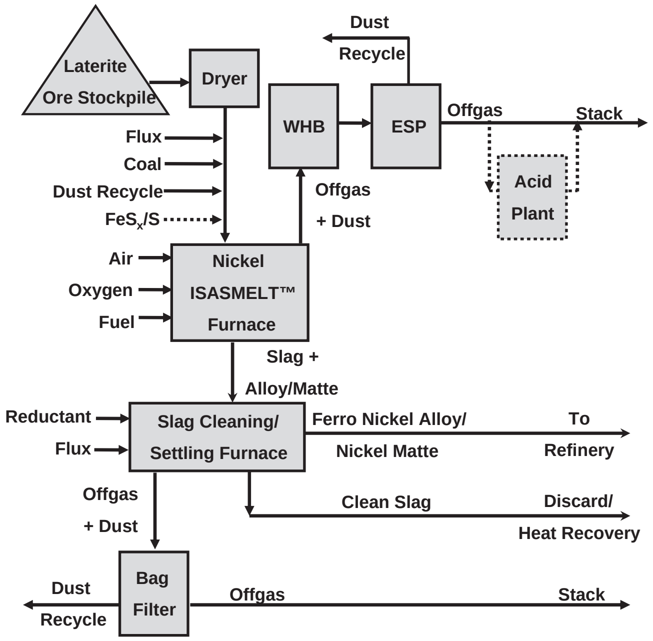


Fig. 7. Process block diagram for a nickel laterite ISASMELT™ plant.

An alternative option to this flowsheet involves the addition of a sulfur-bearing feed (refer to dotted lines and process units in Fig. 6), that can be added to the ISASMELT™ furnace bulk feed to allow for the production of a low-iron matte.

8. Nickel ISASMELT™ pilot plant trials

Several more recent campaigns of trials using an ISASMELT™ pilot plant have been conducted to provide further data regarding the operability of the:

(i) ISASMELT™ process involving the smelting of nickel/copper concentrates to either a high iron primary smelting matte or Bessemer matte;
(ii) ISACONVERT™ process involving the converting of primary high iron smelting nickel/copper matte to Bessemer matte; and
(iii) ISASMELT™ process involving the smelting of nickel laterite ores to a ferro-nickel alloy, involving both pilot and bench-scale tests.

The main objectives of these tests were to determine the product nickel matte/alloy and slag compositions, elemental partition ratios between matte/alloy and slag (particularly nickel and cobalt), and generate slags for subsequent slag cleaning testwork.

8.1. The ISASMELT™ pilot plant

The pilot ISASMELT™ facilities consist of a stationary cylindrical furnace with an internal diameter (within refractory) of approximately 400 mm and a height of approximately 2000 mm. The vessel is lined with chrome–magnesite refractory bricks, and is backed by high alumina insulation bricks. During operation, the furnace contains a molten bath having a maximum depth of about 600 mm. Controlled amounts of air and oxygen are injected into the bath via either a 32 mm (1.25 inch) or 38 mm (1.5 inch) stainless steel lance. The feed is added in known amounts to a calibrated variable speed conveyor belt which drops the feed through a chute at the top of the furnace at a rate typically between 100 kg/h and 250 kg/h. Fuel oil is injected down the lance to control bath temperature. In some cases, solid coal may also be added. Removal of molten products can be achieved by opening one of two tapholes and collecting the materials in cast iron ladles. The process off-gases are cooled and de-dusted before being directed to a caustic soda scrubber for removal of any sulfur-containing gases prior to venting to stack.

8.2. Nickel ISASMELT™ smelting trials

Nickel concentrates were successfully smelted to produce mattes over a wide range of iron contents; from 1.6 wt% Fe to 20.0 wt% Fe. The purpose of the testwork was to demonstrate the process flexibility for the production of matte at any grade required by the customer, ranging in composition from primary smelting matte (Fe in matte greater than 15 wt%) to Bessemer matte (Fe in matte less than 2–4 wt%) from feed concentrates. The typical nickel/copper concentrate feed composition is shown in Table 1.

Table 1
Nickel/copper concentrate feed composition.

Element	Average content (wt%)
Ni	17.1
Cu	4.3
Co	0.4
Fe	31.0
S	27.0
SiO_2	9.8
MgO	6.5
Al_2O_3	1.0
CaO	0.7

Table 2
Range of ISASMELT™ slag compositions.

Element	Range (wt%)
Ni	0.8–7.0
Cu	0.3–1.5
Co	0.2–0.3
Fe	32.0–37.6
SiO_2	25.6–34.9
MgO	6.7–10.3
Al_2O_3	3.8–5.1
CaO	1.4–2.5

The concentrate was pelletised with pre-determined amounts of silica flux (according to the planned slag composition) and charged to the furnace at a rate equivalent to 150 kg/h of wet concentrate. Granular coal was typically added to the feedbelt at a rate equivalent to 5% of the concentrate feedrate to provide additional smelting heat, due to the heat loss term associated with the size of the pilot-scale furnace. Smelting air and oxygen were metered through separate rotameters at a ratio to yield 40–45 v/v% total oxygen. Fuel oil was injected down the lance as trim fuel to maintain the bath temperatures between 1310 °C and 1450 °C, depending on the test conditions.

Fluid slags were produced under all test conditions. The iron–silica ratio in ISASMELT™ slag generated during the tests ranged from 0.9 to 1.4. Compositional ranges for the ISASMELT™ slag for the tests are shown in Table 2.

The distribution coefficients for nickel, as defined by Eq. (1) below, are shown as a function of iron content in bulk matte in Fig. 7, both for the ISASMELT™ and ISACONVERT™ tests (discussed later in the ISACONVERT™ trials section). The bulk matte and slag were analysed from spoon samples taken during tapping.

$$L_X^{s/m} = (\text{wt\%} \times \text{in slag})/(\text{wt\%} \times \text{in matte}) \tag{1}$$

Included in Fig. 7 for comparison are the laboratory results of Henao (2003) and the results obtained during a sampling campaign of a Peirce–Smith blow on nickel-copper matte at the Xstrata Nickel (XNi) Falconbridge smelter (Bustos et al. (1988)) and at the Vale Inco Thompson smelter (Diakow et al. (1975)). In Henao's work, a matte-slag melt was equilibrated at 1500 °C or 1600 °C for a given time period at a specific fixed oxygen (Po_2) and sulfur dioxide (Pso_2) atmosphere (set by S_2/SO_2 and CO/CO_2 ratio control). At the end of each equilibrated test, quenched samples of matte and slag were taken for assay. The XNi Falconbridge and Vale Inco Thompson smelter data were obtained by sampling the matte and slag in the converter at the end of each individual blow. Interestingly, all the data in Fig. 7 compare very well.

The distribution coefficients for cobalt, as defined by Eq. (1), are shown in Fig. 8 as a function of the iron content of matte for the ISASMELT™ and ISACONVERT™ tests. Included in Fig. 8 for comparison are the laboratory results of Font (1999) and, as per Fig. 7 the industrial data from Bustos et al. (1988) and Diakow et al. (1975). In Font (1999)s work, a matte-slag melt was equilibrated (as in Henao (2003)s later work noted above) for a given time at a specific fixed oxygen (Po_2) and sulfur dioxide (Pso_2) atmosphere (set by S_2/SO_2 and CO/CO_2 ratio control). At the end of each test, quenched samples of matte and slag were taken for assay.

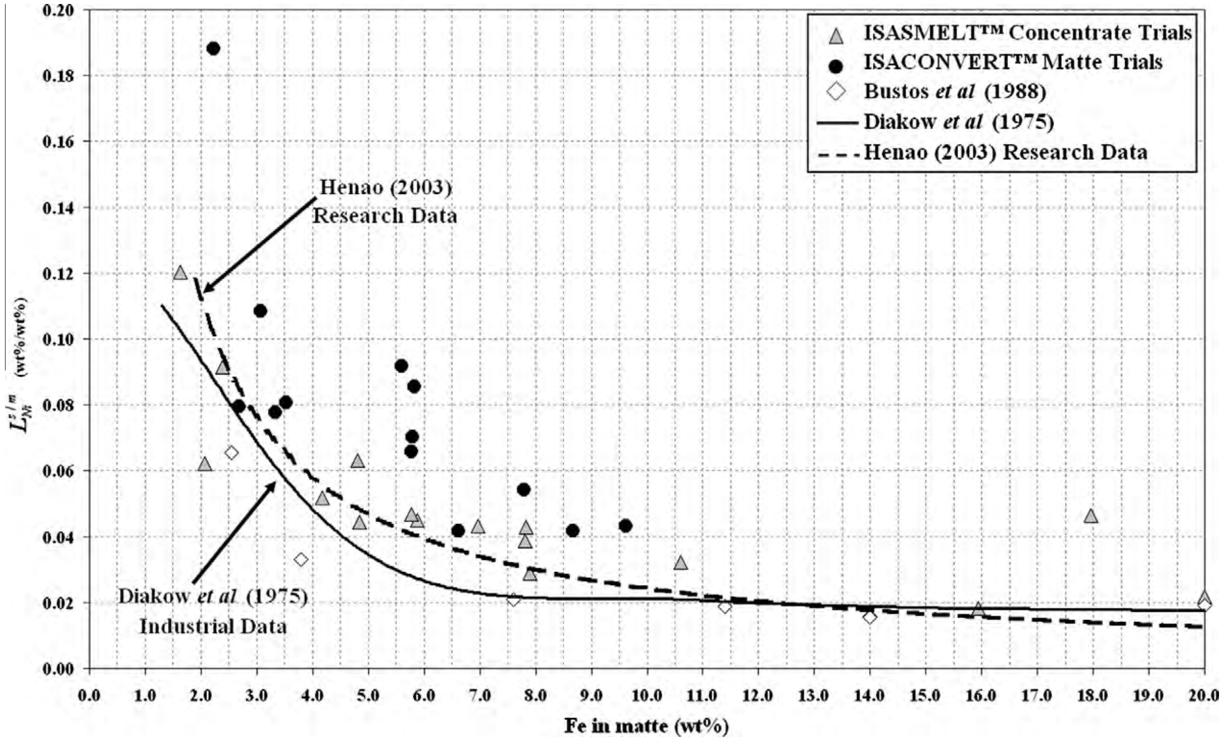


Fig. 8. Nickel in slag as a function of iron in ISASMELT™ matte.

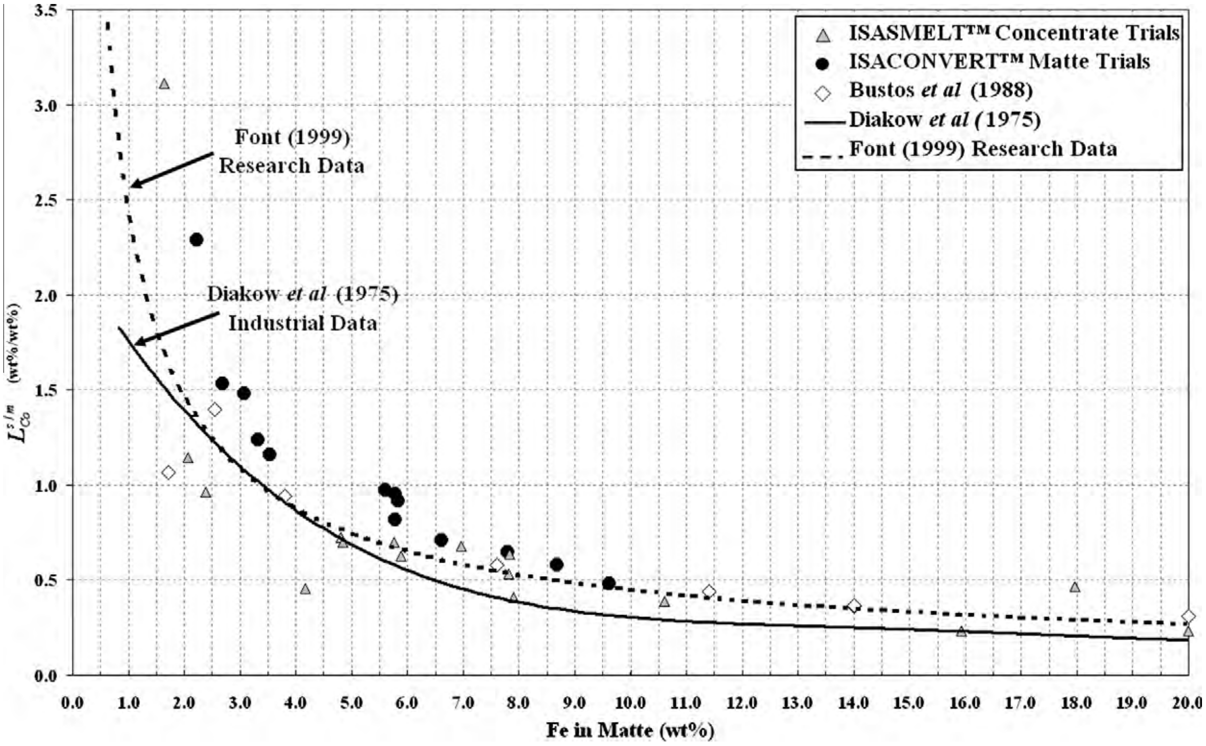


Fig. 9. Cobalt distribution as a function of iron in ISASMELT™ matte.

The XNi Falconbridge and Vale Inco Thompson smelter data here were obtained by the same method as described for Fig. 7. As with Fig. 7, it is of interest to note that all data in Fig. 8 for cobalt distribution between slag and matte compare very well.

Figs. 7 and 8 shows that with decreasing iron levels in matte, both the nickel and cobalt slag/matte distribution coefficients increase. This is as expected due to the higher oxygen potential (Po_2) of the system associated with lower iron levels in matte. Further as noted, the data in Figs. 7 and 8 shows that there is good agreement between the pilot plant results for both the ISASMELT™ and ISACONVERT™ trials and previously published data for somewhat similar matte-slag systems.

In order to provide information on the slag liquidus temperatures for slags encountered in the pilot plant trials, the FactSage™ (Bale et al. (2002)) package was used to evaluate phase equilibria in the $NiO–MgO–FeO–Fe_2O_3–SiO_2–Al_2O_3–CaO$ slag system. The results of the FactSage™ calculations at the conditions of: $Po_2 = 10^{-7.6}$ atm, $Al_2O_3 = 4$ wt%, CaO = 1.5 wt% and MgO = 10 wt% are shown on the pseudo-ternary $NiO–FeO–SiO_2$ diagram in Fig. 9. The FactSage™ liquidus calculations and slag compositions from the pilot trials were plotted by normalising to the axes of the pseudo-ternary graph. It can be seen that all slags encountered in the tests were predicted to lie within the olivine primary phase field (p.p.f.).

8.3. Nickel ISACONVERT™ trials

The purpose of the converting testwork was to demonstrate the ISACONVERT™ process for converting solid high iron primary smelting matte feed to low iron Bessemer mattes. The typical primary smelting matte composition is shown in Table 3. As will be shown, the primary smelting mattes were successfully converted to produce finished mattes with iron contents from 2.2 wt% Fe to 9.6 wt% Fe.

In the tests, matte with a pre-determined amount of high silica flux (set according to the target iron to silica ratio in slag) was charged to the furnace at a rate equivalent to 200 kg/h of “as received” solid matte. Granular coal was added to the feedbelt at a rate equivalent to 5% of the matte feed rate, due to similar considerations as mentioned in pilot-smelting. Converting air and oxygen were metered through separate rotameters at a ratio to yield 35–40 v/v% total oxygen enrichment. As with the smelting tests, fuel oil was injected down the lance as trim fuel to maintain the bath temperatures between 1310 °C and 1380 °C depending on the test conditions.

It was found that fluid slags were produced under all test conditions over a range of iron–silica ratios in ISACONVERT™ slag from 1.1 to 2.3. Compositional ranges for the ISACONVERT™ slag for the tests are shown in Table 4.

Table 3
Primary smelting nickel matte composition.

Element	Average content (wt%)
Ni	44.5
Cu	9.7
Co	2.9
Fe	25.3
S	17.2
SiO_2	0.3

Table 4
Range of ISACONVERT™ slag compositions.

Element	Range (wt%)
Ni	2.2–6.4
Cu	0.6–1.2
Co	1.3–2.5
Fe	37.8–48.0
SiO_2	25.6–34.9

As noted above, the level of nickel in slag and the slag/matte cobalt distribution coefficient for the ISACONVERT™ tests are included in Figs. 7 and 8. The nickel ISACONVERT™ tests aimed to achieve low iron content in matte. As evident in Fig. 7, the Ni level in slag rises significantly towards the low Fe matte composition. On account of the more oxidized conditions at low iron levels in matte, the slag under these conditions has a higher liquidus. However in this case, the Fe/SiO_2 can be trimmed to lower the liquidus temperature.

As with the smelting tests, the FactSage™ (Bale et al. (2002)) package was also used for the converting conditions to predict the liquidus temperatures in the slag system using the simplified $NiO–FeO–Fe_2O_3–SiO_2–Al_2O_3–CaO–MgO$ system. The predictions were projected, through normalisation of the data, onto the pseudo-ternary $NiO–FeO–SiO_2$ system in Fig. 10. The following conditions were used for the FactSage™ calculations: $Po_2 = 10^{-7.6}$ atm, $Al_2O_3 = 2.5$ wt%, CaO = 1.5 wt%, and MgO = 2.5 wt%. The compositions of the final ISACONVERT™ slags are plotted together with the liquidus predictions in Fig. 10. Similar to the ISASMELT™ slags, most of the ISACONVERT™ slags were within the olivine primary phase field (p.p.f.).

8.4. Nickel laterite ISASMELT™ trials

The purpose of the laterite smelting testwork was to demonstrate the production of a ferro-nickel alloy at both the bench and pilot-scale. The campaigns were undertaken on two separate laterite deposits that were not related by geographical location. The typical compositional ranges for the two laterite ores used for the work are shown in Table 5. These laterite ores were successfully smelted in the tests under reducing conditions to produce a ferro-nickel alloy and a low nickel slag. The bench-scale trials produced ferro-nickel alloy grades between 15 and 40 wt% Ni with a starting SiO_2/MgO and Fe/Ni ratios between 1.2–3.1 and 10–20 respectively. On the other hand, the pilot-scale trials produced ferro-nickel alloy grades between 60 and 85 wt% Ni with a starting SiO_2/MgO and Fe/Ni ratios between 2.0–3.6 and 5–12 respectively.

The bench-scale tests were carried out batch-wise in small crucibles (holding approximately 1 kg of melt). After initial melting, the bath was reduced with the addition of coal over time. Due to the preliminary nature of the tests, a flux – high grade silica and/or limestone – was added to the ore to ensure the formation of a liquid slag. It is recognised that in practice, fluxing would not normally be carried out. For the pilot-scale trials, the laterite ore with silica and limestone flux were charged to the furnace at a rate equivalent to 150–200 kg/h of ore. Solid coal was added to the feedbelt for both reduction and to provide a portion of the required smelting energy. The balance of the heat to maintain the molten bath was supplied by fuel oil injected down the lance.

Table 5
Nickel laterite feed compositions.

Element	Bench-scale testwork Average content (wt%)	Pilot-scale testwork Average content (wt%)
Ni	0.6–1.8	0.3–3.6
Fe	5.0–25	8.4–20.9
SiO_2	22–46	40–46
MgO	12–25	12–22
Al_2O_3	2.5–8.5	0.7–3.0
CaO	0.1–1.1	0.2–2.5

It was found that fluid slags were produced under all test conditions; the compositional ranges for the slags for both the bench-scale tests and the pilot-scale trials are shown in Table 6.

The trend of weight percent nickel in slag as a function of time, as the coal was added to the crucible melt, in the bench-scale tests is shown in Fig. 11. The bench-scale trials were targeted to achieve selected ferro-nickel grades, whilst at the same time producing low nickel contents in the slag. It was also found that the nickel level in slag was highly affected by the settling time. Operating temperature and flux addition can be optimized to minimise the settling requirement and maximise ferro-nickel alloy recovery.

The FactSage™ (Bale et al. (2002)) package was used to predict the liquidus temperatures for the laterite smelting slag conditions. The simplified $FeO–Fe_2O_3–SiO_2–Al_2O_3–CaO–MgO–Cr_2O_3$ system was used to represent these slags. Due to the low levels of nickel oxide in the slag after reduction, nickel was not taken into account in the FactSage™ calculations. The predictions were projected, through normalisation of the data, onto the pseudo-ternary

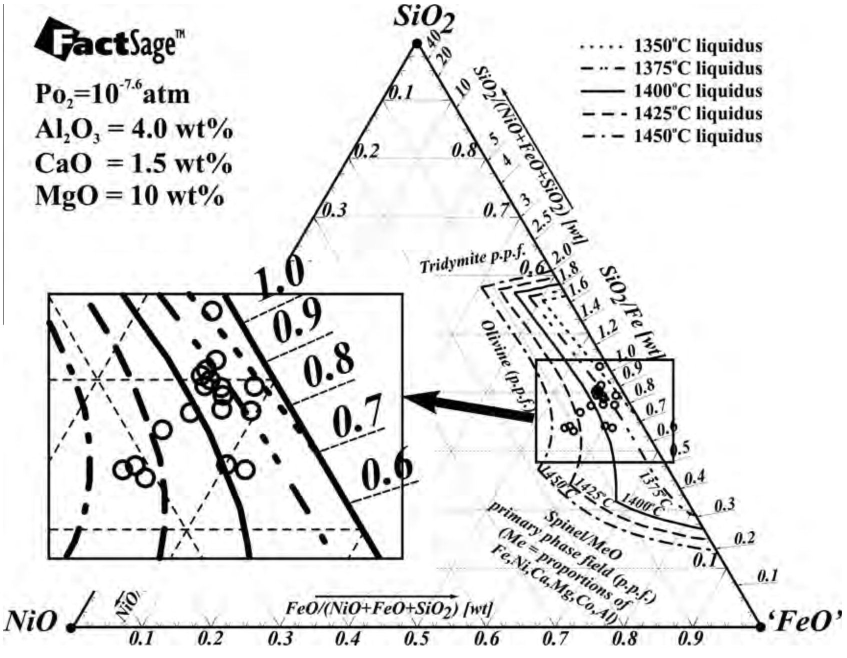


Fig. 10. Liquidus in the system NiO–'FeO'–SiO_2–Al_2O_3–CaO–MgO at $Po_2 = 10^{-7.6}$ atm, $Al_2O_3 = 4.0$ wt%, CaO = 1.5 wt% and MgO = 10 wt%.

Table 6
Range of ISASMELT™ nickel laterite slag compositions.

Element	Batch-scale testwork Average content (wt%)	Pilot-scale testwork Average content (wt%)
Ni	0.03–0.3	–
Fe	7.1–21	18–25
SiO_2	35–50	44–50
MgO	20–27	14–18
Al_2O_3	5.0–6.9	2.6–3.3
CaO	2.8–8.6	0.2–2.3

$MgO–FeO–SiO_2$ system in Fig. 12. The following conditions were used for the FactSage™ calculations: $Po_2 = 10^{-11.1}$ atm, $Al_2O_3 = 4.0$ wt%, CaO = 0.2 wt%, and $Cr_2O_3 = 1.0$ wt%. The range of laterite crucible-scale and pilot-scale compositions for the ISASMELT™ slags are plotted as a grey circle together with the FactSage™ liquidus predictions in Fig. 12. Depending on bulk slag composition and temperature, the primary phase field of the slag system in the highlighted region can be pyroxene, olivine or tridymite/cristobalite (see Fig. 13).

9. Future development

Xstrata Technology has demonstrated to date that the ISASMELT™ process is an attractive option for both primary and secondary copper and lead smelting. More than 9,000,000 tonnes per year of feed materials are processed in these ISASMELT™ furnaces in copper and lead smelters at smelters around the world. The features that make ISASMELT™ attractive for copper and lead smelting can be applied equally to nickel smelting and converting operations. The use of the ISASMELT™ process for copper/nickel smelting has already been successfully demonstrated on a commercial scale.

Recent ISASMELT™ and ISACONVERT™ pilot and batch-scale testwork described here has been successfully completed on the:

(i) ISASMELT™ process for producing primary high iron smelting matte and Bessemer matte directly from concentrates;
(ii) ISACONVERT™ processes for production of Bessemer matte from high iron matte; and

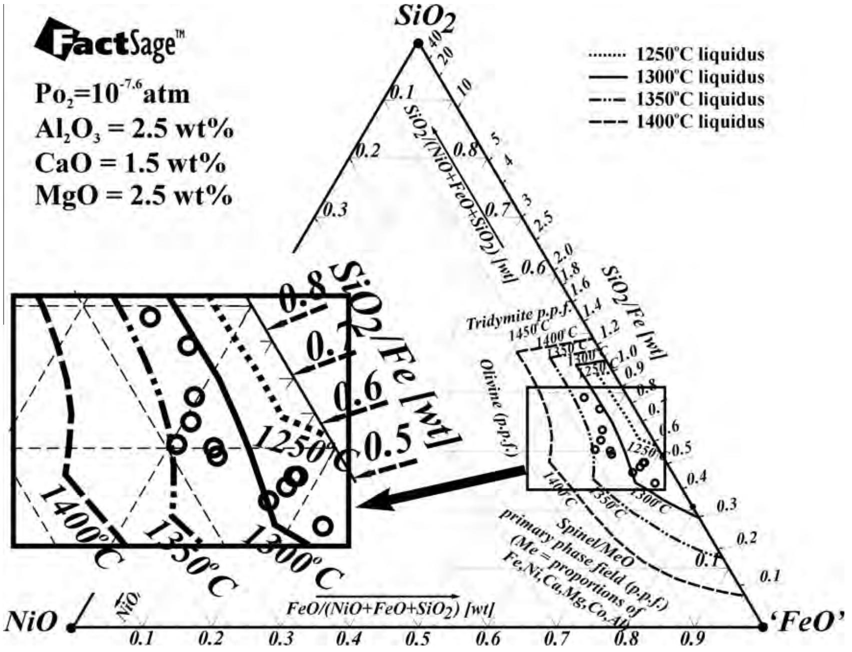


Fig. 11. Liquidus in the system NiO–'FeO'–SiO_2–Al_2O_3–CaO–MgO at $Po_2 = 10^{-7.6}$ atm, Al_2O_3 = 2.5 wt%, CaO = 1.5 wt% and MgO = 2.5 wt%.

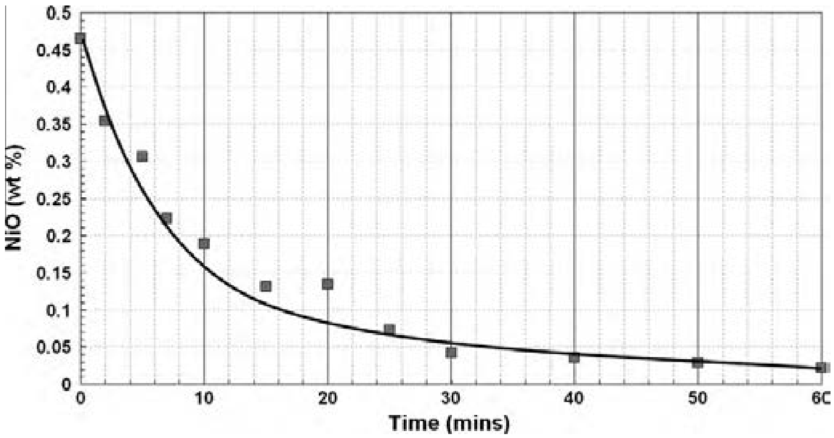


Fig. 12. Reduction of nickel from slag bath from Bench-scale testing.

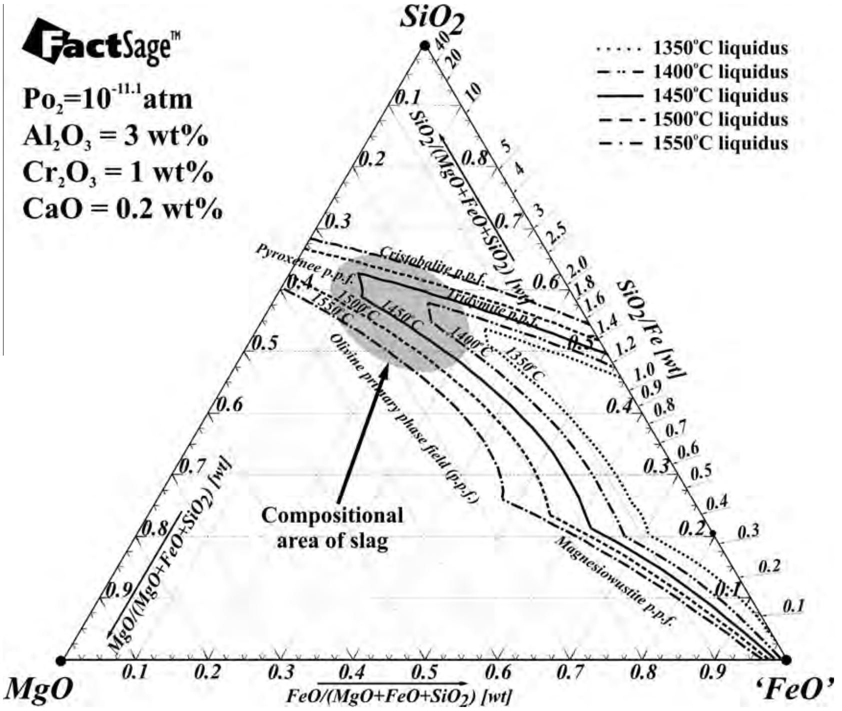


Fig. 13. Liquidus in the system FeO–Fe_2O_3–SiO_2–Al_2O_3–CaO–MgO–Cr_2O_3 at $Po_2 = 10^{-11.1}$ atm, Al_2O_3 = 4.0 wt%, CaO = 0.2 wt%, and Cr_2O_3 = 1.0 wt%.

(iii) ISASMELT™ process for producing ferro-nickel alloys of various grades from nickel laterite ores.

Future development of the nickel ISASMELT™ process may require additional improvements to be made to the existing furnace designs. These include:

(i) Furnace cooling – operation at temperatures above 1300 °C will require furnace cooling systems to ensure sufficiently long campaign life;
(ii) Tapping blocks – tapping of nickel matte/alloy will require improved tapping block design;
(iii) Pneumatic injection – exploitation of dry concentrate injection through the ISASMELT™ lance to minimise fuel requirements; and
(iv) Optimisation of the ISASMELT™ slag cleaning process to maximise payable metal recoveries.

These development issues are currently being addressed.

Acknowledgement

The authors would also like to thank Xstrata Technology for permission to publish this paper.

References

Arthur, P.S., Hunt, S.P., 2005. ISASMELT™ – 25 Years of Continuous Evolution. In: Nilmani, M., Rankin, W.J. (Eds.), John Floyd International Symposium on Sustainable Developments of Metals Processing, NCS Associates, Melbourne, Australia, pp. 73–94.

Bakker, M.L., Alvear, G.R.F., Kreuh, M., 2009. ISASMELTâ¢ TSL â€Making a splash for nickel. In: The 48th Annual Conference of Metallurgists 2009 â€ Pyrometallurgy of Nickel and Cobalt, Sudbury, Canada, 2009.

Bale, C.W., Chartrand, P., Decterov, S.A., Eriksson, G., Hack, K., Mahfoud, R.B., Melançon, J., Pelton, A.D., Petersen, S., 2002. CALPHAD 26, 189–228.

Burford, B., 2009. The ISASMELT™ Technology Package: Over 30 Years of Innovation, AusIMM Bulletin, No. 1, February 2009, pp. 26–30.

Bustos, A.A., Ip, S.W., O'Connell, G., Kaiura, G.H., Toguri, J.M., 1988. In: Tyroler, G.P., Landolt, C.A. (Eds.), Converting Simulation at Falconbridge Limited, Extractive Metallurgy of Nickel and Cobalt, The Metallurgical Society, Inc., AIME, USA, pp. 335–354.

Diakow, J.S., Mak, Y.F., Orr, R.G., 1975. Metallurgy of the Converting Process in the Thompson Smelter, Paper Presented to the 14th Annual Conference of Metallurgists, Edmonton, Alberta, August 1975, The CIM, Montreal, Quebec, Canada, 1975.

Donald, J.R., Warner, A.E.M., Dalvi, A.D., Scholey, K., Ross, A., Harris, C., 2005. In: Ross, A., Warner, T., Scholey, K., (Eds.), Risks and Opportunities in Continuous Converting for Nickel at Inco, Converter and Fire Refining Practices, The Minerals, Metals & Materials Society, USA, 2005, pp. 295–307.

Edwards, J.S., Alvear, G.R.F., 2007. Converting using ISASMELT™ Technology, Copper 07, vol. III, Book 2: The Carlos Diaz Symposium on Pyrometallurgy. In: Warner, A.E.M., Newman, C.J., Vahed, A., George, D.B., Mackey, P.J., Warczok, A. (Eds.), MetSoc, Toronto, CA, 2007, CIM, Montreal, 2007, vol. 3, Book 2, pp. 17–28.

Font, J.M., 1999. Phase equilibrium and minor elements distribution between slag and matte phases in nickel smelting, Ph.D. Thesis, Tohoku University, 1999.

Henao, H.M.Z.,2003. Phase equilibrium between Ni–S or Ni–Fe melt and slag in nickel smelting, Ph.D. Thesis, Tohoku University, 2003.

Mäkinen, T., Taskinen, P., 2006. The State of the Art in Nickel Smelting: Direct Outokumpu Nickel Technology. In: Kongoli F., Reddy, R.G., (Eds.), Sohn International Symposium Advanced Processing of Metals and Materials, The Minerals, Metals & Materials Society, USA, 2006, pp. 313–325.

Nikolic, S., Edwards, J.S., Burrows, A.S., Alvear, G.R.F., 2009. ISACONVERT™ – TSL Continuous Copper Converting Update. In: Kapusta, J. and Warner, T. (Eds.), In: International Peirce–Smith Converting Centennial: Held During TMS 2009 Annual Meeting and Exhibition. The Minerals, Metals & Materials Society, USA, February 15–19, pp. 407–414.

Victorovich, G.S., 1993. Oxygen Flash Converting for Production of Copper, Extractive Metallurgy of Copper, Nickel and Cobalt. In: Reddy, R.G., Weizenbach, R.N., (Eds.), In: Proceedings of the Paul E. Queneau International Symposium. The Metallurgical Society, Inc., AIME, USA, 1993, vol. 1, pp. 501–529.

Warner, A.E.M., Diaz, C.M., 2003. An Overview of the Metallurgy of Nickel–Copper Matte Converting. In: Kongoli, F., Itagaki, K., Yamauchi, C., Sohn, H.Y., (Eds.), Yazawa International Symposium Metallurgical and Materials Processing: Principles and Technologies. The Minerals, Metals & Materials Society, USA, 2003, pp. 113–129.

Warner, A.E.M., Diaz, C.M., Dalvi, A.D., Mackey, P.J., Tarasov, A.V., 2006. JOM World Nonferrous Smelter Survey Part III: Nickel: Laterite, JOM, April 2006, pp. 11–20.

Warner, A.E.M., Diaz, C.M., Dalvi, A.D., Mackey, P.J., Tarasov, A.V., Jones, R.T., 2007. JOM World Nonferrous Smelter Survey Part IV: Nickel: Sulfide, JOM, April 2007, pp. 58–72.

Технология продувки методом погружения сверху (TSL) ISASMELT™ – Применение в никелевой промышленности

М.Л. Беккер[a,*], С. Николич[3], П.Дж. Макки[b]
[a] Xstrata Technology, Level 4, 307 Queen Street, Brisbane 4000, Australia
[b] PJ Mackey Technology Inc„ Kirkland, Quebec, Canada
* Ответственный автор:
Эл. почта: mbakker@xstratatech.com.au (М.Л. Беккер).

ИНФОРМАЦИЯ О СТАТЬЕ

История статьи: Опубликована онлайн 25 октября 2010 года
Ключевые слова: Экстракционная металлургия, Пирометаллургия, Сульфидные руды, Оксидные руд

АВТОРЕФЕРАТ

Процесс ISASMELT™ представляет собой технологию плавки в жидкой ванне с продувкой методом погружения сверху (TSL), которая развивается и оптимизируется на протяжении последних 25 лет. К концу 2011 года общая установленная мощность технологии ISASMELT превысит 9 млн. тонн в год питания, перерабатываемого на медe- и свинцовоплавильных заводах по всему миру. Технология в равной мере эффективна при плавке никелевых сульфидных концентратов, конвертировании никелевого штейна и производстве ферроникеля из латеритных руд. В настоящей статье описывается, каким образом особенности, которые делают технологию ISASMELT™ привлекательной для использования в плавке меди и свинца, могут быть применены при плавке и конвертировании никеля. Представлены концептуальные технологические схемы, основанные на результатах недавних полупромышленных и лабораторных испытаний.

1. Введение

Применение фурмы позволило использовать стационарные печи простой конструкции и с очень высокой скоростью реакции. Предшествующая технология была основана на подаче газов в жидкие шлаки или штейны, как правило, через фурмы, что усложняло конструкцию и приводило к проблемам с огнеупорами. Технология плавки в жидкой ванне с продувкой методом погружения сверху (TSL) была разработана и внедрена в промышленном масштабе на металлургическом комплексе Маунт Айза в Австралии в начале 1990-х годов и впоследствии получила название ISASMELT™. Развитие технологии на протяжении последних 30 лет было направлено преимущественно на плавку свинцовых и медных концентратов или вторичных материалов. Промышленные печи ISASMELT™, эксплуатируемые компанией Xstrata и ее лицензиатами, в настоящее время способны переплавлять свыше 9 млн. тонн питания в год.

2. Технология ISASMELT™

В технологии ISASMELT™ используется закрытая конструкции печи, предотвращающая выбросы в окружающую среду. Технология продувки через погружную фурму обеспечивает высокоэффективное перемешивание и протекание реакций между компонентами питания в расплавленной шлаковой ванне. Использование совершенных систем управления технологическим процессом позволяет в значительной степени автоматизировать работу печи. Поскольку печь является вертикальной, установка занимает очень малую площадь и может быть с легкостью включена в технологическую схему существующего металлургического завода, дополнив или заменив применяющуюся технологию. Принципиальная схема процесса показана на рисунке 1.

За 30 лет разработки и эксплуатации технологии продувки через погружную фурму на крупных заводах были внесены существенные технические усовершенствования, в частности, в конструкцию печи, системы подготовки питания, системы утилизации отходящих газов, стратегии эксплуатации и управления технологическим процессом, обслуживание огнеупоров и обучение операторов. Результатом накопленного опыта стал продукт, сегодня хорошо известный рынку как технологический пакет ISASMELT™, лицензии на который предлагаются всем заинтересованным клиентам, - Burford (2009). Многие из усовершенствований, внесенных операторами установок, были доведены до сведения и внедрены другими лицензиатами. Обмен идеями и техническими усовершенствованиями происходит в форме отдельных поездок на объекты других лицензиатов, а также на регулярных семинарах для лицензиатов, организуемых Xstrata Technology. На рисунке 2 показано расположение действующих лицензионных промышленных установок.

3. Никелевая установка ISASMELT™ компании Agip

Менее известно, что процесс ISASMELT™ был также адаптирована для переработки никельсодержащего питания на самой ранней стадии своей разработки. В 1980-е годы на никелевых месторождениях, принадлежащих Mount Isa Mines Ltd., были проведены масштабные полупромышленные испытания. Испытания были также выполнены по заказу компании AGIP Australia Pty Ltd., которой принадлежало месторождение Радио Хилл в Западной Австралии, - Bakker et al. (2009). В 1991 году компания AGIP приняла решение построить полупромышленную установку ISASMELT™ для производства никель-медного штейна из концентрата. Это была первая установка ISASMELT™, построенная и сданная в эксплуатацию по заказу стороннего клиента. Ранее в 1991 году на комбинате Маунт Айза была сдана в эксплуатацию первая промышленная установка ISASMELT™, предназначенная для производства чернового свинца из концентратов. Год спустя были сданы в эксплуатацию две промышленные установки ISASMELT™, предназначенные для производства медного штейна из концентратов: одна на комбинате Маунт Айза, а другая в Майами, штат Аризона.

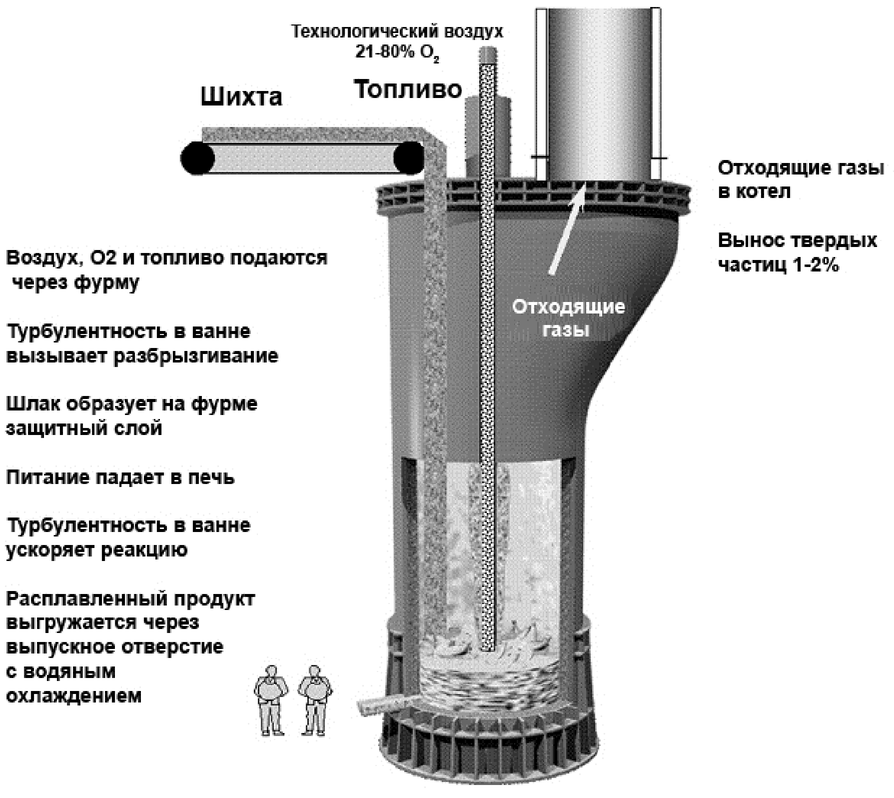
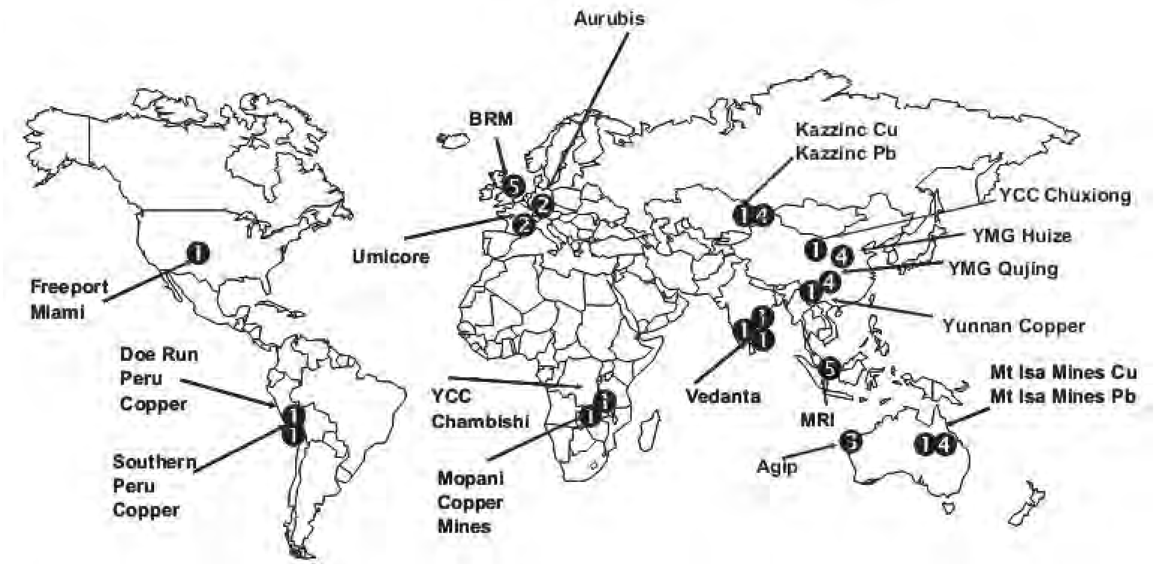


Рисунок 1. Принципиальная схема ISASMELT™.



УСЛОВНЫЕ ОБОЗНАЧЕНИЯ:
1. Металлургический завод, первичная медь
2. Металлургический завод, вторичная медь
3. Металлургический завод, первичный никель
4. Металлургический завод, первичный свинец
5. Металлургический завод, вторичный свинец

Рисунок 2. Расположение установок ISASMELT™.

Промышленная никелевая установка ISASMELT™ компании AGIP была введена в эксплуатацию в сентябре 1991 года и в течение 3 месяцев вышла на проектную мощность 7,5 т концентрата в час - Arthur and Hunt (2005). Она производила никель-медный штейн с содержанием 45% масс. из концентрата, содержавшего примерно 7% масс. никеля и 3,5% масс. меди. Установка показана на рисунке 3 - Bakker et al. (2009). На тот момент проект был признан успешным с технической и эксплуатационной точек зрения, однако из-за большого падения цен на никель объект пришлось закрыть в конце 1991 года. Затем установка была демонтирована.

4. Процесс ISACONVERT™

Результаты опытной эксплуатации полупромышленных установок непрерывного конвертирования меди и никеля в совокупности с результатами промышленной эксплуатации установок периодического конвертирования показывают, что печь ISASMELT™ хорошо подходит для непрерывного конвертирования, - печь ISACONVERT™ - Edwards and Alvear [7].

В технологии ISASMELT™ используется такая же закрытая конструкции печи, как и в ISASMELT™, предотвращающая выбросы в окружающую среду. Процесс использует технологию продувки TSL, обеспечивающую высокоэффективное смешивание и взаимодействие твердого штейна и флюса, которые могут подаваться через свод печи. Как и печь ISASMELT™, ISACONVERT™ занимает небольшую площадь и, как правило, может быть с легкостью установлена на действующих производствах.

Последние прикладные исследования в совокупности с новаторскими полупромышленными испытаниями сделали возможным внедрение процесса ISACONVERT™ в промышленных масштабах - Nikolic et al. (2009). Установка ISACONVERT™ в разрезе показана на рисунке 4.

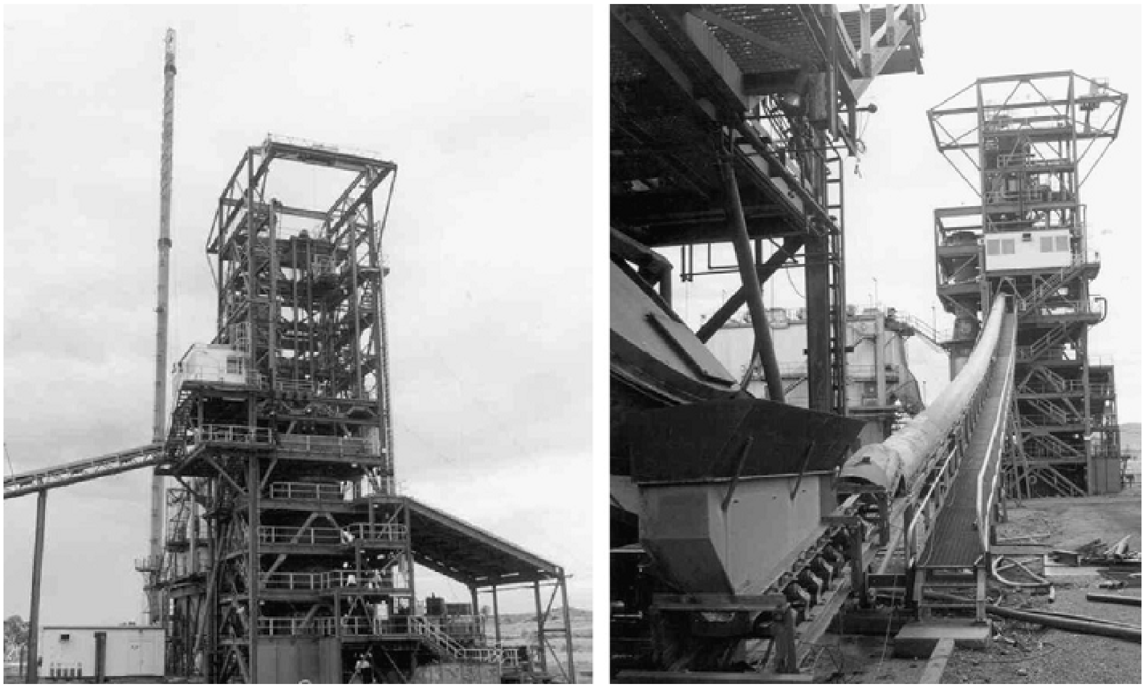


Рисунок 3. Фотографии никелевой установки ISASMELT™ компании AGIP.

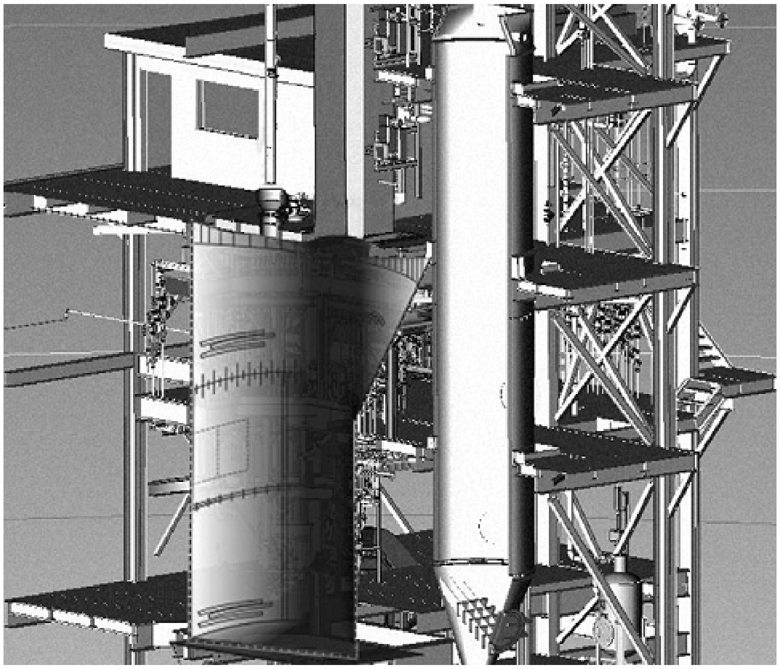


Рисунок 4. Конструкция печи ISACONVERT™ в разрезе.

Применение печей ISASMELT™ и ISACONVERT™ в медной промышленности значительно сократить занимаемую площадь по сравнению с обычным медеплавильным заводом. То обстоятельство, что процессы выплавки и конвертирования очень похожи и легко управляются с использованием проверенной системы управления технологическим процессом, также упрощает эксплуатацию металлургического завода и логистику, позволяя осуществлять все операции из одного центрального пункта управления. Эти факторы снижают стоимость эксплуатации нового завода по сравнению с традиционной технологией, основанной на использовании конвертеров Пирса-Смита. Значительно меньший объем отходящих газов в процессе ISACONVERT™ позволяет сократить капитальные и эксплуатационные затраты на системы сбора и очистки отходящих газов.

5. Принцип действия процесса плавки сульфида никеля ISASMELT™

Питанием металлургического завода, перерабатывающего сульфид никеля, как правило, является никель-медный концентрат, который также может содержать небольшое количество кобальта и металлов платиновой группы. Продуктом плавки такого питания является первичный штейн с высоким содержанием железа, который подвергается дополнительной обработке, как правило, в конвертерах Пирса-Смита с получением конечного высокосортного штейна, часто называемого "файнштейн".

Проведенное TMS в 2007 году исследование металлургических заводов, перерабатывающих сульфид никеля, показало, что на долю взвешенной плавки приходится примерно 70% от мирового объема производства никеля (исходя из данных отчетности за 2004-2005 годы), а остальной объем производства - на долю плавки в электропечах - Warner et al. (2007). Однако, те самые особенности, которые делают процесс ISASMELT™ привлекательным для плавки меди и свинца, столь же применимы в условиях плавки никеля - Bakker et al. (2009).

Принципиальная технологическая схема основной плавки никеля с использованием процесса ISASMELT™ показана на рисунке 4. В никелевой печи ISASMELT™ происходит непрерывная переработка концентрата, флюсов и рециркулируемого выноса печи. Образующиеся жидкий штейн и шлак периодически выпускают из печи ISASMELT™ в отдельную печь отжига шлака через одну летку. Отходящие газы печи ISASMELT™ направляют в котел-утилизатор для утилизации тепла и обеспыливания с использованием электростатического осадителя, а затем на установку серной кислоты для извлечения серы. Штейн основной плавки переносят ковшами из печи осаждения в конвертеры Пирса-Смита для получения файнштейна. Также возможно применение непрерывного конвертирования (см. ниже). Печь осаждения также периодически освобождают от утилизируемого отработанного шлака. Вся пыль, собранная системами очистки газов, возвращается в печь ISASMELT™.

Возможность применения в производстве никеля процесса ISASMELT™ для получения файнштейна с низким содержанием железа непосредственно из концентрата также показана пунктирной линией на рисунке 4. Это полностью устраняет необходимость в конвертерах Пирса-Смита. Прямой процесс выплавки никеля ISASMELT™ аналогичен прямому процессу выплавки никеля Outokumpu (DON), в котором ценные металлы, содержащиеся в шлаке, извлекают в печи отжига шлака. Процесс DON был впервые применен на заводе Харьявалта в Финляндии в 1995 году, а затем

в гораздо меньших масштабах на заводе Форталеза в Бразилии в 1998 году - Makinen and Taskinen (2006). Таким образом, процесс выплавки никеля ISASMELT™ позволяет производить либо штейн первичной плавки (содержание железа в штейне выше 15% масс.), либо файнштейн (содержание железа в штейне ниже 4% масс.) в зависимости от потребностей заказчика.

6. Принцип действия процесса выплавки никеля ISACONVERT™

Штейн, полученный в результате основной плавки никелевого концентрата, почти полностью перерабатывают в файнштейн в нескольких конвертерах Пирса-Смита. Исключение составляют металлургический завод Ватервал компании Anglo Platinum в ЮАР, на котором применяется процесс конвертирования Anglo Platinum (ACP), и металлургический завод Stillwater Mining Company в штате Монтана, США, на котором установлены вращающиеся конвертеры верхней продувки (TBRC). В обоих случаях в соответствующих процессах конвертирования перерабатывается штейн основной плавки, но только процесс ACP имеет непрерывную подачу питания.

Непрерывное конвертирование никеля не является новой технологией и исследовано с точки зрения повышения производительности и снижения уровня выбросов по сравнению с традиционными периодическими конвертерами Пирса-Смита. Как отмечено выше, основные принципы процесса уже реализованы в промышленном масштабе в установке ACP. Кроме того, компания Vale Inco в течение уже более чем десяти лет осуществляет масштабную программу исследований и разработок, предметом которой является непрерывное конвертирование никеля, - Warner and Diaz (2003). В течение этого периода компания Vale Inco изучила три подхода к непрерывному конвертированию: собственная технология взвешенной плавки (Victorovich (1993)), технология конвертирования в ванне с верхней продувкой кислородом и донным перемешиванием азотом и технология конвертирования в ванне типа Noranda/El Teniente (Donald at al. (2005)). Необходимо отметить, что основная цель программы заключалась в разработке технологии непрерывного конвертирования, применимой на никелевом заводе Коппер Клифф, Онтарио, где требования рафинировочного завода предполагают содержание железа в штейне не выше 0,5% масс. Хотя это технически осуществимо, испытания показали (Donald et al. 2005), что окисленный шлак, получаемый при столь низком содержании железа, является источником определенных проблем. Поэтому предпочтение было отдано "двухстадийному" подходу с непрерывным конвертированием до содержания железа в штейне на уровне 2-3% с последующей периодической доводкой для получения конечного штейна желаемого сорта.

Нишей, в которой, как представляется, может успешно применяться процесс ISACONVERT™, является производство файнштейнов с содержанием железа менее 4% масс. Принципиальная технологическая схема плавки никеля с использованием процесса ISACONVERT™ показана на рисунке 5. Следует отметить, что процесс плавки никеля ISACONVERT™ является процессом непрерывного конвертирования с непрерывной подачей в ванну штейна и воздуха/кислорода. Ванна всегда состоит из продуктов конвертирования, так как технологический режим фактически удерживает процесс в точке, соответствующей точке окончания реакций конвертирования в конвертерах Пирса-Смита.

В никелевой печи ISACONVERT™ переработка штейна основной плавки, флюсов и рециркулируемого штейна печи отжига шлака и выноса происходит непрерывно. Продукт, жидкий файнштейн, периодически выпускают через летку в: (1) систему гранулирования или (2) систему медленного охлаждения (в зависимости от металлургического завода). Шлак выпускают из отдельной летки печи ISACONVERT™ в отдельную печь отжига шлака, в которой ценные металлы извлекают путем добавления восстановителя (кокса и/или концентрата) с получением отдельного штейна и очищенной шлаковой фазы. Штейн из печи отжига шлака гранулируют и возвращают в печь ISACONVERT™, а шлак утилизируется. Отходящие газы печи ISACONVERT™ направляют в котел-утилизатор для утилизации тепла и обеспыливания с использованием электростатического осадителя, а затем на установку серной кислоты для извлечения серы. Вся пыль, собранная системами обработки газов, возвращается в печь ISACONVERT™.

Описанный процесс конвертирования никеля ISACONVERT™ обладает двумя основными преимуществами по сравнению с традиционным периодическим конвертированием в конвертере Пирса-Смита.

Во-первых, процесс ISACONVERT™ отличается постоянным объемным расходом отходящих газов с высоким содержанием SO_2, которые могут перерабатываться в обычной установке серной кислоты. Это важное преимущество, учитывая строгие экологические нормы, регулирующие современные и будущие промышленные выбросы и промышленную гигиену. Хотя установка камина остается одним из возможных вариантов поддержания необходимо уровня SO_2 в отходящих газах конвертера Пирса-Смита, при этом также требуется второй камин для сбора выбросов, миновавших первый камин, что, как правило, увеличивает общую стоимость конвертеров Пирса-Смита. В процессе ISACONVERT™ конвертирование происходит в один этап, что позволяет использовать максимальное обогащение кислородом с минимальным разбавлением воздухом и меньшим объемом отходящих газов, которые без дополнительной подготовки могут быть направлены в установку серной кислоты.

Во-вторых, процесс ISACONVERT™ отличается гибкостью, позволяющей использовать в качестве питания твердый штейн, что устраняет необходимость в подаче расплавленного штейна в ковше, а также уменьшает риск неконтролируемых выбросов и улучшает состояние промышленной гигиены на заводе. Кроме того, использование твердого или гранулированного питания позволяет развязать стадии плавки и конвертирования, что повышает гибкость процесса и упрощает обслуживание и эксплуатацию завода.

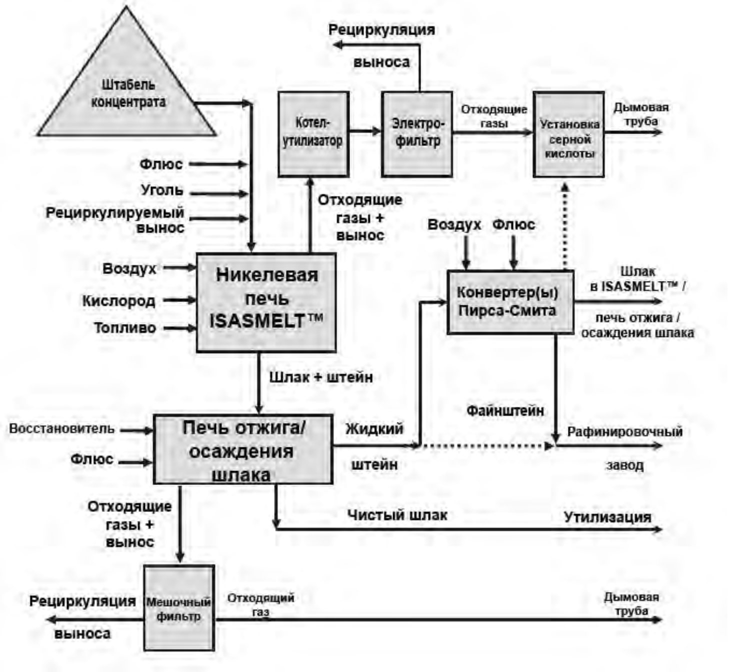


Рисунок 5. Технологическая схема никелевой установки ISASMELT™.

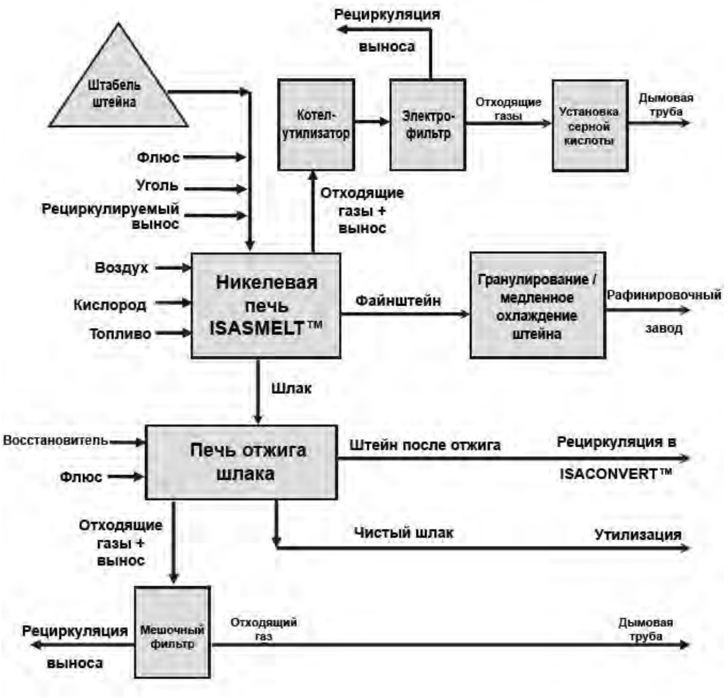


Рисунок 6. Технологическая схема никелевой установки ISACONVERT™.

7. Принцип действия процесса плавки латерита ISASMELT™

Проведенное TMS в 2007 году исследование металлургических заводов, перерабатывающих латеритный никель, показало, что для плавки/восстановления никеленосных латеритов применяется практически исключительно технология вращающейся печи - электропечи (RKEF) - Warner et al. (2006). Процесс RKEF, как правило, используется для переработки сапролитов, образованных силикатами магния и алюминия с низким содержанием железа (или иногда "переходного" материала); продуктами плавки в электропечи являются ферро-никелевый сплав, поставляемый на рынок (после рафинирования), и утилизируемый шлак. В некоторых случаях в печь восстановления добавляют серу для получения в электропечи штейна с низким содержанием серы, из которого обычным конвертированием получают файнштейн. Технология ISASMELT™ была недавно изучена с точки зрения плавки никелевых латеритных руд этого типа.

Принципиальная технологическая схема основной плавки латеритного никеля с использованием процесса ISASMELT™ показана на рисунке 7. Печь

ISASMELT™ в непрерывном режиме плавит сухое латеритное питание и рециркулируемую пыль. Источником тепла для плавки являются уголь и обогащенный кислородом воздух. В зависимости от перерабатываемой руды может потребоваться добавление флюса. Образующиеся жидкий штейн и шлак периодически выпускают из печи ISASMELT™ в отдельную печь отжига/осаждения шлака через одну летку. Как вариант, печь ISASMELT™ может использоваться для плавки/предварительного восстановления с целью максимального использования энергии топлива, при этом основной объем операции восстановления, а следовательно и образования никелевого сплава, происходит в печи отжига шлака. Другой вариант заключается в том, чтобы ввести стадию прокаливания, например, в печи прокалки перед ISASMELT™. Возможен дожиг отходящих газов ISASMELT™ в верхней части печи (с добавлением, при необходимости, некоторого объема воздуха, обогащенного кислородом) с передачей тепла поступающему питанию. Отходящие газы печи ISASMELT™ направляют в котел-утилизатор для утилизации тепла и обеспыливания с использованием электростатического осадителя, а затем в дымовую трубу. Получаемый жидкий ферро-никелевый сплав периодически выпускают из электропечи в: (i) систему грануляции или (ii) систему литья (в зависимости от спецификаций продукции) или (iii) в ковш для дальнейшего рафинирования, если это необходимо. Печь осаждения также периодически освобождают от отработанного шлака, который утилизируют или используют для регенерации тепла. Вся пыль, собранная системами очистки газов, возвращается в печь ISASMELT™.

Альтернативный вариант этой технологической схемы заключается в добавлении питания, содержащего серу (см. пунктирные линии и технологические блоки на рисунке 6), к общему объему питания печи ISASMELT™ для получения штейна с низким содержанием железа.

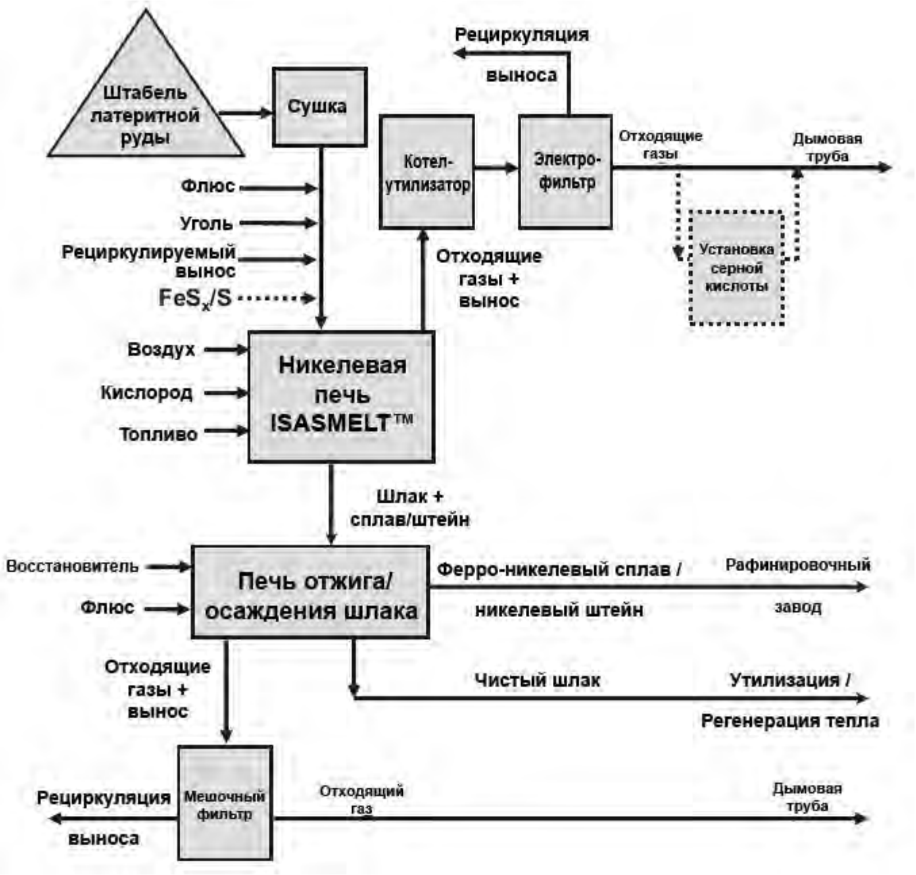


Рисунок 7. Технологическая схема установки ISASMELT™ для плавки латеритного никеля.

8. Испытания пилотной никелевой установки ISASMELT™

Недавно было проведено еще несколько испытаний пилотных установок ISASMELT™ для сбора дополнительных данных относительно возможности применения:

(i) процесса ISASMELT™ для плавки никелевых/медных концентратов с получением либо первичного штейна с высоким содержанием железа, либо файнштейна;

(ii) процесса ISACONVERT™ для конвертирования первичного штейна с высоким содержанием железа в файнштейн; и

(iii) процесса ISASMELT™ для плавки латеритных никелевых руд с получением ферро-никелевого сплава (как пилотные, так и лабораторные испытания).

Основные задачи этих испытаний заключались в том, чтобы определить состав получаемого никелевого штейна/сплава и шлака, коэффициенты распределения элементов между штейном/сплавом и шлаком (в частности, никеля и кобальта), а также получить шлаки для последующих испытаний отжига шлака.

8.1. Пилотная установка ISASMELT™

Пилотная установка ISASMELT™ состоит из стационарной цилиндрической печи с внутренним диаметром (по огнеупору) приблизительно 400 мм и высотой примерно 2000 мм. Печь выложена хромомагнезитовым огнеупорным кирпичом и теплоизоляционным кирпичом с высоким содержанием глинозема. Во время работы печь содержит расплавленную ванну с максимальной глубиной около 600 мм. Контролируемый объем воздуха и кислорода вводят в ванну с помощью фурмы 32 мм (1,25 дюйма) или 38 мм (1,5 дюйма) из нержавеющей стали. Известное количество питания подается на калиброванный ленточный конвейер с изменяемой скоростью ленты, который сбрасывает питание в желоб в верхней части печи, как правило, с расходом от 100 кг/ч до 250 кг/ч. Через фурму подается мазут для поддержания желаемой температуры ванны. В некоторых случаях может также добавляться однородный уголь. Расплавленные продукты могут выгружаться путем открытия одной из двух леток и сбора в чугунные ковши. Отходящие газы процесса охлаждаются и очищаются от выноса печи, а затем поступают в щелочной скруббер для удаления серосодержащих газов перед выпуском в атмосферу.

8.2. Испытания плавки никеля в печи ISASMELT™

В результате плавки никелевых концентратов были получены штейны с различным содержанием железа: от 1,6 % масс. до 20,0% масс. Цель испытаний заключалась в том, чтобы подтвердить гибкость процесса, позволяющую получать из концентрата штейн любого сорта в соответствии с требованиями заказчика - от штейна первичной плавки (содержание железа выше 15% масс.) до файнштейна (содержание железа ниже 2-4% масс.). Типичный состав никелевого/медного концентрата представлен в таблице 1.

Таблица 1 - Состав никелевого/ медного концентрата.

Элемент	Среднее содержание (% масс.)
Ni	17.1
Cu	4.3
Co	0.4
Fe	31.0
S	27.0
SiO_2	9.8
MgO	6.5
Al2O3	1.0
CaO	0.7

Таблица 2 - Состав шлака ISASMELT™.

Элемент	Диапазон (% масс.)
Ni	0.8-7.0
Cu	0.3-1.5
Co	0.2-0.3
Fe	32.0-37.6
SiO2	25.6-34.9
MgO	6.7-10.3
Al2O3	3.8-5.1
CaO	1.4-2.5

Концентрат был гранулирован вместе с заранее определенным количеством кремнеземистого флюса (в соответствии с планируемым составом шлака) и загружен в печь с расходом, равным 150 кг/ч влажного концентрата. Гранулированный уголь подавался на питающий конвейер с расходом, равным 5% расхода концентрата, в качестве дополнительного источника тепла плавки, учитывая коэффициент потери тепла в соответствии с размерами пилотной печи. Воздух и кислород для плавки подавались через отдельные ротаметры при соотношении с выходом 40-45% об. от общего обогащения кислородом. Мазут подавался по фурме для поддержания температуры ванны между 1310°C и 1450°C в зависимости от условий проведения испытаний.

Жидкие шлаки были получены при всех условиях проведения испытаний. Соотношение железа и кремния в шлаке ISASMELT™, полученном при испытаниях, составляло от 0,9 до 1,4. Состав шлаков ISASMELT™, полученных в ходе испытаний, показан в таблице 2.

Коэффициенты распределения никеля, определенные в соответствии формулой (1) ниже, показаны на рисунке 7 в зависимости от содержания железа в массовом штейне как для тестов ISACONVERT™, так и для тестов ISACONVERT™ (будет рассмотрено ниже в разделе, посвященном испытаниям процесса ISACONVERT™). Анализ массового штейна и шлака был выполнен на ложечных пробах, отобранных при выпуске продукта.

$Lx^{s/m}$ = (% масс. x в шлаке)/(% масс. x в штейне) (1)

На рисунке 7 для сравнения представлены лабораторные результаты, описанные в Henao (2003), и результаты, полученные при отборе проб при продувке никель-медного штейна в конвертере Пирса-Смита на заводе Фальконбридж Xstrata Nickel (XNi) (Bustos et al. (1988)) и на заводе Томпсон Vale Inco (Diakow et al. (1975)). В исследовании Энао расплав штейна и шлака был уравновешен при 1500°C или 1600°C в течение определенного периода времени при заданном парциальном давлении кислорода (Po_2) и диоксида серы (Pso_2) (задается соотношениями S_2/SO_2 и CO/CO_2). В конце каждого равновесного теста для анализа отбирались охлажденные образцы штейна и шлака. Данные на заводах Фальконбридж XNi и Томпсон Vale Inco были получены путем отбора проб штейна и шлака в конвертере в конце каждой продувки. Интересно, что все данные на рисунке 7 очень хорошо согласуются.

Коэффициенты распределения кобальта, определенные в соответствии формулой (1), показаны на рисунке 8 в зависимости от содержания железа в штейне как для тестов ISACONVERT™, так и для тестов ISACONVERT™. На рисунке 8 для сравнения представлены лабораторные результаты, описанные в Font (1999), и, в соответствии с рисунком 7, промышленных данные, описанные в Bustos et al. (1988) и Diakow et al. (1975). В исследовании Font (1999) расплав штейна и шлака был уравновешен (как и в более поздней работе Henao (2003), указанной выше) в течение определенного периода времени при заданном парциальном давлении кислорода (Po_2) и диоксида серы (Pso_2) (задается соотношениями S_2/SO_2 и CO/CO_2). В конце каждого теста для анализа отбирались образцы застывшего штейна и шлака.

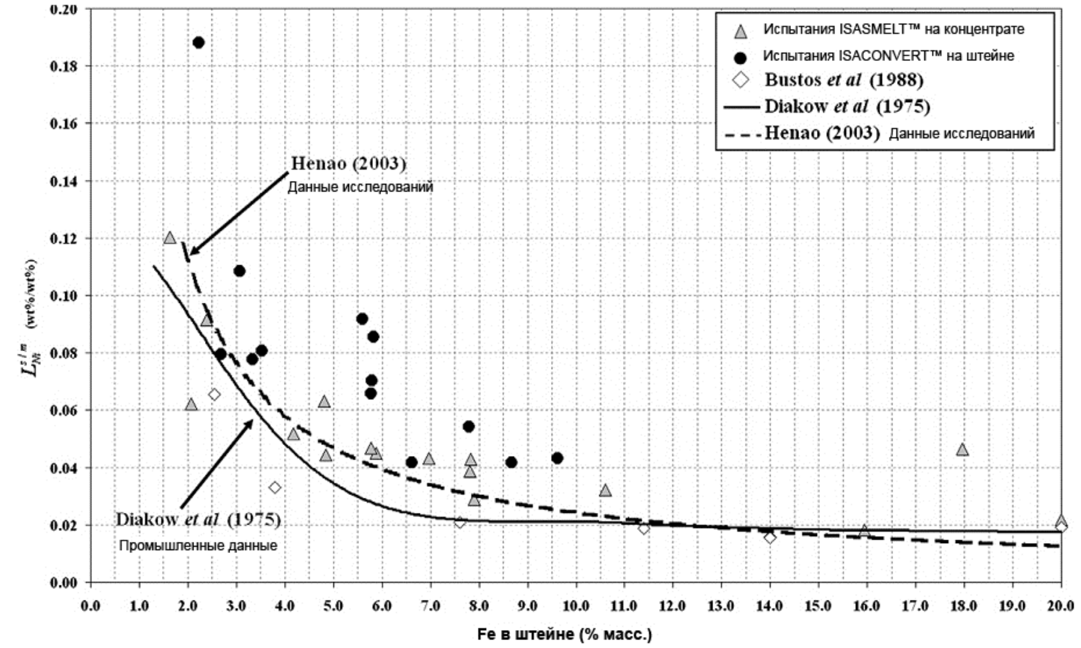


Рисунок 8. Содержание никеля в шлаке в зависимости от содержания железа в штейне ISASMELT™.

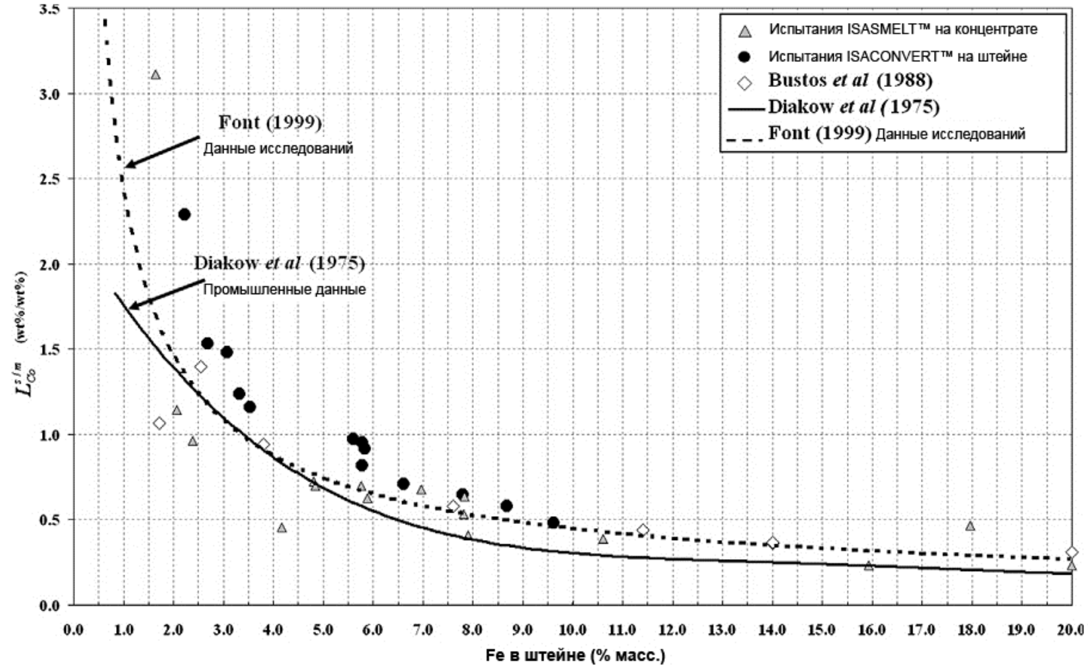


Рисунок 9. Распределение кобальта в зависимости от содержания железа в штейне ISASMELT™.

Данные заводов Фальконбридж XNi и Томпсон Vale Inco были получены тем же методом, что и данные, представленные на рисунке 7. Как и на рисунке 7, интересно отметить, что все данные о распределении кобальта между шлаком и штейном на рисунке 8 очень хорошо согласуются.

Из рисунков 7 и 8 следует, что по мере уменьшения содержания железа в штейне коэффициенты распределения как никеля, так и кобальта между шлаком и штейном увеличиваются. Это, как предполагается, связано с более высоким кислородным потенциалом (Po_2) системы вследствие более низкого содержания железа в штейне. Далее, как отмечено, данные на рисунках 7 и 8 показывают, что результаты полупромышленной эксплуатации как ISASMELT™, так и ISACONVERT™ хорошо согласуются с ранее опубликованными данных по подобным системам штейна и шлака.

Чтобы представить информацию о температуре ликвидуса шлаков, полученных при испытаниях пилотных установок, для оценки фазовых равновесий в шлаковой системе $NiO\text{-}MgO\text{-}FeO\text{-}Fe_2O3\text{-}SiO2\text{-}Al_2O3\text{-}CaO$ был использован пакет FactSage™ (Bale et al. (2002)). Результаты расчетов в FactSage™ в следующих условиях: $Po_2 = 10^{-7.6}$ атм., Al_2O_3 = 4% масс., CaO = 1,5% масс. и MgO = 10% масс. представлены на псевдотроичной диаграмме $NiO\text{-}FeO\text{-}SiO_2$ на рисунке 9. Результаты расчета в FactSage™ ликвидуса и состава шлака при пилотных испытания были представлены графически путем приведения к осям псевдотроичной диаграммы. Можно заметить, что, как и ожидалось, все шлаки, полученные в ходе испытаний, находились в пределах области первичной фазы оливина (p.p.f.).

8.3. Испытания процесса ISACONVERT™ в конвертировании никеля

Цель испытаний конвертирования заключалась в том, что испытать процесс ISACONVERT™ в условиях конвертирования твердого штейна первичной плавки с высоким содержанием железа в файнштейн с низким содержанием железа. Типичный состав штейна первичной плавки представлен в таблице 3. Как будет показано ниже, штейны первичной плавки были успешно конвертированы в конечные файнштейны с содержанием железа от 2,2 до 9,6% масс..

При проведении испытаний штейн с заранее определенным количеством обогащенного кремнеземистого флюса (в соответствии с желаемым соотношением железа и кремнезема в шлаке) загружали в печь с расходом, равным 200 кг/ч необработанного твердого штейна. Гранулированный уголь добавляли к питанию с расходом, равным 5% от расхода штейна, из соображений, описанных выше в разделе о полупромышленной плавке. Воздух и кислород подавались через отдельные ротаметры при соотношении с выходом 35-40% об. от общего обогащения кислородом. Как и при плавке, мазут подавался по фурме для поддержания температуры ванны между 1310°C и 1380°C в зависимости от условий проведения испытаний.

Было установлено, что жидкие шлаки были получены при всех условиях проведения испытаний в диапазоне соотношений железа и кремнезема в шлаке ISACONVERT™ от 1,1 до 2,3. Состав шлаков ISACONVERT™, полученных в ходе испытаний, показан в таблице 4.

Таблица 3 - Состав никелевого штейна первичной плавки.

Элемент	Среднее содержание (% масс.)
Ni	44.5
Cu	9.7
Co	2.9
Fe	25.3
S	17.2
SiO2	0.3

Таблица 4 - Состав шлака ISACONVERT™.

Элемент	Диапазон (% масс.)
Ni	2.2-6.4
Cu	0.6-1.2
Co	1.3-2.5
Fe	37.8-48.0
SiO2	25.6-34.9

Как отмечено выше, содержание никеля в шлаке и коэффициент распределения кобальта между шлаком и штейном при испытаниях ISACONVERT™ показаны на рисунках 7 и 8. Испытания процесса ISACONVERT™ применительно к никелю были направлены на достижение низкого содержания железа в штейне. Как видно на рисунке 7, содержание Ni в шлаке значительно возрастает при снижении содержания Fe в штейне. Учитывая более окисленную среду при низком содержании железа в штейне, шлак в этих условиях имеет более высокий ликвидус. Однако в данном случае Fe/SiO_2 можно скорректировать, чтобы понизить температуру ликвидуса.

Как и при испытаниях плавки, для оценки условий конвертирования и прогнозирования температур ликвидуса в упрощенной шлаковой системе NiO-FeO-Fe2O_3-SiO2-Al_2O_3-CaO-MgO был использован пакет FactSage™ (Bale et al. (2002)). Прогнозы были экстраполированы путем нормализации данных на псевдотроичную систему NiO-FeO-SiO_2 (рисунок 10). В расчетах FactSage™ были использованы следующие условия: $Po_2 = 10^{-7,6}$ атм., Al_2O_3 = 2,5% масс., CaO = 1,5% масс. и MgO = 2,5% масс. Состав конечных шлаков ISACONVERT™ показан на рисунке 10 с расчетным ликвидусом. Как и шлаки ISASMELT™, большинство шлаков ISACONVERT™ находились в пределах области первичной фазы оливина (p.p.f.).

8.4. Испытания плавки латеритного никеля в печи ISASMELT™

Цель испытаний плавки латеритного никеля заключалась в том, чтобы показать возможность получения ферро-никелевого сплава как в лабораторных, так и в полупромышленных условиях. Испытания были проведены на рудах двух латеритных месторождений, имеющих разное географическое расположение. Типичный состав двух латеритных руды, использованных для проведения испытаний, приведен в таблице 5. Эти латеритные руды успешно прошли плавку в восстанавливающих условиях с получением ферро-никелевого сплава и шлака с низким содержанием никеля. При лабораторных испытаниях были получены ферро-никелевые сплавы, содержащие от 15 до 40% масс. Ni, при исходных соотношениях SiO_2/MgO и Fe/Ni 1,2-3,1 и 10-20, соответственно. В то же время, при пилотных испытаниях были получены ферро-никелевые сплавы, содержащие от 60 до 85% масс. Ni, при исходных соотношениях SiO_2/MgO и Fe/Ni 2,0-3,6 и 5-12, соответственно.

Таблица 5 - Состав питания при плавке латеритного никеля.

Элемент	Среднее содержание, лабораторные испытания (% масс.)	Среднее содержание, пилотные испытания (% масс.)
Ni	0.6-1.8	0.3-3.6
Fe	5.0-25	8.4-20.9
SiO_2	22-46	40-46
MgO	12-25	12-22
Al_2O_3	2.5-8.5	0.7-3.0
CaO	0.1-1.1	0.2-2.5

Лабораторные испытания проводились порционно в небольших тиглях (вмещающих около 1 кг расплава). После первоначальной плавки ванну восстанавливали постепенным добавлением угля. В связи с тем, что испытания имели предварительный характер, к руде добавляли флюс - высокосортный кремнезем и/или известняк, - чтобы обеспечить образование жидкого шлака. При этом учитывается, что на практике флюсование обычно не применяется. При проведении пилотных испытаний латеритные руды с добавлением кремнеземистого и известнякового флюса подавали в печь с расходом 150-200 кг руды в час. На конвейер также подавался однородный уголь для восстановления и в качестве источника части необходимой энергии плавки. Оставшуюся часть тепла, необходимого для поддержания ванны в расплавленном состоянии, обеспечивал мазут, подаваемый через фурму.

Было установлено, что жидкие шлаки были получены при всех условиях проведения испытаний; состав шлаков как лабораторных, так и пилотных испытаний приведен в таблице 6.

Тренд весового процентного содержания никеля в шлаке в ходе лабораторных испытаний по мере добавления угля в расплав в тигле показан на рисунке 11. Лабораторные испытания были направлены на получение выбранных сортов ферро-никелевых сплавов при низком содержании никеля в шлаке. Было также установлено, что уровень никеля в шлаке сильно зависит от времени осаждения. Представляется возможным оптимизировать рабочую температуру и расход флюса для получения минимального времени осаждения и максимального извлечения ферро-никелевого сплава.

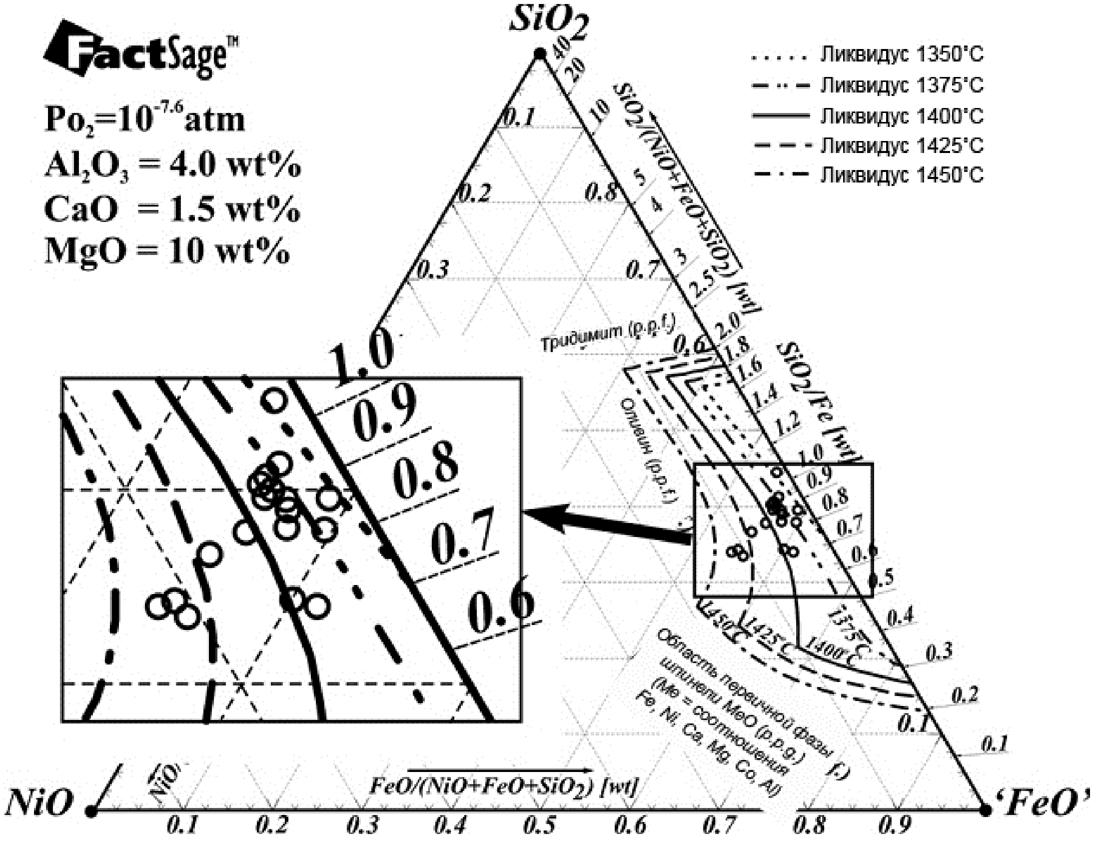


Рисунок 10. Ликвидус в системе NiO-'FeO'-SiO_2-Al_2O3-CaO-MgO при Po_2 = 10 $^{-7.6}$ атм., Al_2O_3 = 4,0% масс., CaO = 1,5% масс. и MgO = 10% масс.

Таблица 6 - Состав шлака латеритного никеля в процессе ISASMELT™.

Элемент	Среднее содержание, лабораторные испытания (% масс.)	Среднее содержание, пилотные испытания (% масс.)
Ni	0.03-0.3	-
Fe	7.1-21	18-25
SiO_2	35-50	44-50
MgO	20-27	14-18
Al_2O_3	5.0-6.9	2.6-3.3
CaO	2.8-8.6	0.2-2.3

Для расчета температур ликвидуса в условиях шлака, образующегося при плавке латеритного никеля, был использован пакет FactSage™ (Bale et al. (2002)). Шлаки были представлены упрощенной системой $FeO-Fe_2O_3-SiO_2-Al_2O_3-CaO-MgO-Cr_2O3$. Из-за низкого содержания оксида никеля в шлаке после восстановления никель не учитывался в расчетах FactSage™. Прогнозы были экстраполированы путем нормализации данных на псевдотроичную систему $MgO-FeO-SiO_2$ (рисунок 12). В расчетах FactSage™ были использованы следующие условия: $Po_2 = 10^{-11,1}$ атм., Al_2O_3 = 4,0% масс., CaO = 0,2% масс. и Cr_2O_3 = 1,0% масс. Состав шлаков ISASMELT™ в условиях лабораторных и пилотных испытаний показан на рисунке 12 в виде серого круга вместе с расчетами ликвидуса в FactSage™. В зависимости от состава и температуры массового шлака, область первичной фазы шлаковой системы шлака в показанных границах может соответствовать пироксену, оливину или тридимиту/кристобалиту (см. рисунок 13).

9. Перспективы развития

Компания Xstrata Technology доказала, что процесс ISASMELT™ является перспективной технологией плавки первичной и вторичной меди и свинца. На медеплавильных и свинцовоплавильных заводах по всему миру в печах ISASMELT™ перерабатывается более 9 млн. тонн питания в год. Особенности, которые делают технологию ISASMELT™ привлекательной для использования в плавке меди и свинца, могут быть также применены при плавке и конвертировании никеля. Процесс ISASMELT™ уже успешно опробован в промышленном масштабе в качестве технологии плавки меди/никеля.

В настоящей работе описаны последние успешные лабораторные и пилотные испытания процессов ISASMELT™ и ISACONVERT™:

(i) испытания процесса ISASMELT™ в качестве технологии получения штейна первичной плавки с высоким содержанием железа и файнштейна непосредственно из концентратов;

(ii) испытания процесса ISACONVERT™ в качестве технологии получения файнштейна из штейна с высоким содержанием железа; и

(iii) испытания процесса ISASMELT™ в качестве технологии получения ферро-никелевых сплавов различных марок из никелевых латеритных руд.

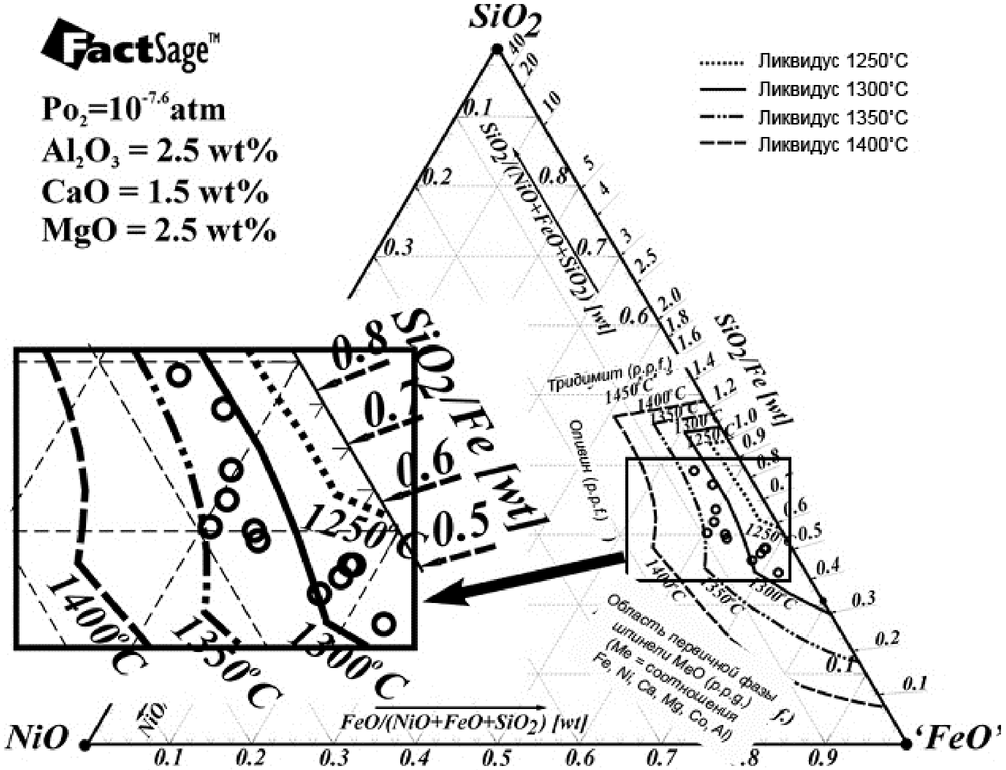


Рисунок 11. Ликвидус в системе NiO-'FeO'-SiO2-Al2O3-CaO-MgO при $Po_2 = 10^{-7,6}$ атм., Al_2O_3 = 2,5% масс., CaO = 1,5% масс. и MgO = 2,5% масс.

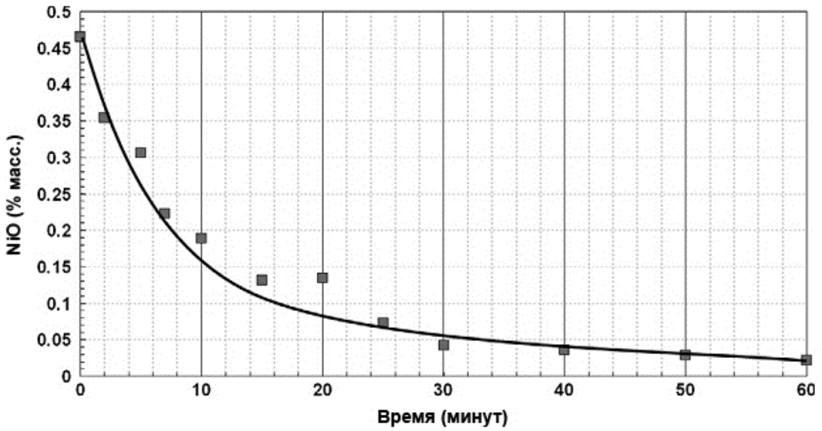


Рисунок 12. Восстановление никеля из шлаковой ванны при лабораторных испытаниях

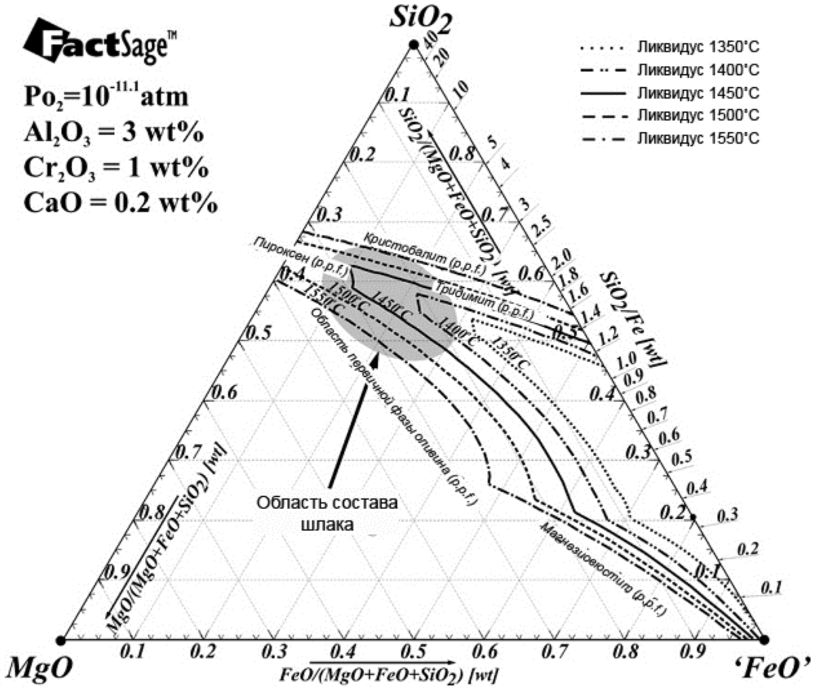


Рисунок 13. Ликвидус в системе FeO-Fe_2O_3-SiO_2-Al_2O_3-CaO-MgO-Cr_2O_3 при P_{O2}=10 $^{-11.1}$ атм., Al_2O_3 = 4,0% масс., CaO = 0,2% масс. и Cr_2O_3 = 1,0% масс.

Для дальнейшего развития процесса ISASMELT™ для никелевой промышленности могут потребоваться улучшения существующей конструкции печи, в частности:

(i) Охлаждение печи - эксплуатация при температурах выше 1300°C требует наличия систем охлаждения печи, обеспечивающих достаточно долгую продолжительность кампании;

(ii) Блоки выпускного отверстия - для выпуска никелевого штейна/сплава необходимо доработать блоки выпускного отверстия;

(iii) Пневматическая подача - применение сухого способа подачи концентрата через фурму ISASMELT™, чтобы свести к минимуму потребность в топливе; и

(iv) Оптимизация процесса отжига шлака ISASMELT™ для максимального извлечения ценных металлов.

В настоящее время ведутся работы, необходимые для внесения этих усовершенствований.

Благодарности

Авторы выражают признательность компании Xstrata Technology за разрешение опубликовать настоящую статью.

Список использованной литературы

Arthur, P.S., Hunt, S.P., 2005. ISASMELT™ - 25 Years of Continuous Evolution. In: Nilmani, M., Rankin, W.J. (Eds.), John Floyd International Symposium on Sustainable Developments of Metals Processing, NCS Associates, Melbourne, Australia, pp. 73-94.

Bakker, M.L., Alvear, G.R.F., Kreuh, M., 2009. ISASMELT TSL Making a splash for nickel. In: The 48th Annual Conference of Metallurgists 2009. Pyrometallurgy of Nickel and Cobalt, Sudbury, Canada, 2009.

Bale, C.W., Chartrand, P., Decterov, S.A., Eriksson, G., Hack, K., Mahfoud, R.B., Melanqon, J., Pelton, A.D., Petersen, S., 2002. CALPHAD 26, 189-228.

Burford, B., 2009. The ISASMELT™ Technology Package: Over 30 Years of Innovation, AusIMM Bulletin, No. 1, February 2009, pp. 26-30.

Bustos, A.A., Ip, S.W., O'Connell, G., Kaiura, G.H., Toguri, J.M., 1988. In: Tyroler, G.P., Landolt, С.Л. (Eds.), Converting Simulation at Falconbridge Limited, Extractive Metallurgy of Nickel and Cobalt, The Metallurgical Society, Inc., AIME, USA, pp. 335-354.

Diakow, J.S., Mak, Y.F., Oit, R.G., 1975. Metallurgy of the Converting Process in the Thompson Smelter, Paper Presented to the 14th Annual Conference of Metallurgists, Edmonton, Alberta, August 1975, The CIM, Montreal, Quebec, Canada, 1975.

Donald, J.R., Warner, A.E.M., Dalvi, A.D., Scholey, K., Ross, A„ Harris, C, 2005. In: Ross, A„ Warner, T., Scholey, K., (Eds.), Risks and Opportunities in Continuous Converting for Nickel at Inco, Converter and Fire Refining Practices, The Minerals, Metals & Materials Society, USA, 2005, pp. 295-307.

Edwards, J.S., Alvear, G.R.F., 2007. Converting using ISASMELT™ Technology, Copper 07, vol. Ill, Book 2: The Carlos Diaz Symposium on Pyrometallurgy. In: Warner, A.E.M., Newman, C.J., Vahed, A., George, D.B., Mackey, P.J., Warczok, A. (Eds.), MetSoc, Toronto, CA, 2007, CIM, Montreal, 2007, vol. 3, Book 2, pp. 17- 28.

Font, J.M., 1999. Phase equilibrium and minor elements distribution between slag and matte phases in nickel smelting, Ph.D. Thesis, Tohoku University, 1999.

Henao, H.M.Z.,2003. Phase equilibrium between Ni-S or Ni-Fe melt and slag in nickel smelting, Ph.D. Thesis, Tohoku University, 2003.

Makinen, T„ Taskinen, P., 2006. The State of the Art in Nickel Smelting: Direct Outokumpu Nickel Technology. In: Kongoli F., Reddy, R.G., (Eds.), Sohn International Symposium Advanced Processing of Metals and Materials, The Minerals, Metals & Materials Society, USA, 2006, pp. 313-325.

Nikolic, S„ Edwards, J.S., Burrows, A.S., Alvear, G.R.F., 2009. 1SACONVERT™ - TSL Continuous Copper Converting Update. In: Kapusta, J. and Warner, T. (Eds.), In: International Peirce-Smith Converting Centennial: Held During TMS 2009 Annual Meeting and Exhibition. The Minerals, Metals & Materials Society, USA, February 15-19, pp. 407-414.

Victorovich, G.S., 1993. Oxygen Flash Converting for Production of Copper, Extractive Metallurgy of Copper, Nickel and Cobalt. In: Reddy, R.G., Weizenbach, R.N., (Eds.), In: Proceedings of the Paul E. Queneau International Symposium. The Metallurgical Society, Inc., AIME, USA, 1993, vol. 1, pp. 501- 529.

Warner, A.E.M., Diaz, C.M., 2003. An Overview of the Metallurgy of Nickel-Copper Matte Converting. In: Kongoli, F., Itagaki, K., Yamauchi, C., Sohn, H.Y., (Eds.), Yazawa International Symposium Metallurgical and Materials Processing: Principles and Technologies. The Minerals, Metals & Materials Society, USA, 2003, pp. 113-129.

Warner, A.E.M., Diaz, C.M., Dalvi, A.D., Mackey, P.J., Tarasov, A.V., 2006. JOM World Nonferrous Smelter Survey Part III: Nickel: Laterite, JOM, April 2006, pp. 11-20.

Warner, A.F.M., Diaz, C.M., Dalvi, A.D., Mackey, P.J., Tarasov, A.V., Jones, R.T., 2007. |OM World Nonferrous Smelter Survey Part IV: Nickel: Sulfide, JOM, April 2007, pp. 58-72.

Свинец-Цинк 2010
Под редакцией: А.Зигмунда, Л. Центомо, Г. Гинен, Н. Пирет, Г. Ричардс и Р. Стефенс
TMS (Ассоциация Металлов, Минералов и Материалов), 2010 г

ТЕХНОЛОГИЯ ISASMELT ™ ДЛЯ ВТОРИЧНОЙ ПЕРЕРАБОТКИ СВИНЦА

*Билл Эрингтон[1], Питер Хокинз[2], Эндрю Лим

[1]*Xstrata Technology*
Level 4, 307 Queen Street
Brisbane, Australia 4000
(*Ответственный автор:berrington@xstratatech.com)

[2]*Xstrata Zinc,*
Botany Road
Northfleet,
Кент, Соединенное Королевство.

[3]*Metal Reclamation (Industries) SDN, BHD,*
Pulau Indah Industrial Park
West Port
Селангор, Малазия

РЕФЕРАТ

ISASMELT™ является хорошо зарекомендовавшей себя технологией плавки первичных медных и свинцовых концентратов. Менее известно её применение для вторичной переработки отработанных свинцовых аккумуляторов. На сегодняшний день построено две установки ISASMELT™ для вторичной переработки свинцовых аккумуляторов с получением мягкого свинца с низким содержанием сурьмы плюс свинцово-сурьмянистый шлак, из которого можно производить свинцово-сурьмянистый сплав и силикатный шлак с низким содержанием свинца. Утилизация серы происходит либо путем использования известкового скруббера, либо путем десульфуризации пасты до начала процесса плавки. Опыт, полученный при эксплуатации этих установок, был использован для установки большего масштаба (>300 000 т/г).

ВВЕДЕНИЕ

Технология ISASMELT™ в настоящее время является хорошо зарекомендовавшей себя стандартной технологией плавки первичных медных концентратов [1]. На сегодняшний день эксплуатируется семь установок для переработки первичной меди, а еще три установки находится на стадии строительства. На трех действующих установках перерабатывается свыше 1 000 000 тонн медных концентратов в год, а максимальная производительность, достигнутая на сегодняшний день, составляет примерно 1 400 000 тонн концентратов в год.

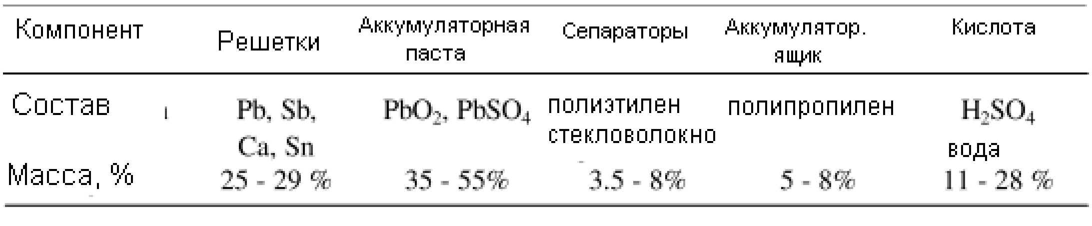
Все большее распространение получает технология первичной плавки свинца, с использованием которой в эксплуатации находится одна свинцовая установка производительностью 80 000 т/г [2], а еще две находятся на стадии строительства. Кроме того, технология ISASMELT™ является основой системы Kayser для вторичной переработки меди на производстве Люнен, Германия, и на заводе Umicore Precious Metals, Хобокен, Бельгия [3].

Применение технологии ISASMELT™ для вторичной переработки свинцовых аккумуляторов менее известно. В данной статье обсуждается разработка технологии, использование установок на производствах Britannia Refined Metals в Великобритании и Metal Reclamation Industries в Малайзии, и возможное применение технологии для большой установки по вторичной переработке свинца (>300,000 т металла в год).

РАЗРАБОТКА ПРОЦЕССА ВТОРИЧНОЙ ПЕРЕРАБОТКИ СВИНЦА

Автомобильные и промышленные свинцово-кислотные аккумуляторы, вместе с аккумуляторами электровелосипедов, составляющих быстро растущий компонент сырьевых запасов в Азии, являются основными источниками сырья для заводов по вторичной переработке свинца. Обычно аккумуляторы дробят и разделяют на компоненты перед дальнейшей переработкой. В таблице 1 показаны основные компоненты свинцово-кислотного аккумулятора, из которых нельзя повторно перерабатывать только сепараторы.

Таблица 1 - Компоненты свинцово-кислотного аккумулятора



Компонент	Решетки	Аккумуляторная паста	Сепараторы	Аккумулятор. ящик	Кислота
Состав	Pb, Sb, Ca, Sn	PbO_2, $PbSO_4$	полиэтилен стекловолокно	полипропилен	H_2SO_4 вода
Масса, %	25 - 29 %	35 - 55%	3.5 - 8%	5 - 8%	11 - 28 %

Разработка процесса ISASMELT™ для вторичной переработки свинца была сконцентрирована на проектировании наиболее эффективного процесса для переработки аккумуляторной пасты и решеток. С этой целью, на производстве Mount Isa, Австралия были проведены пилотные испытания с использованием пилотной установки ISASMELT™ производительностью 250 кг. Процесс, который был разработан и, в конечном итоге, запатентован, подразумевал применение одной печи для прямой плавки при низких температурах, в результате которого напрямую производится относительно чистый мягкий свинец (<0.05% Sb) из аккумуляторной пасты с использованием шлака с высоким содержанием свинца в качестве химически активного среды. Примеси в пасте, такие как сурьма и кремнезем, постепенно накапливаются в шлаке. При достижении определенного уровня примесей в шлаке, шлак выпускается с целью последующей переработки для производства свинцового/сурьмянистого сплава. Шлаковая ванная с высоким содержанием свинца затем восстанавливается в печи, далее продолжается производство мягкого свинца.

Помимо пилотных испытаний, был проведен ряд экспериментов с использованием тигельной плавки в Организации по научным и производственным исследованиям Австралии, Мельбурн, для лучшего понимания разделения металлов и кинетики восстановления [4].

Оказалось, что процесс легко управляемый, полностью герметичный для улавливания свинцовых возгонов с высокой производительностью в относительно небольшой печи. Положительные результаты пилотных испытаний придали компании MIM (в настоящее время Xstrata) достаточно уверенности в применении технологии ISASMELT™ на её дочернем производстве Britannia Refined Metals (BRM) в Нортфлит, Великобритания.

УСТАНОВКА ДЛЯ ВТОРИЧНОЙ ПЕРЕРАБОТКИ СВИНЦА НА ЗАВОДЕ BRM

Введение

Основной деятельностью завода Britannia Refined Metals (BRM) является рафинирование первичного свинца, получаемого на заводе Mount Isa, Австралия. Однако до 2004 года на этой же площадке находился рафинировочный завод по переработке вторичного свинца. До 1991 года на рафинировочном заводе производили 10 000 тонн в год свинца с использованием короткой вращающейся печи. В 1991 году BRM провели модернизацию завода для производства 30 000 тонн в год рафинированного свинца и свинцовых сплавов. Эта реконструкция позволила BRM удовлетворить их требования по сокращению эксплутационных затрат, снижению выбросов свинца и уменьшению количества конечного шлака, требующего утилизации.

Выбранная технология основывается на следующих процессах:

- Механическое дробление и разделка аккумуляторов и десульфуризация пасты с использованием процесса CX компании Engitec;
- Плавка аккумуляторной пасты и решеток в печи ISASMELT™ для производства мягкого свинца и свинцово-сурьмянистого шлака;
- Использование действующей вращающейся печи для восстановления шлака с целью производства свинцово/сурьмянистого сплава и конечного шлака.

Схема производства BRM по переработке вторичного свинца показана на рисунке 1.

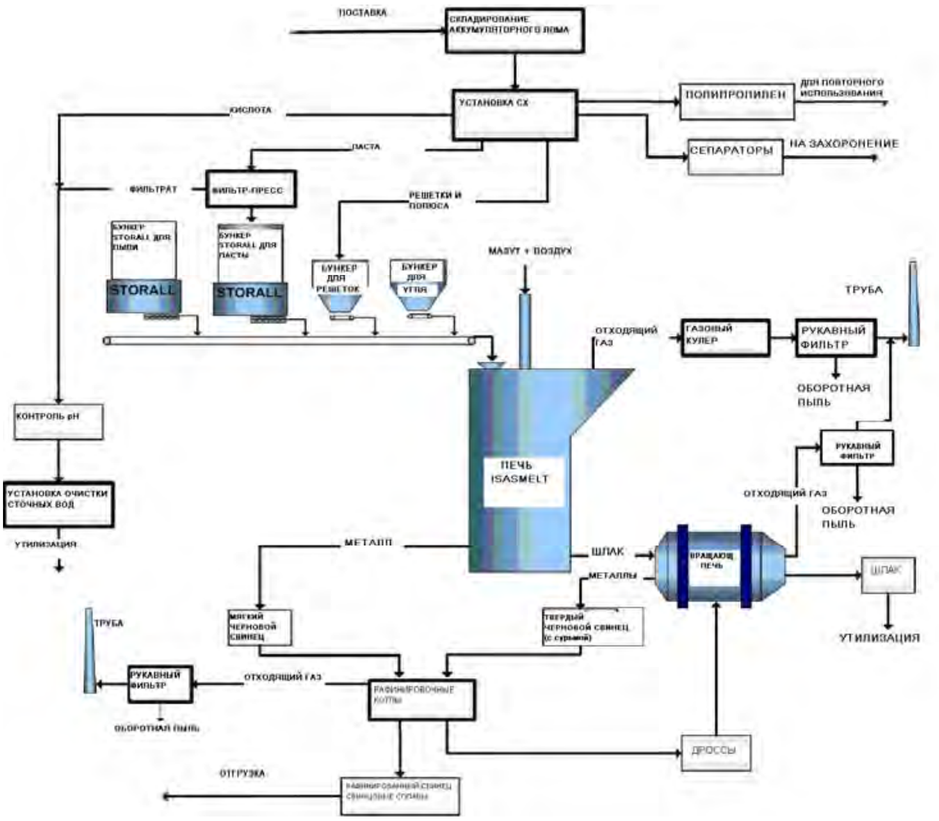


Рисунок 1 - Технологическая схема завода Britannia Refined Metals по переработке вторичного свинца

Дробление и разделение отработанных аккумуляторов

Система СХ компании Engitec была спроектирована для дробления и переработки целых автомобильных аккумуляторов с неслитым электролитом. Перед первичным дроблением в молотковой мельнице, свободная кислота, полученная из хранилища отработанных аккумуляторов, собиралась, фильтровалась и нейтрализовалась до передачи на очистные сооружения. Дробленные аккумуляторные материалы подвергались мокрому грохочению для отделения пасты от металлических и других компонентов. Паста проходила через грохот с размером ячеек 0.6 мм в сборный резервуар для сгущения, а затем посредством дозирующего транспортера в емкости для десульфуризации. Свинец металлический, материал корпуса и сепараторы загружались в двухстадийный гидродинамический сепаратор (осаждение/флотирование), из которого каждую фракцию можно получать отдельно. Полипропилен продавали, тогда как металлические решетки либо переплавлялись в печи ISASMELT™ либо во вращающейся печи.

Десульфуризация пасты

Паста откачивалась из бака-накопителя в любой из двух резервуаров для десульфуризации. Раствор гидроксида натрия (50%) смешивался с пастой, в результате реакции образовывался оксид свинца и раствор сульфата натрия. Оксид свинца затем извлекался путем фильтрации с использованием фильтр-пресса и доставлялся во вращающийся бункер Storall, из которого паста извлекалась шнековым конвейером и загружалась при контролируемой скорости в печь ISASMELT™ посредством конвейерных весов.

Работа печи ISASMELT™

Печь ISASMELT™ на заводе BRM представляет собой вертикальный цилиндр с огнеупорной футеровкой, с внутренним диаметром примерно 1,8 м, с отверстиями в своде для подачи шихты, для фурмы и отвода отходящих газов. Мягкий свинец и свинцовый шлак с высоким содержанием сурьмы выпускают из печи с использованием одного выпускного отверстия и комбинации поворотного и опрокидывающегося желоба, чтобы направлять продукты в один из трех ковшей, расположенных на рельсах под выпускной площадкой.

Аккумуляторная паста и материалы решетки обычно перерабатывались отдельными кампаниями для упрощения последующего процесса рафинирования. В цикле переработки пасты, сначала загружалась аккумуляторная паста, для образования начальной ванны. Затем паста загружалась в печь вместе с добавлением кокса или угля в качестве восстановителя. Добавление восстановителя рассчитывалось с целью плавки всей пасты в металл, без учета первоначально загруженной пасты. Основная часть сурьмы, кремнезема, железа и других неосновных компонентов переходила в шлаковую фазу. Свинец в пасте восстанавливался для получения ванны мягкого свинца с низким содержанием сурьмы (0.01-0.1%), который выпускался из печи через определенные промежутки времени в ковши, из которых он транспортировался в расплавленном состоянии в рафинировочные котлы. Рабочая температура в печи составляла примерно 810 °C.

Производство мягкого свинца продолжалось до тех пор, пока в печь не загружалось около 150 тонн пасты. К этому моменту содержание сурьмы в шлаке достигало очень высокого уровня при содержании оксида свинца 55-65%. Хотя этот шлак можно было восстанавливать непосредственно в печи ISASMELT™ для получения свинцового/сурьмянистого сплава, пропускная способность ISASMELT™ была максимально увеличена посредством использования вращающейся печи для восстановления свинцового шлака, и этот процесс с двумя реакторами стал обычной практикой работы. Во вращающейся печи также перерабатывали дроссы из рафинировочных котлов. На рисунке 2 показано распределение свинца в процессе.

Печи ISASMELT™ обычно работают с использованием дутья, обогащенного кислородом. На производстве BRM установка была спроектирована для переработки 7,7 тонн пасты в час без обогащения кислородом, но обычно на ней перерабатывали 12 тонн пасты в час. Решетки переплавлялись в ходе отдельной компании при скорости загрузки до 35 тонн в час для производства мягкого свинца (но с большим содержанием сурьмы, чем в цикле переработки пасты). Эту пропускную способность можно было бы удвоить путем обогащения воздуха кислородом до 30%.

В ранние годы работы производительность была ограничена задержками, связанными с трудностями транспортировки шихты, а также с требованиями техобслуживания установки CX для дробления аккумуляторов. После модернизации конвейерной системы и усовершенствования установки CX проектные характеристики печи ISASMELT™ были превышены в 1995 году, и развитее производства продолжалось до 2004 года, когда Xstrata Zinc приняла решение о прекращении деятельности по переработке вторичного свинца.

Существовало много положительных аспектов производства, включая значительные усовершенствования, которые были достигнуты в отношении эксплуатационных характеристик огнеупорной футеровки печи, несмотря на агрессивный характер шлака, содержащий оксиды свинца. Полная замена огнеупорной кладки требовалась только после производства 60-70,000 тонн свинца, хотя частичный ремонт был необходим после производства примерно 20-22,000 тонн свинца. Простота и эксплуатационная надежность процесса была продемонстрирована тем, что после периода пусконаладочных работ и усовершенствования некоторого оборудования количество персонала для обслуживания производства по вторичной переработке свинца было сокращено с поддержанием такого же объема производства свинца. Бесперебойному производству значительно способствовало использование двух бункеров-питателей Storall (поставляемых компанией Mitchell Engineering), которые использовались для подачи контролируемых количеств пасты и пыли, смешанной с водой, в печь ISASMELT™. Затраты на топливо сократились в результате использования регенерированной нефти вместо дистиллята для фурмы и природного газа для горелки поддержания температуры.

Отрицательным аспектом вторичного производства была невозможность производить пасту с низким содержанием натрия после десульфуризации, несмотря на использование NaOH в качестве реагента. Остаточный натрий в пасте (до 1,5%) приводил к образованию двухслойного шлака в печи ISASMELT™, который состоял из шлака с содержанием окиси свинца в пределах 55-85% и натриевого шлака низкой плотности, содержащего до 35% натрия. Несмотря на присущие трудности данной ситуации, операторы научились справляться с этим путем использования отдельного верхнего выпускного отверстия для натриевого шлака. Сульфат натрия периодически выпускался из печи для обеспечения постоянного контроля расплава свинцового шлака и процесса плавки мягкого свинца.

Содержание натрия в пасте можно было бы значительно сократить путем вложения инвестиций в процесс промывки и фильтрации пасты, но это в свою очередь, повлекло бы за собой необходимость инвестиций в очистные сооружения.

Более подробное описание завода BRM представлено в статье авторами Ramus и Hawkins [5].

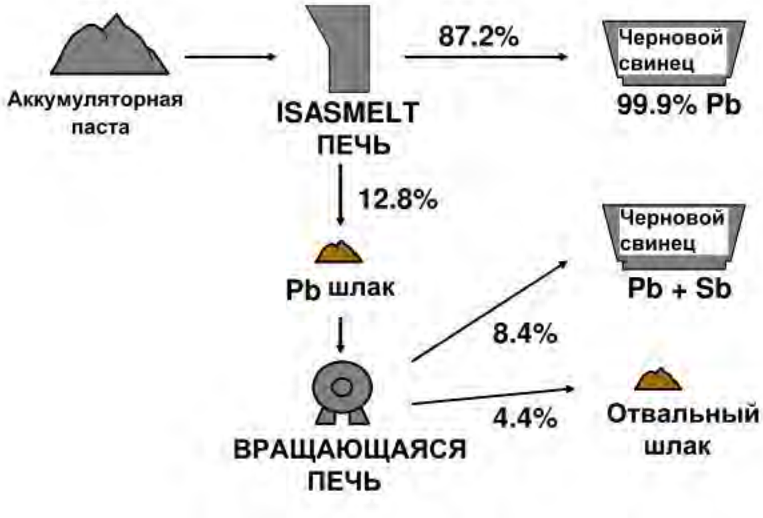


Рисунок 2 - Распределение свинца в процессе на заводе BRM

ЗАВОД ПО ВТОРИЧНОЙ ПЕРЕРАБОТКЕ СВИНЦА КОМПАНИИ MRI

Введение

Компания Metal Reclamation (Industries) Sdn. Bhd. (MRI) эксплуатирует завод по вторичной переработке свинца в Малайзии с 1972 года. В 1995 году технология основывалась на использовании короткой вращающейся печи, в которой производили около 1 000 тонн в месяц рафинированного свинца и свинцовых сплавов. Было ясно, что такой объем производства не будет соответствовать требованиям рынка, и таким образом, был подготовлен долгосрочный план с целью увеличения мощности и соответствия рыночным требованиям. С этой целью, компания MRI приобрела землю за пределами города Куала-Лумпур для переноса и расширения своего предприятия. MRI выбрала технологию ISASMELT™ как наиболее соответствующую технологию, которая позволит им соблюдать любые будущие требования по охране окружающей среды в отношении выбросов серы и утилизации шлака.

Описание завода

Завод MRI спроектирован для производства примерно 40,000 тонн в год рафинированного свинца и свинцовых сплавов из загружаемых материалов, содержащих свыше 70,000 тонн в год аккумуляторного лома. Аккумуляторы дробятся и разделяются в дробилке местного производства, мощностью 40 т/ч. Пластмассовые компоненты из дробленных аккумуляторов упаковываются и продаются для повторной переработки. Аккумуляторная паста смешивается с оборотной пылью и доставляется в бункер Storall с целью контролируемой загрузки в печь ISASMELT™. Решетки загружаются из бункера с использованием ленточного питателя с регулируемой скоростью. Дополнительные бункеры-питатели используются для загрузки угля, дроссов и флюсов при контролируемой скорости в печь ISASMELT™.

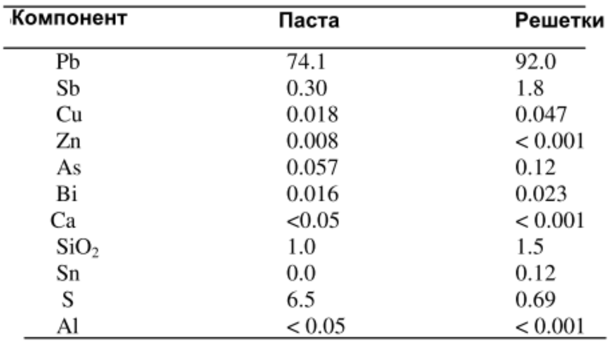
Печь ISASMELT™ аналогична печи BRM, но её внутренний диаметр составляет примерно 2.5 м. Мягкий свинец, производимый в печи, периодически выпускается в один из двух ковшей вместимостью 50 тонн, расположенных рядом с печью. В ковше поддерживается слой холодного свинца в нижней части для охлаждения поступающего свинца. Охлажденный свинец затем перекачивается в рафинировочные котлы. Шлак из печи выпускается через специальное отверстие для выпуска шлака, гранулируется и обезвоживается с использованием системы грануляции в чане с мешалкой компании Paul Wurth.

Отходящие газы из печи ISASMELT™ охлаждаются на двух стадиях испарительного охлаждения, очищаются с использованием рукавного фильтра и затем промываются в десульфуризаторе топочных газов Chiyoda (ДТГ). В ДТГ газ вдувается в воду, образуя тонкий слой пузырьков, в котором SO_2 абсорбируется, окисляется вдуваемым воздухом и затем нейтрализуется с использованием измельченной известковой пульпы. Полученный гипс обезвоживается на ленточном фильтре и продается.

Описание работы печи ISASMELT™

Печь ISASMELT™ была спроектирована для переработки примерно 31,000 тонн в год аккумуляторной пасты и 19,000 тонн в год решеток, имеющих состав, показанный в таблице 2.

Таблица 2 - Состав загружаемых материалов MRI



Компонент	Паста	Решетки
Pb	74.1	92.0
Sb	0.30	1.8
Cu	0.018	0.047
Zn	0.008	< 0.001
As	0.057	0.12
Bi	0.016	0.023
Ca	<0.05	< 0.001
SiO_2	1.0	1.5
Sn	0.0	0.12
S	6.5	0.69
Al	< 0.05	< 0.001

Первоначальная схема технологического процесса включала следующие этапы:

- Плавка аккумуляторной пасты для образования начальной ванны
- Плавка смеси пасты, решеток и оборотной пыли плюс уголь для получения мягкого свинца
- Периодический выпуск мягкого свинца с интервалом в несколько часов
- Примерно через 10-12 часов, завершение цикла путем выпуска и грануляции большей части шлака с высоким содержанием свинца, в котором также присутствует сурьма и другие примеси. Шлак складируется для дальнейшей переработки
- Повторное образование начальной ванны и плавка

Первоначальная схема технологического процесса предполагала достаточно времени для восстановления складированного шлака кампаниями. Этапы восстановления включали загрузку шлака в печь при непрерывных восстановительных условиях для получения примерно 15-20 % свинца в шлаке по мере добавления железных (вторичная окалина) и известковых флюсов. По достижению достаточной глубины расплава, начиналось периодическое восстановление с целью получения <1% свинца в конечном шлаке.

На рисунке 3 показано распределение свинца на основе данного подхода. Происходит значительное восстановление конечного шлака в сравнении с вращающейся печью. Однако, в настоящее время, MRI не имеет возможности сбыта этого шлака и, таким образом, она переходит на контролируемую утилизацию.

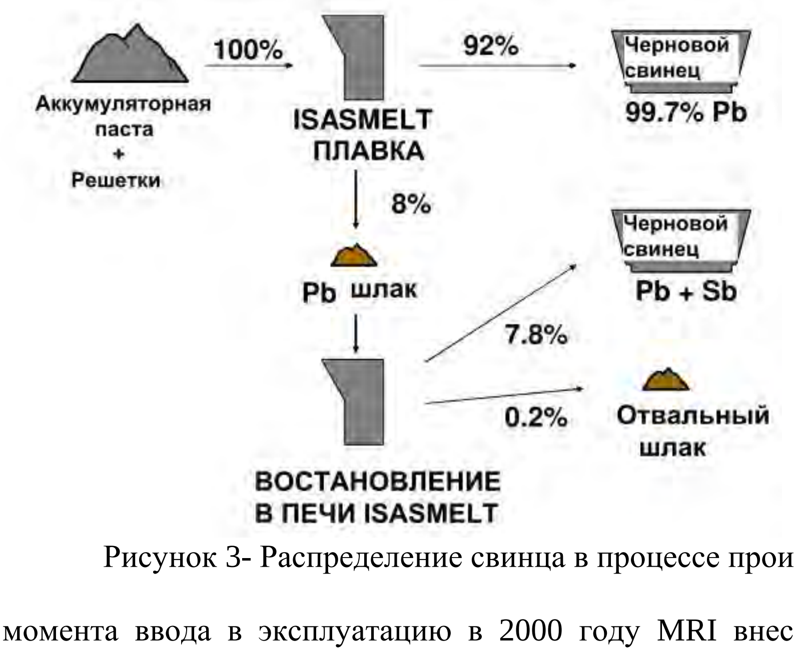


Рисунок 3- Распределение свинца в процессе производства MRI

С момента ввода в эксплуатацию в 2000 году MRI внесла много изменений и улучшений в производство. Они посчитали более экономичным проводить периодическое восстановление в конце каждого цикла плавки, таким образом, избегая необходимость складировать и переплавлять шлак с высоким содержанием свинца.

MRI продемонстрировала гибкость их установки ISASMELT™ путем её использования для плавки первичных свинцовых концентратов (в пределах их промывной мощности), когда это стало выгодным в соответствии с рыночными условиями. В процессе плавки первичных свинцовых концентратов MRI показала, что концентраты можно использовать для частичного восстановления шлака с высоким содержанием свинца с последующим добавлением угля для завершения восстановления.

Дополнительные модификации, сделанные MRI, также включали использование до 20 тонн кислорода в сутки для увеличения производительности установки. Последним усовершенствованием было преобразование установки с целью использования природного газа в качестве топлива вместо мазута, что приводит к значительной экономии затрат.

В целом, в сравнении с производством Britannia Refined Metals, MRI продемонстрировала преимущества процесса промывки газа в сравнении с десульфуризацией пасты. Однако это преимущество зависит от наличия рынка для гипса, производимого в процессе. В настоящее время большая часть гипса, производимая на заводе MRI, продается цементной промышленности.

Фотография завода MRI показана на рисунке 4. Технологическая схема завода MRI показана на рисунке 5.

Рисунок 4 - Завод по вторичной переработке свинца MRI

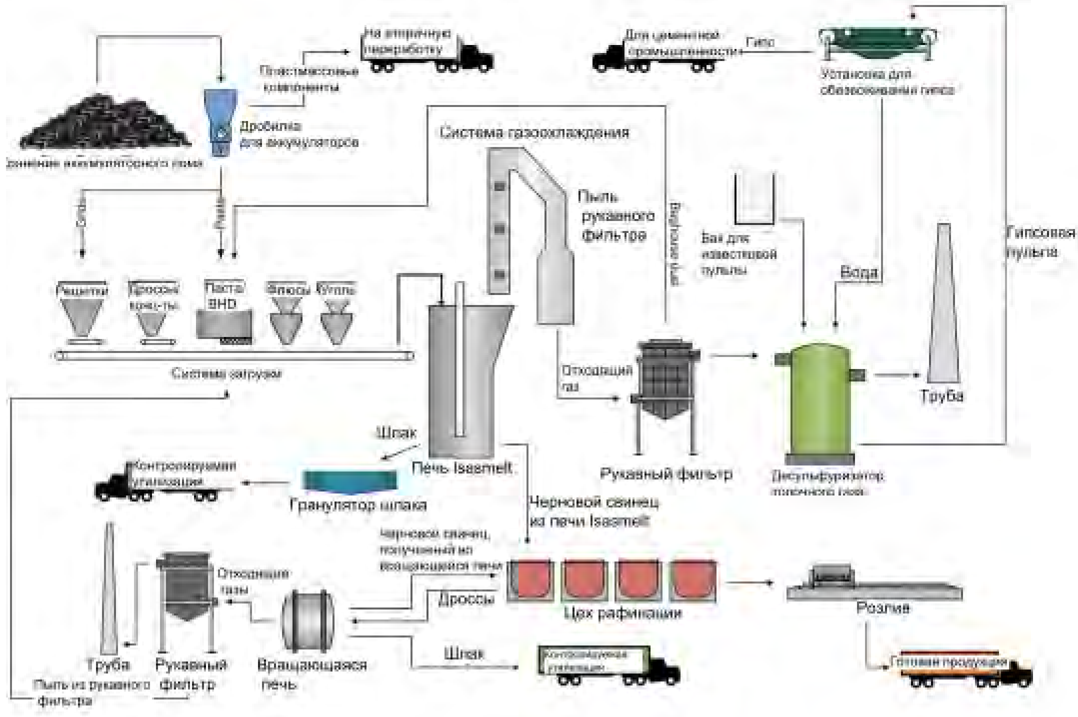
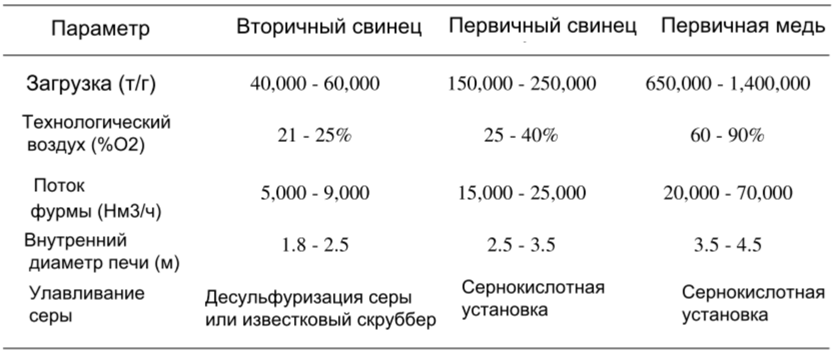


Рисунок 5 - Технологическая схема производства Metal Reclamation Industries

УСТАНОВКА ДЛЯ ВТОРИЧНОЙ ПЕРЕРАБОТКИ СВИНЦА МОЩНОСТЬЮ 300,000 ТОНН В ГОД

В таблице 3 показано сравнение некоторых эксплуатационных и проектных параметров для действующих установок ISASMELT™. Следует отметить большие различия в эксплуатационных параметрах и относительно небольшие размеры печи, требуемые для приспособления к различным технологическим схемам.

Таблица 3 - Сравнение параметров установки ISASMELT™



Параметр	Вторичный свинец	Первичный свинец	Первичная медь
Загрузка (т/г)	40,000 - 60,000	150,000 - 250,000	650,000 - 1,400,000
Технологический воздух (%O2)	21 - 25%	25 - 40%	60 - 90%
Поток фурмы (Нм3/ч)	5,000 - 9,000	15,000 - 25,000	20,000 - 70,000
Внутренний диаметр печи (м)	1.8 - 2.5	2.5 - 3.5	3.5 - 4.5
Улавливание серы	Десульфуризация серы или известковый скруббер	Сернокислотная установка	Сернокислотная установка

Как видно из таблицы, производительность установок по переработке вторичного свинца относительно небольшая в сравнении с установками по переработке первичного свинца. Это отражает характер промышленности по переработке вторичного свинца. Несмотря на то, что вторичный свинец составляет более 50% от ежегодного производства свинца, данное направление деятельности рационально применяется только на относительно крупных заводах в Северной Америке и Европе, при этом самым крупным заводом по вторичной переработке является завод Doe Run Buick в Миссури, с объемом производства примерно 145,000 тонн в год. В других странах имеется много мелких производителей. Например, в Индии зарегистрировано свыше 240 заводов по переработке вторичного свинца, и примерно 100 заводов зарегистрировано в Китае. Кроме того, в большинстве развивающихся стран существует, в основном, неконтролируемая неофициальная отрасль по переработке вторичного свинца, которая конкурирует с контролируемой промышленностью. Дополнительным важным фактором является Базельская Конвенция, запрещающая транспортировку за границу опасных отходов, включая аккумуляторный лом. Например, Китай официально не импортирует аккумуляторный лом из других стран.

Однако когда мы смотрим в будущее, мы можем ожидать ужесточение экологических стандартов и вытекающее из этого закрытие мелких производств, сопровождающееся увеличением требований к переработке вторичного свинца. Например, в Китае ожидается увеличение производства вторичного свинца вдвое, которое превысит 2, 000,000 тонн свинца в год к 2015 году [6]. В соответствии с этим сценарием и при повышении внимания к сокращению опасных выбросов доминирующими станут крупные перерабатывающие заводы, на которых используется современная технология, как например ISASMELT™.

На рисунке 6 показан пример технологической схемы для печи ISASMELT™, в которой можно производить свыше 300,000 тонн в год мягкого свинца из сырья, аналогичного по составу сырью, показанному в таблице 2.

Основные характеристика данного производства следующие:

- Обогащение технологического дутьевого воздуха кислородом до 40%
- Непрерывное производство мягкого свинца, содержащего 0,2% Sb или меньше
- Низкотемпературный процесс (<850°C) в результате переработки низкоплавкого шлака $PbO\text{-}Sb_2O_3$.
- Относительно небольшая печь (внутренний диаметр 3,0-3,5 м)
- Производство отходящих газов с относительно высоким содержанием SO2, подходящих для производства серной кислоты.

Шлак с высоким содержанием свинца, получаемый в процессе, можно либо складировать и перерабатывать кампаниями в печи ISASMELT™, либо передавать в расплавленном состоянии во вторую небольшую (<2 м внутренний диаметр) печь ISASMELT™ для извлечения свинца и получением отвального шлака с низким содержанием свинца плюс свинцово/сурьмянистый сплав.

Следует отметить, что, несмотря на относительно большой объем производства мягкого свинца, эта установка относительно небольшая в сравнении с установками первичной плавки, где потоки фурмы составляют менее одной пятой от максимального объема, используемого в процессе плавки первичной меди, а требование по объему печи составляет примерно половину от объема крупных медных установок.

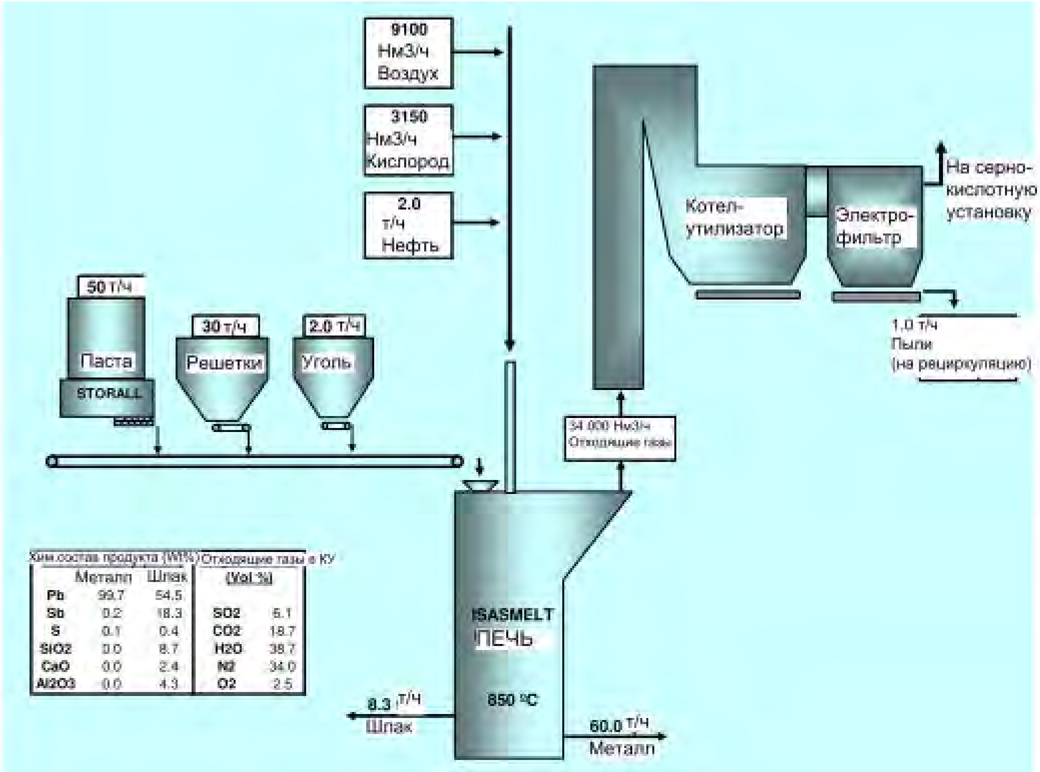


Рисунок 6 - Технологическая схема для установки ISASMELT™ для переработки вторичного свинца производительностью 300,000 тонн в год

В заключение, применение технологии ISASMELT™ для производства вторичного свинца оказалось очень положительным со следующими преимуществами, которые продемонстрировала технология в сравнении с работой традиционной вращающейся печью:

- Прямое производство мягкого свинца и свинцово-сурьмянистого сплава, что дает гибкость смешивания
- Возможность получения нерастворимых силикатных отвальных шлаков с низким содержанием свинца
- Хорошие гигиенические условия процесса благодаря герметичному реактору, работающему под разрежением
- Работа одной печи, допускающая любые возможные увеличения производительности в будущем

СПРАВОЧНАЯ ЛИТЕРАТУРА

1. G. Alvear, P. Arthur and P. Partington, “Feasibility to Profitability with Copper ISASMELT™ ”, Proceedings of Copper 2010, Hamburg, Germany, June 2010, GDMB.

2. B. Errington, P. Arthur, J. Wang and Y. Dong, “The ISA-YMG Lead Smelting Process”, Proceedings of the International Symposium on Lead and Zinc Processing, Kyoto, Japan October 2005, T. Fujisawa et al., Eds., MMIJ, pp. 581-599.

3. F. Vanbellen and M. Chintinne, “The Precious Art of Metals Recycling”, Advanced Processing of Metals and Materials, F. Kongoli and R.G. Reddy, Eds., TMS, Warrendale, Pennsylvania, 2006, Vol. 1, 43-52.

4. S. Wright, S. Jahanshahi and W.J. Errington, “Reduction Kinetics of Slags produced from Recycling of Lead Batteries”, Pyrometallurgy for Complex Materials and Wastes, Melbourne June 1994, The Minerals, Metals and Materials Society, pp 121-132.

5 K. Ramus and P. Hawkins, “Lead/acid Battery Recycling and the new ISASMELT™ Process”, Journal of Power Sources, 42, (1993), pp 299-313.

6. C. Zhang and R. Zhang, “Current Status and Outlook on Chinese Secondary Lead Industry”, International Secondary Lead Conference, Macau, August/September 2009, Paper 1.3.

驰 宏 锌 锗
CHIHONG Zn & Ge
驰骋天下・宏图高远

产品商标

CHIHONG Zn & Ge

云南驰宏锌锗股份有限公司
YUNNAN CHIHONG ZINC&GERMANIUM CO., LTD

公司简介

云南驰宏锌锗股份有限公司成立于2000年7月18日，总部位于云南省曲靖市经济技术开发区，公司主要从事铅、锌、锗系列产品的开采、加工、进出口业务和国内贸易。具备探矿、采矿、选矿、冶炼、化工和科研为一体化的生产经营配套能力。公司（代码：600497）于2004年4月20日在上海证券挂牌上市交易。

公司的主要发起人云南会泽铅锌矿，始建于1951年1月5日，是我国第一个五年计划156个重点建设项目之一，也是国内第一个采用烟化法富集技术处理低品位共生难选矿，能同时处理铅锌氧化矿和硫化矿的企业，是中国最早从氧化铅锌矿中提取锗的企业。

公司采用了先进，成熟的技术工艺和大型自动化的设备，粗铅引进澳大利亚艾萨炉富氧吹熔炼技术，解决了二氧化硫对环境的污染问题。锌冶炼采用国内最大的沸腾焙烧炉和世界先进的自动剥锌技术，整个生产系统运用自动控制网络系统进行管理。公司引进、使用X射线荧光分析仪、光电直读光谱分析仪、电感耦合等离子体发射光谱分析仪、原子吸收光谱分析仪、原子荧光分析仪等大型分析仪器检测设备，严格保证了产品质量。

公司铅、锌、锗系列产品覆盖全国十余个省市，远销美、德、日、韩、澳等国家，锗产品产量和质量居全国同行业之首；白银即将完成在LBMA的注册，公司铅锌产品均为英国伦敦金属交易所（LME）注册产品，同时公司还是伦敦金属交易所批准的中国第一家，也是目前唯一一家锌检测机构。

公司于1998年通过ISO9002质量标准体系认证，2007年通过了中国质量认证中心（CQC）审核，取得了质量、环境、职业健康安全“三标一体”化体系认证。

公司属云南冶金集团控股企业，是集团重点发展的五大板块之一——铅锌板块。在云南省内会泽、曲靖、昭通、建水、澜沧、龙陵，省外黑龙江大兴安岭、内蒙古呼伦贝尔、四川宁南和国外澳大利亚、加拿大等地拥有矿山和控股公司。公司当前具有年采选矿石80万吨、电锌18万吨、电铅10万吨、锗产品30吨、白银150吨、黄金120千克、硫酸32万吨的综合生产能力，至“十二五”末，公司产能将达到年产锌70万吨、铅30万吨、白银400吨、黄金800千克、锗40吨、铟30吨、镉800吨、铋300吨、硫酸130万吨的规模。届时，公司将跨国跨区域形成多个生产基地。当前“银鑫”牌锌锭、“金沙”牌铅锭、“银晶”牌锗系列产品、“华达”牌硫酸已在市场上获得较高的知名度和美誉度，为进一步实施品牌战略，公司将自2010年起将不同产地产品统一使用“驰宏锌锗”商标(▲)。

展望未来，公司将秉承“科学发展、和谐共荣的发展理念；市场为先、理性繁荣的市场理念；善待自然、和谐发展的环保理念”，把公司建设成为环保、高效、可持续发展的现代化企业。

"Extreme makeover ": UPMR's Hoboken plant

Francis Vanbellen and Mathias Chintinne

Umicore Precious Metals Refining
A. Greinerstraat 14
B-2660 Hoboken, Belgium

Abstract

For technology-based companies, continuous improvement has become as natural as evolution itself. On top of this, innovative breakthroughs are necessary to keep a high performance in an ever more rapidly changing world. This paper describes the complete turn-around of Umicore's Hoboken plant, from a complex Pb-Cu-Ni concentrate smelter and refiner, to the world's largest recycling facility for precious metals. This turn-around was realised through a series of drastic improvements, which will be remembered at Umicore Precious Metals Refining as Hoboken's "extreme makeover".

The makeover allowed for treatment of complex recycled materials from the world's largest open-pit mine: the consumer world. The European environmental legislation is based on a life-cycle approach, assessing the sustainability of products through the cycle of production, use and recycling. The WEEE Directive, the Battery Directive are examples of the new legislation, setting out collection and recycling targets for valuable scattered waste streams.

Based on a long tradition in complex non-ferrous smelting, UPMR masters the complexity of many precious metals containing waste streams. The new flow sheet has great flexibility in terms of physical aspect of the recycled streams; it is able to treat dry, fine, coarse, wet or lumpy materials, making it perfectly suited for recycling. Apart from these technical realisations, UPMR's strategy is to sustain its business by being a reliable partner for its customers, socially responsible towards the community and caring for the environment, making "Materials for a better life" sound more true every day.

1. The open-pit mine of recycled consumer goods

European environmental legislation

For a long time, environmental policies have been focused on large point sources of pollution, such as industrial emissions and household waste. These policies have contributed to reduce pollution and encouraged the use of 'clean' technologies. However, smaller and more scattered waste streams are not treated in a specific way. As an example, in the EU in 2003, 90% of the waste of electrical and electronic equipment (WEEE) was landfilled or incinerated. To address the challenge of the scattered waste streams, an Integrated Product Policy (IPP) has been developed as a guideline for future environmental legislation.

This IPP is based on the concept of Life-Cycle thinking. The environmental impact of a product is a sum of the impacts during production, use and recycling. Each part of the life-cycle has an impact on the environment, but also determines the impact in subsequent phases of the life-cycle. For example, the design of a product will strongly determine the dismantling after use. The IPP ensures that environmental impacts are addressed in that part of the life-cycle where they must be solved, and not just shifted from one part of the life-cycle to another.

This new approach implies that recycling is not just the responsibility of recyclers. The producer's responsibility for his product reaches until the end of the life-cycle. This has led to the concept of Extended Producer Responsibility (EPR). The producer is responsible for the recycling, by organizing take-back of his discarded products. By this policy instrument, product groups such as packaging waste or WEEE are diverted from the household waste streams, and treated in an environmentally sound way, such as re-use or recycling. The extra cost for the take-back and recycling will initially be transferred to the consumer, but can be minimized by design improvements and material choice, thus making the recyclability a competitive advantage. Policy makers can influence the market by granting a label to excellent environmental performers, again creating competitive advantage.

European Directives, such as the WEEE Directive or the Battery Directive are based on the abovementioned principles. In the Directives, objectives are defined in the form of collection rates and recycling rates of the targeted waste streams. For example, PC's should have a recovery rate (including heat recovery) of 75% by weight, and re-use and recycling of components or materials of 65% by weight.

The open-pit mine for precious metals

Precious metals are being used for their beauty and nobility (jewellery), but more and more for their technical properties. Examples are excellent electric conductivity (electronics) and interesting catalytic properties (car catalysts, petrochemical catalysts). A catalyst promotes chemical reactions, so that a same result is realised with less energy usage. The development of catalytic processes, often with precious metals, helps creating more sustainable technologies.

Thanks to their limited reactivity, they are endlessly recyclable, without loss of mass or quality. Precious metals recovered from recycled waste streams have exactly the same properties as those originating from primary production (mining). There is one big difference though: recycling of precious metals uses 10 times less energy than the primary production ! In the context of sustainability, recycling of precious metals should be maximised. A reason for this favourable energy balance, is the relative richness of "ores" such as waste electronic scrap en spent catalyst compared with natural resources.

2. Hoboken's Makeover: 10 Years Of Innovation

Flow Sheet before 1997

Historically, the Hoboken plant has been specializing in the treatment of complex Pb-Cu concentrates. The old flow sheet is shown in figure 1. In the roasting step, a primary desulphurisation of the sulphidic concentrates is carried out. In the sinter plant, the concentrate is agglomerated to a mechanically strong porous sinter, ready to be fed in the Pb blast furnace. The sinter is then reduced to Pb bullion, speiss, matte and slag, all of which need further processing. The Lead Refinery produces pure Pb, Sb, Sn and Bi. The speiss containing Ni and As is purified in the Cu electro-winning plant. The slag is suited for disposal after treatment in the electric furnace. Finally, the matte is fed to a second blast furnace (the Cu blast furnace), producing a rich Cu matte. The blister from the Hoboken convertors is taken to the Cu-refinery (electrolysis), and the Cu-containing converter slag is recycled internally.

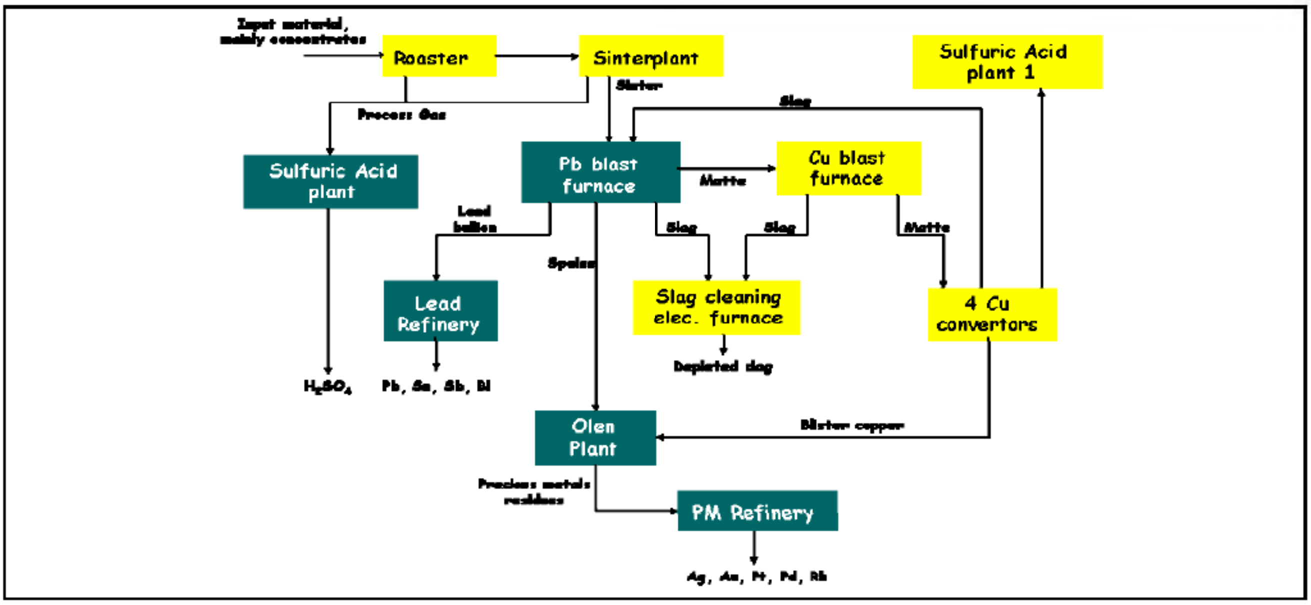


Figure 1: The Hoboken flow sheet until 1997 focused on complex PbCu concentrates

The Makeover

In 1995, the strategic decision was taken to innovate the processes in order to treat a more diversified feedstock. The core of this technological quantum leap was the new PbCu smelting process. This process uses submerged lance technology; fuel and enriched air are injected into the bath. A novel two-step batch process was developed in-house to separate the feed and form a Pb-rich slag and a Cu bullion. The Smelter is able to cope with all forms of feed material, be it wet or dry, fine or lumpy, making it perfectly suited for the processing of recycled materials from various sources, without extensive pre-processing of the feed. The closed vessel allows for good gas cleaning and optimal work floor hygiene. The Smelter replaced following installations: Roasting plant, Sinter plant, 1 of 2 Sulphuric Acid plants, Cu blast furnace and 4 Hoboken Converters. The result on the plant lay-out (and cleanliness) is spectacular:

Figure 2: Aerial view of the Hoboken plant (2003) after clean-up of old installations

In the same period 1995 - 1998, the precious metals (PM) refinery was revamped using in-house developed technologies and processes for flexible refining of all kinds of precious metals mixes. A completely new Leaching and Electro-winning (LEW) installation was commissioned in 2003, bringing this technology in-house and reducing throughput significantly. In the LEW, the Cu bullion is leached to produce Cu sulphate. The precious metals do not leach and are collected in the tank house slimes, which are subsequently treated in the PM refinery. The Cu sulphate is electrolytically deposited, resulting in 99.5% pure Cu cathodes.

The last breakthrough (until now) was realized in 2003-2004, through the integration of Precious Metals Group (PMG), the former Degussa Dmc², which specialized in precious-metals added value products. Integration of both operations offers a high added value output for UPMR, and a powerful recycling service to the former PMG and its customers.

Today's Flow Sheet

Today's flow sheet illustrates UPMR's integrated approach towards the recycling of precious metals: mastering complexity without giving up on flexibility.

The Smelter acts as an entrance gate to the flow sheet for most of the feed. Recycled materials, such as industrial by-products and consumer products, are processed and concentrated in 2 liquid phases: a Pb silicate slag and a Cu bullion. The aim is to collect precious metals (Ag, Au, PGM's) in a Cu bullion and impurities (As, Sb, Sn, Ni, Bi) in a Pb silicate slag. The feed is thus separated upfront in the flow sheet into a precious metals stream (Cu as collector) and a base metal stream (Pb and Ni as collectors).

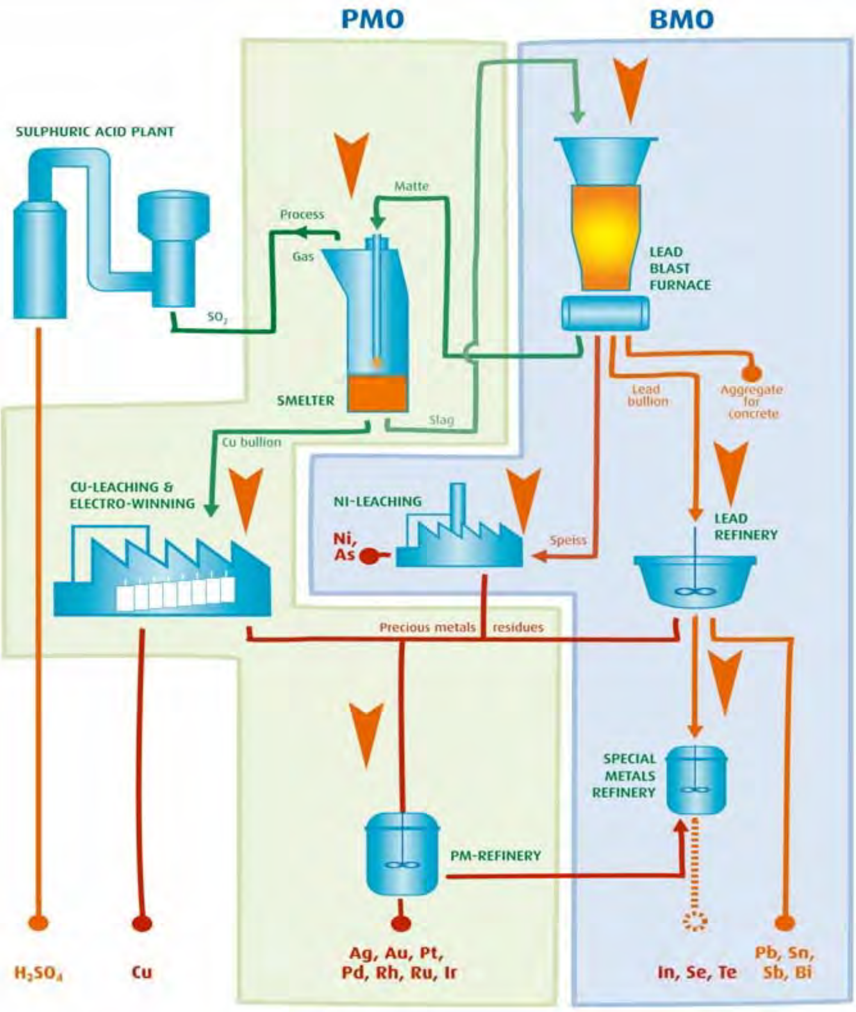


Figure 3: UPMR's flowsheet, highlighting the two major operational lines: Precious Metals Operations and Base Metals Operations

Consequently, the plant has been divided in two major operational lines: Precious Metals Operations (PMO) focuses on the rapid processing of precious-metals bearing product streams, from the feed to the pure metal. Base Metals Operations (BMO) flexibly processes the by-products of the PMO, taking care of the major impurities, while recovering the last part of the precious metals to realize an excellent overall precious metals yield.

The PMO stream is composed of Smelter, Leaching and Electro-winning (LEW) and PM refining. The Smelter uses a unique in-house developed two-step batch process to treat the complex feedstock. In the first step, a Cu matte and Pb silicate slag are produced. The slag is fed to the blast furnace, while the matte remains in the Smelter and undergoes a converting step, producing a Cu bullion and a Cu oxide slag. The gas cleaning of the Smelter is state of the art. Halogens and Mercury are collected in the wet gas cleaning while SO2 is transformed to H2SO4 in the acid plant. The gas cleaning includes post combustion, heat recovery, fast adiabatic quenching, dust removal in a five field hot electrostatic precipitator, quenching, venturi washing, water condensation, wet electrostatic precipitation, star cooling, wet electrostatic precipitation and acid production in the acid plant. The acid plant has been adapted with extra heat exchangers to cope with changing off gas compositions encountered when operating in batch mode.

The Cu bullion from the Smelter is granulated and refined in the LEW. Pure Cu cathodes are produced, while the precious metals and impurities are collected in the tank house residues (slimes). These are subsequently processed in the PM refining, realizing the separation of Ag, Au, Pt, Pd, Rh, Ru and Ir.

The BMO stream is composed of Blast Furnace, Lead Refinery and Special Metals Refinery. The Pb silicate slag from the Smelter is reduced in the blast furnace to form a Pb bullion. The Blast Furnace slag is environmentally clean and sold as an aggregate for construction, for example in concrete applications. The Blast Furnace matte is returned to the Smelter, while the Blast Furnace speiss (an arsenic matte, the name “matte” being reserved for sulphur mattes) is leached. The precious metals containing leaching residue is treated in the PM refinery. The Pb bullion also contains precious metals. In the Lead Refinery, pure Pb is produced as well as sodium antimonate. A precious metals concentrate is sent to the PM refinery, whereas the special metals (In, Se, Te) are further refined in the special metals refinery.

3. Mastering Complexity

Complexity

Precious metals containing materials are an interesting recyclable, because they have a high value per tonne of recycled material . Unfortunately for the gold rushers looking for easy money, the first law of the Universe remains applicable: “No pain, no gain”. In industrial by-products as well as in consumer products, the precious metals tend to be associated with less attractive elements: As, Sb, Sn, Se, Te, In, Cd, Hg, Bi. Apart from causing health hazards if not dealt with properly, the separation of these impurities from the precious metals requires an appropriate flow sheet. Traditional non ferrous smelters and refiners consider these impurities as pollutants for their flow sheet, because they tend to build up and contaminate the valuable productions. UPMR being an integrated smelter, the abovementioned impurities are dealt with as part of the overall process. They are either converted into useful products (sodium antimonate, indium, selenium and tellurium) or disposed of in a safe and controlled way (arsenic, cadmium and mercury).

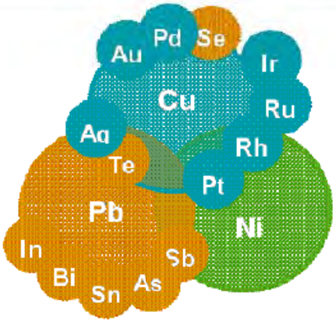


Figure 4: The complex mix of precious metals and impurities in a Pb-Cu-Ni flow sheet

The feed consists mainly of recycled materials, such as industrial by-products and consumer products. They show a great variability in analysis: per year, more than 5000 different lots enter the plant. The industrial by-products are (or should be) somewhat predictable, but the consumer products are more likely to be "catch of the day".

Apart from the analysis, the physical aspect can be just anything, dry lumps, slurries, plastic boards or dry powders. Some examples are shown in figure 5. Recycling processes, requiring an extensive pre-processing of the feedstock are vulnerable compared to flexible processes. The Smelter has the great advantage to enable treatment of all types of feed, ranging from 0 to 95% humidity. Plastics are used as a reducing agent and replace valuable reductants such as coke and fuel. The gas cleaning is designed to take care of the volatile impurities associated with the burning of plastics (VOC's, dioxins).

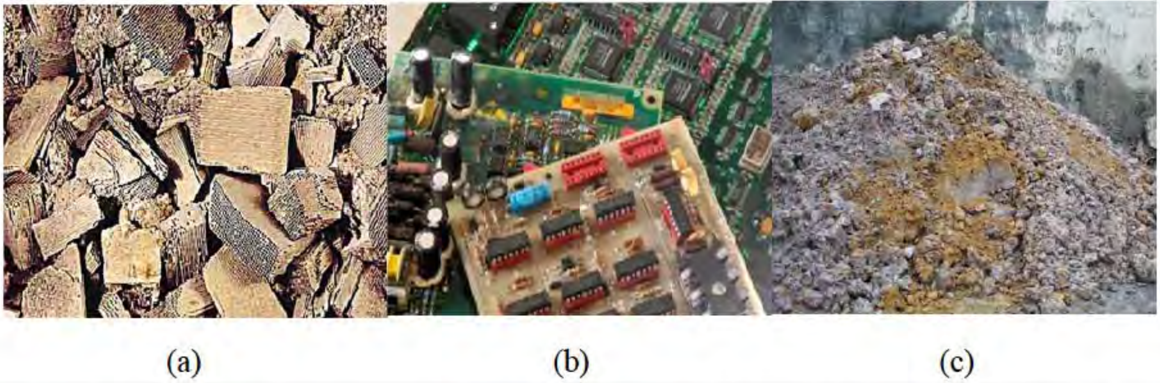

(a) (b) (c)

Figure 5: Some of the precious metals containing recyclables: (a) car catalysts, (b) electronic scrap and (c) zinc leaching residue

Conclusion: the mix of challenging chemical elements (Ni, Pb, As, Te, Sn, Cd), together with a large variance in analysis and physical aspect, make up the “complexity” a precious metals recycler has to deal with.

4. Ensuring reliability

UPMR is recognized as a world player, not only for the quality of its technologies and products, but also for the outstanding service level and business ethics. Sampling and commercial services are key success factors for sustainable precious metals recycling.

Sampling and Assaying

Sampling and assaying is of crucial importance for the recycling job. An accurate determination of the composition of the incoming materials is the basis for the refining agreement with the customer. Especially in our business, a reliable recycling partner is ‘precious’ ! Also, knowledge of the exact composition of the feed allows to tune the process and define the optimal processing route for every feed material.

The sampling facilities in Hoboken are state-of-the-art and able to treat all raw materials. Dedicated installations exist for e-scrap, automotive catalysts, fine free flowing, metallic, wet or lumpy materials. Customer-specific demands are satisfied by an “A la carte” approach, in which a sampling procedure is developed in close collaboration with the material supplier. Today, more than 150 different sampling procedures exist and are continuously improved, with regards to minimization of human intervention and shortening of lead times.

Figure 6: UPMR's dedicated sampling procedure for e-scrap

UPMR's analytical laboratory is internationally recognized in the field of precious metals and controls the assaying process. Many analytical techniques are available for precious metals and trace metals analysis as well as environmental analysis techniques (e.g. for dioxins).

Metals Management

UPMR offers a complete range of commercial and financing services for an optimal match with the customer. The logistic service provides for transport of raw materials and refined products, and assists the customer in taking care of the environmental and administrative regulations, as set out by the Basel Convention or the OECD (for Europe). On the commercial level, risk management solutions are offered. Other services such as leasing, swaps, VAT regulation or optimal metal valorisation can be tailored for every customer individually.

5. Business Sustainability

Being "able to sustain" the endless business of precious metals recycling requires a long-term vision, based on following core competencies: Master Complexity, Foster Flexibility and Ensure Reliability.

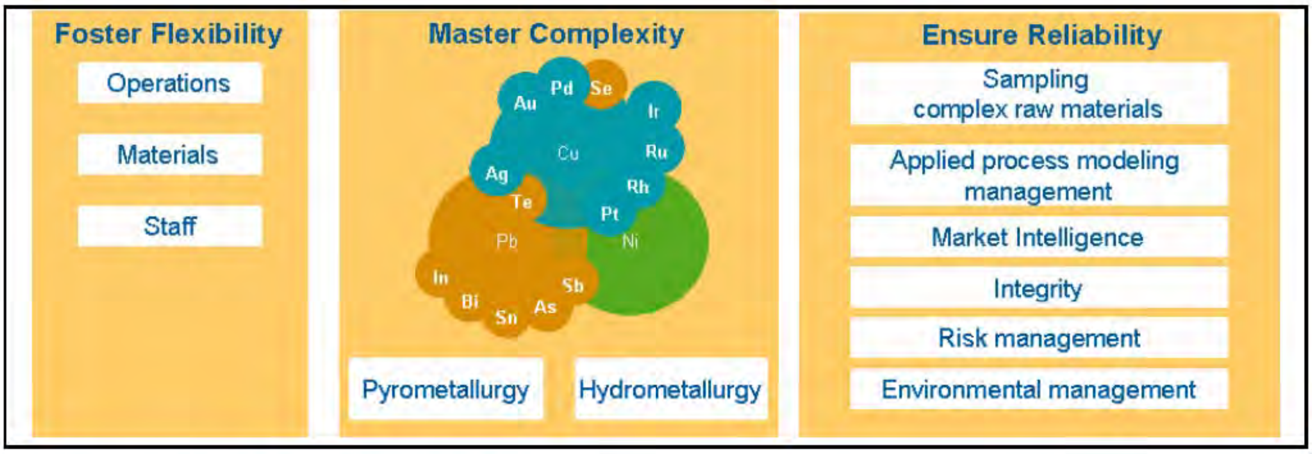


Figure 7: UPMR's core competencies

CONVERTING USING ISASMELT^{TM} TECHNOLOGY

J.S. Edwards and G.R.F. Alvear F.
xstrata technology
Level 4, 307 Queen Street
Brisbane, Queensland 4000
Australia
jedwards@xstratatech.com.au

ABSTRACT

The ISASMELT^{TM} process is recognized worldwide as a modern, flexible, environmentally-friendly and low capital cost smelting technology. Due to the inherent simplicity, and the advanced process control that has been developed, personnel become competent within weeks of being introduced to the ISASMELT^{TM} technology. This compares with the years required to become proficient with alternative technologies. Copper smelters in Australia, USA, India, China, Peru and Zambia have installed ISASMELT^{TM} technology under licence from Xstrata Technology.

In the secondary copper business, plants in Belgium and Germany are processing copper and/or lead materials in ISASMELT^{TM} furnaces in a two stage process. In a single furnace smelting is followed by converting. Both plants have exceeded design capacities and are continuously improving their performance.

The next evolutionary step for the technology is the implementation of the continuous copper converting process on a commercial scale.

The present work describes the design status of the continuous converting process - ISACONVERT^{TM}.

INTRODUCTION

Over the last 25 years the ISASMELT^{TM} process has been developed into a process suited to various roles in the non-ferrous smelting industry including primary and secondary copper smelting, and primary and secondary lead smelting. Furnace capacities have increased from a feed rate of 12 tph of concentrate on the first copper demonstration plant to a feed rate of 180 tph at Mt Isa, or more than 300,000 tonne per annum of cathode copper equivalent from the Vedanta furnace in India. Development of the process continues for different feed qualities, higher throughputs and different processes.

The rate of adoption of the ISASMELT^{TM} furnace has been highest in the primary copper smelting industry with five furnaces in operation, one furnace due to be commissioned in 2007 and detail engineering commencing on two furnaces due for commissioning in 2009. Based on the survey of operating copper smelters in 2003 [1] the fraction of the world primary copper produced in ISASMELT^{TM} furnaces will have increased to approximately 12 % when the Southern Peru Copper Corporation furnace reaches full production in 2007.

The ISASMELT^{TM} furnace is also suited to continuous and batch converting of copper matte to blister copper. Continuous converting has been performed successfully in ISASMELT^{TM} pilot plants; blister copper was produced using a calcium ferrite slag with a matte feed rate of 250 kg/h. Batch converting in the ISASMELT^{TM} furnace is performed by two smelters in Europe. Umicore Precious Metals in Hoboken, Belgium, has been using the ISASMELT^{TM} furnace for converting since 1997, and in 2002 NA AG, Lünen, Germany, commissioned the second ISASMELT^{TM} furnace to be used for converting.

BATCH CONVERTING

Copper Converting at Umicore

Umicore is a speciality materials group with activities centred in Precious Metal Services, Precious Metal Products and Catalysts, Advanced Materials and Zinc Specialities [2] . The company, under its former name Metallurgie Hoboken Overpelt (MHO), initially developed the Hoboken converter, but during the 1990s came to realise that it could not meet the strict environmental regulations coming into force in the 21st century. As a result, Umicore elected to collaborate with Mount Isa Mines (now Xstrata) on further developing the ISASMELT^{TM} process to smelt and convert mixed copper feed materials.

A demonstration plant was designed by Xstrata and operated for some months at the Hoboken smelter site. Following successful operation of the demonstration plant a full scale commercial ISASMELT^{TM} furnace was designed and constructed. The commercial smelter was commissioned at the end of 1997 and is now treating up to 300,000 tonnes per year of secondary feed materials. Figure 1 shows a general view of the ISASMELT^{TM} plant at Hoboken [2].

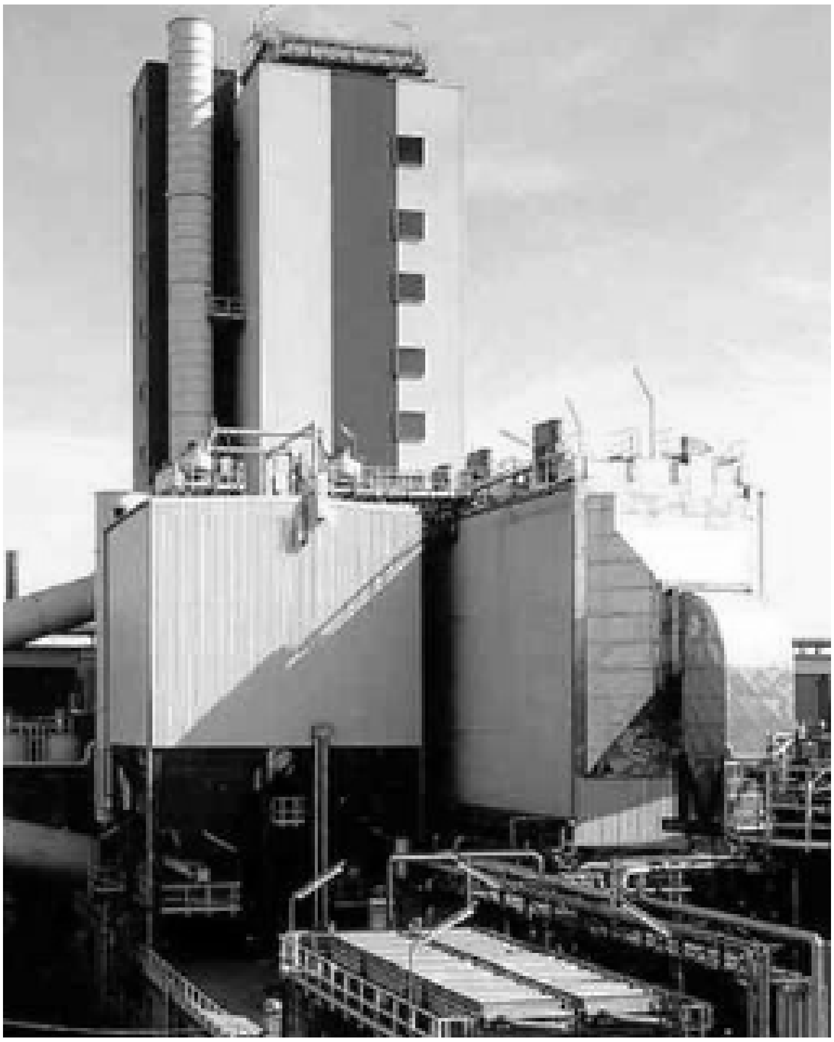


Figure 1 - ISASMELT^{TM} plant at Hoboken

Figure 2 shows the Precious and Base Metals operation process flowsheet at Hoboken [3]. The ISASMELT^{TM} furnace plays a key role in the overall metallurgical process. Complex lead/copper materials, containing valuable minor elements are treated in the ISASMELT^{TM} furnace in a two step smelting and converting process.

The smelting stage involves the oxidation of the feed to form a copper matte and a smelting slag by injection of oxygen enriched air through the lance. The silica-based Pb-rich slag produced during smelting is tapped and the remaining copper matte is converted to blister copper. All the precious minor elements are captured into the copper phase [4].

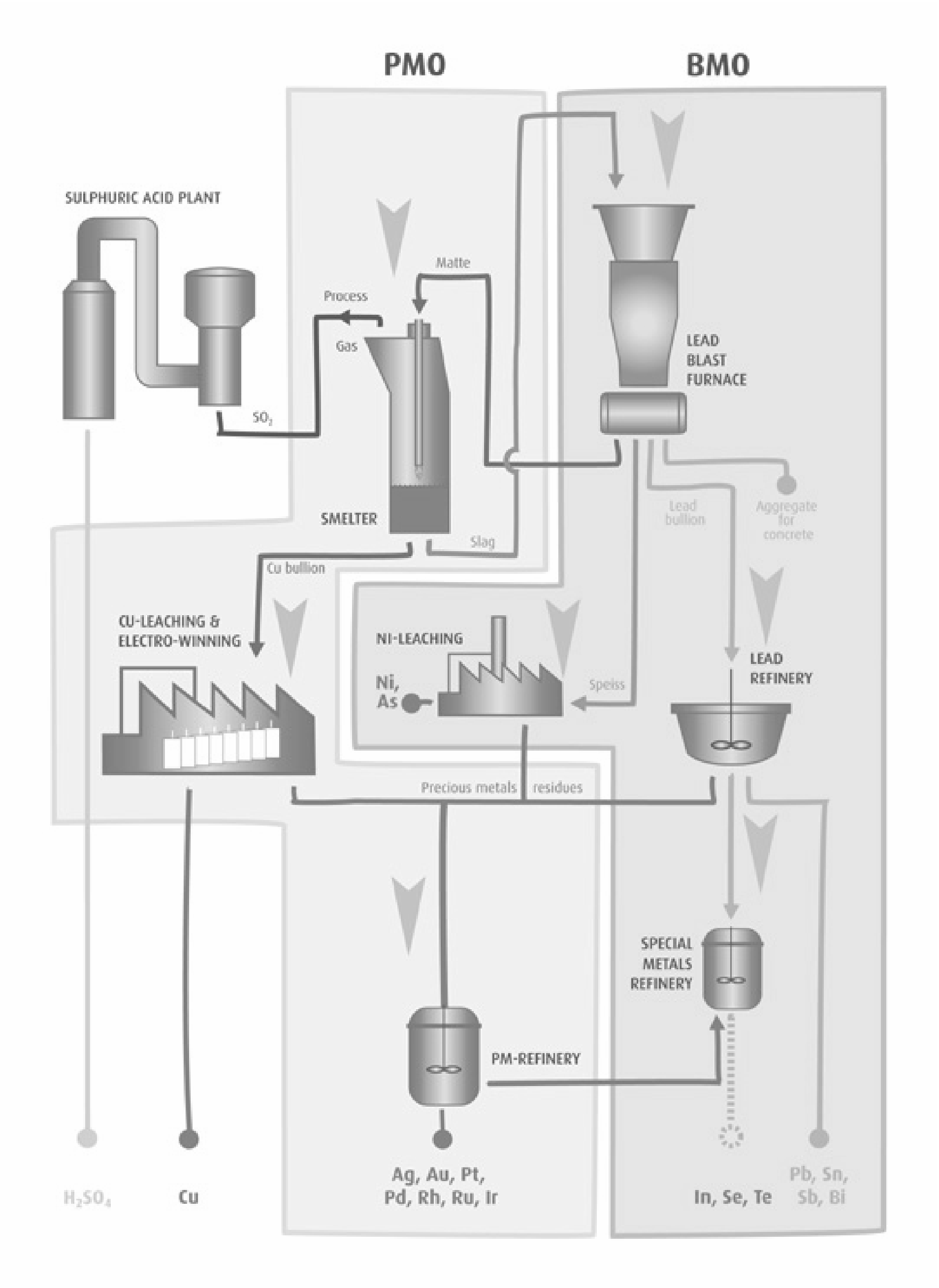


Figure 2 – Precious Metals and Base Metals Operations flowsheet at Hoboken

From the above process description it can be noted that the ISASMELT^{TM} furnace is operating in two very different environments: low oxygen partial pressure for the smelting and high oxygen potential partial pressure for converting. The different oxygen partial pressure conditions provide a harsh test for the refractory. Even in these extreme operational conditions, the ISASMELT^{TM} furnace is achieving a campaign life of 15 months [5].

Copper Smelting and Converting at Norddeutsche Affinerie - Hüttenwerke Kayser

Norddeutsche Affinerie AG operates an ISASMELT^{TM} plant for secondary copper smelting within a smelting and refining operation in Lünen, Germany [6]. Figure 2 shows an overall view of the plant at Lünen. The ISASMELT^{TM} furnace replaced three blast furnaces and one Peirce Smith converter used for smelting residues and scrap copper and plays an essential role in the copper recycling process: Kayser Recycling System (KRS).

Figure 2 - ISASMELT^{TM} Plant at NA in Lünen, Germany

Figure 3 shows the flowsheet for the KRS process. The smelting and converting of the copper bearing materials are carried out in the ISASMELT^{TM} furnace in batch charges. Copper residues and scrap containing 1 to 80 % copper are charged into the ISASMELT^{TM} furnace in a reductive smelting stage. A black copper phase and a silica-based slag that contains very low residual economic metals are produced during the reductive smelting. The slag is tapped and granulated.

After the tapping of slag the black copper phase is converted to produce blister copper with a copper content around 95%. A Sn-Pb-enriched converting slag is also generated and processed in a separate furnace. Due to the nature of the KRS process, the ISASMELT^{TM} furnace operates in a wide range of oxygen partial pressure in each cycle providing a harsh test for the refractory.

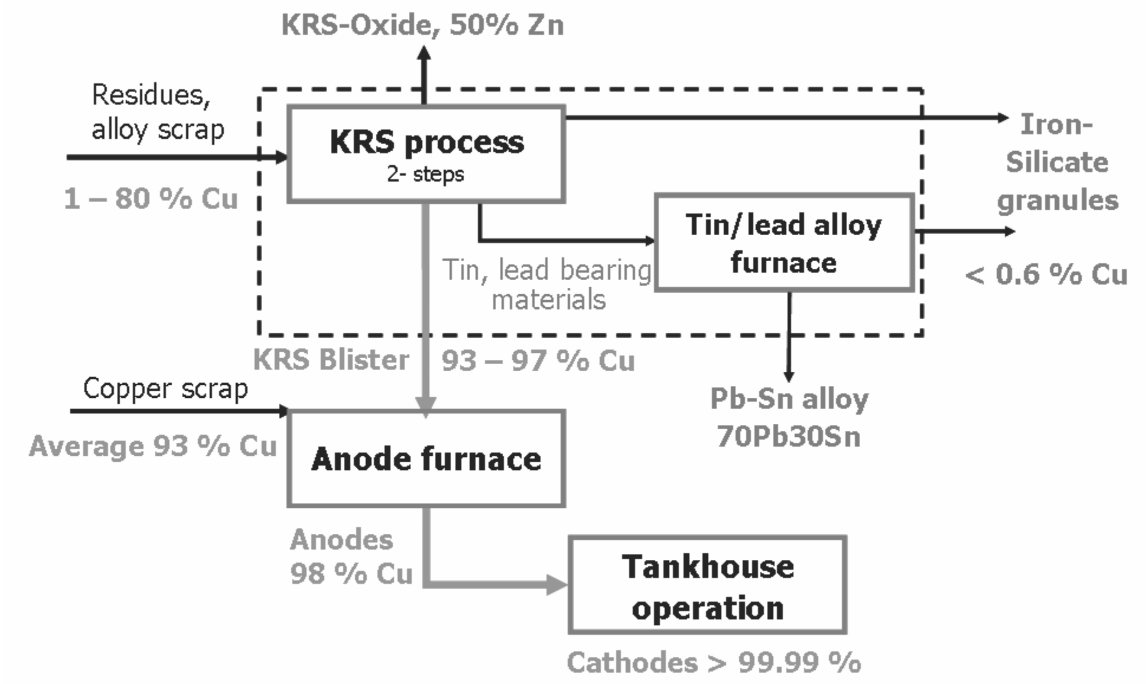


Figure 3 - Kayser Recycling System

The implementation of the ISASMELT^{TM} technology as a main component of the KRS process has allowed NA to achieve an outstanding position in productivity, energy efficiency and environmental protection. The main benefits claimed by NA AG, Lünen, after the successful implementation of the ISASMELT^{TM} process can be summarized as follows

- Improved overall copper recovery of the plant by achieving lower copper content in the dischargeable slag produced at the end of the smelting stage.
- Reduced number of furnaces in operation
- Significant decrease in off gas volume

- Production capacity exceeds the original design by 40%
- Decrease in energy consumption by more than 50%
- Decrease in CO_2 emissions by more than 64%
- Overall emissions reduced by 90%

CONTINUOUS CONVERTING

A combination of pilot plant continuous converting trials and the commercial batch converting operations show the ISASMELT^{TM} furnace to be well suited to the duty of continuous converting [7]. Areas of operation requiring consideration in designing a copper smelter using an ISASMELT^{TM} furnace for continuous converting – an ISACONVERT^{TM} furnace - are the slag chemistry, desulfurization of the blister copper, containment of the melt and smelter layout.

Slag Chemistry

The ISASMELT^{TM} process either in its smelting or converting mode involves the injection of oxygen enriched air into the molten slag. Control of the chemical composition and physical properties of the molten slag, such as the viscosity, is required.

Since the beginning of the development of the ISACONVERT^{TM} process it was noted a gap existed in the phase equilibria information available for the "Cu_2O"-CaO-"Fe_2O_3" system. Much of the existing information was under oxygen potential conditions that were not applicable for the continuous converting process conditions. Therefore, a cooperative research program was established with the Pyrometallurgy Research Centre (PYROSEARCH) at the University of Queensland in order to evaluate the copper converting slag systems and their potential applications to the ISACONVERT^{TM} process. Several experiments were carried out to determine phase equilibria in the following slag systems as a function of temperature and oxygen partial pressure:

- $CaO\text{-}FeO\text{-}Fe_2O_3$
- $Cu_2O\text{-}CaO\text{-}FeO\text{-}Fe_2O_3$ at copper saturation
- $CaO\text{-}FeO\text{-}Fe_2O_3\text{-}SiO_2$
- $Cu_2O\text{-}CaO\text{-}FeO\text{-}Fe_2O_3\text{-}SiO_2$ at copper saturation

The experimental work allowed for accurate determination of areas of interest in the above mentioned slag systems [8] [9]. Based on this fundamental work and a series of continuous converting pilot tests it was possible to define the slag composition region in which the ISACONVERT^{TM} process will operate. Figure 4 shows the slag composition region for the ISACONVERT^{TM} process, in the Cu_2O"-CaO-"Fe_2O_3" system at copper saturation at 1250 °C [8] The continuous line on the spinel saturation region shows the effect of the oxygen partial pressure on the copper oxide concentration on the slag phase. The grey filled area indicates the expected slag composition region in which the ISACONVERT^{TM} process will be carried out, at a Fe/CaO ratio of approximately 2.3.

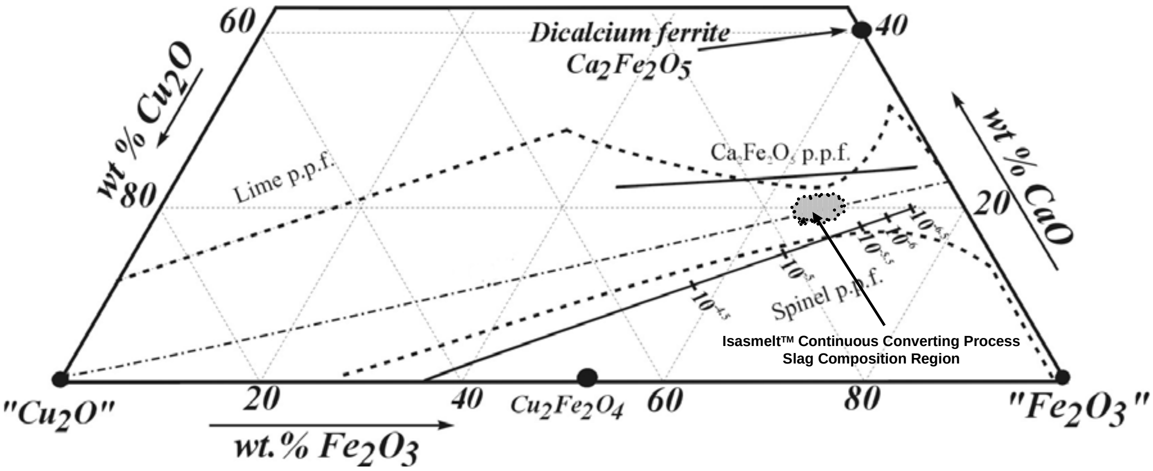


Figure 4 - “Cu2O”-CaO-“Fe_2O_3” system at copper saturation at 1250 °C

Anode Furnace Operation

Anode Refining Implications of Continuous Converting

Blister copper produced by continuous converting processes contains higher concentrations of sulfur compared with blister copper produced by batch processes such as Peirce-Smith converter. In continuous converting processes that involve three (3) condensed phases in equilibrium with a gas phase, such as the Noranda Continuous Converting process, the oxygen and sulfur potentials are fixed by the coexistence of liquid Cu_2S as a single phase at the desired temperature. This thermodynamic condition raises the sulfur content in blister copper to levels close to 1 %.

In the case of converting processes involving two (2) condensate phases, such as the ISACONVERT^{TM} process, the oxygen partial pressure is selected aiming to control the copper oxide content in the converting slag between 12 to 18 % resulting in sulfur contents between 0.3 and 0.4 %. Alternatives for shortening the desulfurization time when operating a continuous converting process include modifications to anode furnaces such as adding extra tuyeres, and/or porous plugs to enhance the rate of desulfurization via addition of extra oxygen, and homogenization of temperature and oxygen content in the bath.

With the ISACONVERT^{TM} process, either a direct transfer to an anode furnace or a pre-treatment in a retention vessel can be adopted. In both cases, it will be preferable to have porous plugs to enhance the homogenization of the bath (thermal and chemical) and at least four tuyeres for air and/or oxygen enriched air injection to enhance oxidation of the bath. The size of the anode furnace will need to be determined by the processing time required.

From the minor elements elimination point of view, a relatively high Pb level in the blister copper produced by the ISACONVERT^{TM} furnace may be expected, compared with blister produced in a batch converting process. This is due to the relatively low Pb capacity of calcium ferrite slag compared with silica based slag. Depending upon the anode quality specifications, a slagging stage with silica addition may be required.

Melt Containment

For continuous copper converting, given the aggressive nature of the lime ferrite slag, the ISACONVERT^{TM} furnace would be designed with water-cooled copper at the slag line. A great deal of experience has been gained in the non-ferrous industry and the ferroalloy industry in water cooling system design over the past 20 years. Water cooling system designs are now available from a number of suppliers. Options available for the furnace design include copper elements behind bricks, copper elements interleaved with brick, or operating on water-cooled panels with an initial working lining of castable.

The proposed design for an ISACONVERT^{TM} furnace performing continuous copper converting is a steel shell with a brick hearth and vertical water-cooled copper panels in the slag line. The panels will be installed with a working lining of castable and will be designed to anchor slag as a wear lining as the castable wears.

Smelter Layout

Figure 5 shows a block diagram for a plant with an ISASMELT^{TM} furnace for smelting of concentrates followed by an ISACONVERT^{TM} furnace for converting of copper matte to blister copper. As already practised in the Kennecott flash converting plant matte is granulated and stockpiled in front of the converting furnace [10]. The use of a small stockpile decouples the smelting and converting furnaces. Feeding solid matte to the converting furnace increases the heat load on the furnace which is balanced by the removal of nitrogen from the system through the use of tonnage oxygen resulting in a high SO_2 tenor of the offgas.

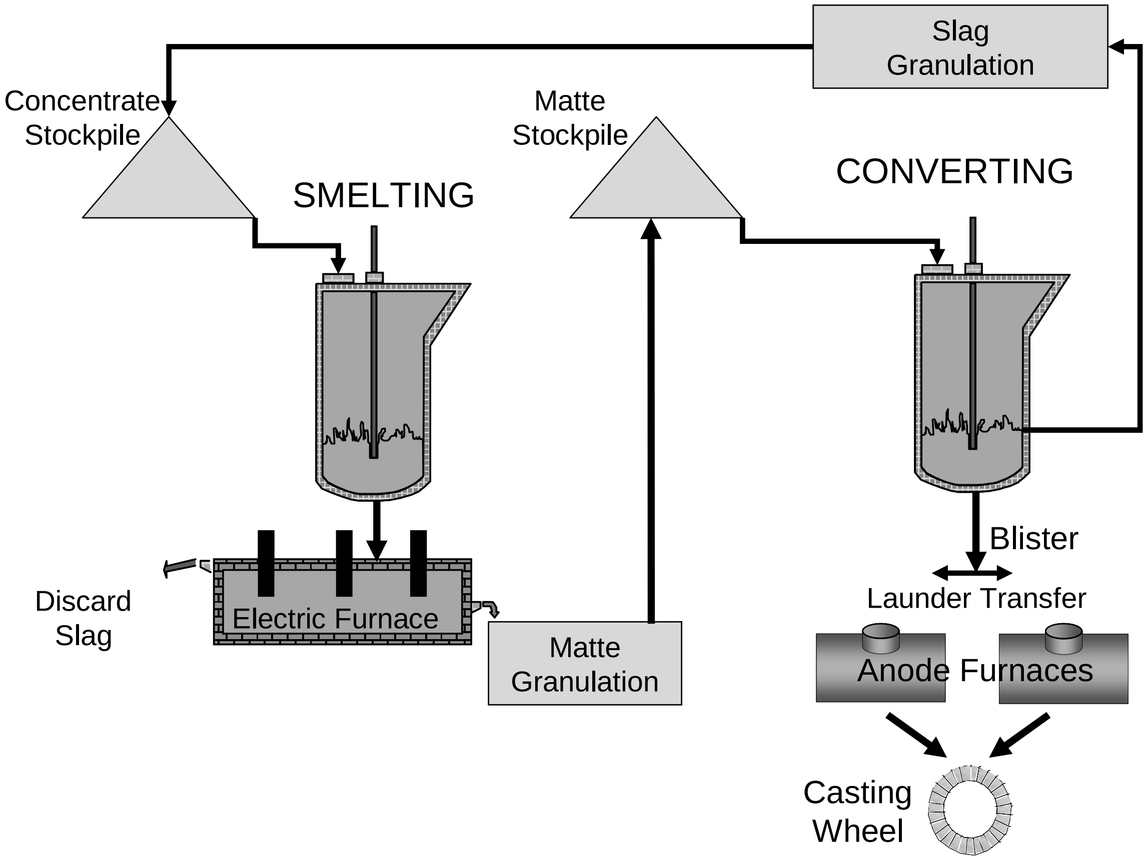


Figure 5 – Continuous converting plant flowsheet

Solid matte and flux will be dropped in through the roof of the converting furnace to be incorporated into the bath in the converting furnace. Slag will be tapped from the converting furnace to a granulation system, and the blister will be tapped to either a holding furnace or an anode furnace.

In a greenfield smelter where anode furnaces have been sized to treat blister containing 0.3 to 0.4 % sulfur the blister may be tapped directly from the converting furnace into an anode furnace. For existing smelters where anode furnaces have been sized to treat blister containing 0.05 % sulfur a holding furnace may be used after the converting furnace to allow pre-treatment of the blister copper before fire refining. The holding furnace would also act as surge capacity between the converting furnace and the anode furnaces.

Converter slag would be tapped from the ISACONVERT^{TM} furnace to a granulation system. The granulated converter slag would be fed to the primary ISASMELT^{TM} furnace. From previous experience solid slag fed to the ISASMELT^{TM} furnace is readily incorporated into the bath and acts as an oxygen source.

The layout for a smelter using an ISASMELT^{TM} furnace and an ISACONVERT^{TM} furnace will be very compact and will require significantly less area than conventional copper smelters. The fact that the smelting and converting processes are very similar and easy to control via a proven robust process control system will also simplify operations and logistics within the smelter, allowing all operations to be carried out from one central control room. These factors will lower the operating costs of the new smelter compared with the conventional Peirce-Smith technology. The significantly reduced offgas volume from the converting process will result in lower capital and operating costs for offgas collection and cleaning systems.

SUMMARY

- Copper ISASMELT^{TM} furnaces are operating successfully in 7 countries with capacities up to 180 tph of feed or 300,000 tpa of cathode copper equivalent.
- Batch converting of copper matte in commercial ISASMELT^{TM} furnaces has been performed since 1997 in Europe.
- Continuous converting in an ISACONVERT^{TM} furnace using a calcium ferrite slag has been successfully performed at a pilot scale.
- Slag systems for copper converting have been researched extensively and a lime ferrite slag appears the most practical for implementation of the continuous process in an ISACONVERT^{TM} furnace.
- Containment of the lime ferrite slag will be achieved using water-cooled systems in critical areas of the furnace only.
- Furnace layout will allow a very small footprint to be achieved for new copper smelters. There will be advantages in operations and logistics resulting in reduced operating costs.
- The significantly lower offgas volumes from the continuous converting process will result in lower capital and operating costs for offgas collection and cleaning.

ACKNOWLEDGEMENTS

Xstrata Technology gave permission to publish this paper.

Information on the Kayser Recycling System was published with the permission of Norddeutsche Affinerie AG - Hüttenwerke Kayser.

Xstrata thanks Umicore for technical discussions.

REFERENCES

---

1. V. Ramachandran, C. Diaz, T. Eltringham, C.Y. Jiang, T. Lehner, P.J. Mackey, C.J. Newman and A.V. Tarasov, “Primary Copper Production – A Survey of Operating World Copper Smelters”, Volume IV: Pyrometallurgy of Copper

Hermann Schwarze Symposium, Book 1: Smelting Operations, Ancillary Operations and Furnace Integrity, C. Diaz, J. Kapusta and C. Newman, Eds., The Canadian Institute of Mining, Metallurgy and Petroleum, Montreal, Canada, 2003, 3-106.

2. C. Hageluken, “Recycling of electronic scrap at Umicore’s integrated metals smelter and refinery”, Proceedings of European Metallurgical Conference 2005, GDMB, Clausthal-Zellerfeld, 2005, Vol. 1, 307-324.

3. F. Vanbellen and M. Chintinne, “The Precious Art of Metals Recycling”, Advanced Processing of Metals and Materials, F. Kongoli and R.G. Reddy, Eds., TMS, Warrendale, Pennsylvania, 2006, Vol. 1, 43-52.

4. C. Hageluken, “Recycling of electronic scrap at Umicore Precious Metals Refining”, Acta Metallurgica Slovaca, 12, 2006, (111-120)

5. Umicore Press Release, Third Quarter Update, October 27th 2006.

6. S. Schmidt, Norddeutsche Affinerie AG-Hüttenwerke Kayser, unpublished, March 2006.

7. J.S. Edwards and S. Jahanshahi, “Copper Converting”, United States Patent, No. 5,888,270, 30 March 1999.

8. S. Nikolic et al, “Liquidus temperatures in calcium ferrite slags equilibrated with molten copper at fixed partial pressures”, The 6th International Copper/Cobre Conference, Canadian Institute of Mining, Metallurgy and Petroleum, Toronto, 2007.

9. E. Jak et al, “Experimental Investigations of Phase Equilibria for Copper Smelting and Converting Silicate Slags in the “Cu2O”-FeO-Fe2O3-CaO-SiO2 System at controlled Oxygen Partial Pressures”, The 6th International Copper/Cobre Conference, Canadian Institute of Mining, Metallurgy and Petroleum, Toronto, 2007.

10. D.B. George, R.J. Gottling and C.J. Newman, “Modernization of Kennecott Utah Copper Smelter”, Proceedings of the Copper 95 – Cobre 95 – International Conference, Pyrometallurgy of Copper, W.J. Chen, C. Diaz, A. Luraschi, and P.J. Mackay, Eds., Canadian Institute of Mining, Metallurgy and Petroleum, Montreal, Canada, 41-52.

КОНВЕРТИРОВАНИЕ С ИСПОЛЬЗОВАНИЕМ ТЕХНОЛОГИИ ISASMELT™

Дж.С. Эдвардс и Дж.Р.Ф. Альвеар Ф.
Xstrata Technology
Level 4, 307 Queen Street
Brisbane, Queensland 4000 Australia
jedwards@xstratatech.com.au

АВТОРЕФЕРАТ

Технология ISASMELT™ признана во всем мире современной, гибкой, экологически чистой и не требующей значительных капитальных затрат технологией плавки. Благодаря простоте, присущей технологии, и совершенной системе управления процессом, операторы приобретают все необходимые знания и навыки в течение нескольких недель после начала обучения технологии ISASMELT™. Для сравнения, на овладение альтернативными технологиями уходят годы. Технология ISASMELT™ применяется по лицензии компании Xstrata Technology на медеплавильных заводах в Австралии, США, Индии, Китае, Перу и Замбии.

Заводы в Бельгии и Германии перерабатывают лом меди и/или свинца в печах ISASMELT™, используя двухстадийный процесс. За плавкой следует конвертирование в одной печи. Оба завода уже превысили проектную мощность и последовательно улучшают показатели эффективности.

Следующим шагом эволюции технологии является реализация процесса непрерывного конвертирования меди в промышленном масштабе.

Настоящая работа описывает ход и состояние разработки процесса непрерывного конвертирования ISACONVERT™.

Введение

За последние 25 лет процесс ISASMELT™ стал технологией, применяемой в цветной металлургии для решения различных задач, включая плавку первичной и вторичной меди, а также первичного и вторичного свинца.

Производительность печей по питанию, составлявшая на первой демонстрационной установке 12 тонн концентрата в час, выросла до 180 тонн в час на комбинате Маунт Айза, или свыше 300 тысяч тонн меди в катодном эквиваленте в год (комбинат Веданта, Индия). Совершенствование технологии продолжается с целью сделать возможной переработку питания различного качества, достичь более высокой производительности и реализовать различные процессы.

Темпы внедрения печи ISASMELT™ стали самыми высокими в выплавке первичной меди: пять печей находятся в эксплуатации, одна печь будет пущена в эксплуатацию в 2007 году, начато рабочее проектирование двух печей, ввод которых в эксплуатацию ожидается в 2009 году. По данным исследования действующих медеплавильных заводов, проведенного в 2003 году [1], доля первичной меди, полученной в печах ISASMELT™, в мировом производстве меди достигнет примерно 12% после выхода на полную мощность печи на медеплавильном заводе Southern Peru Copper Corporation в 2007 году.

Печь ISASMELT™ также может использоваться для непрерывного и периодического конвертирования медного штейна с получением черновой меди. Непрерывное конвертирование успешно применяется в пилотных установках ISASMELT™; черновую медь получают с использованием шлака на основе феррита кальция с расходом штейна 250 кг/ч. В Европе периодическое конвертирование в печи ISASMELT™ применяется на двух заводах. На заводе Umicore Precious Metals (Хобокен, Бельгия) печь ISASMELT™ применяется для конвертирования с 1997 года. Вторая печь ISASMELT™, применяемая для конвертирования, была пущена в 2002 году на заводе компании NA AG (Люнен, Германия).

ПЕРИОДИЧЕСКОЕ КОНВЕРТИРОВАНИЕ

Конвертирование меди на заводе Umicore

Umicore является производителем специализированных материалов со следующими основным направлениями деятельности: услуги в области драгоценных металлов, драгоценные металлы и катализаторы, улучшенные материалы и специальные цинковые продукты [2]. Компания, которая тогда называлась Metallurgie Hoboken Overpelt (MHO), сперва разработала конвертер Хобокен, но в 1990-е годы поняла, что он не сможет соответствовать строгим экологическим нормам, которые должны были вступить в силу в XXI веке. Поэтому компания Umicore решила начать сотрудничество с Mount Isa Mines (в настоящее время – Xstrata) с целью дальнейшей разработки процесса ISASMELT™ для выплавки и конвертирования смешанных видов медной шихты.

Xstrata спроектировала демонстрационную установку, которая затем в течение нескольких месяцев эксплуатировалась на металлургическом заводе в Хобокене. По результатам успешной эксплуатации демонстрационной установки была спроектирована и построена промышленная печь ISASMELT™. Промышленный завод был введен в эксплуатацию в конце 1997 года и в настоящее время перерабатывает до 300 тысяч тонн вторичного питания в год. На рисунке 1 показан общий вид установки ISASMELT™ на металлургическом заводе в Хобокене [2].

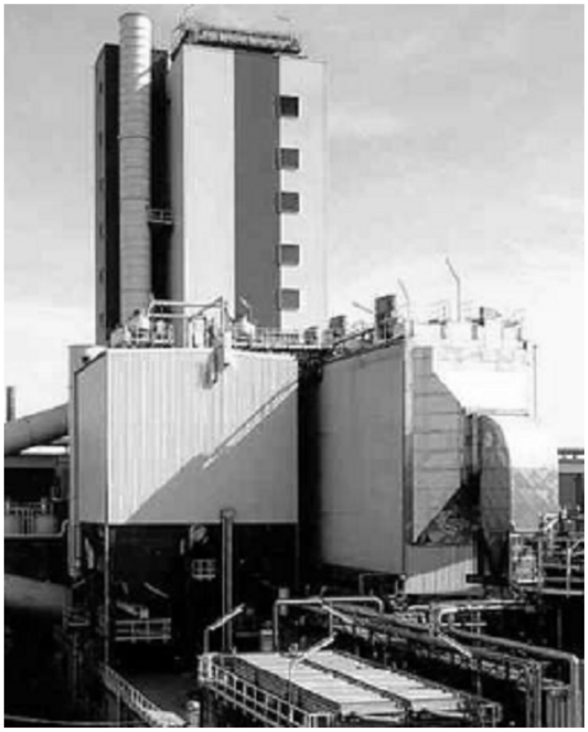


Рисунок 1 - Установка ISASMELT™ на металлургическом заводе в Хобокене

На рисунке 2 показана технологическая схема производства драгоценных и цветных металлов на металлургическом заводе в Хобокене [3]. Печь ISASMELT™ играет ключевую роль в металлургическом процессе. Сложное свинцовое/медное сырье, имеющее небольшое содержание ценных элементов, перерабатывается в печи ISASMELT™ в два этапа (плавка и конвертирование).

На этапе плавки происходит окисление питания путем подачи обогащенного кислородом воздуха через фурму с образованием медного штейна и шлака. Получаемый в процессе плавки богатый свинцом кремнистый шлак сливают, а оставшиеся медные штейны конвертируют в черновую медь. Все второстепенные драгоценные металлы попадают в медную фазу [4].

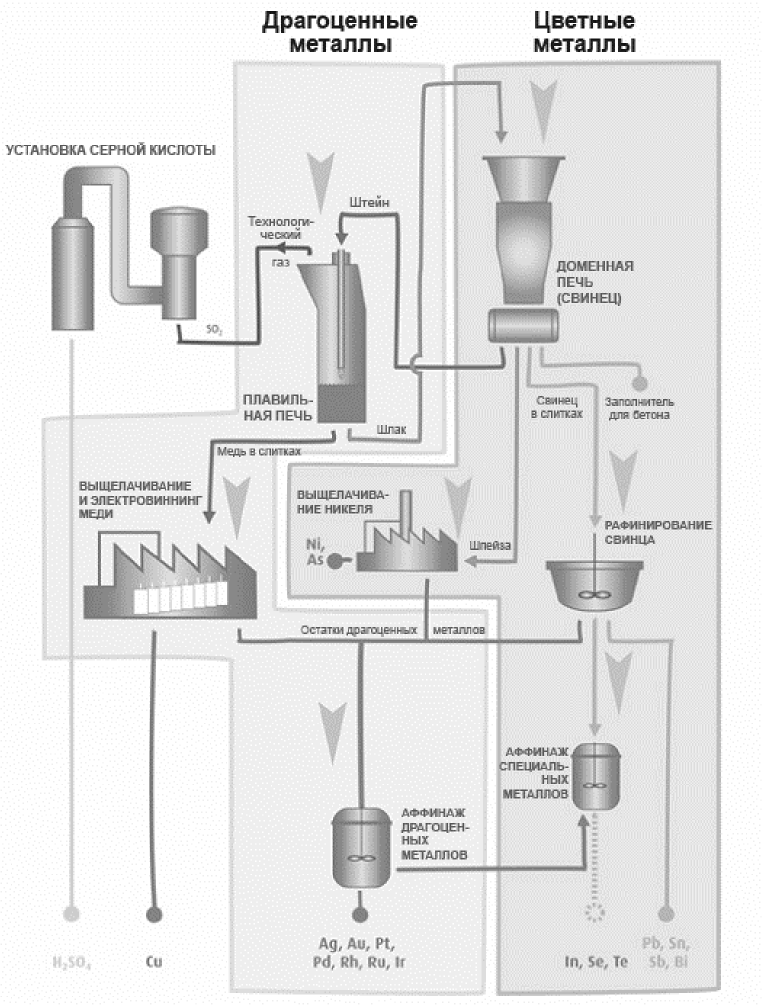


Рисунок 2 - Технологическая схема получения драгоценных и цветных металлов на металлургическом заводе в Хобокене

Из приведенного выше описания технологии можно заметить, что печь ISASMELT™ работает в двух очень разных режимах: низкое парциальное давление кислорода при плавке и высокое парциальное давление кислорода при конвертировании. Изменение парциального давления кислорода предъявляет жесткие требования к огнеупорной футеровке. Даже в столь сложных условиях работы продолжительность кампании печи ISASMELT™ составляет 15 месяцев [5].

Плавка и конвертирование меди на заводе Norddeutsche Affinerie - Hüttenwerke Kayser

Компания Norddeutsche Affinerie AG применяет технологию ISASMELT™ для плавки вторичной меди на медеплавильном и рафинировочном заводе в Люнене, Германия [6]. На рисунке 3 показан общий вид установки на заводе в Люнене. Печь ISASMELT™ заменила три доменные печи и один цилиндрический конвертер (конвертер Пирса-Смита), применявшиеся для плавки остатков и лома меди, и играет важную роль в процессе переработки меди - системе переработки Kayser (KRS).

Рисунок 3 - Установка ISASMELT™ на заводе NA, Люнен, Германия

Технологическая схема процесса KRS показана на рисунке 4. Плавка и конвертирование медьсодержащих материалов производится в печи ISASMELT™ с периодической загрузкой. Медные остатки и отходы, илинсодержащие от 1 до 80% меди, загружаются в печь ISASMELT™ на стадии восстановительной плавки. В результате восстановительной плавки получают оксид меди и кремнистый шлак, имеющий очень низкое остаточное содержание промышленных металлов. Шлак сливают и гранулируют.

После слива шлака оксид меди конвертируют для получения черновой меди с содержанием меди около 95%. Также образуется обогащенный Sn и Pb конвертерный шлак, который загружается в отдельную печь. В связи с особенностями процесса KRS печь ISASMELT™ работает в каждом цикле в широком диапазоне значений парциального давления кислорода, что обуславливает высокую нагрузку на огнеупоры.

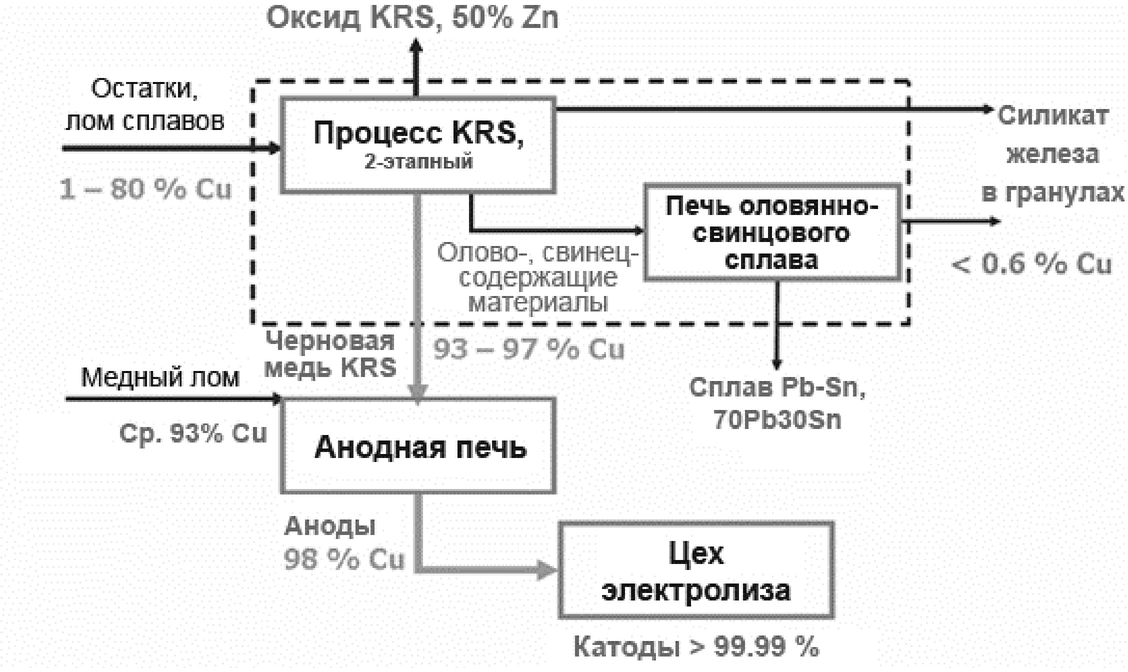


Рисунок 4 - Система переработки Kayser

Внедрение технологии ISASMELT™ в качестве основного компонента процесса KRS позволило NA добиться выдающихся показателей производительности, энергоэффективности и экологичности. Основные преимущества, полученные от успешного внедрения ISASMELT™, о которых сообщает завод NA AG в Люнене, можно обобщить следующим образом

- Более высокие общие показатели извлечения меди благодаря более низкому содержанию меди в выгружаемом шлаке в конце стадии плавки.
- Меньшее количество эксплуатируемых печей
- Значительное уменьшение объема отходящих газов
- Производственная мощность превышает проектную на 40%
- Снижение потребления энергии более чем на 50%
- Снижение выбросов CO2 более чем на 64%
- Снижение общих выбросов на 90%

НЕПРЕРЫВНОЕ КОНВЕРТИРОВАНИЕ

Результаты опытной эксплуатации пилотных установок непрерывного конвертирования в совокупности с результатами промышленной эксплуатации установок периодического конвертирования показывают, что печь ISASMELT™ хорошо подходит для непрерывного конвертирования [7]. При проектировании медеплавильного завода с использованием печи ISASMELT™ для непрерывного конвертирования - печи ISACONVERT™ - необходимо учитывать химический состав шлака, обессеривание черновой меди, способ удержания расплава и компоновочную схему медеплавильного завода.

Химический состав шлака

Процесс ISASMELT™ в режиме как плавки, так и и конвертирования предполагает подачу воздуха, обогащенного кислородом, в расплавленный шлак. Необходим контроль химического состава и физических свойств расплавленного шлака, таких как вязкость.

С самого начала разработки процесса ISACONVERT™ известно, что существует пробел в имеющихся данных о фазовых равновесиях в системе "Cu_2O"-CaO-"Fe_2O_3". Большая часть существующих данных была собрана в условиях кислородного потенциала, неприменимых в условиях непрерывного конвертирования. Таким образом, была начата программа совместных с Научно-исследовательским центром пирометаллургии (PYROSEARCH) Университета Квинсленда исследований с целью оценить шлаковые системы при конвертировании меди и их потенциальную применимость в условиях процесса ISACONVERT™. Было проведено несколько экспериментов, чтобы определить фазовые равновесия в следующих шлаковых системах в зависимости от температуры и парциального давления кислорода:

- CaO-FeO-Fe_2O_3
- Cu_2O-CaO-FeO-Fe_2O_3 при насыщении меди
- CaO-FeO-$Fe2O_3$-SiO_2
- Cu_2O-CaO-FeO-Fe_2O_3-SiO_2 при насыщении меди

Экспериментальные работы позволили точно определить интересующие области указанных шлаковых систем [8] [9]. На основе этих фундаментальных работ и ряда полупромышленных испытаний непрерывного конвертирования стало возможным определить область состава шлака, в котором будет протекать процесс ISACONVERT™. На рисунке 5 показан участок состава шлака, соответствующий процессу ISACONVERT™, в системе Cu_2O"-CaO-"Fe_2O_3 при насыщении меди при 1250°C [8] Сплошная линия в области насыщения шпинели показывает влияние парциального давления кислорода на концентрацию оксида меди в шлаковой фазе. Заштрихованная область обозначает область состава шлака, в которой, как ожидается, процесс ISACONVERT™ будет протекать при соотношении Fe/CaO около 2.3.

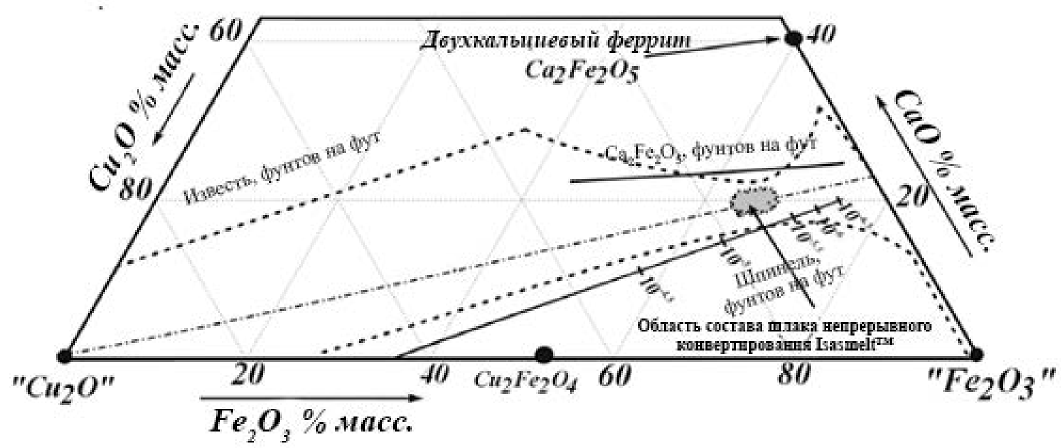


Рисунок 5 - Система "Cu_2O"-CaO-"Fe_2O_3" при насыщении меди при 1250˚C

Эксплуатационные показатели анодной печи

Влияние непрерывного конвертирования на рафинирование анодной меди

Черновая медь, полученная путем непрерывного конвертирования, имеет более высокое содержание серы по сравнению с черновой медью, полученной посредством периодических процессов, например в конвертере Пирса-Смита. В процессах непрерывного конвертирования, в которых три (3) конденсированные фазы находятся в равновесии с газовой фазой, таких как процесс непрерывного конвертирования Noranda, кислородный и серный потенциалы поддерживаются за счет присутствия жидкого Cu_2S в однофазном состоянии при желаемой температуре. Данное термодинамическое условие повышает содержание серы в черновой меди до уровня, близкого к 1%.

В процессах конвертирования с двумя (2) конденсированными фазами, таких как процесс ISACONVERT™, парциальное давление кислорода задается с целью поддержания содержания оксида меди в конвертерном шлаке в пределах от 12 до 18%, что обеспечивает содержание серы от 0,3 до 0,4%. К альтернативным способам сокращения времени десульфуризации в непрерывном процессе конвертирования относится внесение изменений в конструкцию анодных печей, например, установка дополнительных фурм и/или пористого днища для увеличения скорости десульфурации путем подачи дополнительного кислорода, а также гомогенизация температуры и содержания кислорода в ванне.

При использовании процесса ISACONVERT™ возможна либо подача непосредственно в анодную печь, либо предварительная обработка в печи для выдержки. В обоих случаях желательно использовать пористое днище для лучшей гомогенизации (термической и химической) ванны и не менее четырех фурм для подачи воздуха и/или воздуха, обогащенного кислородом, чтобы обеспечить более полное окисление ванны. Размер анодной печи определяется в зависимости от необходимого времени нахождения.

С точки зрения удаления второстепенных элементов, можно ожидать относительно высокого содержания Pb в черновой меди, производимой печью ISACONVERT™, по сравнению с черновой медью, полученной в результате периодического конвертирования. Это связано с относительно низкой емкостью по Pb шлака на основе феррита кальция по сравнению с кремнистым шлаком. В зависимости от требований к качеству анодов, может потребоваться шлакоудаление с добавлением кремнезема.

Удержание расплава

Учитывая агрессивный характер шлака на основе феррита кальция, печь ISACONVERT™, предназначенная для непрерывного конвертирования меди, должна проектироваться с водоохлаждаемой медной секцией в шлаковой линии. За последние 20 лет в цветной металлургии и промышленности ферросплавов накоплен значительный опыт проектирования систем водяного охлаждения. В настоящее время системы водяного охлаждения различной конструкции предлагают ряд поставщиков. При проектировании печей могут использоваться такие варианты, как медные элементы за кирпичом, медные элементы, чередуемые с кирпичом, или водоохлаждаемые панели с основной рабочей футеровкой из литого огнеупора.

Для печи ISACONVERT™, применяемой для непрерывного конвертирования меди, предлагается конструкция, представляющая собой стальной кожух с кирпичным подом и вертикальными водоохлаждаемыми панелями в шлаковой линии. Панели устанавливаются на рабочую футеровку из литого огнеупора, что предполагает осаждение шлака в качестве износной футеровки по мере износа литого огнеупора.

Компоновочная схема металлургического завода

На рисунке 6 показана технологическая схема, в которой для плавки концентратов используется печь ISASMELT™, после которой установлена печь ISACONVERT™ для конвертирования медного штейна в черновую медь. В соответствии с практикой, принятой на установке взвешенного конвертирования Кеннекотт, штейн гранулируется и складируется перед конвертерной печью [10]. Применение небольшого склада устраняет зависимость между печами плавки и конвертирования. Подача твердого штейна в конвертерную печь увеличивает тепловую нагрузку на печь, которая уравновешивается удалением азота из системы с помощью технического кислорода, что обуславливает высокое содержание SO2 в отходящих газах.

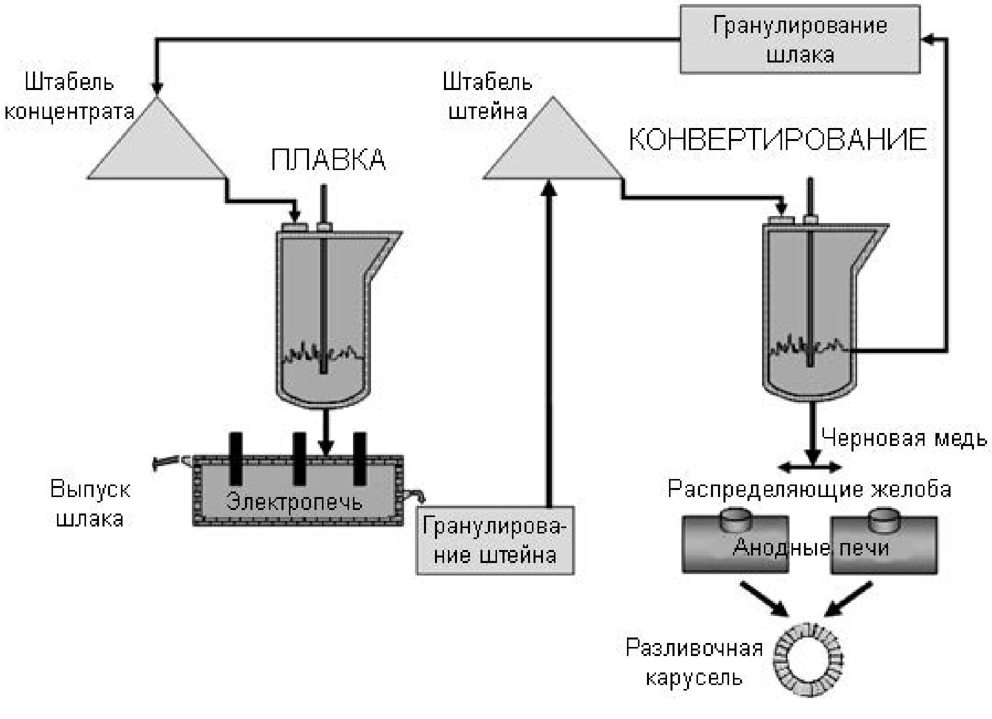


Рисунок 6. Технологическая схема установки непрерывного конвертирования

Твердый штейн и флюс поступают в ванну через свод конвертерной печи. Шлак выпускают из конвертерной печи в систему грануляции, а черновую медь – в печь выдержки или в анодную печь.

В условиях нового металлургического производства, если анодные печи рассчитаны на переработку черновой меди, содержащей от 0,3% до 0,4% серы, черновая медь может поступать из конвертерной печи непосредственно в анодную печь. В условиях действующего металлургического производства, если анодные печи рассчитаны на переработку черновой меди, содержащей 0,05% серы, для предварительной обработки черновой меди после конвертерной печи перед огневым рафинированием может быть использована печь выдержки. Печь выдержки также будет выступать в качестве буферной емкости между конвертерной печью и анодными печами.

Конвертерный шлак из печи ISACONVERT™ выпускают в систему грануляции. Гранулированный конвертерный шлак подается в первичную печь ISASMELT™. Как показывает опыт, твердый шлак, подаваемый в печь ISASMELT™, с легкостью растворяется в ванне и служит источником кислорода.

Схема металлургического завода с использованием печей ISASMELT™ и ISACONVERT™ очень компактна и занимает значительно меньшую площадь по сравнению с традиционным медеплавильным заводом. То обстоятельство, что процессы выплавки и конвертирования очень похожи и легко управляются с использованием проверенной системы управления технологическим процессом, также упрощает эксплуатацию металлургического завода и логистику, позволяя осуществлять все операции из одного центрального пункта управления. Эти факторы снижают стоимость эксплуатации нового завода по сравнению с традиционной технологией, основанной на использовании конвертеров Пирса-Смита. Значительно меньший объем отходящих газов конвертирования позволит сократить капитальные и эксплуатационные затраты на системы сбора и очистки отходящих газов.

ВЫВОДЫ

> Медеплавильные печи ISASMELT™ успешно применяются в 7 странах и имеют производительность до 180 тонн питания в час, или 300 тысяч тонн меди в катодном эквиваленте в год.

> Периодическое конвертирование медного штейна в промышленных печах ISASMELT™ применяется в Европе с 1997 года.

> Непрерывное конвертирование в печи ISACONVERT™ с использованием шлака на основе феррита кальция успешно опробовано в экспериментальном масштабе.

> Шлаковые системы при конвертировании меди хорошо изучены, и представляется, что шлак на основе феррита кальция является наиболее целесообразным для реализации процесса непрерывного конвертирования в печи ISACONVERT™.

> Для удержания шлака на основе феррита кальция системы водяного охлаждения будут применяться только в критических зонах печи.

> Конструкция печи обеспечивает высокую компактность новых медеплавильных заводов. Преимущества с точки зрения эксплуатации и логистики обеспечат снижение эксплуатационных расходов.

> Значительно меньший объем отходящих газов процесса непрерывного конвертирования позволит сократить капитальные и эксплуатационные затраты на системы сбора и очистки отходящих газов.

БЛАГОДАРНОСТИ

Данная статья опубликована с разрешения компании Xstrata Technology.

Информация о системе переработки Kayser опубликована с разрешения компании Norddeutsche Affinerie AG - Hüttenwerke Kayser.

Xstrata выражает Umicore благодарность за предоставленные технические пояснения.

СПИСОК ИСПОЛЬЗОВАННОЙ ЛИТЕРАТУРЫ

1. V. Ramachandran, C. Diaz, T. Eltringham, C.Y. Jiang, T. Lehner, P.J. Mackey, C.J. Newman and A.V. Tarasov, “Primary Copper Production - A Survey of Operating World Copper Smelters”, Volume IV: Pyrometallurgy of Copper Hermann Schwarze Symposium, Book 1: Smelting Operations, Ancillary Operations and Furnace Integrity, C. Diaz, J. Kapusta and C. Newman, Eds., The Canadian Institute of Mining, Metallurgy and Petroleum, Montreal, Canada, 2003, 3-106.
2. C. Hageluken, “Recycling of electronic scrap at Umicore’s integrated metals smelter and refinery”. Proceedings of European Metallurgical Conference 2005, GDMB, Clausthal-Zellerfeld, 2005, Vol. 1, 307-324.
3. F. Vanbellen and M. Chintinne, “The Precious Art of Metals Recycling”, Advanced Processing of Metals and Materials, F. Kongoli and R.G. Reddy, Eds., TMS, Warrendale, Pennsylvania, 2006, Vol. 1, 43-52.
4. C. Hageluken, “Recycling of electronic scrap at Umicore Precious Metals Refining”, Acta Metallurgica Slovaca, 12, 2006, (111-120)
5. Umicore Press Release, Third Quarter Update, October 27th 2006.
6. S. Schmidt, Norddeutsche Affinerie AG-Huttenwerke Kayser, unpublished, March 2006.
7. J.S. Edwards and S. Jahanshahi, “Copper Converting”, United States Patent, No. 5,888,270, 30 March 1999.
8. S. Nikolic et al, “Liquidus temperatures in calcium ferrite slags equilibrated with molten copper at fixed partial pressures”, The 6th International Copper/Cobre Conference, Canadian Institute of Mining, Metallurgy and Petroleum, Toronto, 2007.
9. E. Jak et al, “Experimental Investigations of Phase Equilibria for Copper Smelting and Converting Silicate Slags in the “Cu20”-Fe0-Fe203-Ca0-Si02 System at controlled Oxygen Partial Pressures”, The 6th International Copper/Cobre Conference, Canadian Institute of Mining, Metallurgy and Petroleum, Toronto, 2007.
10. D.B. George, R.J. Gottling and C.J. Newman, “Modernization of Kennecott Utah Copper Smelter”, Proceedings of the Copper 95 - Cobre 95 - International Conference, Pyrometallurgy of Copper, W.J. Chen, C. Diaz, A. Luraschi, and P.J. Mackay, Eds., Canadian Institute of Mining, Metallurgy and Petroleum, Montreal, Canada, 41-52.

USING METAL-RICH WEEE PLASTICS AS FEEDSTOCK / FUEL SUBSTITUTE FOR AN INTEGRATED METALS SMELTER

November 2006

Prepared by:
Johan Brusselaers,
Dr. Frank E. Mark,
Lein Tange

A Technical Report
produced by:
Plastics*Europe* in cooperation with Umicore and EFRA

Plastics*Europe*
Association of Plastics Manufacturers

USING METAL-RICH WEEE PLASTICS AS FEEDSTOCK / FUEL SUBSTITUTE FOR AN INTEGRATED METALS SMELTER

November 2006

Prepared by :

Johan Brusselaers, Umicore Precious Metals Refining,
www.preciousmetals.umicore.com
Frank E. Mark, Dow Europe, www.dow.com and
Plastics*Europe* www.plasticseurope.org
Lein Tange, Eurobrom ICI-IP www.iclfr.com and
EFRA www.cefic-efra.eu

Table of Contents

1. Executive summary

European legislation fixes obligations for end of life management of waste electrical and electronic equipment (WEEE), with ambitious targets, both for equipment reuse or materials recycling, and for energy recovery.

Economic and environmentally appropriate valorisation routes for WEEE materials with high plastics contents therefore have to be identified. These must be adapted to plastics with and without containing flame retardants, which are used in many electrical and electronic products to improve fire safety, either to meet mandatory safety standards or because of manufacturers' voluntary product safety objectives.

Integrated metal smelters represent one recycling route for such WEEE materials: metals can be recovered and recycled, and the plastics content can serve as an energy source and reducing agent.

The European Flame Retardants Association (EFRA), the plastics producing industry federation PlasticsEurope, and Umicore Precious Metals Refining carried out a full scale trial to assess the technical feasibility of including mixed WEEE materials with high plastics contents in the feed to an integrated metals smelter. The trial ran for a total of 8 days, using a sample of 217 tonnes of household WEEE. Impacts on plant performance, metals recovery, emissions and energy balance were assessed.

Results show that including approximately 6% WEEE (mixed metals – plastics materials, approx 3/4 plastics) in the smelter feed at this plant resulted in no operating problems, and that environmental emissions were not affected. One tonne of WEEE plastics enabled a reduction of consumption of nearly 0.6 tonnes of coke and 0.1 tonnes of fuel oil: approximately 2/3 of the plastics' energy content was used to replace consumption of coke and fuel oil (reducing agent and energy source) and approximately 1/3 was recovered as increased steam production.

At this plant, use of plastics rich WEEE at levels significantly higher than 6% of feed would require control and/or plant modifications to increase coke and fuel oil replacement as well as energy recovery in steam. The relatively low levels of flame retardants (FR) present in the typical household WEEE used in this trial posed no problem to use as smelter feed.

2. Introduction and objectives

The EU WEEE Directive (Waste Electrical and Electronic Equipment) 2002/96/EC requires Member States to ensure that producers of such equipment achieve ambitious targets both for equipment reuse or materials recycling and for energy recovery as of December 2006. The targets will be reviewed in 2008.

Economic and environmentally appropriate waste valorisation routes for WEEE plastics therefore have to be identified.

A significant part of these plastics from information technology, communication and some consumer sectors contain flame retardants, based on bromine, antimony, phosphorus, nitrogen and/or minerals. Annex II of the EU WEEE Directive specifies that WEEE plastics containing brominated flame retardants need to be identified and separately treated.

Integrated metal smelters equipped with state-of-the-art gas cleaning systems, potentially offer advantages for recycling of WEEE materials containing plastics: the metal content can be recovered for recycling (copper, precious metals used in electronic circuitry, solders ...) and the polymer content can serve as an energy source and reducing agent for the smelter process. Organic contaminants are eliminated in the high temperature smelter process and in the smelter emission controls systems.

Therefore, a full scale trial was carried out at the Umicore integrated metals smelter, Hoboken, Belgium. The objectives were to demonstrate the technical feasibility of including WEEE materials with high plastics content in the smelter feed and to assess the effects on the smelter performance and metals recovery (substitution potential for coke as a reducing agent and fuel oil as an energy source).

3. WEEE plastics

A wide variety of plastics are used in Electrical and electronic equipment (E&E), in order to meet the performance specifications defined by manufacturers and required by consumers.

Flame retardants are often included in WEEE plastics to improve fire safety. The main flame retardants used are magnesium and aluminium hydroxides, brominated flame retardants (in particular, reacted TBBPA and DecaBDE in WEEE), antimony compounds, phosphorus-based flame retardants, melamine compounds.

The type and content of flame retardants found in WEEE plastics varies widely depending on the type of equipment, collection and sorting. The amount and type of flame retardants in the WEEE plastics used in this trial was typical of that collected in The Netherlands .

4. The WEEE materials used in the trial

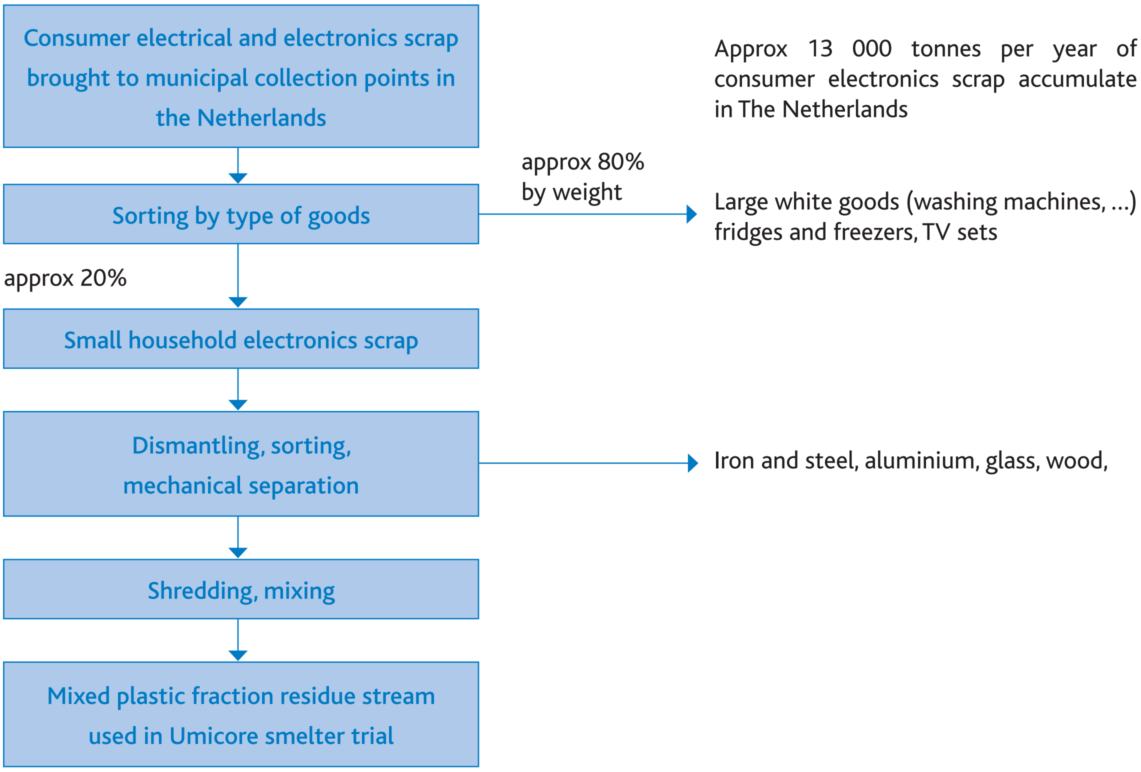
The WEEE (waste electrical and electronic equipment) material used was a plastics rich fraction of consumer electronics scrap, derived from partial dismantling of mixed small household appliances. brown goods (audio ...), grey goods (computers ...), etc.

The WEEE was supplied by a collecting organisation in The Netherlands and was from small household electrical and electronic goods brought voluntarily to municipal collection points by consumers in 2004. Large electrical equipment, white goods and TV sets were collected separately and were not included in the WEEE supplied.

A total sample of 290 tonnes of WEEE materials was provided via the recycling company PHB, consisting mainly of plastics and mixed metals, of which 217 tonnes were used in the trial.

As shown below, the actual plastic content of this WEEE residue is approximately 76%.

4.1. Origin and processing of the WEEE material



The 290 tonnes of WEEE delivered were divided in 3 equal lots. From each of these lots a representative 400 kg sample was taken for analysis to establish the elemental composition.

4.2. Composition of the WEEE material used in the trial

	% by weight		
Organic content (plastics)	Approx. 76%	Oxygen	8.4 - 13.5 %
		Carbon	58.6 - 63.7 %
		Hydrogen	6.2 - 7.2 %
		Nitrogen *	2 %
Metals	Approx. 10%	Including:	
		Copper (Cu)	3.9 - 6.7 %
		Silver (Ag)	0.01 %
		Lead (Pb)	0.2 - 0.4 %
		Iron (Fe)	0.3 - 1.1 %
		Aluminium (Al) *	2 - 3.4 %
		Antimony (Sb) *	0.1 - 0.25 %
		Magnesium (Mg) *	0.14 - 0.28 %
Other		Halogens	
		Bromine (Br) *	0.2 - 0.24 %
		Chlorine (Cl)	2.2 - 3.2 %
		Phosphorus (P) *	0.05 - 0.2 %
		Inert (eg. silica Si)	Approx 10%

* = *elements present in WEEE principally because of flame retardant content*

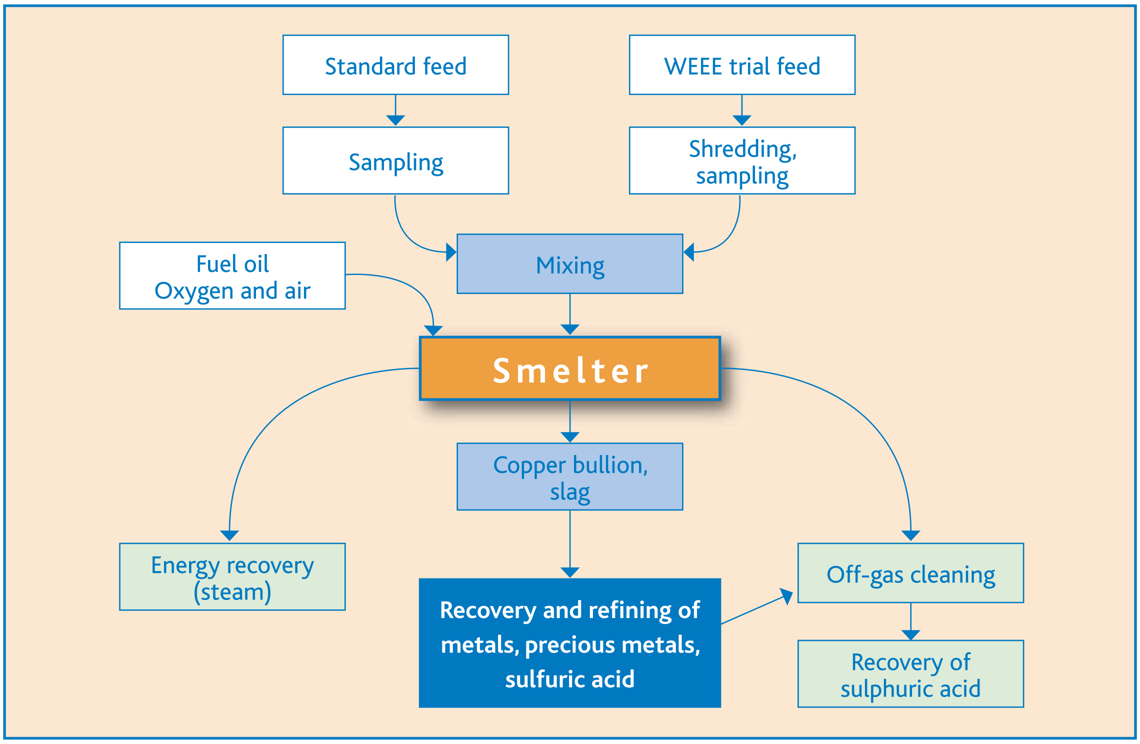
5. The Umicore plant at Hoboken

The trial was carried out in 2004 at the Umicore Precious Metals Refining plant, operating at Hoboken near Antwerp, Belgium. This is an integrated metals smelter and refinery. The plant recovers, separates and purifies a wide range of precious metals, special metals, base metals, antimony and sulphur (as sulphuric acid).

Feed materials are various complex industrial by-products (drosses, slags, flue dusts, process sludges, filter cakes, leachates, …) as well as consumer recyclables such as end-of-life automotive catalysts and printed circuit boards / electronic components. The plant treated some 250,000 tonnes of different materials per year in 2004 (300,000 t/y today), of which electronic scrap is at present up to 10%. These 25 000 tonnes/year of electronic scrap are mainly printed circuit boards, containing on average 25-30% of plastics.

The recycling operations at the plant are based on complex lead / copper / nickel metallurgy and the use of these base metals as collectors for precious and special metals.

5.1. Simplified flow chart of the integrated smelter process



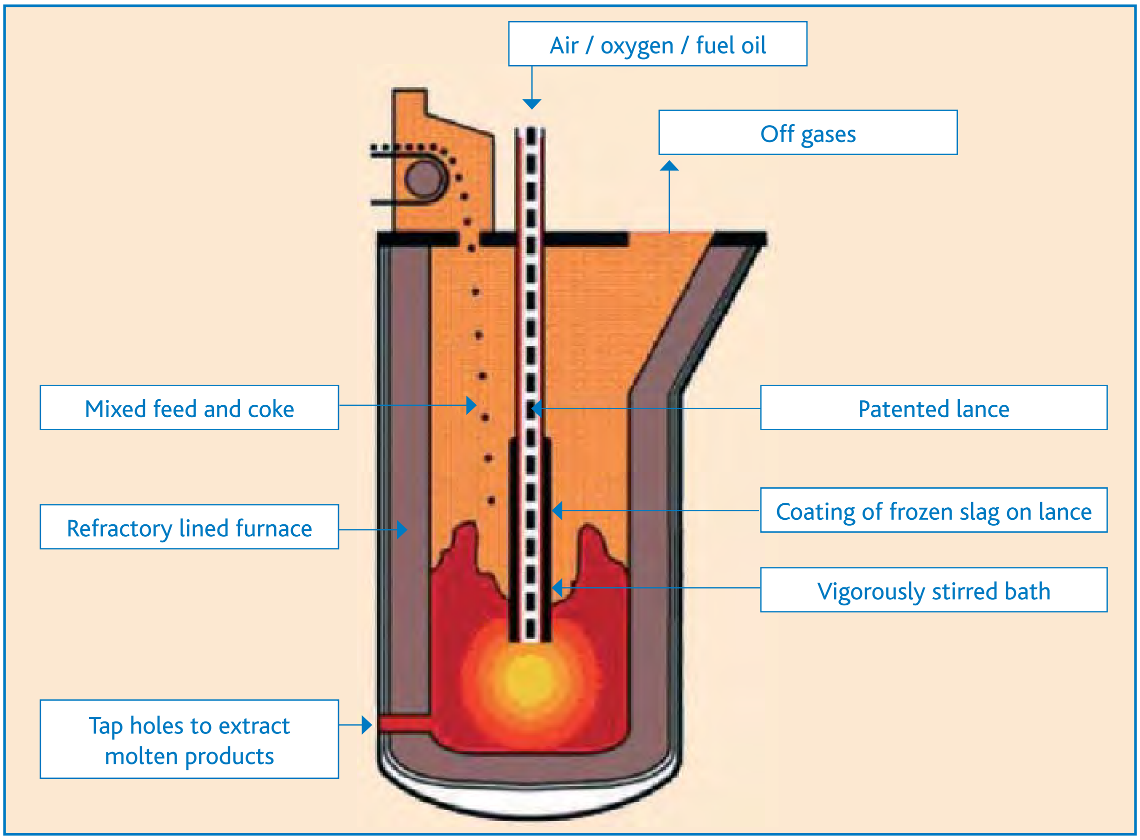
This integrated plant comprises two linked operations:

- the Precious Metals Operation, which includes the main processing steps: smelter, copper leaching and electro-winning, precious metals refinery. The smelter collects precious metals in a copper bullion, while most other metals are concentrated in a lead slag and further treated at the Base Metals Operation. After leaching out the copper in the leaching and electro-winning plant, the precious metals are collected in a residue that is further refined at the precious metals refinery.
- the Base Metals Operation includes a lead blast furnace, the lead refinery and a special metals plant.

The plant was comprehensively refurbished between 1997 and 2003. Around one quarter of the total costs concerned installations for safety and environmental protection, including state-of-the-art air and water treatment facilities.

5.2. Diagram of the Hoboken smelter

The core of the plant is the smelter, commissioned in 1997. The smelter uses the IsaSmelt Top Submerged Lance (TSL) combustion technique.

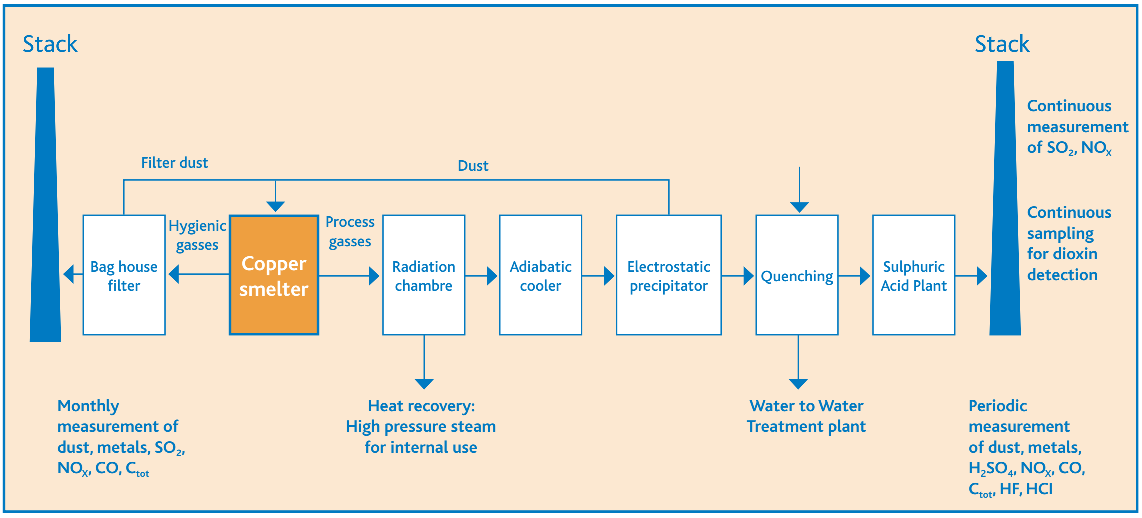


The ISA smelter (see diagram) is an upright tube, approx 15 metres high, with ceramic lining. A concentric lance feeds air, oxygen and fuel, in order to smelt and to reduce the material fed to the furnace, in the molten bath. Coke is added as a reducing agent for the metals. Material is fed to the furnace from the top by means of a metallic feed conveyor. The process gases leave the furnace also by the top where they are collected and sent to the gas-cleaning installation. The smelter offers considerable flexibility and a high process efficiency, which results in maximum metal recovery rates and optimal precious metal yields via different process steps. The process used is able to cope with all forms of feed material, wet or dry, fine or lumpy and all types of materials, metallic, oxidised, hydroxide, sulphide or sulphate, and also with the whole range of metals and plastics present in WEEE.

The smelters' role is to separate lead from copper and precious metals. The smelter operates in a temperature range between 1050 and 1200 °C. During this trial, the temperature was always close to 1200 °C.

5.3. Off-gas emission control installations at Umicore's Hoboken site

The Hoboken plant is equipped with an extensive flue gas cleaning system (see diagram), resulting in very low emissions. The following are monitored on a regular basis: dust, metals, dioxins, total organic carbon (TOC), HCl, HF, HBr, NO_x, and SO_2.



6. Trial run using WEEE in the smelter feed stream

6.1. Parameters of trial run and control

In order to measure and assess the impacts of including WEEE plastics in the smelter feed, a trial run with around 6% (by weight) WEEE plastics was compared to a "control" run with feed not containing such materials.

For the control run, a feed mix typical of daily operation at the Hoboken plant was used, but specifically excluding electronic scrap (whereas the plant currently regularly includes WEEE in its feed, mainly in the form of printed circuit boards, see above). This control feed mix included around 3.7% (by weight) coke necessary to ensure the reduction of the metal content of the waste stream. This run was carried out over about 3 days of continuous smelter operation.

A transition run was then performed (using the trial run feed as below), in order to ensure that trial run data was fully representative of the trial run conditions and not carry-over from the control run.

For the trial run, the same average typical feed mix was used, but with 6.4% (by weight) WEEE (of which approx 76% plastics, as described above), and only 0.9% (by weight) coke. This run was carried out over about 4 days of continuous operation.

In total, the two runs and the transition run were carried out over about 8 days of continuous smelter operation.

During the whole experiment (control run and trial run), the standard feed composition was changed as little as possible (other than the WEEE content in the trial run). The different waste types mixed in the feed stream were adjusted to maintain a fairly constant average composition although, because of the large volumes treated, the feed composition could not be kept 100% identical.

6.2. Specifications of feed mixtures

		Control run	Transition	Trial run
Duration (approx)		3 days	1 day	4 days
Tonnage standard feed used		1 530 tonnes	893 tonnes	2 733 tonnes
Feed composition				
As % of standard feed	WEEE	None	6.4%	
	Coke	3.7%	0.9%	
	Fuel oil	4%	3.3%	
Standard feed : feed mix typical of daily smelter operation at the plant, but specifically excluding electronic scrap				

7. Trial results

7.1. Energy balance

The energy balance calculation is given below. Energy losses are accounted into the steam energy production figure. This is based on the known energy content of coke, fuel oil and steam production, and the measured changes in energy input (coke, fuel oil) and output (steam production). From these, the useful energy content of the WEEE used in the feed is calculated.

	Inputs			Outputs
	WEEE	Coke	Fuel Oil	Steam production (20°C -> 1180°C)
Energy content (GJ/tonne)		30	43	2.8
Control run				
Tonnes / tonne standard feed		3.7%	4.0%	0.467 tonnes
MJ / tonne standard feed		1102 MJ	1700 MJ	- 1 313 MJ
Trial run				
Tonnes / tonne standard feed	6.4%	0.9%	3.3%	0.658 tonnes
MJ / tonne standard feed	Calculated as: 1629 MJ	272 MJ	1 435 MJ	- 1 847 MJ
MJ / tonne standard feed		Reduction in coke and fuel consumption: 1095 MJ		Increased steam production: - 534 MJ

Note: The detailed calculations are to be seen in the appendix.

From the above assessment of the plant performance data approximately 2/3 of the recovered energy (1095 MJ) resulted in this trial run in reduced coke and fuel consumption (reducing agent, energy resource), and 1/3 in increased steam production (534 MJ). The steam production is used as an energy source for various processes in the Umicore plant.

Observation suggests that this is because the plastic is not all coming into the molten bath zone of the smelter, but a part of it is decomposed by the smelter heat during loading into the top of the smelter, volatilises and burns higher up in the furnace, transferring its energy into steam production.

It can therefore be concluded, that approximately 2/3 of the energy content of the WEEE plastics included in the smelter feed stream in this trial replaces consumption of virgin coke and fuel oil (reducing agent and energy source) and approximately 1/3 is used in increased steam production.

7.2. Environmental performance

During the trial run, plant emissions measured after the existing off-gas cleaning system for chlorinated dioxins/furans remained well below the legal emission limit of 0.1 ngTEQ/Nm^3, and no significant difference was observed between the control and trial periods. Brominated dioxin/furan congeners were not detected during the air emission measurements. The overall effects on solid waste production and water emissions have not been measured.

It can be assumed that organic compounds (such as brominated flame retardants TBBA, DecaBDE, organic phosphorus flame retardants, melamine compounds …) are essentially destructed at the high smelter operating temperature (>1100°C). However, exact destruction efficiency calculations could not be carried out in the industrial operating context of this trial. Bromine from brominated flame retardants will be converted to a large extent to soluble acid (HBr) and transferred to the plant's waste water treatment system.

Other flame retardant compounds: at the Hoboken plant, over 70% of antimony (Sb) in feed is recovered and recycled as a as commercial antimony salt (sold for industrial applications); aluminium, magnesium and phosphorus are transferred mainly to slags which are reused in the concrete industry as a substitute for gravel ; some phosphorus may be transferred to phosphoric acid in the plant's air scrubbing system.

Reliable and meaningful mass balances could not be measured due to the relatively short term and the industrial scale of this trial as well as the integration of the integrated smelter into the whole plant operations at the Hoboken site.

7.3. Smelter operational performance

The trial shows that, in the current smelter configuration, the proportion of WEEE plastics which can be included in the feed stream is limited by the steam production and by the increase in temperature at the top of the furnace.

8. Conclusions and recommendations

8.1. Technical conclusions

This trial showed that in the Umicore smelter at Hoboken:

- inclusion of approximately 6% plastics rich WEEE (mixed metals – plastics materials, approx 3/4 plastics) in the smelter feed resulted in no operating problems. The process stability and environmental emissions were not affected as far as measured.
- input levels of plastics rich WEEE significantly higher than normally today are limited by steam production capacity and would require control and/or modifications to prevent overheating of the furnace
- in this trial, one tonne of WEEE plastics enabled a reduction of consumption of approximately 0.6 tonnes of coke and 0.1 tonnes of fuel oil: approximately 2/3 of the plastics' energy content was used to replace consumption of coke and fuel oil (reducing agent and energy source) and approximately 1/3 was recovered as increased steam production.
- flame retardants present in the typical household WEEE posed no problem to use as smelter feed within the % limits indicated above
- there was no measurable increase in air emissions from dioxin/furan like compounds from chlorine or bromine containing flame retardants at the levels present in this household WEEE

8.2. Recommendations

- Implementation of the EU WEEE Directive (2002/96/CE) will result in considerably increased arisings of collected WEEE-scrap materials.
- Integrated metals smelters can provide an appropriate route for certain plastics and metal rich WEEE residue materials, enabling recovery of different precious and non-ferrous metals, including antimony in flame retardants, and the use of the plastics fraction to replace coke and fuel oil (reducing agent and energy source)..
- The Hoboken smelter plant has the ability to treat WEEE plastic containing fractions in conformity to applicable legislation, and without requiring meticulous sorting or complex pre-processing of WEEE-scrap. It therefore provides an appropriate solution for complex WEEE-scrap fractions, in particular circuit boards and small electronic apparatus such as mobile phones, portable music players, calculators, telephones, modems and computer components....

9. Appendix

	Inputs			Outputs
	WEEE	Coke	Fuel Oil	Steam production (20°C -> 1180°C)
Energy content (GJ/tonne)		30	43	2.8
Control run				
Tonnes / tonne standard feed		3.7%	4.0%	0.467 tonnes
MJ / tonne standard feed		1102 MJ	1700 MJ	- 1 313 MJ
Trial run				
Tonnes / tonne standard feed	6.4%	0.9%	3.3%	0.658 tonnes
MJ / tonne standard feed	Calculated as: 1629 MJ	272 MJ	1 435 MJ	- 1 847 MJ
MJ / tonne standard feed		Reduction in coke and fuel consumption: 1095 MJ		Increased steam production: - 534 MJ

From the table above:

One tonne of WEEE plastics input corresponds to a reduction in consumption of 0.57 tonnes of coke and 0.1 tonnes of fuel oil.

Calculation:
6.4% WEEE x 76% plastics = 4.9% WEEE plastics
< --- > (3.7 – 0.9)% coke and (4.0 – 3.3)% fuel oil

The smelter's inherent energy consumption and the energy content of the standard feed input are assumed to be the same for the two runs. Therefore, the energy provided by the WEEE included in the feed of the trial run can be calculated as follows:

Calculation:
X = MJ energy in WEEE / tonne standard feed
X + 272 + 1435 -1847 = 1102 + 1700 – 1313
Therefore : X = 1629 MJ / tonne standard feed

Given that the WEEE fraction used contained approximately 76% plastics (see composition table, the rest being metals and inerts), this means that its plastic content delivered around 33 MJ of recovered energy per kg plastic content:

Calculation:
6.4% WEEE / tonne standard feed -> 1629 MJ / tonne standard feed
(0.76 x 6.4%) tonnes WEEE plastic -> 1629 MJ
Therefore : 1 tonne WEEE plastic -> 33 000 MJ

This amount for the delivered energy from the WEEE plastics in the trial run (33 000 MJ/tonne above) is coherent with the theoretical calorific value of plastics (which ranges from around 25 000 – 45 000 MJ/tonne) and with analysis of samples of the WEEE plastics used which were variable but within this range.

10. Acknowledgements

Technical support and assistance: NVMP Stichting Nederlandse Verwijdering Metalektro Producten (Netherlands Association for Disposal of Metalectro Products) www.nvmp.nl

Financial support: IAOIA (International Antimony Oxide Industry Association) www.iaoia.org

The WEEE used in trial was supplied via the recycling company PHB (Plastic Herverwerking Brakel B.V.) The Netherlands www.phb-recycling.com

Photos: PHB, Umicore

PlasticsEurope
Association of Plastics Manufacturers

Plastics*Europe*
Avenue E. Van Nieuwenhuyse 4 - Box 3
B-1160 Brussels, BELGIUM
Tel: +32(0)2 675 32 97
Tel: +32(0)2 675 39 35
www.plasticseurope.org

8047/GB/12/06

Sohn International Symposium ADVANCED PROCESSING OF METALS AND MATERIALS
VOLUME 8 - INTERNATIONAL SYMPOSIUM ON SULFIDE SMELTING 2006
Edited by F. Kongoli and R.G. Reddy ***TMS (The Minerals, Metals & Materials Society), 2006***

COPPER ISASMELT – DEALING WITH IMPURITIES

Gerardo R. Alvear[1], Simon P. Hunt[1], Bangqi Zhang[2]
[1]Xstrata Technology, 87 Wickham Terrace, Brisbane, 4000, Australia
[2]Yunnan Copper Co., Ltd., Wang Jia Qiao, Wuhua District, Kunming 650102, P.R. China

Keywords: ISASMELT, arsenic, distribution, volatilisation

Abstract

The ability to efficiently remove impurities contained in concentrates is a key point to consider when selecting copper smelting technology for new plants or modernisation projects. Volatilisation of impurities such As and Sb should occur as early as possible in the process, and preferably in the smelting furnace, so that they do not impact on downstream unit operations.

The ISASMELT™ process has been demonstrated to remove impurities efficiently, through high levels of volatilisation. This behaviour is promoted by attributes of the process such as the strong bath agitation, and the flexible nature of the process design, allowing the operator to choose the optimal matte grade for impurity partitioning, and the positive effect of the moisture content in the feed.

This paper presents the distribution of minor elements in the copper ISASMELT™ furnace. Distribution impurities of As, Sb, Pb, Zn and Co observed in the ISASMELT™ furnace at Mount Isa Smelter are reported. Then, based on plant results and thermodynamic considerations, the potential application of ISASMELT™ technology for processing complex concentrates is discussed.

Introduction

Xstrata is a global mining and processing company whose main products are copper, ferro-alloys, coal and zinc. Xstrata's mining, smelting and refining operations are based in Australia, South Africa, Spain, Germany, England and Argentina, while the company's headquarters are in Switzerland. In 2003, Xstrata acquired the Australian company MIM Holdings Limited (MIM), including the technologies that MIM had developed over more than 80 years of operating copper, lead and coal plants in Australia and Europe. This suite of technologies included the ISASMELT™ process.

The ISASMELT™ process is a simple, cost-effective, non-ferrous smelting process. Today, it is designed by, marketed by, and supported by Xstrata Technology. It is one of the most efficient and flexible smelting technologies available in the world today.

ISASMELT™ was developed from pilot plant to commercial scale during the 1980's and 1990's at MIM's smelting operations in Mount Isa, Australia. It rapidly became a crucial part of MIM's smelting operations. Eight companies in seven different countries are now using the technology in large-scale operations, and two more plants, in Zambia and Peru, are going to be commissioned during 2006. The history of the process has been documented in a number of published papers [1-9].

The process is able to treat a wide range of feed materials, ranging from mineral concentrates to scrap metals and recycled smelter materials. Feed preparation requirements are simple, and concentrates do not need to be dried as required by some other copper smelting technologies.

A key component of the ISASMELT™ process is the unique ISASMELT™ lance. The lance is inserted into a molten slag bath contained within the ISASMELT™ furnace, which is a stationary, vertical, refractory-lined vessel. The lance is used to inject air – or oxygen-enriched air – into the slag, resulting in a highly turbulent molten bath. A layer of slag frozen on the outer surface of the ISASMELT™ lance protects it from the molten bath. A swirler inside the lance increases the velocity of the process air inside the lance, thus increasing the heat transfer rate from the steel of the lance body to the process air stream. The swirler is specially designed to produce only a small pressure-drop, and this allows the lance to operate with a total pressure-drop of less than 80 kPa. A single stage blower can thus be used to provide process air to the lance. Lance construction is simple and robust, and normal lance repairs can be carried out within 6 hours. Lance operating life is typically 10 – 20 days.

Feed material falling into the turbulent bath from above reacts rapidly, resulting in extremely high productivity for a relatively small bath volume. The copper ISASMELT™ furnace at Mount Isa has smelted up to 194 tonnes per hour of copper-bearing feed (concentrate, reverts, and other internal smelter recycle materials) in a total bath volume of approximately 15 m^3.

Since the beginning of the development of the ISASMELT™ technology it was noted its capacity to remove impurities efficiently, through high levels of volatilisation (3,4). This behaviour is promoted by attributes of the process such as the strong bath agitation, the flexible nature of the process design, allowing the operator to choose the optimal matte grade for impurity partitioning, and the positive effect of the moisture content in the feed.

The present work presents the distribution of minor elements observed in the copper ISASMELT™ furnace at Mount Isa Copper Smelter. Then, based on the observed minor element distributions at Mount Isa, additional plant data from the ISASMELT™ furnace at Yunnan Copper Corporation and thermodynamic considerations, the potential application of ISASMELT™ technology for processing complex concentrates is analysed.

Mount Isa Copper Smelter

Plant Description

Figure 1 shows a flowsheet of the Copper ISASMELT™ plant (CIP) at Mount Isa. The primary smelting area of the copper smelter consists of a feed storage and delivery system, the ISASMELT™ Furnace, the offgas-handling system and two Rotary Holding Furnaces (RHF).

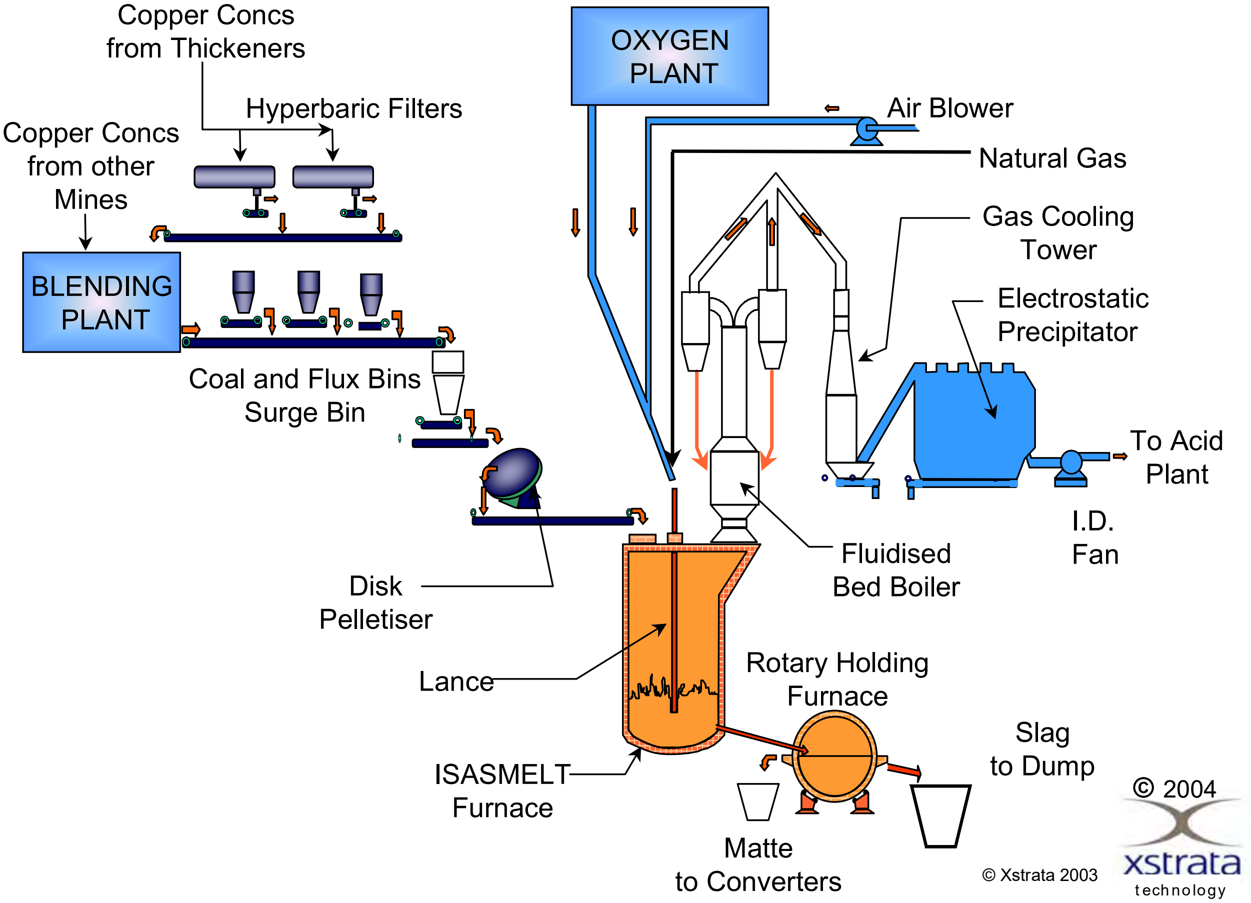


Figure 1- Flowsheet of the Copper ISASMELT^{TM} plant at Mount Isa.

Table I shows the typical process parameters for the CIP at Mount Isa. The copper ISASMELT^{TM} furnace treats about one million tonnes of copper concentrate per year with a normal target feed rate of 170 t/h (dry, unfluxed basis), a moisture content of 8% and a SiO_2/Fe ratio target of 0.85. The copper matte target is 60%. Typical oxygen content in the enriched air injected through the lance is 60%. Bath temperature is controlled to 1180 °C +/- °5 C.

Table II shows typical raw materials composition fed into the ISASMELT^{TM} furnace at Mount Isa. As can be noted, lead is the main impurity present in the concentrate with a typical concentration of 2,000 ppm. Zinc, cobalt and arsenic are normally present in levels between 1,000 and 1,500 ppm. Bismuth is present in levels close to 100 ppm while antimony levels are near 30 ppm.

Table I- Typical Process Parameters for Copper ISASMELT^{TM} Plant at Mount Isa

Parameter	**Value**	**Unit**
Maximum Feed rate	194	dry t/h
Normal target feed rate	170	dry t/h
Average Cu content in concentrate	25.8	%
Moisture	7 - 8	%
Average silica flux feed rate	3.7	dry t/h
Average coal feed rate	1.2	dry t/h
Average reverts feed rate	3.0	dry t/h
Average copper matte grade	59.5	%
SiO_2/Fe in slag	0.85 –	0.90
Average total lance flow rate	11.9	Nm^3/s
Oxygen Enrichment	60	%
Bath Temperature	1180	°C

Table II- Typical Process Parameters for Copper ISASMELT^{TM} Plant at Mount Isa

Element/Species	**Typical Value**	**Unit**
Cu	25.8	%
Fe	24.0	%
S	25.0	%
Pb	2,000	ppm
Zn	1,000	ppm
Co	1,100	ppm
As	1,500	ppm
Sb	30	ppm
Bi	120	ppm
SiO_2	16	%
CaO	0.8	%
MgO	1.0	%
Fe_3O_4	1.0	%
Al_2O_3	0.9	%

Minor Element Distribution

Distributions of As, Pb, Zn and Co between matte, slag and gas phases in the ISASMELT^{TM} furnace were calculated from the monthly mass balance and average chemical compositions drawn from plant data collected since 2002. Copper content in the feed during this period was almost constant. The distribution of minor elements to the gas phase was determined by ratio between the mass balance of the minor element in the feed and the condensate phases. In case of Bi and Sb, due to the low concentration in the feed, a four day measurement campaign was performed to accurately determine their concentration in the molten phase.

Figure 2 shows the arsenic distribution between the matte, D_{matte}, slag, D_{Slag}, and gas, D_{Gas}, phases, at a copper matte grade of 60%, bath temperature of 1180 °C and 60% oxygen enrichment, for arsenic contents in the feed between 0.10 and 0.26%. The symbols correspond to the calculated values (circles: matte, triangles: slag) while the lines shows the regression determined from the calculated values (continuous line: matte, dotted line: slag). The volatilisation of arsenic increases with increasing arsenic content in the feed achieving values close to 90%. The arsenic deportment to the slag phase decreases by increasing the arsenic content in the feed.

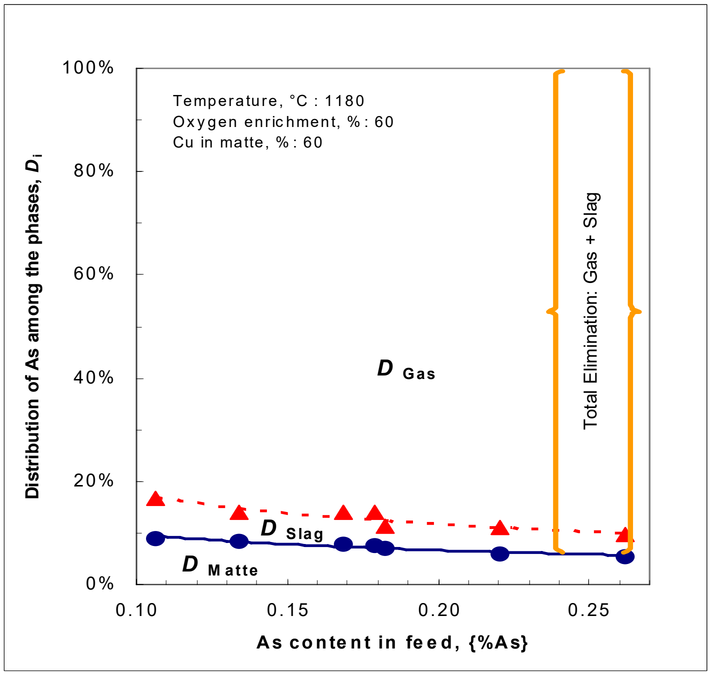


Figure 2 - Arsenic distribution between the matte, slag and gas phases (Cu matte grade: 60%, *T*: 1180 °C, Oxygen enrichment: 60%, $\{\%As\}_{feed}$: 0.10 and 0.26%.

Figure 3 shows the relationship between the arsenic content in the feed and the arsenic in the copper matte (full line, primary y-axis) and in the slag (dotted line, secondary y-axis). As can be noted, the arsenic content in the matte increases with increasing arsenic content in the feed. A similar behaviour is observed for the arsenic content in the slag phase.

Figure 4 shows the calculated lead distributions between the matte, slag and gas phases, at a copper matte grade of 60%, bath temperature of 1170 °C and 60% oxygen enrichment, for lead contents in the feed between 0.10% to 0.26 %. The symbols correspond to the determined distribution values while the lines show the calculated regression curves for the matte and slag distributions. The results obtained show a slight decrease in the lead deportment to the matte as the lead content in the feed increases. Removal rates of 35% - 40% are achieved for lead.

Figure 5 shows the relationship between lead content in the feed and its content in the matte (continuous line) and slag (broken line) phases, respectively. Lead content in the matte and slag phases increases when the lead content in feed increases.

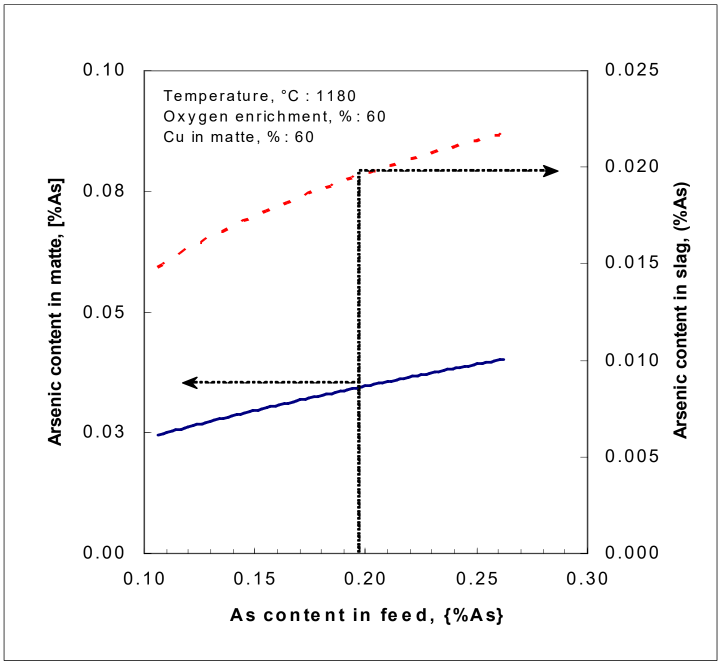


Figure 3 - Relationship between arsenic content in the feed and its content in copper matte (full line, primary y-axis) and slag (dotted line, secondary y-axis)

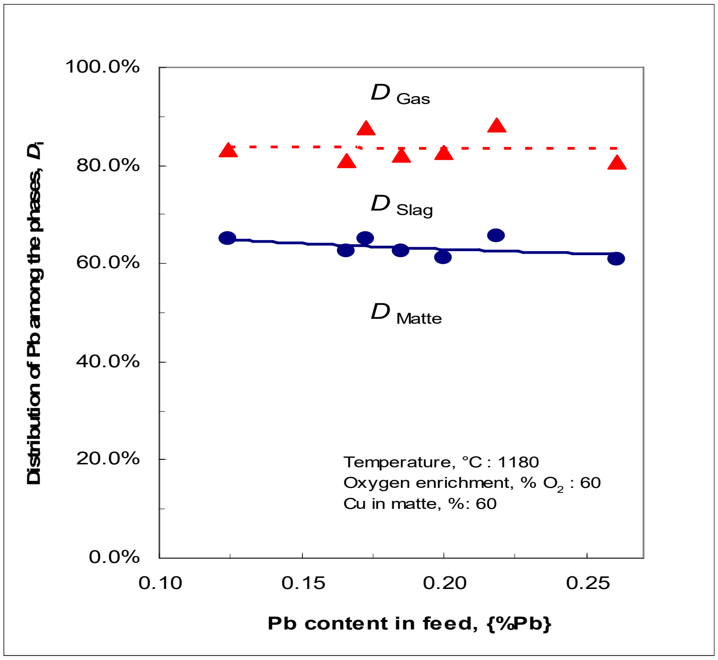


Figure 4 - Lead distribution between the matte, slag and gas phases (Cu matte grade: 60%, *T*: 1180 °C, Oxygen enrichment: 60%, $\{\%Pb\}_{feed}$: 0.10 and 0.26%.

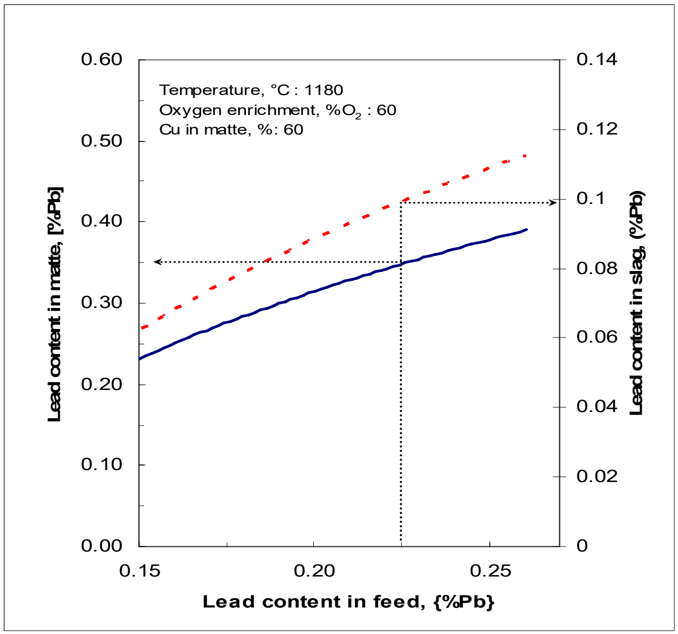


Figure 5 -Relationship between lead content in the feed and its content in copper matte (full line, primary y-axis) and slag (dotted line, secondary y-axis).

Figure 6 shows the relationship between the zinc content in the feed and its distribution to the matte, slag and gas phases, respectively, at a copper matte grade of 60%, bath temperature of 1180 °C and 60% oxygen enrichment, for zinc contents in the feed between 0.025% and 0.175%. The symbols correspond to the calculated values (circles: matte, triangles: slag) while the lines shows the regression determined from the calculated values (continuous line: matte, dotted line: slag). The zinc distribution to the matte, slag and gas phases remain constant by increasing the zinc content in the feed. Zinc is mainly removed to the slag phase. Figure 7 shows the relationship between zinc content in the feed and its concentration in the matte (continuous line) and slag (broken line) phases, respectively.

The relationship between the cobalt content in the feed and its distribution to the matte, slag and gas phases is shown in figure 8, for 60% copper matte grade, bath temperature of 1170 °C and 60% oxygen enrichment, for cobalt contents in the feed between 0.006% and 0.012%. The symbols correspond to the calculated values (circles: matte, triangles: slag) while the lines shows the regression determined from the calculated values (continuous line: matte, dotted line: slag). The calculated distribution shows that cobalt removal to the matte phase and slag phases increases by increasing its content in the feed. Figure 9 shows the relationship between cobalt content in the feed and its concentration in the matte (continuous line) and slag (broken line) phases, respectively.

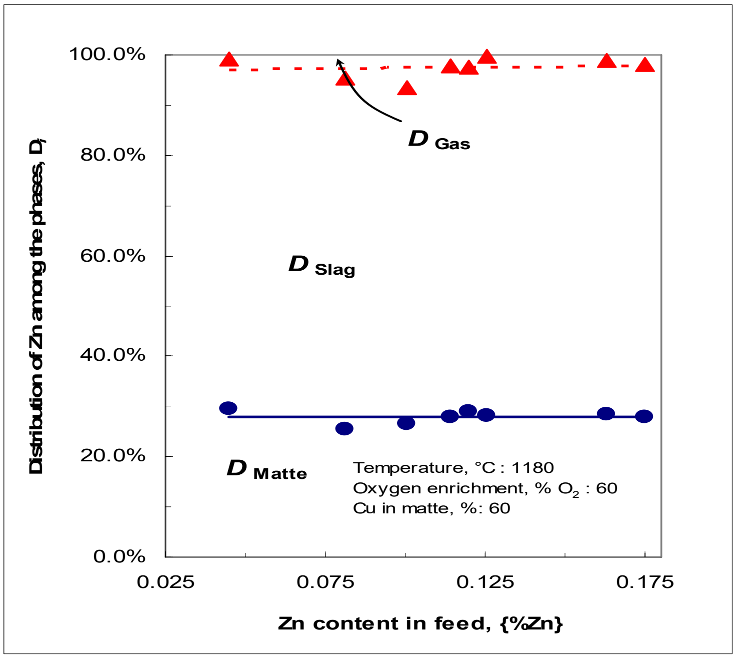


Figure 6 - Zinc distribution between the matte, slag and gas phases (Cu matte grade: 60%, *T*: 1180 °C, Oxygen enrichment: 60%, $\{\%Zn\}_{feed}$: 0.025 and 0.175%.

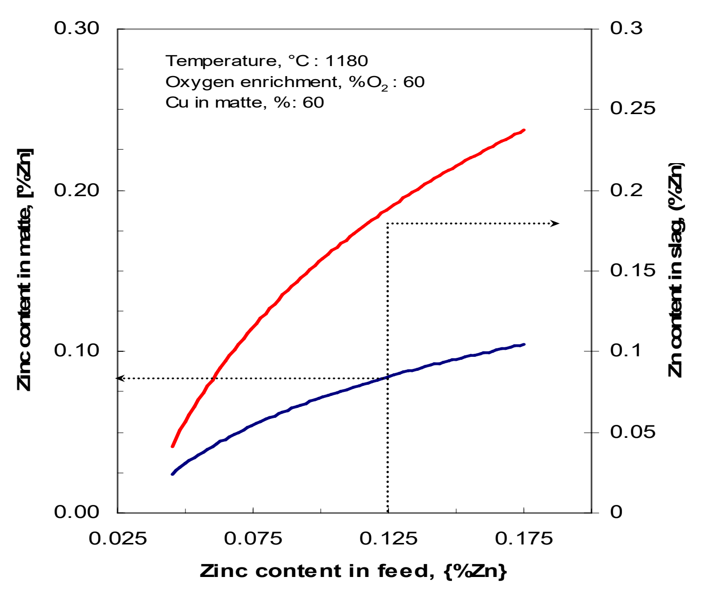


Figure 7 -Relationship between zinc content in the feed and its content in copper matte (full line, primary y-axis) and slag (dotted line, secondary y-axis).

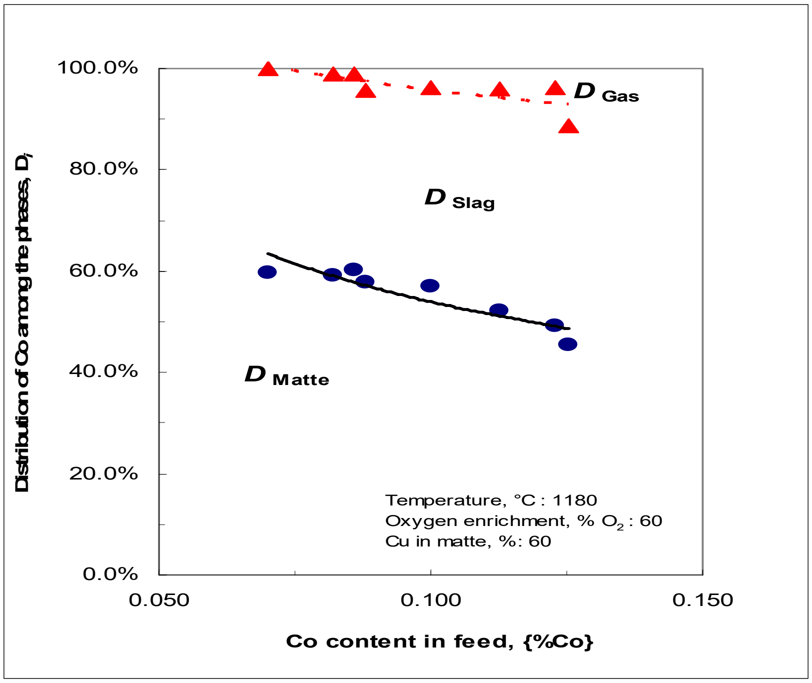


Figure 8 - Cobalt distribution between the matte, slag and gas phases (Cu matte grade: 60%, *T*: 1180 °C, Oxygen enrichment: 60%, $\{\%Co\}_{feed}$: 0.006% and 0.012%.

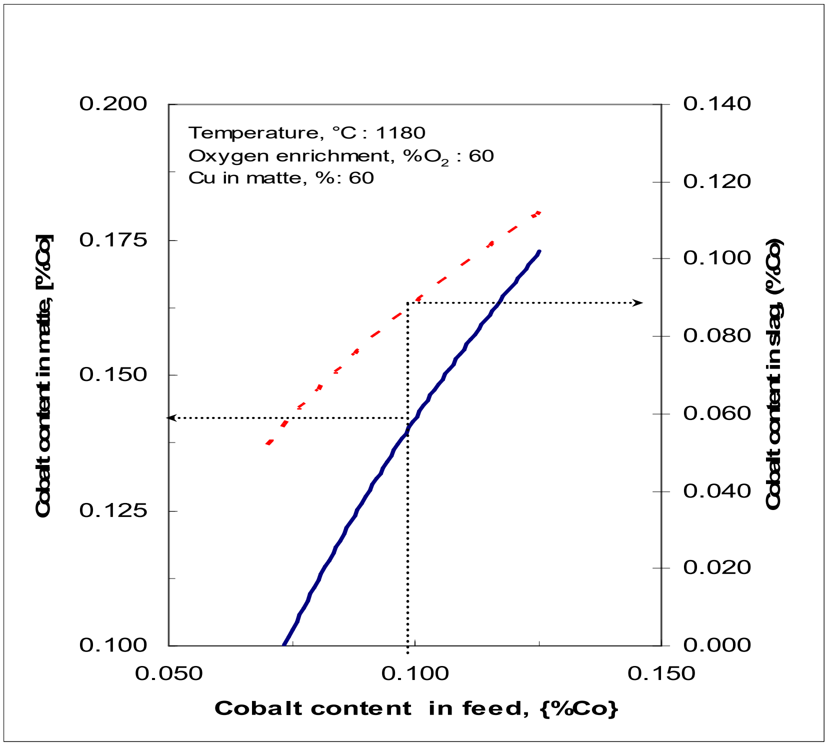


Figure 9 -Relationship between cobalt content in the feed and its content in copper matte (full line, primary y-axis) and slag (dotted line, secondary y-axis).

The concentration of antimony in the feed to the ISASMELT^{TM} at Mount Isa is usually too low to obtain accurate distribution values. Chemical analysis precision at 1 ppm level are required to ensure a proper determination of its concentration in the matte and slag phases, respectively. To be able to determine the distribution of antimony and bismuth in the ISASMELT^{TM} furnace, two measurement campaigns were carried out at Mount Isa Smelter in Australia and Yunnan Copper Corporation (YCC) in China. Table 3 shows the operational parameters for each ISASMELT^{TM} furnace, under the measurements carried out, and the results achieved.

Table III - Distribution of antimony between matte, slag and gas phases observed at Mount Isa and YCC

Parameter	**Unit**	**Mount Isa**	**YCC**
Concentrate Feed Rate	tph	140.0	103.0
Cu content, concentrate	%	23.0	22.5
Moisture content	%	8.0	10.0
Oxygen Enrichment, O_2	%	61.2	50.0
Copper matte grade	%	60.9	55.0
Sb content in feed	%	0.0035	0.0150
Sb Distribution			
D_{matte}	%	19.0	31.0
D_{Slag}	%	9.0	3.0
D_{Gas}	%	72.0	66.0

Volatilisation of antimony varies between 66 and 72%. Itagaki reported an antimony volatilisation of 62% under air blowing conditions [10]. These values seem to be higher than normally expected volatilisation values. However, the antimony volatilisation is enhanced by the contribution to the gas volume of the moisture content in the feed.

ISASMELT^{TM} for Treating High Arsenic Concentrates

Copper smelting relies on effective minor element control. This is required because of the mineralogical nature of sulfide ores which are normally accompanied with a number of minor elements in addition to the gangue.

For new projects that might consider the installation of smelting processes, it will be necessary to evaluate the ability of each smelting technology to volatilize or remove to the slag phase the minor elements present in the copper concentrates. Moreover, special attention should be paid to the slag cleaning and settling processes that might result in back contamination of the recovered copper matte due to the reducing conditions used in the slag cleaning reactor.

A potential option to overcome these difficulties while producing a copper matte of acceptable quality for the converting stage is the combination of the ISASMELT^{TM} furnace for smelting concentrates and the RHF for the gravity separation of matte and slag. The settling process performed in the RHF can be performed under a neutral atmosphere with minimum addition of reductant. This help to minimize any back contamination of the copper matte with impurities and therefore, ensure an effective removal of the impurities to the gas phase (in the ISASMELT™ furnace) and the slag phase.

As was shown in figure 2, the elimination of arsenic to the gas phase is very high in the ISASMELT™ furnace. Table 4 presents a comparison for the arsenic removal between ISASMELT, Flash Furnace, Mitsubishi Process, and Teniente Converter technologies.

Table IV – Arsenic Removal Comparison for Different Copper Smelting Technologies

Technology	**Ref.**	**As in Feed, {%As}**	**Oxygen Enrich-ment, %O_2**	**Matte Grade, %Cu**	D_{Matte}	D_{Slag}	D_{Gas}
Teniente Converter	[11]	0.30	36	70 - 75	12-20	13 - 38	75 - 42
Mitsubishi Process	[12]	0.30	48	68%	19.0	27.2	53.8
Flash Furnace	[11]	0.30	60	60 - 65	35	32	33
ISASMELT™	PW*	0.26	60	60	5.5	4.2	90.3

* Present Work

Although the matte grade, temperature and oxygen enrichment conditions are different for each process, the following statements can be made:

≠ Despite its lower matte grade and higher oxygen enrichment, the ISASMELT™ furnace shows the largest arsenic elimination to the gas phase

≠ High arsenic deportment to the slag phase followed by pyrometallurgical copper recovery processes under reducing conditions, as occurs with the Teniente Converter process, will result in back contamination of the copper matte

≠ In case of the ISASMELT™ furnace, even a settling process under reducing conditions in an electric furnace will not excessively back contaminate the copper matte with arsenic due to the very low distribution to the slag phase.

The high arsenic volatilisation achieved in the ISASMELT™ furnace is attributed to the strong bath agitation that provides a thorough mixing of the matte and gas phases allowing arsenic to saturate the gas phase. In addition, the positive impact of the moisture content in the feed increases the volume of gas available to remove arsenic from the bath.

The capability of the ISASMELT™ furnace to treat high arsenic content concentrates was evaluated based on the values observed at Mount Isa. Table 5 shows the expected arsenic distribution for the ISASMELT™ furnace as a function of the arsenic content in the feed for 60% copper matte, 60% oxygen enrichment at 1180°C (Assuming 25% of copper in concentrate). The results presented in table 5 indicate that the ISASMELT™ furnace an excellent reactor for treating high arsenic content copper concentrates.

Table V – Expected ISASMELT^{TM} furnace arsenic distribution for high As concentrates

Arsenic in the feed, {%As}	Estimated As distribution for ISASMELT^{TM}			Estimated As distribution for Teniente Converter S = 0.7			Estimated As distribution for Flash Furnace S = 0.3		
	D_{Matte}	D_{Slag}	D_{Gas}	D_{Matte}	D_{Slag}	D_{Gas}	D_{Matte}	D_{Slag}	D_{Gas}
0.26	5.5	4.2	90.3						
0.50	4.0	3.0	92.0	22	22	56	32	42	26
1.00	3.0	2.0	95.0	15	15	70	26	36	38
3.00	1.0	1.0	98.0	8	6	86	16	20	64

In table 5 the estimated arsenic distributions for the ISASMELT^{TM} furnace are compared with reported arsenic distributions for the Teniente Converter (36% oxygen enrichment, 1270 °C, 75% matte grade) and the Flash Furnace (55% oxygen enrichment, 1290 °C, 65% matte grade) (ref). Table 5 shows that the ISASMELT^{TM} furnace has a higher elimination of arsenic to the gas phase than both the Teniente converter and the Flash Furnace. The ISASMELT^{TM} furnace also has a lower deportment of arsenic to the slag phase than the Teniente converter. This means that there is less risk of back contamination of the matte during the slag cleaning process. Finally, it should be noted that the high arsenic removal expected in the ISASMELT^{TM} furnace will reduce the impurity removal duty for the converting and anode refining processes compared with the other two technologies.

Conclusions

1. Distribution impurities for As, Sb, Pb, Zn and Co have been calculated for the ISASMELT^{TM} furnace at the Mount Isa Copper Smelter

2. The ISASMELT^{TM} process has been demonstrated to remove impurities efficiently, through high levels of volatilisation. This behaviour is promoted by attributes of the process such as the strong bath agitation, the flexible nature of the process design allowing the operator to choose the optimal matte grade for impurity partitioning, and the positive effect of the moisture content in the feed

3. The ISASMELT^{TM} furnace has a higher arsenic elimination to the gas phase than the Flash Furnace or the Teniente converter

4. The ISASMELT^{TM} furnace has a lower arsenic deportment to the slag phase than the Teniente converter resulting in less risk of back contamination of the matte in the slag cleaning process

5. The high arsenic removal expected in the ISASMELT^{TM} furnace will reduce the impurity removal duty for the converting and anode refining processes compared with the other two technologies

Acknowledgements

The authors would like to thank Mount Isa Mines and Yunnan Copper for permission to publish the plant data. Also, special thanks are due to Marta Bailey and Lauren Fung at Mount Isa for their assistance with data collection.

References

(1) W.J. Errington, J.H. Fewings, V.P.Keran, and W.T. Denholm, "The Isasmelt Lead Smelting Process." *Transactions of the Institution of Mining and Metallurgy*, Section C, 96 (1987), 1-6.

(2) M.D. Coulter and C.R. Fountain, "The Isasmelt Process for Copper Smelting", *Non-ferrous Smelting Symposium*, (Port Pirie, South Australia, 17-21 September 1989, Aus.I.M.M., Melbourne, 1989), 237-240.

(3) C.R. Fountain, M.D. Coulter, and J.S. Edwards, "Minor Element Distribution in the Copper Isasmelt Process," *Copper '91, Vol. IV*, ed. C. Diaz, C. Landolt, A Luraschi, and C.J. Newman (Pergamon Press, New York, 1991), 359-373.

(4) R. Player, C.R. Fountain, T.V. Nguyen, and F.R. Jorgensen, "Top-entry Submerged Injection and the Isasmelt Technology," *Proceedings of the Savard/Lee International Symposium on Bath Smelting*, ed. J.K. Brimacombe, P.J. Mackey, G.J.W. Kor, C. Bickert and M.G. Ranade (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 1992), 215-229.

(5) J.L. Cribb, J.S. Edwards, C.R. Fountain, and S.P. Matthew, "Isasmelt Technology for the Smelting of Copper," *15th CMMI Congress*, Vol. 2 (SAIMM, Johannesburg, 1994), 99-103.

(6) R. Player, "Copper ISASMELT - Process Investigations", *The Howard Worner International Symposium on Injection in Pyrometallurgy*, ed. M. Nilmani and T. Lehner, Melbourne, (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 1996), 439-446.

(7) W.J. Errington, P.S. Arthur, C.R. Fountain, "ISASMELT - Clean Efficient Smelting", *GME'99 Global Metals Environment Conference, Beijing, 24-27 May 1999*, (Nonfer-rous Metals Society of China, Beijing, 1999), 164-172.

(8) J.S. Edwards, C.R. Fountain, and R.L. Morland, "ISASMELT—Extending the Envelope"; *Proceedings of the Brimacombe Memorial Symposium, October 1 to 4, 2000)* ed. by W. Poole: Poster Session Proceedings, (CIM, October 2000).

(9) J.L. Bill, T.E. Briffa, A.S. Burrows, C.R. Fountain, D. Retallick, J.M.I. Tuppurainen, J.S. Edwards, and P. Partington, "ISASMELT—Mount Isa Copper Smelter Progress Update", *Sulfide Smelting 2002*, ed. R.L. Stephens and H.Y. Sohn, (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 2002), 181-193.

(10) K. Itagaki, "Thermodynamic Evaluation of Distribution Behaviour of VA Elements and Effect of the Use of Oxygen in Copper Smelting", *Metallurgical Review of the Mining and Materials Processing Institute of Japan, MMIJ*, 3 (1986), 87-100.

(11) J. Font, G. Alvear, A. Moyano and C. Caballero, "Fractional distribution of arsenic in the Teniente Continuos Converting Process", *Arsenic Metallurgy*, ed. R. G. Reddy and V. Ramachadran, (TMS USA, 2005), 195-205.

(12) S. Surpunt and N. Hasegawa, "Distribution behavior of arsenic, antimony and bismuth in the smelting stage of the Mitsubishi process", *Proceedings of the Yazawa International Symposium, Vol I,* ed. F. Kongoli, K. Itagaki, C. Yamauchi, H. Sohn, (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 2003), 375-381.

Международный симпозиум по передовым технологиям переработки металлов и материалов. Ред. Флориан Конголи
TMS (Общество по минералам, металлам и материалам), 2006

ТЕХНОЛОГИЯ ВЫПЛАВКИ МЕДИ ISASMELT: УДАЛЕНИЕ ПРИМЕСЕЙ

Жерардо Р. Алвеар[1], Саймон П. Хант[1], Банци Чжан[2]
[1]Xstrata Technology, 87 Wickham Terrace, Brisbane, 4000, Australia
isasmelt@xstratatech.com.au
[2]Yunnan Copper Co., Ltd., Wang Jia Qiao, Wuhua District, Kunming 650102, P.R. China

Ключевые слова: ISASMELT, мышьяк, распределение, испарение примесей

Автореферат

Способность эффективно удалять примеси, содержащиеся в концентратах, является ключевым критерием при выборе технологии выплавки меди для новых или модернизируемых производств. Желательно, чтобы испарение таких примесей, как мышьяк и сурьма, происходило как можно раньше в процессе плавки, и желательно в плавильной печи, чтобы примеси не влияли на последующие технологические операции.

Технология ISASMELT™ позволяет эффективно удалять примеси путем интенсивного испарения. Этому способствуют такие особенности технологии, как интенсивное перемешивание ванны, а также гибкость технологического процесса, позволяющая оператору выбирать оптимальное для отделения примесей содержание штейна, и положительная роль влажности питания.

В данной работе рассматривается распределение второстепенных примесей в медеплавильной печи ISASMELT™. В первой части работы представлены данные о распределении примесей As, Sb, Pb, Zn и Co в печи ISASMELT™ на медеплавильном заводе Маунт Айза. Во второй части, исходя из данных, полученных на заводе, и соображений термодинамики, рассматриваются возможности применения технологии ISASMELT™ для переработки сложных концентратов.

Введение

Xstrata является международной компанией горно-добывающей и перерабатывающей промышленности, производящей преимущественно медь, ферросплавы, уголь и цинк. Рудники, металлургические и рафинировочные заводы Xstrata расположены в Австралии, ЮАР, Испании, Германии, Англии и Аргентине, а главный офис компании находится в Швейцарии. В 2003 году Xstrata приобрела австралийскую компанию MIM Holdings Limited (MIM) вместе с технологиями, разработанными MIM на протяжении более чем 80 лет эксплуатации предприятий по добыче и производству меди, свинца и угля в Австралии и Европе. Одной из этих технологий была технология ISASMELT^{TM}.

ISASMELT^{TM} - это простая и экономически эффективная технология выплавки цветных металлов. В настоящее время технологию развивает, продает и поддерживает компания Xstrata Technology. Данная технология относится к числу самых эффективных и гибких пирометаллургических технологий среди существующих.

Разработка технологии ISASMELT^{TM} от пилотной установки до промышленного применения была проведена в 1980-е и 1990-е годы на медеплавильном заводе MIM Маунт Айза в Австралии. Она быстро стала ключевым компонентом металлургических предприятий MIM. В настоящее время технология применяется в промышленном масштабе восемью компаниями в семи странах мира, а в 2006 году ожидается пуск еще двух заводов в Замбии и Перу. История разработки технологии описана в ряде публикаций [1-9].

Технология способна перерабатывать различные виды питания, от минеральных концентратов до металлолома и вторичных материалов металлургического производства. Требования к подготовке питания просты и, в отличие от некоторых других технологий выплавки меди, не предполагают сушки концентрата.

Важнейшим элементом технологии ISASMELT^{TM} является специальная фурма ISASMELT^{TM}. Фурма вставляется в ванну с расплавом печи ISASMELT^{TM}, представляющей собой стационарную вертикальную емкость с огнеупорной футеровкой. Фурма используется для продувки расплава воздухом (или воздухом, обогащенным кислородом), создающей высокую турбулентность в ванне с расплавом. Слой шлака, застывшего на наружной поверхности фурмы ISASMELT^{TM}, защищает ее от воздействия расплава. Установленный внутри фурмы завихритель увеличивает скорость технологического воздуха внутри фурмы, таким образом увеличивая интенсивность теплопередачи между сталью корпуса фурмы и струей технологического воздуха. Завихритель имеет особую конструкцию, обеспечивающую минимальное падение давления, что позволяет поддерживать общее падение давления в фурме на уровне не выше 80 кПа. Поэтому для подачи технологического воздуха в фурму достаточно одноступенчатого нагнетателя. Фурма имеет простую и надежную конструкцию. Текущий ремонт фурмы занимает не более 6 часов. Срок службы фурмы составляет, как правило, 10 - 20 дней.

Питание, загружаемое в турбулентную ванну сверху, быстро вступает в реакцию, что обеспечивает чрезвычайно высокую производительность при относительно небольшом объеме ванны. На медеплавильной печи ISASMELT^{TM}, установленной на заводе Маунт Айза, за час подачи медьсодержащего питания (концентрат, оборотные материалы, иные внутренние оборотные материалы завода) выплавлялось до 194 т продукта при полном объеме ванны около 15 м^3.

С момента начала разработки технологии ISASMELT^{TM} была отмечена ее способность эффективно удалять примеси за счет интенсивного испарения (3, 4). Этому способствуют такие особенности технологии, как интенсивное перемешивание ванны, а также гибкость технологического процесса, позволяющая оператору выбирать оптимальное для отделения примесей содержание штейна, и положительная роль влажности питания.

В настоящей работе представлены данные о распределении примесей в медеплавильной печи ISASMELT^{TM} на заводе Маунт Айза. Затем, исходя из данных о распределении примесей, полученных на заводе Маунт Айза, дополнительных эксплуатационных данных печи ISASMELT^{TM}, установленной на предприятии Yunnan Copper Corporation, и соображений термодинамики, рассматриваются возможности применения технологии ISASMELT^{TM} для переработки сложных концентратов.

Медеплавильный завод Маунт Айза

Описание завода

На рисунке 1 показана технологическая схема медеплавильного завода ISASMELT^{TM} (CIP) Маунт Айза. Цех первичной плавки состоит из системы хранения и подачи питания, печи ISASMELT^{TM}, системы отведения печных газов и двух вращающихся миксеров (RHF).

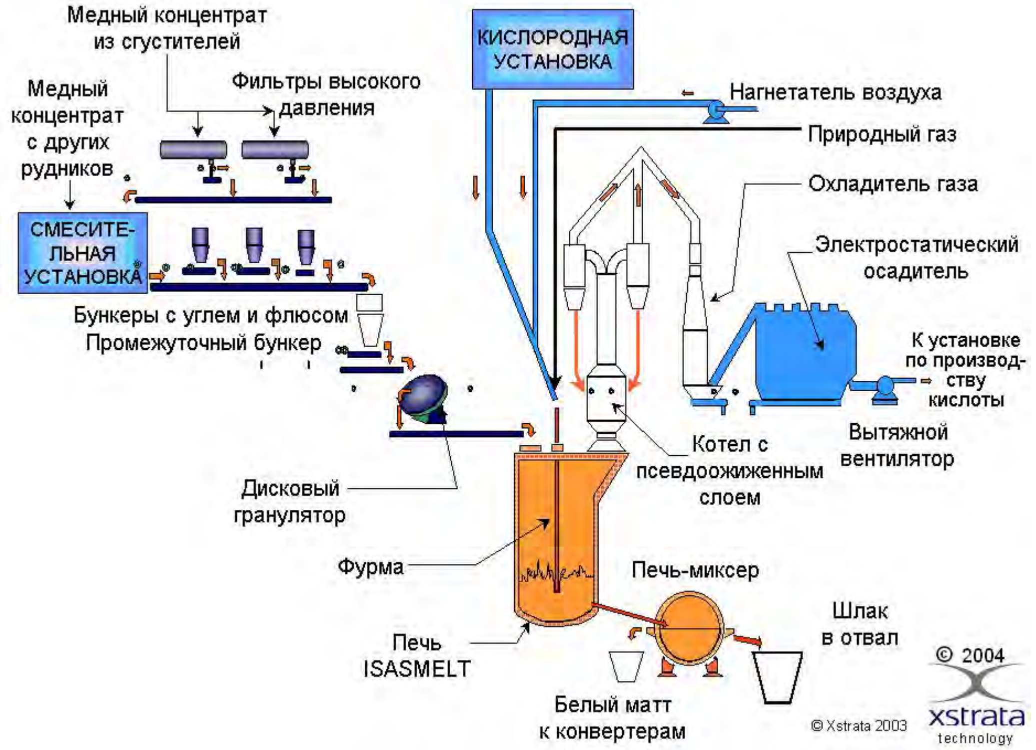


Рисунок 1 - Технологическая схема медеплавильного завода ISASMELT^{TM} Маунт Айза.

В таблице I показаны характерные технологические параметры медеплавильного завода CIP Маунт Айза. Медеплавильная печь ISASMELT^{TM} позволяет переплавлять примерно 1 млн. тонн медного концентрата в год при стандартной интенсивности питания 170 т/ч (в пересчете на сухое питание без учета флюса), содержании влаги 8% и стандартном отношении SiO_2/Fe 0,85. Стандартное содержание меди в штейне составляет 60%. Содержание кислорода в обогащенном воздухе, подаваемом через фурму, составляет 60%. Температура в ванне поддерживается на уровне 1180˚С +/- 5˚С.

В таблице II показан обычный материальный состав питания печи ISASMELT^{TM} на заводе Маунт Айза Как можно заметить, основной примесью в концентрате является свинец, содержание которого составляет, как правило, 2 000 м.д. Содержание цинка, кобальта и мышьяка составляет обычно 1 000 - 1 500 м.д. Содержание висмута близко к 100 м.д., сурьмы - к 30 м.д.

Таблица I - Характерные технологические параметры медеплавильной печи ISASMELT^{TM} на заводе Маунт Айза

Характеристика	**Значение**	**Ед. изм.**
Максимальная интенсивность питания	194	т/ч (сухое вещество)
Стандартная интенсивность питания	170	т/ч (сухое вещество)
Среднее содержание Cu в концентрате	25.8	%
Содержание влаги	7 - 8	%
Средняя интенсивность подачи кремнеземистого флюса	3.7	т/ч (сухое вещество)
Средняя интенсивность подачи угля	1.2	т/ч (сухое вещество)
Средняя интенсивность подачи оборотного материала	3.0	т/ч (сухое вещество)
Среднее содержание меди в штейне	59.5	%
Отношение SiO_2/Fe в шлаке	0.85 –	0.90
Средний расход воздуха в фурме	11.9	нм^3/с
Содержание кислорода	60	%
Температура ванны	1180	°C

Таблица II - Характерные технологические параметры медеплавильной печи ISASMELT^{TM} на заводе Маунт Айза

Хим. элемент/соединение	**Характерное значение**	**Ед. изм.**
Cu	25.8	%
Fe	24.0	%
S	25.0	%
Pb	2,000	м.д.
Zn	1,000	м.д.
Co	1,100	м.д.
As	1,500	м.д.
Sb	30	м.д.
Bi	120	м.д.
SiO_2	16	%
CaO	0.8	%
MgO	1.0	%
Fe_3O_4	1.0	%
Al_2O_3	0.9	%

Распределение примесей

Распределение As, Pb, Zn и Co между штейном, шлаком и газообразной фазой в печи ISASMELT^{TM} было определено на основании данных ежемесячного масс-баланса и среднего химического состава, собираемых на заводе с 2002 года. Содержание меди в питании на протяжении всего периода было практически постоянным. Доля примесей в газообразной фазе определялась отношением масс-балансов примесей в питании и в конденсате. Применительно к Bi и Sb, из-за низкой концентрации в питании для точного определения концентрации в расплаве измерения проводились на протяжении четырех дней.

На рисунке 2 показано распределение мышьяка между штейном, $D_{\text{штейн}}$, шлаком, $D_{\text{шлак}}$, и газообразной фазой, $D_{\text{газ}}$, при содержании меди в штейне 60%, температуре ванны 1180°C, содержании кислорода в обогащенном воздухе 60% и содержании мышьяка в питании от 0,10% до 0,26%. Символы соответствуют расчетным значениям (круг - штейн, треугольник - шлаг), а кривые обозначают регрессию, построенную согласно расчетным значениям (сплошная линия - штейн, пунктирная линия - шлак). Испарение мышьяка увеличивается с увеличением содержания мышьяка в питании, достигая значений, близких к 90%. Переход мышьяка в шлак уменьшается по мере увеличения содержания мышьяка в питании.

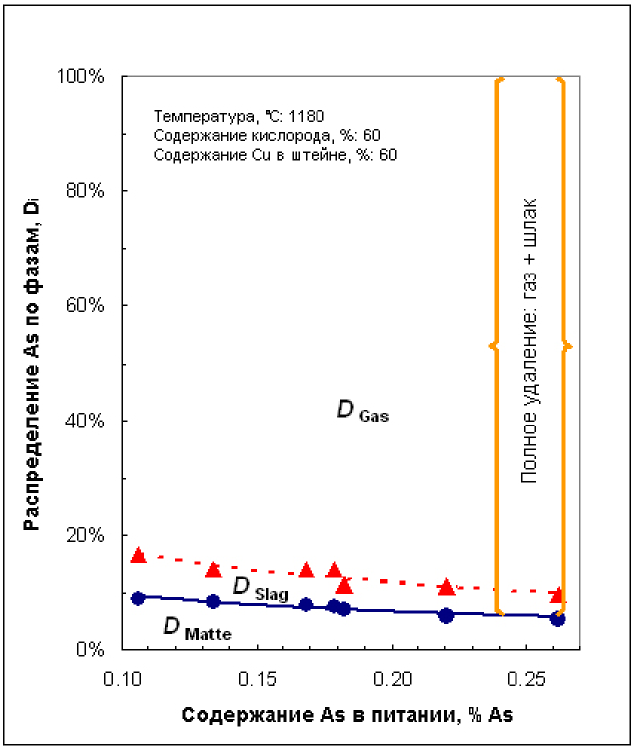


Рисунок 2 - Распределение мышьяка между штейном, шлаком и газообразной фазой (содержание Cu в штейне: 60%, *T*: 1180˚C, содержание кислорода: 60%, содержание As в питании: 0,10 и 0,26%.

На рисунке 3 показана зависимость между содержанием мышьяка в питании и содержанием мышьяка в медном штейне (сплошная линия, основная ось y) и в шлаке (пунктирная линия, дополнительная ось y). Как можно заметить, содержание мышьяка в штейне увеличивается по мере увеличения содержания мышьяка в питании. Аналогичная зависимость наблюдается применительно к содержанию мышьяка в шлаке.

На рисунке 4 показано расчетное распределение свинца между штейном, шлаком и газообразной фазой при содержании меди в штейне 60%, температуре ванны 1170˚C, содержании кислорода в обогащенном воздухе 60% и содержании свинца в питании от 0,10% до 0,26%. Символы соответствуют определенным значениям распределения, а кривые соответствуют кривым регрессии, построенным согласно определенным значениям. Полученные результаты указывают на небольшое уменьшение перехода свинца в штейн по мере увеличения содержания свинца в питании. Достигнутые показатели удаления свинца составляют 35-40%.

На рисунке 5 показана зависимость между содержанием свинца в питании и содержанием свинца в штейне (сплошная линия) и в шлаке (пунктирная линия). Содержание свинца в штейне и шлаке увеличивается по мере увеличения содержания свинца в питании.

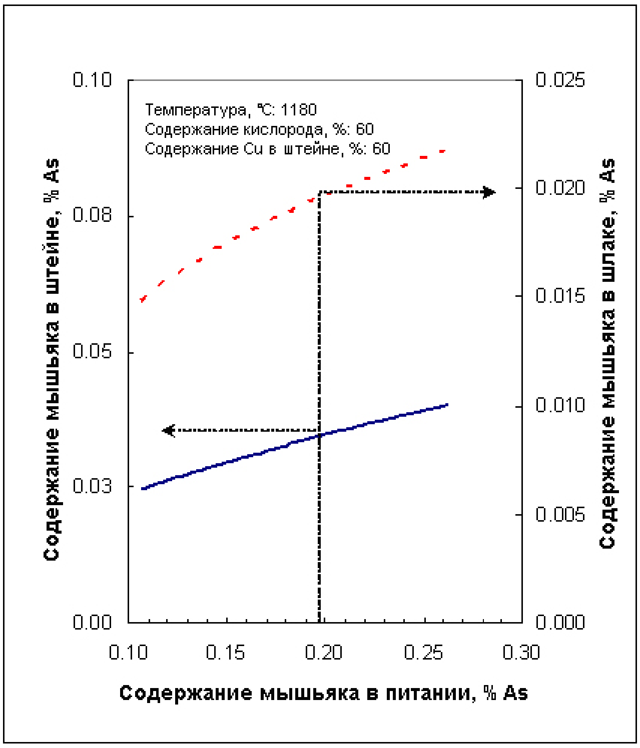


Рисунок 3 - Зависимость между содержанием мышьяка в питании и содержанием мышьяка в медном штейне (сплошная линия, основная ось y) и в шлаке (пунктирная линия, дополнительная ось y)

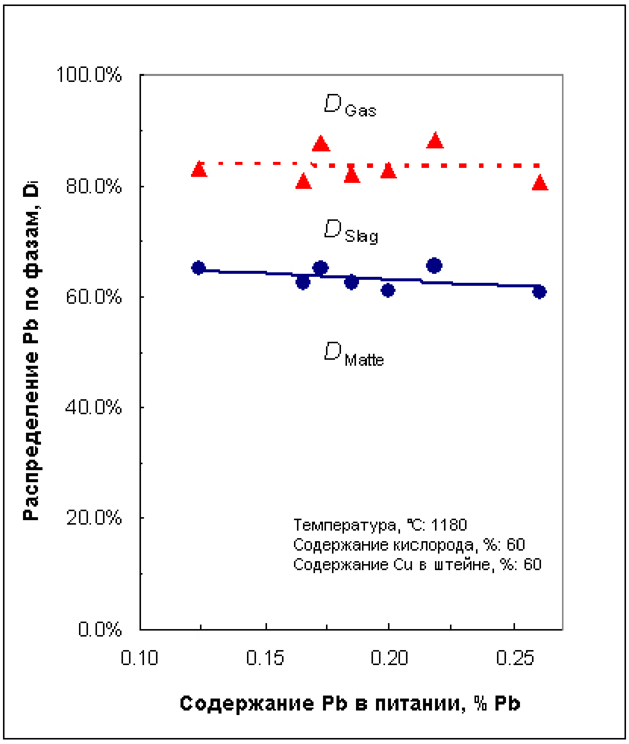


Рисунок 4 - Распределение свинца между штейном, шлаком и газообразной фазой (содержание Cu в штейне: 60%, T: 1180°C, содержание кислорода: 60%, содержание Pb в питании: 0,10 и 0,26%.

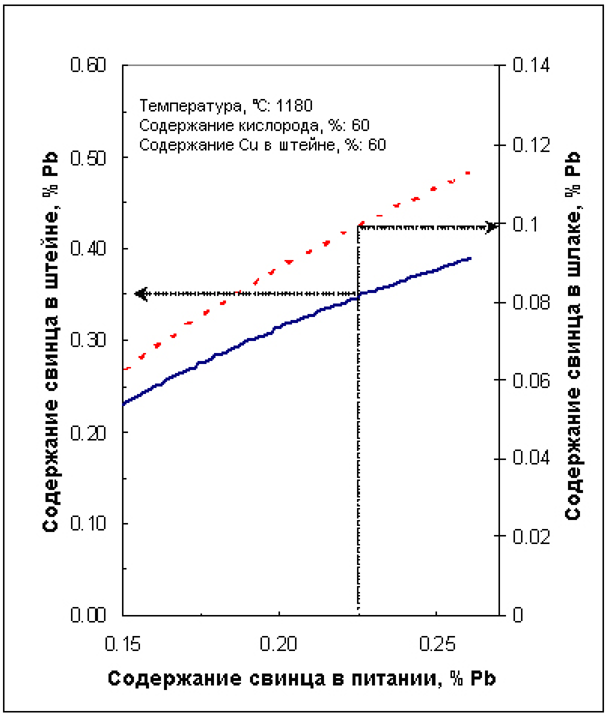


Рисунок 5 - Зависимость между содержанием свинца в питании и содержанием свинца в медном штейне (сплошная линия, основная ось у) и в шлаке (пунктирная линия, дополнительная ось у)

На рисунке 6 показана зависимость между содержанием цинка в питании и распределением цинка между штейном, шлаком и газообразной фазой при содержании меди в штейне 60%, температуре ванны 1180˚С, содержании кислорода в обогащенном воздухе 60% и содержании цинка в питании от 0,025% до 0,175%. Символы соответствуют расчетным значениям (круг - штейн, треугольник - шлаг), а кривые обозначают регрессию, построенную согласно расчетным значениям (сплошная линия - штейн, пунктирная линия - шлак). Распределение цинка между штейном, шлаком и газообразной фазой остается неизменным при увеличении содержания цинка в питании. Цинк преимущественно переходит в шлак. На рисунке 7 показана зависимость между содержанием цинка в питании и содержанием цинка в штейне (сплошная линия) и в шлаке (пунктирная линия).

На рисунке 8 показана зависимость между содержанием кобальта в питании и распределением кобальта между штейном, шлаком и газообразной фазой при содержании меди в штейне 60%, температуре ванны 1170˚С, содержании кислорода в обогащенном воздухе 60% и содержании кобальта в питании от 0,006% до 0,012%. Символы соответствуют расчетным значениям (круг - штейн, треугольник - шлаг), а кривые обозначают регрессию, построенную согласно расчетным значениям (сплошная линия - штейн, пунктирная линия - шлак). Расчетное распределение указывает на то, что переход кобальта в штейн и шлак увеличивается по мере увеличения его содержания в питании. На рисунке 9 показана зависимость между содержанием кобальта в питании и содержанием кобальта в штейне (сплошная линия) и в шлаке (пунктирная линия).

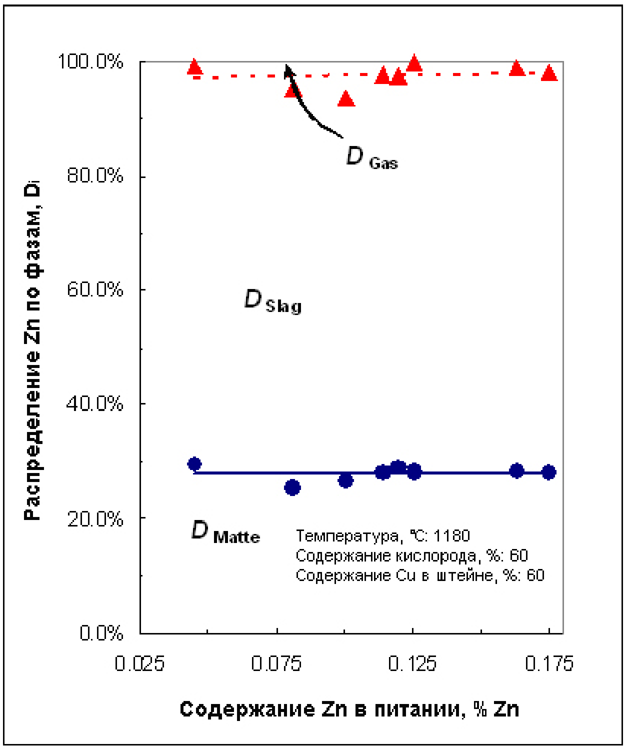


Рисунок 6 - Распределение цинка между штейном, шлаком и газообразной фазой (содержание Cu в штейне: 60%, Т: 1180˚С, содержание кислорода: 60%, содержание Zn в питании: 0,025 и 0,175%.

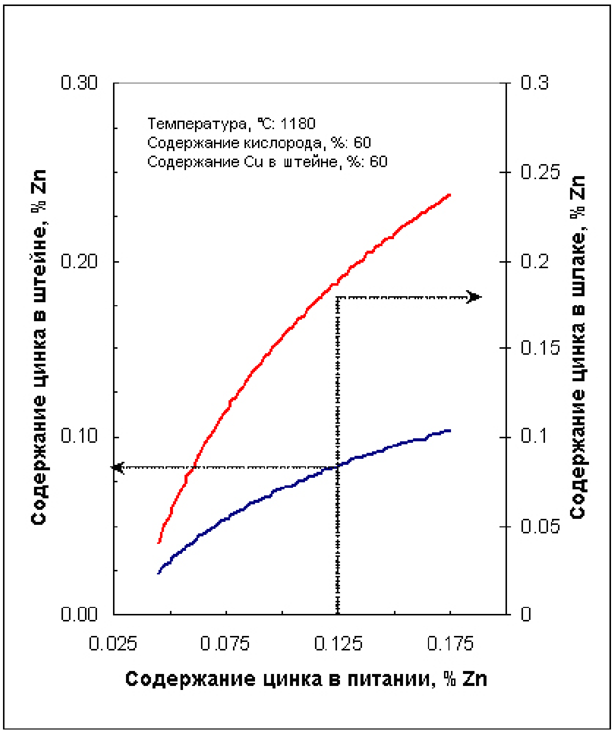


Рисунок 7 - Зависимость между содержанием цинка в питании и содержанием цинка в медном штейне (сплошная линия, основная ось y) и в шлаке (пунктирная линия, дополнительная ось y)

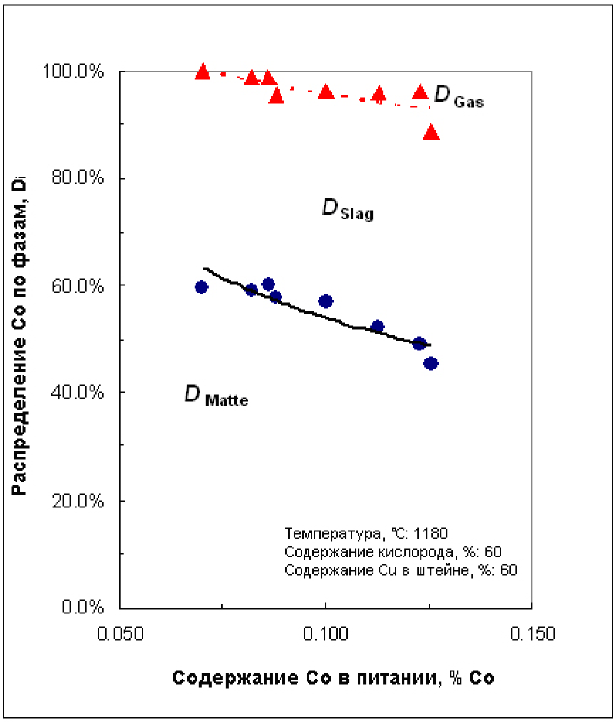


Рисунок 8 - Распределение кобальта между штейном, шлаком и газообразной фазой (содержание Cu в штейне: 60%, T: 1180˚C, содержание кислорода: 60%, содержание Co в питании: 0,006 и 0,012%.

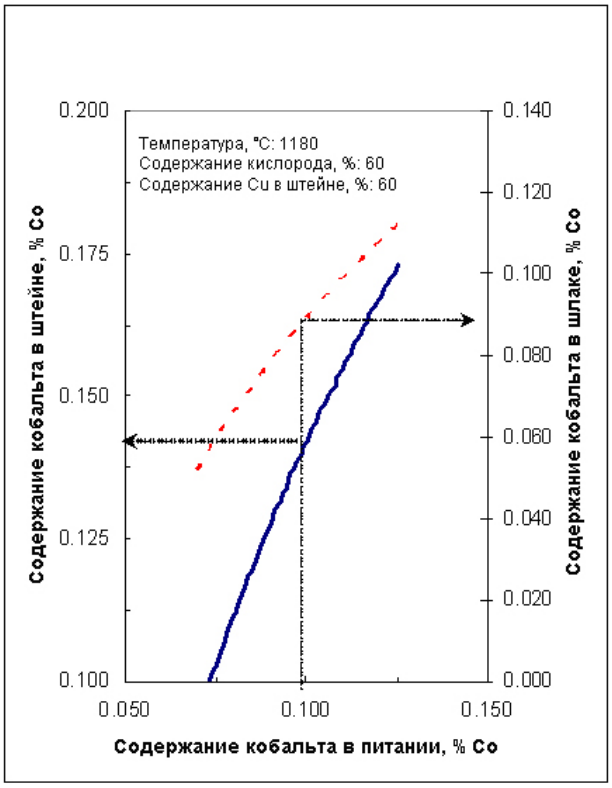


Рисунок 9 - Зависимость между содержанием кобальта в питании и содержанием кобальта в медном штейне (сплошная линия, основная ось y) и в шлаке (пунктирная линия, дополнительная ось y)

Концентрация сурьмы в питании печи ISASMELT^{TM} на заводе Маунт Айза, как правило, слишком мала для получения точных данных о распределении. Для определения его концентрации в штейне и шлаке требуется химический анализ с точностью до 1 м.д. Чтобы определить распределение сурьмы и висмута в печи ISASMELT^{TM}, были выполнены две программы измерений на медеплавильном заводе Маунт Айза в Австралии, и на заводе Yunnan Copper Corporation (YCC) в Китае. В таблице 3 показаны эксплуатационные параметры каждой из печей ISASMELT^{TM} и их значения.

Таблица III - Распределение сурьмы между штейном, шлаком и газообразной фазой на заводах Маунт Айза и YCC

Характеристика	**Ед. изм.**	**Маунт Айза**	**YCC**
Интенсивность подачи концентрата	т/ч	140.0	103.0
Содержание Cu в концентрате	%	23.0	22.5
Содержание влаги	%	8.0	10.0
Содержание кислорода, O_2	%	61.2	50.0
Содержание меди в штейне	%	60.9	55.0
Содержание Sb в питании	%	0.0035	0.0150
Распределение Sb			
$D_{штейн}$	%	19.0	31.0
$D_{шлак}$	%	9.0	3.0
$D_{газ}$	%	72.0	66.0

Испарение сурьмы составляет от 66 до 72%. В работе Итагаки сообщается об испарении сурьмы 62% при продувке воздухом [10]. Эти показатели выше обычных ожидаемых показателей испарения. Однако испарению способствует увеличение объема газа благодаря содержанию влаги в питании.

ISASMELT^{TM} при плавке концентратов с высоким содержанием мышьяка

При выплавке меди необходимо эффективное удаление примесей. Это требование связано с минералогической природой сульфидных руд, которые, как правило, помимо жильных минералов содержат целый ряд примесей.

Рассматривая возможность применения пирометаллургических технологий в новых проектах, необходимо оценить способность технологии испарять или удалять в шлак примеси, содержащиеся в медных концентратах. Кроме того, особое внимание следует уделять процессам очистки и осаждения шлака, при которых возможно повторное загрязнение полученного медного штейна из-за восстановительных условий в реакторе, используемом для очистки шлака.

Возможным решением этой проблемы, позволяющим получать медный штейн приемлемого качества на этапе конвертерной плавки, является использование печи ISASMELT^{TM} для плавки концентратов и миксера для гравитационного разделения штейна и шлака. Процесс осаждения, происходящий в миксере, может протекать в нейтральных условиях с добавлением минимального количества восстановителя. Это позволяет свести к минимуму загрязнение медного штейна примесями, а следовательно, обеспечивает эффективное удаление примесей в газообразную фазу (в печи ISASMELT™) и в шлак.

Как было показано на рисунке 2, печь ISASMELT^{TM} отличается очень высокими показателями перехода мышьяка в газообразную фазу. В таблице 4 сравниваются показатели удаления мышьяка в печи ISASMELT, печи взвешенной плавки, при использовании процесса Mitsubishi и конвертера Teniente.

Таблица IV - Удаление мышьяка при использовании различных технологий выплавки меди

Технология	**Ист.**	**As в питании, % As**	**Содержание кислорода, %O$_2$**	**Содержание меди в штейне, % Cu**	$D_{штейн}$	$D_{шлак}$	$D_{газ}$
Конвертер Teniente	[11]	0.30	36	70 - 75	35	13 - 38	75 - 42
Процесс Mitsubishi	[12]	0.30	48	68%	19.0	27.2	53.8
Печь взвешенной плавки	[11]	0.30	60	60 - 65	35	32	33
ISASMELT^{TM}	НР*	0.26	60	60	5.5	4.2	90.3

* Настоящая работа.

Хотя содержание штейна, температура и содержание кислорода отличаются, можно сделать следующие выводы:

- Несмотря на меньшее содержание меди в штейне и большее содержание кислорода в обогащенном воздухе, печь ISASMELT^{TM} отличается наиболее высоким показателем перехода мышьяка в газообразную фазу

- Высокий показатель перехода мышьяка в шлак с последующим пирометаллургическим извлечением меди при восстановительных условиях, как в конвертерной технологии Teniente, приведет к повторному загрязнению медного штейна

- При применении печи ISASMELT^{TM} даже осаждение при восстановительных условиях в электропечи не приведет к значительному повторному загрязнению медного штейна мышьяком благодаря очень малому переходу в шлак.

Высокий показатель испарения мышьяка в печи ISASMELT^{TM} предположительно связан с интенсивным перемешиванием ванны, обеспечивающим тщательное перемешивание штейна и газообразной фазы, способствующее насыщению газообразной фазы мышьяком. Кроме того, присутствие влаги в питании увеличивает объем газа, доступного для извлечения мышьяка из ванны.

Пригодность печи ISASMELT^{TM} к переработке концентратов с высоким содержанием мышьяка была оценена на основании данных, полученных на заводе Маунт Айза. В таблице 5 показано ожидаемое распределение мышьяка в печи ISASMELT^{TM} в зависимости от содержания мышьяка в питании при содержании меди в штейне 60%, кислорода 60% и температуре 1180˚C (если принять содержание меди в концентрате равным 25%). Данные таблицы 5 указывают на то, что печь ISASMELT^{TM} идеально подходит для плавки медных концентратов с высоким содержанием мышьяка.

Таблица V - Ожидаемое распределение мышьяка в печи ISASMELT^{TM} при плавке концентратов с высоким содержанием мышьяка

Содержание мышьяка в питании, % As	**Расчетное распределение As в ISASMELT^{TM}**			**Расчетное распределение As в конвертере Teniente S = 0.7**			**Расчетное распределение As в печи взвешенной плавки S = 0.3**		
	$D_{\text{штейн}}$	$D_{\text{шлак}}$	$D_{\text{газ}}$	$D_{\text{штейн}}$	$D_{\text{шлак}}$	$D_{\text{газ}}$	$D_{\text{штейн}}$	$D_{\text{шлак}}$	$D_{\text{газ}}$
0.26	5.5	4.2	90.3						
0.50	4.0	3.0	92.0	22	22	56	32	42	26
1.00	3.0	2.0	95.0	15	15	70	26	36	38
3.00	1.0	1.0	98.0	8	6	86	16	20	64

В таблице 5 сравнивается расчетное распределение мышьяка в печи ISASMELT^{TM} с распределением мышьяка в конвертере Teniente (содержание кислорода 36%, 1270˚C, содержание штейна 75%) и в печи взвешенной плавки (содержание кислорода 55%, 1290˚C, содержание меди в штейне 65%) (ист.). Из таблицы 5 следует, что печь ISASMELT^{TM} отличается большим переходом мышьяка в газообразную фазу по сравнению как с конвертером Teniente, так и с печью взвешенной плавки. Печь ISASMELT^{TM} также отличается меньшим переходом мышьяка в шлак по сравнению с конвертером Teniente. Это означает меньший риск повторного загрязнения штейна при очистке шлака. Наконец, необходимо отметить, что высокие ожидаемые показатели удаления мышьяка в печи ISASMELT^{TM} уменьшают количество примесей, требующих удаления при конвертировании и рафинировании анодной меди, по сравнению с двумя другими технологиями.

Заключение

1. Было определено распределение примесей As, Sb, Pb, Zn и Co в печи ISASMELT^{TM} на медеплавильном заводе Маунт Айза

2. Технология ISASMELT^{TM} позволяет эффективно удалять примеси путем интенсивного испарения. Этому способствуют такие особенности технологии, как интенсивное перемешивание ванны, а также гибкость технологического процесса, позволяющая оператору выбирать оптимальное для отделения примесей содержание штейна, и положительная роль влажности питания

3. ISASMELT^{TM} имеет более высокие показатели перехода мышьяка в газообразную фазу по сравнению с печью взвешенной плавки и конвертером Teniente

4. ISASMELT^{TM} отличается меньшим переходом мышьяка в шлак, чем конвертер Teniente, что уменьшает риск повторного загрязнения штейна при очистке шлака

5. Высокие ожидаемые показатели удаления мышьяка в печи ISASMELT^{TM} уменьшают количество примесей, требующих удаления при конвертировании и рафинировании анодной меди, по сравнению с двумя другими технологиями

Благодарности

Авторы выражаю признательность компаниям Mount Isa Mines и Yunnan Copper за разрешение на публикацию производственных данных. Также авторы выражают особую благодарность сотрудникам завода Маунт Айза Марте Бейли и Лорен Фанг за помощь в сборе данных.

Список использованной литературы

(1) W.J. Errington, J.H. Fewings, V.P.Keran, and W.T. Denholm, “The Isasmelt Lead Smelting Process.” *Transactions of the Institution of Mining and Metallurgy*, Section C, 96 (1987), 1-6.
(2) M.D. Coulter and C.R. Fountain, “The Isasmelt Process for Copper Smelting”, *Non-ferrous Smelting Symposium*, (Port Pirie, South Australia, 17-21 September 1989, Aus.I.M.M., Melbourne, 1989), 237-240.
(3) C.R. Fountain, M.D. Coulter, and J.S. Edwards, “Minor Element Distribution in the Copper Isasmelt Process,” *Copper ’91, Vol. IV*, ed. C. Diaz, C. Landolt, A Luraschi, and C.J. Newman (Pergamon Press, New York, 1991), 359-373.
(4) R. Player, C.R. Fountain, T.V. Nguyen, and F.R. Jorgensen, “Top-entry Submerged Injection and the Isasmelt Technology,” *Proceedings of the Savard/Lee International Symposium on Bath Smelting*, ed. J.K. Brimacombe, P.J. Mackey, G.J.W. Kor, C. Bickert and M.G. Ranade (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 1992), 215-229.

(5) J.L. Cribb, J.S. Edwards, C.R. Fountain, and S.P. Matthew, “Isasmelt Technology for the Smelting of Copper,” *15th CMMI Congress*, Vol. 2 (SAIMM, Johannesburg, 1994), 99-103.

(6) R. Player, “Copper ISASMELT - Process Investigations”, *The Howard Worner International Symposium on Injection in Pyrometallurgy*, ed. M. Nilmani and T. Lehner, Melbourne, (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 1996), 439-446.

(7) W.J. Errington, P.S. Arthur, C.R. Fountain, “ISASMELT - Clean Efficient Smelting”, *GME’99 Global Metals Environment Conference, Beijing, 24-27 May 1999*, (Nonfer-rous Metals Society of China, Beijing, 1999), 164-172.

(8) J.S. Edwards, C.R. Fountain, and R.L. Morland, “ISASMELT—Extending the Envelope”; *Proceedings of the Brimacombe Memorial Symposium, October 1 to 4, 2000)* ed. by W. Poole: Poster Session Proceedings, (CIM, October 2000).

(9) J.L. Bill, T.E. Briffa, A.S. Burrows, C.R. Fountain, D. Retallick, J.M.I. Tuppurainen, J.S. Edwards, and P. Partington, “ISASMELT—Mount Isa Copper Smelter Progress Update”, *Sulfide Smelting 2002*, ed. R.L. Stephens and H.Y. Sohn, (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 2002), 181-193.

(10) K. Itagaki, “Thermodynamic Evaluation of Distribution Behaviour of VA Elements and Effect of the Use of Oxygen in Copper Smelting”, *Metallurgical Review of the Mining and Materials Processing Institute of Japan, MMIJ*, 3 (1986), 87-100.

(11) J. Font, G. Alvear, A. Moyano and C. Caballero, “Fractional distribution of arsenic in the Teniente Continuos Converting Process”, *Arsenic Metallurgy*, ed. R. G. Reddy and V. Ramachadran, (TMS USA, 2005), 195-205.

(12) S. Surpunt and N. Hasegawa, “Distribution behavior of arsenic, antimony and bismuth in the smelting stage of the Mitsubishi process”, *Proceedings of the Yazawa International Symposium, Vol I*, ed. F. Kongoli, K. Itagaki, C. Yamauchi, H. Sohn, (The Minerals, Metals and Materials Society, Warrendale, Pennsylvania, 2003), 375-381.

Título del
documento:

ISASMELT™ – 25 AÑOS DE EVOLUCIÓN CONTINUA

Documento
presentado en:
Simposio Internacional John Floyd sobre el desarrollo sustentable en el procesamiento de metales. 3 al 6 de julio de 2005, Melbourne.

Autores:
P. S. Arthur, S. P. Hunt, Xstrata Technology

Fecha de la publicación
Julio de 2005

Para obtener más información, comuníquese con nosotros en
isasmelt@xstratatech.com.au

ISASMELT™ – 25 AÑOS DE EVOLUCIÓN CONTINUA

P. S. Arthur, S. P. Hunt

Xstrata Technology

Level 4, 307 Queen Street

Brisbane, Qld. 4000, Australia

ISASMELT@xstrata.com.au

RESUMEN

El proceso ISASMELT^{TM} emergió en la industria mundial de los metales durante la década de los 90 y actualmente está procesando alrededor de cuatro millones de toneladas de concentrados y materiales secundarios cada año. La tecnología de fusión por lanza sumergida produce tanto como plomo metálico, matas de cobre o cobre metálico en las plantas ubicadas en Australia, EE.UU., Bélgica, India, Alemania, Malasia y China. Plantas adicionales se encuentran en construcción en Perú y Zambia.

El proceso se basa en la lanza sumergida con entrada superior refrigerada por aire, Sirosmelt, desarrollado por el Dr. John Floyd en la División CSIRO de Ingeniería de Minerales y Proceso en los años 70. Mount Isa Mines reconoció el potencial que tenía el nuevo concepto de fundición de soplado superior y se embarcó en un programa de desarrollo que sigue continuando hoy, ya por más de 25 años. Las plantas piloto, las plantas de demostración y las plantas a escala comercial han procesado una amplia gama de materiales durante este tiempo. La operación del proceso a escala comercial en la fundición de cobre de Mount Isa ha permitido que ISASMELT^{TM} se haya desarrollado hasta un nivel donde se le reconoce actualmente como el proceso de fundición de cobre más económicamente atractivo disponible. Xstrata licencia la tecnología para compañías en todo el mundo.

Este documento narra la historia del desarrollo del proceso ISASMELT^{TM} y explora algunas de las razones por las que se convirtió en un nuevo proceso de fusión sostenible.

INTRODUCCIÓN

El desarrollo de la lanza Sirosmelt en CSIRO abrió nuevas oportunidades para la industria pirometalúrgica no ferrosa. Antes de su introducción la inyección de gases en la escoria o la mata líquida se lograba predominantemente por medio de toberas, con complicaciones de diseño inherentes y problemas refractarios. Mount Isa Mines Limited (1) se involucró en la tecnología de lanza sumergida durante los 70 y reconoció su potencial para mejorar la eficacia de las operaciones en sus fundiciones de plomo y cobre. La lanza permitió el uso de hornos estacionarios con un diseño simple, pero con índices de reacción muy altos. Luego de la colaboración conjunta inicial con CSIRO, el proceso ISASMELT™ (2) fue desarrollado hasta su éxito comercial en el complejo de fundición de Mount Isa.

HISTORIA DEL PROCESO ISASMELT™

Los primeros pasos

El desarrollo del proceso ISASMELT™ comenzó en el año 1977 cuando se realizaron pruebas con la lanza Sirosmelt para la recuperación de cobre a partir de escorias de los hornos de ánodos y de conversión[1]. Se utilizó una planta piloto con un diámetro interno de 0,8 metros para el tratamiento de combustión sumergida de las escorias de cobre producidas en Mount Isa y Townsville. Se observó que la turbulencia en el baño permitía la reducción rápida por carbón grueso y una rápida digestión del fundente. También se observó que la capacidad de medir y controlar la temperatura era de vital importancia para maximizar la vida refractaria. Los autores concluyeron que la reducción de la escoria de conversión en el horno Sirosmelt tendría ventajas por sobre la práctica tradicional de regresar la escoria a los hornos de reverbero, ya que a menudo tenían problemas operativos causados por la formación de acumulaciones de magnetita. Los resultados de la prueba motivaron al equipo de investigación para realizar más trabajos de prueba utilizando el nuevo proceso.

Desarrollo del plomo en ISASMELT™

A fines de la década del 70 Mount Isa Mines estaba buscando una solución para mejorar y ampliar las operaciones de la fundición de plomo de Mount Isa. A pesar de que se estaban desarrollando una cierta cantidad de nuevos procesos de fundición de plomo en ese entonces, incluidos los procesos de Kivcet, QSL y Kaldo[2], se pensaba que la lanza Sirosmelt podría utilizarse para fundir los concentrados de plomo de Mount Isa de manera económica. En 1978 se inició un proyecto conjunto con CSIRO para investigar la aplicación de la tecnología de combustión sumergida para la fusión de plomo. El trabajo original que incluía el modelamiento computacional de la termodinámica del proceso fue seguido por trabajos de prueba a escala de crisol en 1979[3]. Las pruebas a escala de crisol parecían confirmar las conclusiones del trabajo de modelamiento de que los humos del plomo del baño se limitarían y que la reducción de las escorias de plomo ocurría de manera bastante rápida. Este trabajo se consideró lo suficientemente

---

[1] Mount Isa Mines es propiedad de Xstrata

[2] ISASMELT™ es una marca registrada de Xstrata technology

prometedor como para justificar el trabajo de prueba a una pequeña escala de planta piloto. Se pensaba que a pesar de que un 40% del plomo en la alimentación se reportaba a la fase gaseosa, los índices de volatilización serían significativamente menores a una escala mayor.

Siguieron una serie de pasos de desarrollo de procesos incluida una planta piloto, una planta de demostración y la construcción de una planta a escala comercial. La tabla 1 muestra las plantas de plomo de ISASMELT™ que se construyeron en Mount Isa.

Tabla 1 – Plantas de plomo ISASMELT™ en Mount Isa

Fecha	Tipo de planta	Capacidad de la planta
1980	Planta piloto a pequeña escala	120-250 kg/h de concentrado
1983	Planta de demostración de fundición	5-10 tph de concentrado
1985	Planta de demostración reducción de escoria de plomo	5 tph de escoria de plomo
1991	Fundición de plomo primario	60.000 tpa de plomo en

Las pruebas en las plantas piloto se realizaron entre 1980 y 1982[3]. Se autorizó una planta de 120kg/h en septiembre de 1980. Los objetivos principales eran probar la factibilidad de un proceso de dos etapas para la fusión del concentrado de plomo y la reducción de la escoria para obtener estimaciones razonables para los índices de volatilizacion, el desgaste refractario y los índices de combustibles. La tabla 2 resume los principales hallazgos del trabajo en la planta piloto.

Tabla 2 – Principales hallazgos en la planta piloto

Humos de plomo	Se podría controlar con una amplia gama de composiciones y temperaturas de la escoria
Aumento del contenido de oxigeno en el aire de la lanza	Se disminuyeron los requerimientos de combustible, se disminuyó la producción de humos, no hubo un aumento significativo en el desgaste de la lanza
Desgaste refractario	Menor al esperado, mayor durante la reducción que durante la oxidación
Reducción de la escoria	Se demostró durante la operación en lotes, continuó rápidamente utilizando la adición de carbón grueso
Humos de zinc	Depende mucho de la temperatura

Posteriormente, Mount Isa Mines y CSIRO recibieron una patente para el “proceso de fundición de plomo de alta intensidad”[4]. Este proceso se denominó el proceso ISASMELT™. El término “ISASMELT™” se utilizó luego para describir la combinación del horno y el diseño de la lanza modificada que incorporó el desarrollo y los conocimientos de las operaciones de Mount Isa.

Los resultados de la planta piloto de 120kg/h motivaron a la administración de Mount Isa a construir una planta de demostración para poder evaluar el potencial de una operación comercial del proceso. Se seleccionó una producción total de aproximadamente 5t/h, puesto que correspondía a alrededor de un cuarto a un décimo del tamaño de una planta comercial que permitiría una futura ampliación dentro de los niveles de riesgo aceptables. La intención era que la planta de demostración se utilizaría para proporcionar información adicional sobre el desgaste refractario y de la lanza, el consumo de combustible, volatilización y la higiene. Para mejorar el traspaso de la tecnología al personal de operaciones, la planta de demostración se instaló dentro de la fundición de plomo existente. El objetivo era hacer de la nueva tecnología parte de la operación de fundición de plomo y que el personal de operaciones tuviera el dominio de ésta.

Se decidió construir la planta en dos etapas. La primera etapa, que se puso en servicio en septiembre de 1983, fue un horno con una tasa de alimentación de 5 a 9 t/h. Este horno se utilizó exclusivamente para la fusión oxidante del concentrado. El horno tenía un armazón de acero de 2,7 metros de diámetro. La boca de gases del horno (parte superior) estaba hecha parcialmente de ladrillos y recubierta con material refractario moldeable. El horno fundía aproximadamente el 15% del concentrado tratado en la fundición de plomo de Mount Isa[5], produciendo una escoria de alto contenido de plomo que se granulaba y posteriormente se agregaba a la alimentación de la planta de sinterizado. La capacidad de la planta de sinterizado estaba restringida por la capacidad limitada de combustión de azufre en ese entonces. Al agregar la escoria del horno ISASMELT™ a la alimentación del sinterizado fue posible aumentar la producción total de la planta de sinterizado hasta en un 10%. Las principales diferencias observadas en la planta de demostración fueron que el índice de volatilización fue menor a la mitad que el de la planta piloto y que el desgaste del recubrimiento refractario principal era menor a un décimo que el de la planta piloto.

Mediante el aumento de la producción general de la planta, la administración consideró más fácil justificar el costo de capital de la planta de demostración. Esta estrategia de buscar formas de aumentar la producción total de la planta existente y a la vez desarrollar el nuevo proceso fue una práctica común en la historia del desarrollo de ISASMELT™.

La operación de la planta de demostración llevó a mejoras en el diseño de la planta. Una de las mejoras fue el rediseño de las lanzas para permitir la operación bajo menores presiones de aire. Las lanzas Sirosmelt originales habían sido diseñadas para la operación a aproximadamente 250kPag y el costo del aire comprimido representó una fracción importante del costo operativo de la planta. Se sometieron a prueba varios diseños de lanzas hasta que se diseñó una que podía operar a menos de 100kPag y a la vez exhibía bajos índices de desgaste. Se utilizaron varios combustibles en la planta y se demostró que el carbón grueso o el coque menudo, mezclado con concentrado, se podría utilizar para la mayor parte del combustible usando una pequeña cantidad de petróleo inyectado a través de la lanza para el control de la temperatura. Esta experiencia demostró que no era necesario utilizar los sistemas de inyección neumática para la inyección de combustible sólido, reduciendo el capital total y los costos operacionales de las plantas.

El personal operativo de la fundición estaba involucrado en la operación y el mantenimiento de la planta diariamente y proporcionaba valoraciones y opiniones de gran valor para los diseñadores. Se cree que su insistencia en mantener la operación del proceso y el diseño del equipo lo más sencillo posible es uno de los factores que contribuyó al éxito general del desarrollo del proceso.

Una vez que la primera etapa había demostrado la operación del proceso de fusión por un período superior al año, la segunda etapa de la planta se construyó y se puso en servicio en agosto de 1985. El segundo horno se construyó de manera adyacente al primero, para permitir la transferencia de escoria oxidada fundida entre ellos.

Inicialmente se realizaron más de 150 pruebas de reducción por lotes. Se logró producir escoria de desecho con niveles del 4% de plomo. Las tasas de volatilización de plomo de 7 a 8% se demostraron tanto para la fusión oxidante como para la reducción. Posteriormente, durante 1987, ambos hornos de demostración se operaron de manera simultánea, con continua:

- Fusión del concentrado de plomo en el horno de fundición
- El sangrado y la transferencia de escoria de plomo fundido en el horno de reducción
- La reducción de la escoria con el carbón grueso entre 1.170 a 1.200 ºC y
- El sangrado del plomo crudo y la escoria de desecho juntos a través de una sola placa de sangrado.

Durante estas pruebas de reducción continua el concentrado de plomo se trató a una tasa promedio de 5t/h bajo condiciones operacionales estables. Escorias de desecho con valores entre 2% a 5% de plomo se lograron a partir del procesamiento de concentrados de plomo con 50% Pb. La duración de las lanzas alcanzó valores entre 100 a 200 horas.[5]

La planta de demostración había tratado más de 125.000 toneladas de concentrado de plomo hasta abril de 1989. Al terminar estas pruebas, se tomó la decisión de proceder con la construcción de una planta de plomo ISASMELT^{TM} a escala industrial y de dos etapas en Mount Isa.

Otros desarrollos en Mount Isa

El desarrollo del proceso de fundición de plomo desde la escala a crisol a la de demostración había tardado alrededor de diez años. Durante esta década, Mount Isa Mines había aprendido mucho sobre la tecnología ISASMELT^{TM} en general; ideas que se podrían utilizar al fundir otros materiales que no fueran concentrados de plomo. En paralelo a la operación de la planta de demostración de plomo, la planta piloto se había utilizado para realizar más trabajos de desarrollo, incluidas las investigaciones de la fusión de níquel, el tratamiento de pastas de baterías, el tratamiento de las impurezas del cobre de la fusión de plomo (drosses) y de la fundición de cobre. Los resultados del trabajo de fundición de cobre demostraron un potencial particular para resolver los problemas operacionales en Mount Isa.

Durante la década de los 80 el concentrado de cobre de las minas de Mount Isa se tostó en un tostador de lecho de fluido y siendo la calcina fundida posteriormente en dos hornos de reverbero. Existía la necesidad de reemplazar estas unidades con tecnologías y alternativas más eficaces. Procesos como la fundición flash, reactor Noranda y a la fundición y conversión continua de Mitsubishi fueron evaluados. Cada uno de estos procesos tenía desventajas o requerían de un excesivo gasto de capital y, por lo tanto, las investigaciones comenzaron a utilizar el proceso ISASMELT^{TM} para la fundición de cobre. El trabajo de la planta piloto se realizó en la planta piloto de plomo y en CSIRO. El éxito del trabajo en la planta piloto llevó a la aprobación de la construcción de un horno de demostración de cobre. En abril de 1987, se puso en servicio el nuevo horno de cobre ISASMELT^{TM}, con un diámetro interno de 2,3 metros[6].

La planta de demostración de cobre utilizó originalmente el aire comprimido de la planta disponible de manera inmediata (suministrado a 700 kPag) para la lanza, pero después de la introducción de un nuevo diseño de una lanza se utilizó un soplador con presión de descarga menor a 150kPag. A pesar de que se diseñó inicialmente para operar sólo con aire a través de la lanza ISASMELT^{TM}, las ventajas de la operación con enriquecimiento con oxígeno pronto se volvieron aparentes con la instalación de una planta de oxígeno de 70 toneladas/día (tpd), lo que permitió que la planta funcionara con aire enriquecido con oxígeno de 25 vol% (en adelante denominado como "enriquecimiento con oxígeno de 25%"). A pesar de que originalmente se diseñó para el concentrado de 15t/h, la planta pudo alcanzar las 25 t/h con la adición del oxígeno al aire de la lanza. Una posterior prueba corta que utilizó oxígeno líquido permitió que la planta funcionara con un enriquecimiento de 26% y con tasas de alimentación de 50 t/h. Existían algunas inquietudes iniciales respecto de que la vida útil de la lanza se acortaría al operar a una presión más baja y con un aumento en el enriquecimiento con oxigeno, pero esos problemas se superaron con los diseños mejorados de las lanzas. Con el tiempo, el horno de demostración fundió 20% de la producción total de la fundición de cobre, suplementando la producción de dos hornos de reverbero. Produjo una mata de cobre y escoria que se sangraban en uno de los hornos de reverbero para la separación.

En sus primeros dos años de operación la planta de demostración trató más de 146.000 toneladas de concentrados de cobre y demostró que podía producir una variedad de leyes de mata que iban desde 36 a 65% en peso de cobre. Hacia mayo de 1992 había tratado 512.000 toneladas de concentrado. Se demostró que podía tratar una amplia gama de materiales de alimentación incluido el concentrado de escoria de conversión reciclado con un alto contenido de magnetita que no habían sido adecuados para tratar en los hornos de reverbero.

Las distribuciones de elementos menores se estudiaron durante la demostración de la operación de la planta[7]. La operación demostró que el horno de cobre ISASMELT^{TM} removió de manera consistente más del 90% del arsénico de la alimentación de la mata. Además, se eliminó entre el 80% y el 90% del bismuto y entre el 60% y el 80% del antimonio de la mata.

Durante la operación de la planta de demostración, se puso un particular énfasis en la extensión de la vida refractaria del horno. Se insertaron bloques de cobre enfriados con agua entre el revestimiento refractario y el armazón de acero para reducir el índice de desgaste refractario. Inicialmente se lograron vidas de campaña de sólo unos pocos meses debido a que se realizaron pruebas en diversos tipos de materiales refractarios y técnicas de instalación. La duración de la campaña aumentó en tanto se acumuló experiencia, y dio como resultado la campaña de duración más larga de 90 días. Esta experiencia demostró la dificultad de lograr una vida útil refractaria larga y constante. Sólo a través de la operación extendida del proceso ha sido posible desarrollar un sistema refractario confiable durante los años intermedios. Mount Isa Mines comenzó a acumular esta experiencia durante la operación del horno ISASMELT^{TM} de demostración de cobre y continuó el proceso de aprendizaje en la primera planta comercial.

Implementación comercial

Horno ISASMELT^{TM} de plomo secundario

Siguiendo a la exitosa operación de la planta de demostración de plomo de 5t/h, se realizaron pruebas en la pasta de batería y en los humos de plomo en la planta piloto de 120 kg/h. Como resultado de este trabajo de prueba, se construyó un horno ISASMELT^{TM} de plomo secundario en la planta Britannia Refined Metals de Inglaterra. Esta planta tenía un horno de 2,5 metros de diámetro y comenzó la puesta en servicio en 1991[8,9]. Trató de manera exitosa los desechos de batería durante una vida útil de doce años. Se cerró a fines de 2003 cuando Xstrata Zinc, al asumir la administración de la refinería, decidió descontinuar las operaciones de fundición de plomo secundarias.

Horno ISASMELT™ de plomo de dos etapas

Siguiendo el éxito de la fundición de plomo de demostración, la administración de Mount Isa Mines aprobó la construcción de una planta ISASMELT™ de plomo de dos etapas. Esta planta comenzó su operación en febrero de 1991. El horno de fundición tenía un diámetro interno de 2,5 metros y el horno de reducción tenía un diámetro interno de 3,5 metros. Algunas diferencias entre el diseño de la planta de demostración y la planta comercial eran el uso de una presa de plomo para permitir el sangrado continuo del producto de plomo en lingotes, la instalación de calderas de calor residual y la inyección de carbón a través de la lanza en el horno de reducción en vez de la adición de carbón grueso. El objetivo de inyectar el carbón era principalmente para aumentar los índices de reducción de escoria. En base a los resultados de la planta de demostración, se pensaba que la etapa de reducción del proceso requería una mayor optimización. Los estudios se habían realizado a fines de los 80 en CSIRO acerca de la reducción de las escorias de plomo en una escala de crisol, en un intento para comprender el índice de los mecanismos de control en el proceso de reducción de plomo[10]. Estos estudios demostraron que la cinética de reducción era relativamente rápida en las escorias de plomo disminuyendo hasta aproximadamente 6% en peso de plomo, pero el índice de reducción disminuyó significativamente en menores concentraciones de plomo. La concentración de plomo objetivo para el ISASMELT™ de plomo era entre un 3% y un 5%, por lo tanto, se creía necesario utilizar la inyección neumática como forma de lograr el objetivo.

La planta demostró rápidamente la capacidad de fundir el concentrado de Mount Isa a tasas de 20t/h, produciendo una alta escoria de plomo[8]. La operación de fundición era estable y se lograron largas vidas útiles de la lanza y refractarias. Las presencia de cristales de ferrita de zinc en el producto de escoria de plomo fundido dieron como resultado capas protectoras de ferrita de zinc que se formaban en la lanza y en las paredes refractarias, protegiéndolas así del ataque de la escoria. Las lanzas duraban típicamente más de un mes y había poco desgaste en las paredes del horno. Las condiciones de oxidización y el baño altamente agitado en el horno de fundición lograron suprimir los humos del sulfuro de plomo.

La etapa de reducción sin embargo, demostró ser más problemática, con vidas de la lanza cortas y obstrucciones y desgaste del sistema de inyección de carbón neumático llevando a contenidos erróneos de plomo en la escoria final. Estos problemas se superaron con el rediseño del equipo de inyección de carbón y con la modificación del diseño de la punta de la lanza. El mejor rendimiento ocurrió durante 1993 cuando hubo oxígeno adicional disponible y cuando se fundieron hasta 36 t/h de concentrados utilizando aire de lanza con 33% a 35% de enriquecimiento con oxígeno. Durante este período de fundición y reducción de tasa alta, se logró entre un 2% a 5% de plomo residual en las escorias finales del horno de reducción.

Al operar con los flujos de diseño de gases de proceso, las calderas de calor residual no podían reducir la temperatura de los gases al nivel requerido para el ingreso a los filtros de manga. Por lo tanto, fue necesario agregar un sistema de enfriamiento con aire para lograr las temperaturas requeridas. La restricción de la caldera fue el resultado de una capa cohesiva de polvo que se formó en las uniones de los paneles de los tubos de convección, reduciendo significativamente la tasa de transferencia del calor.

El desgaste refractario fue aceptable en el horno de reducción siendo necesario el reemplazo de los ladrillos después de dos años.

Una disminución en el suministro de concentrado de plomo de la mina de Mount Isa a fines de 1994 hizo que la operación continuada del ISASMELT^{TM} de plomo en paralelo a la planta de sinterizado y al alto horno no fuera económica. Por esta razón, se decidió cerrar la planta ISASMELT^{TM} de plomo a principios de 1995. Había estado sin funcionar desde esa fecha, pero se ha reevaluado regularmente el volver a ponerla en funcionamiento, en función del estado del mercado del plomo y de la disponibilidad de concentrado adecuado.

Fusión de cobre/níquel

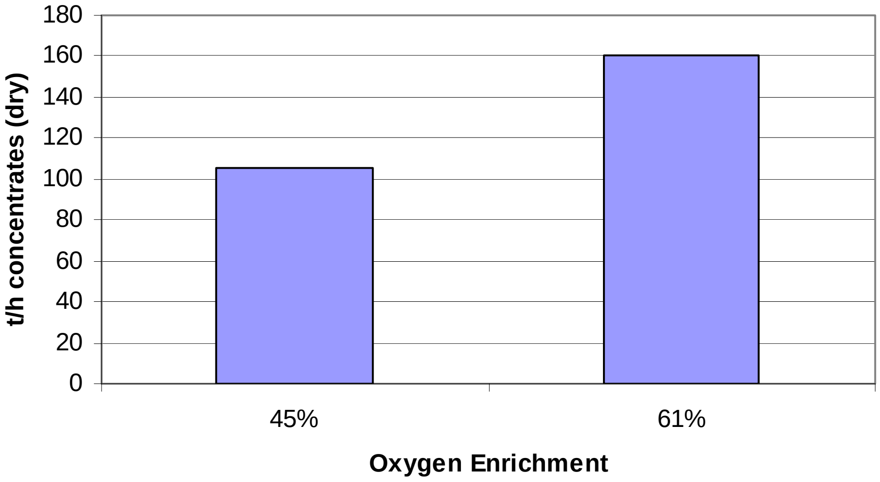
El éxito del horno de demostración de cobre ISASMELT^{TM} y el trabajo de prueba de la planta piloto realizada en la fundición de concentrado de cobre/níquel llevó a Agip Australia a tomar la decisión de instalar el nuevo horno ISASMELT^{TM} para la producción de mata de níquel y cobre en su sitio Radio Hill en Australia occidental[11]. La planta comenzó la puesta en servicio en septiembre de 1991 y dentro de 3 meses estaba funcionando a una capacidad de diseño de 7,5t/h de concentrado. Produjo una mata de 45% de níquel/cobre de un concentrado que contenía aproximadamente 7% de níquel y 3,5% de cobre. Desafortunadamente el precio del níquel, que había estado sobre US$6 por libra en 1989 cayó hasta aproximadamente US$3 por libra hacia 1991.
Por lo tanto, a pesar del éxito técnico del proyecto, Agip cerró la mina, el concentrador y la fundición después de menos de seis meses de operación. Los siguientes propietarios de la mina se han enfocado sólo en la explotación y en el procesamiento y desde entonces la planta de ISASMELT^{TM} se ha desmantelado.

Horno ISASMELT^{TM} de cobre

Continuando con el éxito del horno de demostración de cobre ISASMELT^{TM}, Mount Isa Mines siguió con la construcción de una planta ISASMELT^{TM} de cobre clasificada en 180.000 tpa de cobre en concentrado. El horno, con un diámetro interno de 3,75 metros, comenzó la puesta en servicio en agosto de 1992[12]. La nueva planta comenzó la operación con cerca de un 45% de enriquecimiento con oxígeno y producía una mata que contenía 50 a 60 % en peso de cobre. Se utilizó la adición de carbón grueso a la mezcla de alimentación como el combustible principal, inyectando diesel por la lanza para ajustar la temperatura del baño y para calentar el baño después de las detenciones.

Durante los años la producción del horno aumentó continuamente y en 1998 se instaló una nueva planta de oxígeno, permitiendo que el nivel de enriquecimiento con oxígeno de aire a través de la lanza aumentara en más de un 60%. Como resultado se aumentó la tasa de alimentación de aproximadamente 105 t/h de concentrados a más de 160 t/h de concentrados (base seca) como se ilustra en la Figura 1.

Figura 1 – Efecto del enriquecimiento con oxígeno



Este aumento coincidió con el desarrollo de nuevas minas en Mount Isa y Ernest Henry y una mejora de la refinería de cobre en Townsville. Para reducir el riesgo técnico asociado con el aumento en el enriquecimiento con oxígeno se realizaron dos pruebas de alto índice en 1997. Estas pruebas demostraron que el horno de cobre ISASMELT^{TM} podía tratar los concentrados a una tasa equivalente a 250.000 tpa de cobre contenido13. El horno de cobre ISASMELT^{TM} está tratando actualmente un millón de toneladas de concentrado al año. La Figura 2 muestra las toneladas anuales de alimentación al horno ISASMELT^{TM} de cobre desde que comenzó la puesta en servicio. La Tabla 3 muestra los ingresos típicos al horno de cobre ISASMELT^{TM}.

Figura 2 – Alimentación de concentrado anual al horno ISASMELT^{TM} de cobre

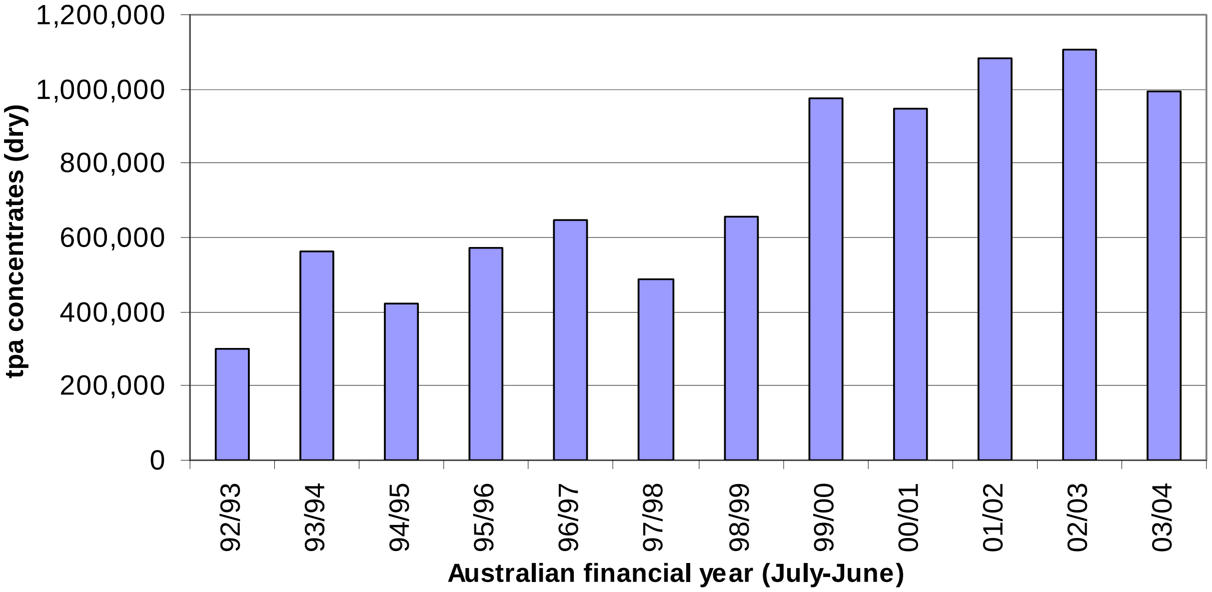


Tabla 3 – Parámetros de operación típicos – ISASMELT^{TM} de cobre

	Tasa	Unidades
Tasa fusión concentrado	160	tph
Contenido cobre (base seca)	23,8	%Cu
Flujo de sílice	3,4	tph
Material Circulante	1,6	tph
Coque menudo	0,7	tph
Gas natural	706	Nm3/h
Aire en la lanza	20.210	Nm3/h
Oxígeno de la lanza (95%)	23.580	Nm3/h
Enriquecimiento con oxígeno de la lanza	60,8	%
Temperatura del baño	1172	Grados C
Ley de la mata	57,0	%

Los índices de desgaste refractario del horno han disminuido gradualmente durante los 12 años de operación. Se decidió durante la fase de construcción que no se utilizaría la refrigeración por agua en el horno a pesar de la experiencia positiva de la planta de demostración. Esta decisión se basó en las inquietudes relacionadas con el uso de la refrigeración por agua en la zona de baño en un horno comercial. Además se creía que debería ser posible, con una selección refractaria cuidadosa, técnicas de instalación y metodología de control, lograr una vida refractaria de dos años sin la refrigeración por agua. Las campañas iniciales tuvieron una duración de entre 12 y 15 meses en cuanto se realizaban pruebas en diversos tipos refractarios y metodologías de instalación. Hacia 1998, no obstante, en base a la experiencia operativa adquirida junto con los avances en los sistemas de control de procesos, se logró un rendimiento refractario significativamente mejorado. El desgaste refractario ahora se monitorea en línea, lo que permite retroalimentación en tiempo real a los operadores de las condiciones dentro del horno. Como resultado, desde 1998 cada campaña ha tenido una duración de al menos 2 años.

La Figura 3 muestra las duraciones de las campañas refractarias del horno de cobre ISASMELT^{TM} a la fecha, mientras que la Figura 4 muestra el índice de desgaste del refractario en el barril del horno para cada una de las campañas desde 1998.

Figura 3 – Historia de la duración de la campaña del horno ISASMELT^{TM} de cobre

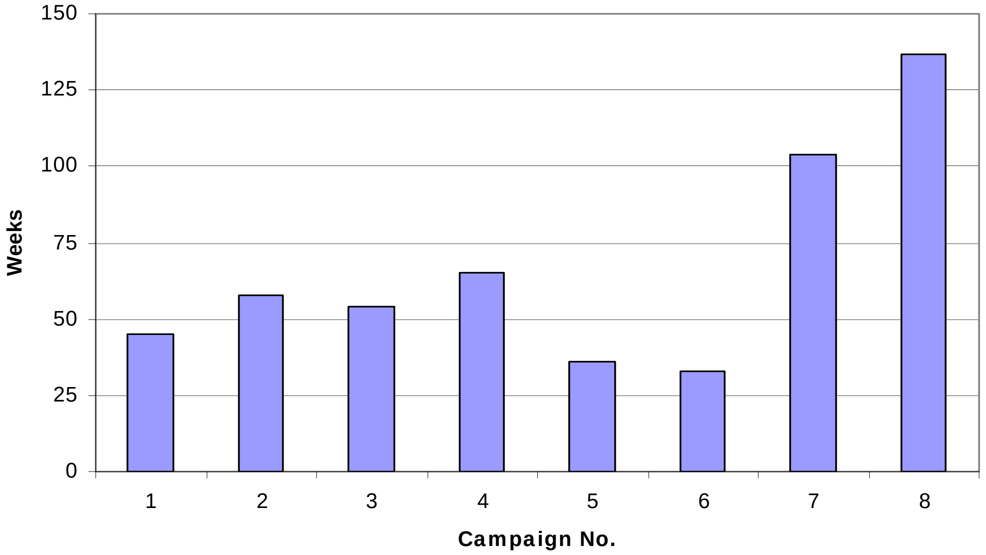
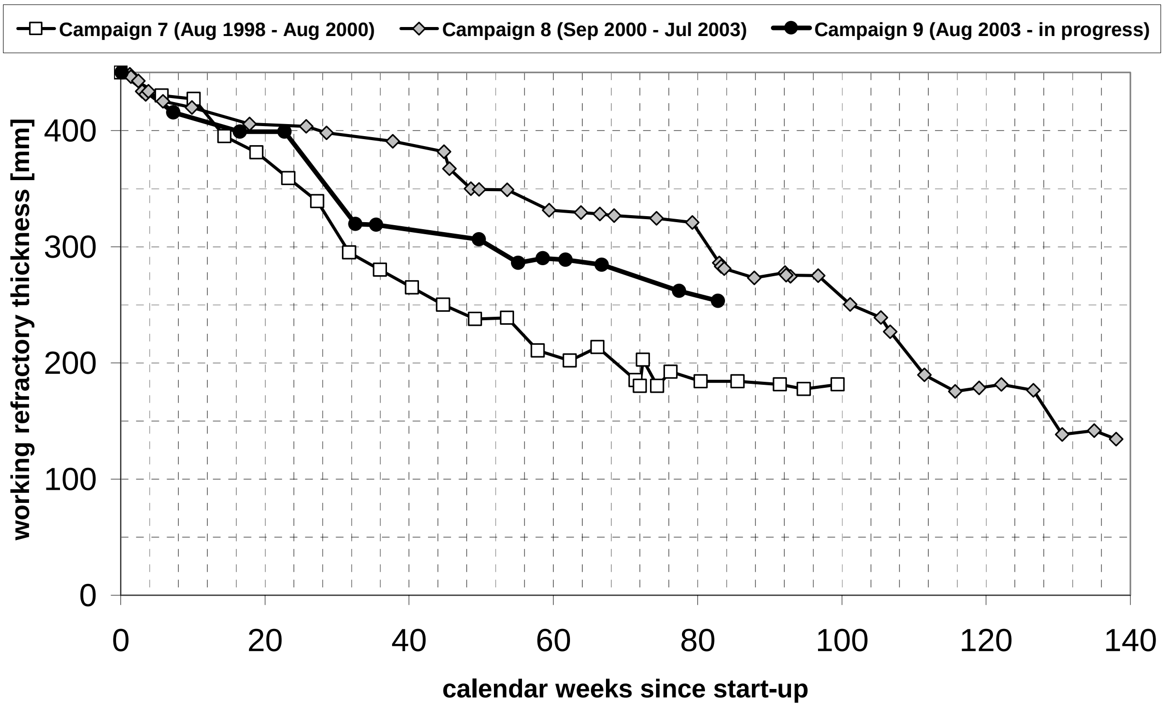


Figura 4 – Desgaste refractario del horno ISASMELT^{TM} de cobre



Fundiciones de cobre para clientes

Las ventajas operacionales de la planta ISASMELT^{TM} de cobre sobre las alternativas suscitó el interés de las compañías en el proceso. Como resultado, Mount Isa Mines decidió licenciar la tecnología a clientes externos hacia fines de los 80 y sigue haciéndolo. La Tabla 4 resume a las fundiciones licenciatarias que están en operación o en construcción.

Tabla 4 – Plantas ISASMELT^{TM} licenciatarias

Fecha de la puesta en marcha	Licenciatario	Ubicación de la planta	Capacidad de la planta
1992	Phelps Dodge Miami	Arizona, EE.UU.	650.000 tpa de concentrado de cobre
1996	Sterlite Industries (India) Ltd	Tuticorin, India	450.000 tpa de concentrado de cobre
1997	Umicore Precious Metals	Hoboken, Belgium	200.000 tpa alimentación mezclada
2000	Metal Reclamation Industries	Pulau Indah, Malasia	40.000 tpa de metal de plomo
2002	Hüttenwerke Kayser	Lünen, Alemania	150.000 tpa de desecho de cobre
2002	Yunnan Copper Corporation	Kunming, China	600.000 tpa de concentrado de cobre
2005 (en	Sterlite Industries (India) Ltd	Tuticorin, India	1.200.000 tpa de concentrado de cobre

2005 (en	Yunnan Metallurgical	Qujing, China	160.000 tpa de concentrado de plomo
2006 (en	Southern Peru Copper	Ilo, Perú	1.200.000 tpa de concentrado de plomo
2006 (en	Mopani Copper Mines	Mufulira, Zambia	650.000 tpa de concentrado de plomo

Fundición de cobre Phelps Dodge Miami

Cyprus Miami Mining Corporation (actualmente Phelps Dodge Miami) seleccionó a ISASMELT^{TM} para su modernización en 1990. La decisión se basó en la observación de la operación de la planta de demostración ISASMELT^{TM} de cobre en Mount Isa. La planta de Miami, ubicada en Claypool Arizona, se diseñó y construyó al mismo tiempo que la planta ISASMELT^{TM} de cobre de 180.000 tpa en Mount Isa. Comenzó la puesta en servicio en junio de 1992, dos meses antes de la planta de Mount Isa. Cyprus seleccionó a ISASMELT^{TM} después de compararla con las tecnologías de Contop, Inco, Mitsubishi, Noranda, Outokumpu y Teniente[14]. Se modificó un horno eléctrico, que anteriormente se utilizaba para fundir concentrados de cobre para actuar como un horno de retención para la mata de cobre y la escoria y para la reducción de la escoria de conversión Peirce Smith.

Cyprus reconoció el riesgo de utilizar la tecnología ISASMELT^{TM} que no se había probado a una escala comercial completa en ese momento y, por lo tanto, instigaron un programa de mitigación de riesgos. El programa incluía un programa de capacitación detallado, enviando 35 empleados de Mount Isa para realizar la capacitación del horno de demostración de cobre ISASMELT^{TM} y en la fundición de plomo. Sin embargo, aún habían muchas características de diseño que diferían de aquellas de la planta de Mount Isa. La primera era el uso del horno eléctrico como un horno de retención, la segunda era el uso de una cubierta inclinada del punto de toma de gas refrigerada con agua seguida por una caldera de calor residual de radiación/convección y la tercera fue el uso de gas natural inyectado a través de la lanza como el combustible principal para el horno. El enriquecimiento con oxígeno era de hasta un 50% de aire de lanza. De estas diferencias, la principal causa del tiempo de inactividad durante los primeros años fue la salida de gases del horno a la caldera, que sufrió serios problemas de corrosión y formación de accreciones. Con el tiempo la campana de gases se reemplazó con un canal de radiación vertical conectado al sistema de la caldera. Los problemas con las pérdidas de cobre mayores a las aceptables en la escoria del horno eléctrico fueron una inquietud hasta fines de la década de los 90 cuando el trabajo realizado en las lanzas de gas natural dieron como resultado un considerable aumento en el rendimiento de limpieza de la escoria[15].

La tasa de alimentación en Miami aumentó a través de los 12 años de operación. La Figura 5 muestra las tasas de alimentación durante ese período. Las campañas refractarias también han mejorado con el tiempo, según se demuestra en la Figura 6.

Figura 5 – Tasa de alimentación de ISASMELT™ en Phelps Dodge Miami

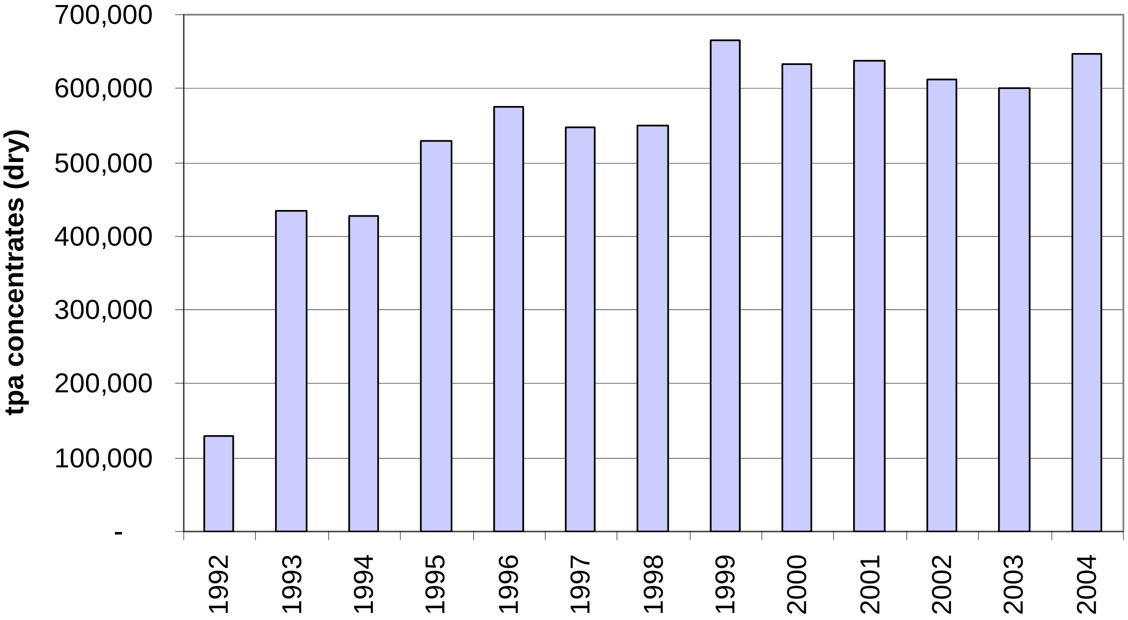
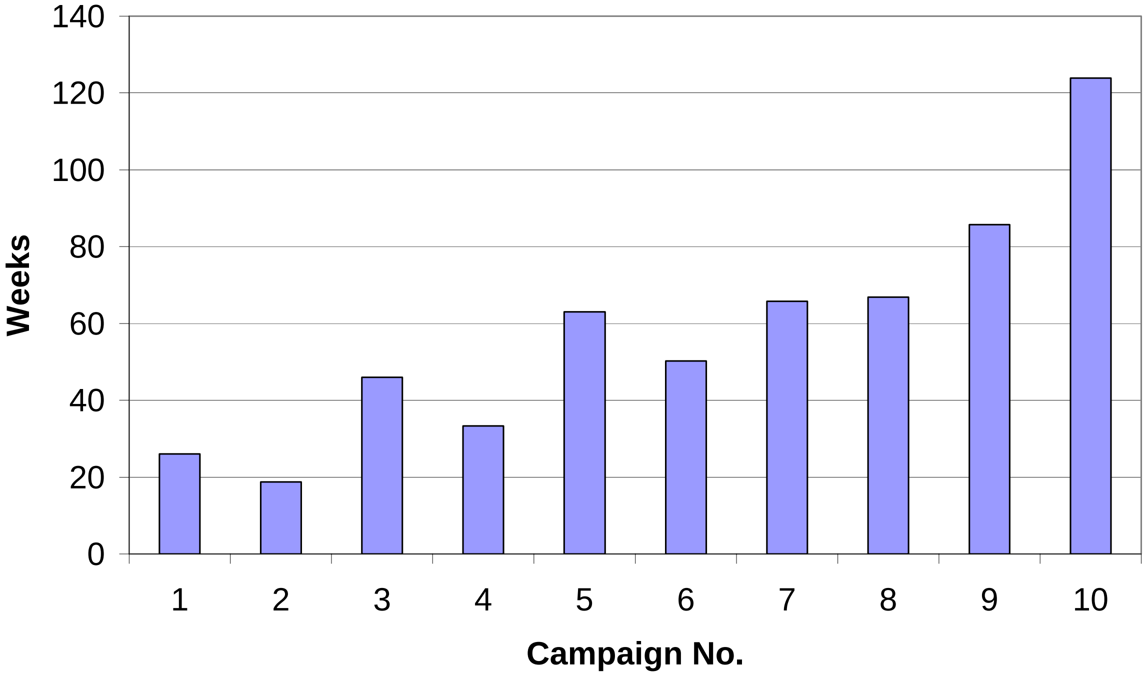


Figura 6 – Historial de duración de la campaña ISASMELT™ en Phelps Dodge Miami



Fundición de cobre Sterlite Industries

La planta ISASMELT™ de cobre de Sterlite Industries está ubicada en Tuticorin al sur de la India. La fundición comenzó la puesta en servicio en 1996 y se construyó en un sitio despejado “green field” para fundir concentrados importados. El horno ISASMELT™ originalmente tenía una capacidad de diseño de 60.000 tpa de cobre en mata, pero ha sido ampliada a 180.000 tpa de cobre. El aumento en la capacidad se logró instalando oxígeno de tonelaje adicional y ampliando la planta asociada y el equipo. Sterlite, que no había operado una fundición de cobre antes la construcción de la planta de Tuticorin, actualmente es un importante productor de cobre.

Sterlite está finalizando la construcción de un nuevo horno ISASMELT^{TM}. Este nuevo horno comenzó la puesta en servicio durante 2005 y tratará a los concentrados que contienen 300.000 tpa de cobre.

Fundición de Umicore Precious Metals

La fundición de Umicore Precious Metals ubicada en Hoboken, Bélgica, utiliza el proceso ISASMELT^{TM} para tratar una variedad de materiales de alimentación primarios y secundarios. El sitio de Hoboken pasó por una modernización drástica a fines de 1990 para poder seguir siendo competitivos mientras operaban bajo regulaciones medioambientales estrictas. Una característica clave de la modernización fue la instalación del horno ISASMELT^{TM}. El horno reemplazó a una gran cantidad de procesos de unidad, permitiéndole a la compañía reducir significativamente los costos operativos mientras reducía las emisiones al medioambiente. La nueva fundición ha estado en funcionamiento desde fines de 1997 y desempeña un papel importante en el negocio de reciclaje de Umicore.

Fundición de Huettenwerke Kayser Copper

Huettenwerke Kayser, una subsidiaria de Norddeutsche Affinerie, opera una planta de ISASMELT^{TM} para la fundición de cobre secundaria dentro de su operación de fundición y refinado cerca de Dortmund en Alemania. El horno ISASMELT^{TM} reemplazó a los dos altos hornos y a los tres convertidores Peirce Smith utilizados para fundir cobre de desecho. La instalación del horno ISASMELT^{TM} ha permitido que la compañía reduzca significativamente los costos operativos y que mejore el rendimiento medioambiental del horno.

Fundición de cobre de Yunnan Copper Corporation

Yunnan Copper Corporation (YCC) comenzó la puesta en servicio de su planta ISASMELT^{TM} de cobre en mayo de 2002 como parte del proyecto de modernización de la fundición de cobre[16,17]. El horno ISASMELT^{TM} trata más de 600.000 tpa de concentrado de cobre. El horno reemplazó a una planta de sinterizado y a dos hornos eléctricos y produce una mata de cobre con un contenido de cobre de 55%. El aire del proceso se enriquece a un contenido de 50% de oxígeno. La instalación del horno ISASMELT^{TM} ha mejorado significativamente el rendimiento medioambiental del horno y redujo los costos operativos, con el aumento de la captura de azufre de 79% a 96% y la disminución en el consumo de energía en un 32%.

Tabla 5 – Cambio en el consumo de energía del horno de YCC

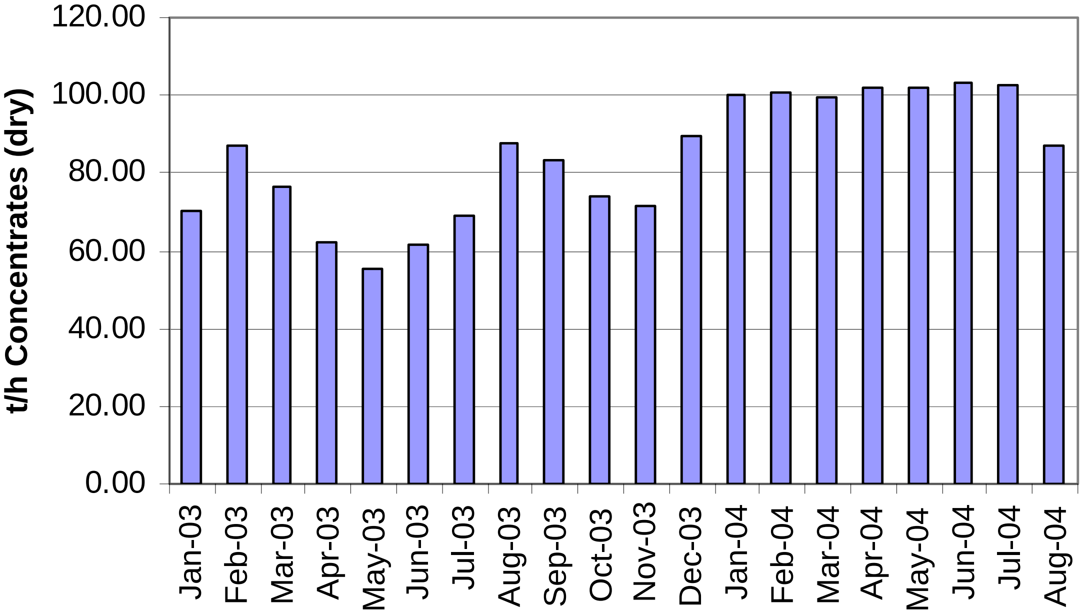
Año	Consumo de energía (concentrado a blister) Toneladas de carbón estándar / tonelada de carbón
2001	0,729
2003	0,523
2004	0,493

La planta de YCC incorporó las mejoras más recientes en el diseño desarrolladas en Mount Isa. Se capacitó al personal de YCC en Mount Isa en un período de 7 meses. El resultado fue que la planta alcanzó la capacidad de diseño (promediada en un período de 7 días) dentro de dos meses a partir del momento en que se alimentó el primer concentrado en el horno.

Siguiendo la exitosa puesta en servicio, el mayor impedimento para alcanzar la capacidad de diseño de manera consistente fue el suministro de concentrado.

La Figura 7 muestra el índice de alimentación mensual promedio desde enero de 2003 hasta fines de la primera campaña operativa. Durante la mayor parte de 2002 y hacia el 2003 YCC no pudo obtener la alimentación suficiente debido al ajustado mercado de concentrados de ese entonces. Una vez que tuvieron disponible una cantidad suficiente de concentrado la planta de YCC pudo exceder la capacidad de diseño de manera constante. La primera campaña duró dos años y cuatro meses desde mayo de 2002 hasta septiembre de 2004, demostrando un rendimiento refractario similar al logrado en Mount Isa.

Figura 7 – Índice de alimentación de ISASMELT^{TM} en YCC



Fundición de cobre de Southern Peru Copper Corporation

Southern Peru Copper Corporation (SPCC) ha comenzado la construcción de un nuevo horno ISASMELT^{TM} en su fundición de cobre ubicada en Ilo, Perú. SPCC investigó una amplia gama de tecnologías alternativas incluida la fundición flash, el convertidor Teniente, Noranda y Mitsubishi, antes de decidir sobre la construcción del horno ISASMELT^{TM}. El horno ISASMELT^{TM} tratará 1.200.000 tpa de concentrado y reemplazará dos hornos de reverbero y un convertidor Teniente. Se utilizarán dos hornos rotatorios de retención para separar los productos de mata y escoria del horno ISASMELT^{TM}.

Se utilizará una caldera de calor residual y un precipitador electroestático para refrigerar y limpiar el gas antes de pasarlo a una nueva planta de ácido sulfúrico. Se utilizarán convertidores Peirce Smith para convertir la mata en cobre blister. La planta comenzará a operar en 2006.

Fundición de cobre de Mopani Copper Mines

Mopani Copper Mines Plc (MCM) ha comenzado la construcción de una nueva planta ISASMELT^{TM} en la fundición de cobre Mufulira en Zambia. MCM decidió instalar un horno ISASMELT^{TM} después de compararlo con tecnologías alternativas incluida la fundición flash, el proceso Mitsubishi y los convertidores Teniente. Concluyeron que la ventaja del proceso ISASMELT^{TM}[18] incluye:

- Su tamaño compacto y pequeña huella, que se podría acomodar fácilmente en la fundición Mufulira.
- El costo de capital relativamente bajo.
- Los bajos costos operativos.
- Los gases de ebullición de alta resistencia que se ajustan bien a la producción de ácido sulfúrico.
- La relativa facilidad de operación.
- La flexibilidad de la tasa de alimentación, adecuada para una operación inicial en 2006 de 650.000 tpa de concentrado con posibilidades para capacidades superiores a 850.000 tpa en el futuro.

El horno ISASMELT^{TM} reemplazará a un horno eléctrico existente. La planta incluirá un nuevo sistema de preparación de alimentación, una caldera de calor residual, horno de retención eléctrico, planta de oxígeno y planta de ácido, así como mejoras al ala del convertidor y las plantas de ánodo. Comenzará a operar en 2006.

Fundiciones de plomo para clientes

Fundición de plomo de Metal Reclamation Industries

Metal Reclamation Industries comenzó la puesta en operación de una fundición de plomo secundario en Pulau Indah, Malasia en el año 2000. La planta produce hasta 40.000 tpa de plomo refinado principalmente de las baterías de ácido de plomo recicladas. La planta se basó en el diseño de la planta de Britannia Refined Metals en Inglaterra y opera un proceso de lotes.

Fundición de plomo YMG

Xstrata Technology diseñó un horno ISASMELT^{TM} para la nueva fundición de plomo de Yunnan Metallurgical Group (YMG) en Quijing, China. El horno ISASMELT^{TM} fundirá 160.000 tpa de concentrado de plomo para producir lingotes de plomo y escoria con alto contenido de plomo. La escoria con alto contenido de plomo se solidificará y se alimentará a un alto horno diseñado por YMG para la reducción. Este proceso, un desarrollo conjunto de Xstrata y YMG, combinará los beneficios del horno ISASMELT^{TM} para fundir con los beneficios del alto horno para la reducción. El horno ISASMELT^{TM} reemplazará de manera eficaz a la planta de sinterizado de un horno de plomo tradicional. El horno ISASMELT^{TM} tiene una ventaja sobre una planta de sinterizado que consiste en que puede convertir una fracción del plomo en alimentación directamente a plomo metálico, disminuyendo así la tarea de la reducción de escoria del alto horno. En la planta de YMG más del 40% del plomo en la alimentación pasará a convertirse directamente en metal de plomo en el horno de fusión. El horno ISASMELT^{TM} tiene la ventaja de que es mucho más pequeño y sencillo que una planta de sinterizado y que se puede conectar rápidamente para eliminar las emisiones. Los gases de ebullición del horno ISASMELT^{TM} tienen un contenido de dióxido azufre relativamente alto, adecuado para la conversión a ácido en una planta de ácido sulfúrico convencional. El producto de la escoria es bajo en azufre en comparación con el sinterizado y así, los gases de ebullición del alto horno contienen una concentración mucho menor de dióxido de azufre que en el caso en que se alimentara un alto horno con sinterizado.

VEINTICINCO AÑOS HACIA UNA TECNOLOGÍA SUSTENTABLE

La puesta en marcha de la planta YCC ISASMELT^{TM} coincidió con el vigésimo quinto aniversario del primer trabajo de prueba de Sirosmelt realizado por Mount Isa Mines y CSIRO. Veinticinco años de trabajo intenso y tenacidad por parte de una amplia gama de personas llevó a la comercialización exitosa de una nueva tecnología de procesamiento de metales realmente sustentable. La planta de YCC incorporó características de diseño que habían sido desarrolladas y optimizadas en las plantas ISASMELT^{TM} anteriores. Después de 25 años del desarrollo de ISASMELT^{TM}, la planta de YCC demostró cómo se puede incorporar el proceso ISASMELT^{TM} de cobre de manera exitosa en las fundiciones de cobre en cualquier lugar del mundo. Basándose en la experiencia obtenida en Mount Isa y en otros lugares, Xstrata actualmente diseña plantas que pueden crecer a la capacidad de diseño rápidamente. Con una combinación de diseño comprobada, programas de capacitación integrales en las operaciones de la fundición de Mount Isa y con la asistencia en la puesta en servicio por parte de los operadores de planta con vasta experiencia en la planta, los clientes tienen una exposición mínima al riesgo al incorporar la tecnología en sus plantas. El proceso ISASMELT^{TM} está ayudando a mejorar el desempeño medioambiental de las fundiciones en todo el mundo. La introducción del horno ISASMELT^{TM} de cobre en Mount Isa ayudó a disminuir el consumo de combustible en la fundición de cobre en más de un 90% entre 1991 y 2002[19].

Se optimizó una gran cantidad de características de diseño durante los últimos 25 años de desarrollo y las mejoras continuarán. Algunas de las funciones cruciales aparecen en la siguiente lista.

Refractarios

Una larga vida refractaria es fundamental para el éxito de cualquier proceso pirometalúrgico. El desarrollo de sistemas refractarios durables toma muchos años, debido al tiempo involucrado en las pruebas, ya sea en ciertos tipos de materiales refractarios, métodos de instalación, sistemas de refrigeración y dispositivos de monitoreo pueden dar como resultado un menor desgaste. Mount Isa Mines se enfocó en los sistemas refractarios mejorados desde el inicio del programa de desarrollo. En conjunto con CSIRO, universidades, AMIRA, proveedores de sistemas refractarios y otros socios de la industria, una serie de programas de investigación estudiaron como optimizar el rendimiento refractario. Se estudiaron los aspectos básicos del desgaste refractario y se pusieron en práctica las implicaciones prácticas de los hallazgos en las plantas operativas.

Como se mencionó anteriormente, ha habido varios intentos de refrigerar las paredes del horno para reducir el desgaste refractario, en particular alrededor de la línea de flotación de la escoria y en los hornos de cobre. Algunas de las plantas tienen bloques de cobre instalados entre el revestimiento refractario y el armazón de acero, mientras que otras tienen una ducha de refrigeración externa instalada. En Mount Isa, el horno de cobre ISASMELT^{TM} no tiene un sistema refrigeración con agua aplicado a las paredes del horno y aun así se logran campañas de más de dos años y medio. Este rendimiento es posible a través del uso de refractarios de alta calidad, procedimientos de instalación cuidadosos, monitoreo de condición en línea y un control muy estricto de la temperatura del baño utilizando sistemas de control que se han desarrollado durante más de una década en la planta operativa. Recientemente, se ha planteado el objetivo de lograr una campaña de cinco años y actualmente se están realizando estudios para determinar la mejor forma de lograrlo.

Sistemas de control

El control del proceso ISASMELT^{TM} para el operador es relativamente directo en la actualidad gracias a la automatización de los parámetros clave del proceso. El desarrollo de esta automatización ha tomado muchos años en la escala industrial. Dado que el horno se alimentaba continuamente, pero que se sangraba por lotes, se hizo necesario desarrollar un método para controlar automáticamente el movimiento de la lanza ISASMELT^{TM} en el horno. El movimiento es necesario para que la punta de la lanza siempre esté correctamente posicionada en el baño, logrando así las tasas de mezcla y reacción necesarias y a la vez minimizar el desgaste de la punta de la lanza. Los sistemas desarrollados en las plantas piloto y de demostración se mejoraron en el horno a escala completa y ahora permiten que se controle automáticamente la inmersión de la punta de la lanza dentro de una tolerancia extremadamente estricta. Durante la operación normal, la posición de la lanza se controla completamente a través del sistema de control de la planta y no es necesaria la intervención humana.

El horno ISASMELT^{TM} es básicamente un reactor CSTR clásico. De tal forma, la composición del baño y la temperatura es muy uniforme a través de la cámara. Esto permite que se pueda controlar la temperatura de manera muy precisa cuando se utiliza una medición de la temperatura y un sistema de retroalimentación confiable. Aquellos sistemas se desarrollaron en los años 90 y actualmente permiten que se pueda controlar la temperatura del baño dentro de un intervalo de aproximadamente ±10 grados Celsius.

Sistemas de tratamiento gases de proceso

El sistema de tratamiento de gases de proceso se debe ver como una parte integral del horno. Los problemas asociados con el enfriamiento de polvos metalúrgicos pegajosos hace que el diseño de los sistemas de tratamiento de los gases de proceso generados en el horno sea especialmente difícil. Como resultado de esto, ha tomado muchos años y experiencia con una variedad de situaciones de planta hasta que se encontró una solución completamente satisfactoria. Las plantas piloto y las de demostración generalmente utilizaban enfriadores por evaporación y ductos con revestimiento refractario. Para las plantas a escala piloto, sin embargo, se consideraba preferible instalar calderas de calor residual para mejorar la eficacia térmica del proceso. Se han probado diversas configuraciones de calderas durante los años con distintos grados de éxito. El horno de la fundición de cobre ISASMELT^{TM} de Mount Isa tiene una caldera de lecho fluidizado circulante que tiene la ventaja de extinguir los gases rápidamente, pero que tiene la desventaja de una caída de presión relativamente alta. La planta de Phelps Dodge tenía originalmente un punto de toma enfriado con agua que sufría de graves problemas con la formación de acumulaciones y corrosión por la condensación de ácido[20]. El horno ISASMELTTM de plomo de Mount Isa tenía una caldera de radiación y convección convencional tanto en el horno de oxidación como en el de reducción, pero tenía una capacidad inadecuada debido a la baja transferencia de calor provocada por la naturaleza de los humos del plomo y del polvo.

En el transcurso de los años y gracias a la cooperación con los principales proveedores, se han encontrado soluciones a estos problemas. Como resultado de ello, Xstrata puede garantizar que los sistemas de gases de proceso del horno ISASMELTTM se pueden diseñar de manera correcta, como lo demostró el rendimiento sin problemas de la caldera YCC desde su puesta en marcha.

Diseño del horno

La patente original de Sirosmelt cubrió principalmente el concepto de la lanza refrigerada con agua. El diseño original del horno parecía ser adecuado en ese entonces para un proceso de lanza sumergida superior. La experiencia que se adquirió durante los años ha demostrado que el diseño original presentaba falencias cuando se utilizaba en una operación comercial. En tanto ha aumentado el tamaño del horno, también ha aumentado el reconocimiento de que existen fuerzas extremadamente altas producidas por la inclinación del baño fundido. Los gases inyectados producen un movimiento ondulante complejo, que imparte fuerzas cíclicas horizontales y verticales en las paredes y en la base del horno[21]. Como resultado de esas fuerzas altas se deben diseñar los cimientos y el sistema de soporte de manera cuidadosa. El método para conectar el horno al sistema de soporte también se tenía que mejorar para garantizar la estabilidad del horno y a la vez permitir la expansión del refractario y de la cubierta de acero.

El techo inclinado del horno original se diseñó originalmente para mejorar la dinámica del flujo de gas, pero en la práctica se descubrió que un techo horizontal era más fácil de construir y mantener, mientras que no sufría de desventajas significativas en comparación con el original. El primer diseño utilizó secciones revestidas de material refractario o de acero refrigerado con agua. Posteriormente se utilizaron los bloques de cobre enfriados con agua, ya que eran más fáciles de mantener. Sin embargo, de manera más reciente, se ha descubierto que para la fundición de cobre una mejor solución es tener un techo plano construido de tubos de caldera y una operación muy por sobre el punto de rocío de los gases de ebullición. Aún existen algunas desventajas con un sistema presurizado, no obstante, todavía se están desarrollando soluciones alternativas.

FACTORES QUE CONTRIBUYEN A UNA INNOVACIÓN EXITOSA

La historia de la innovación tecnológica ha proporcionado muchos ejemplos de ideas y conceptos extraordinarios que no sobrevivieron a la transición de la conceptualización a la implementación a escala industrial. Por ello es importante revisar las principales razones que han contribuido a la implementación exitosa y a la internacionalización de la tecnología ISASMELTTM.

Muchos factores han contribuido al éxito. El desarrollo de la lanza CSIRO fue la innovación inicial. Un elemento clave en el éxito, sin embargo, fue el hecho de que el proceso se desarrolló para la factibilidad comercial en los entornos de hornos operativos combinando los talentos de destacados ingenieros, metalúrgicos y técnicos. Otros factores fueron el impulso emprendedor del personal de Mount Isa que incluía una cantidad de “líderes” de tecnología y el apoyo del gobierno australiano durante los primeros años, a través del Australian Industrial Research

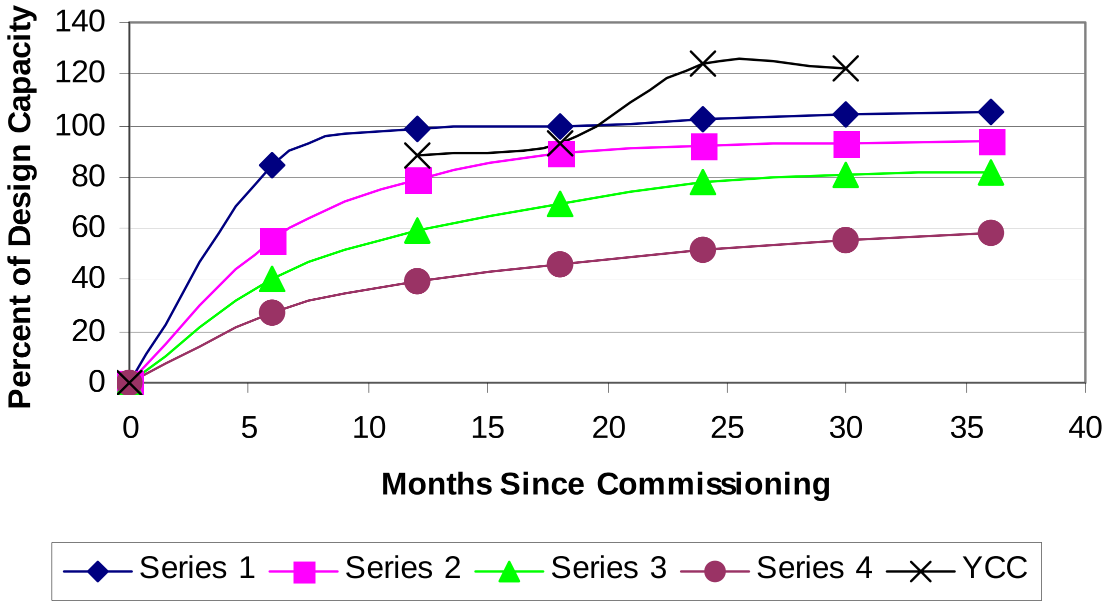
Development Incentives Board (AIRDIB). También fue fundamental que el proceso se probara a través de una serie disciplinada de pasos de desarrollo, incluido el trabajo de laboratorio, las pruebas en la planta piloto, la operación de las plantas de demostración durante períodos prolongados y plantas comerciales a escala completa que posteriormente se expandieron después de realizarse campañas de prueba especializadas.

Finalmente, se cree que la conexión cercana entre el personal de operaciones con el nuevo proceso desempeñó un papel fundamental en su éxito. Es mucho más probable que un proceso sea exitoso si está dirigido por el personal de operaciones que reconoce un beneficio en su utilización y lo ve como una forma de mejorar su calidad de vida en el futuro. En Mount Isa se llevó a los operadores “a bordo” a comienzos del programa de desarrollo y se les incentivó a contribuir al programa. Reaccionaron de manera entusiasta y probaron ofrecer consejos sabios y ser fuentes de consejos pragmáticos extremadamente útiles en las fases críticas del desarrollo. A través de un énfasis constante en la simplificación de las funciones del proceso, el personal de operaciones ayudó a asegurar que resultara un proceso sólido y confiable a nivel industrial.

Como resultado, la puesta en marcha de la planta ISASMELT^{TM} en YCC fue impresionante cuando se comparó con las primeras plantas comerciales ISASMELT^{TM} y con los precedentes históricos. Se ha realizado cierta cantidad de estudios durante las últimas décadas viendo el rendimiento de la puesta en marcha de las plantas de proceso y éstas tienen factores destacados que desempeñan un papel importante en la determinación de qué tan rápido una nueva planta llega a su capacidad completa. En un estudio realizado por Charles River and Associates (CRA)[22], se observó que alrededor del 50% de las fundiciones demostró una producción anual promedio menor a un 70% de la capacidad de diseño desde el primer hasta el tercer año de producción. Concluyeron que una de las principales causas de los problemas en la puesta en marcha era el escalamiento inadecuado de la planta comercial del laboratorio y las etapas de la planta piloto. El alcance del trabajo del laboratorio y de la planta piloto realizado en Mount Isa contribuyó al hecho de que el escalamiento no era un gran problema para el proceso ISASMELT^{TM}. El estudio de CRA concluyó que las principales causas de los retrasos en las operaciones pirometalúrgicas eran las fallas refractarias y las dificultades para manipular los gases calientes. Estos problemas se encontraron ciertamente durante el desarrollo de ISASMELT^{TM}. El sistema de gases de ebullición le generó graves problemas a Phelps Dodge Miami por alrededor de cinco años, dando como resultado un importante rediseño y modificaciones. También fue una causa principal del tiempo de inactividad en Mount Isa durante los primeros 5 a 6 años de operación de la planta comercial. También fueron necesarios alrededor de 5 a 6 años de desarrollo en la escala comercial en Mount Isa hasta que se logró un rendimiento refractario satisfactorio. Ahora que se han superado esos problemas, sin embargo, es posible diseñar el sistema completo del horno y gases ebullición de tal forma que la ampliación de las plantas ISASMELT^{TM} ocurra muy rápidamente.

Un documento de Terry McNulty[23] analizó las historias de casos para 41 plantas de procesos diversas, incluidas seis fundiciones de cobre y níquel. Dividió los proyectos en cuatro series en base al porcentaje de la capacidad de diseño que habían logrado después de 6, 12 y 36 meses después de la puesta en marcha, según se muestra en la Figura 8.

Figura 8 – Índice de acierto del diseño anualizado (después de McNulty)



Las plantas más exitosas, aquellas de la serie 1, lograron un 100% de capacidad de diseño después de 12 meses de la puesta en marcha. Las plantas de la serie 6 lograron en promedio menor al 40% de capacidad de diseño después de 36 meses. De acuerdo con el sistema de McNulty, las fundiciones de cobre de Mount Isa y Phelps Dodge se podrían clasificar como plantas de la serie 2. Las plantas de la serie 2 tienen al menos uno y muchas veces dos o tres de las siguientes características:

- Si el proceso corresponde a tecnología con licencia, fue una de las primeras licencias (este era el caso)
- El equipo especificado para la operación de una unidad era el prototipo en términos de tamaño o aplicación (ambos hornos eran los primeros de su tamaño)
- Los trabajos de prueba a escala piloto estaban incompletos (este no era el caso).
- Las condiciones del proceso en la operación de la unidad clave eran graves, por ejemplo, temperatura alta (este era el caso)
- Las secciones innovadoras de la planta pueden haber funcionado bien, pero es posible que otras operaciones no hayan tenido la ingeniería adecuada (los sistemas de gases de ebullición eran el problema principal, pero se podría decir que estos también eran innovadores)

En la escala de McNulty, y dado el hecho de que la tasa de alimentación de la planta de YCC estaba limitada principalmente por un suministro insuficiente de concentrado, el horno ISASMELT^{TM} de YCC se podría clasificar como una planta de serie 1. Estas plantas se caracterizan generalmente por:

- Apoyarse en una tecnología madura (este era el caso)
- Tener equipos de tamaño y función similares a los utilizados en los proyectos exitosos que se habían realizado anteriormente (este era el caso)
- Si la tecnología tenía licencia existían muchos licenciatarios anteriores (este era el caso)

CONCLUSIONES

Las innovaciones son revolucionarias o evolutivas. La invención de la lanza Sirosmelt por el Dr. John Floyd hace casi 30 años fue definitivamente revolucionaria. El desarrollo de esa idea llevada a un proceso de fundición sustentable ha sido la combinación de muchas innovaciones evolutivas. La combinación de las innovaciones revolucionarias y evolutivas ha dado origen al proceso de fusión de cobre ISASMELT^{TM} que se utiliza actualmente para la producción de cobre en todo el mundo. El proceso ISASMELT^{TM} de plomo está creciendo en popularidad. Los hornos ISASMELT^{TM} secundarios y de desechos se utilizan en las fundiciones de Europa y Asia. Otras aplicaciones para el proceso están aun por ser desarrolladas.

Muchos factores han contribuido al éxito, incluidas las contribuciones de muchas personas, no sólo de CSIRO y Mount Isa, pero también en las compañías licenciatarias de todo el mundo. Se considera que la conexión cercana entre el personal de operaciones con el nuevo proceso desempeñó un papel fundamental en su éxito.

Durante los últimos diez años, ISASMELT^{TM} había tenido la mayor tasa de adopción de cualquier proceso de fusión de metales básicos, con seis plantas ISASMELT^{TM} de cobre y dos plantas ISASMELT^{TM} de plomo diseñadas y construidas. El éxito técnico y económico de estas plantas debería garantizar la adopción continua del proceso por parte de las fundiciones en todo el mundo.

AGRADECIMIENTOS

Los autores quisieran agradecer las contribuciones de las personas que reconocieron el potencial del concepto ISASMELT^{TM} desde sus inicios y a todos los que trabajaron en su desarrollo en los años intermedios. Agradecemos también a la administración de Mount Isa Mines, Phelps Dodge Miami y Yunnan Copper Corporation por otorgar el permiso para publicar datos operativos de las respectivas operaciones. También quisiera mencionar que Patricia Muza nos asistió con la preparación de los datos de operaciones de Mount Isa.

REFERENCIAS

1. J.M. Floyd, G.J. Leahy, R.L. Player, D.J. Wright, “Submerged Combustion Technology Applied to Copper Slag Treatment”, (Tecnología de combustión sumergida aplicada al tratamiento de escoria de cobre) Conferencia AusIMM, North Queensland, 1978.
2. K.R. Barrett, “Lead Smelting into the 90’s” (Fundición de plomo hacia los 90), Conferencia del vigésimo quinto aniversario de ILZIC Silver, Nueva Delhi, 1988.
3. W.J. Errington, J.H. Fewings, V.P. Keran, W.T. Denholm, “The ISASMELT^{TM} lead smelting process” (El proceso de fundición de plomo de ISASMELT^{TM}), Trans. Instn. Min. Metall., Sección C, vol 96, 1987
4. W. T.Denholm, J.M.Floyd, W.J.Errington y A.N.Parry, asignados a Mount Isa Mines Ltd, “High Intensity Lead Smelting Process” (Proceso de fundición de alta intensidad) Patente EE.UU. 4514222, noviembre 1983.
5. S.P. Matthew, G.R. McKean, R.L. Player y K.E. Ramus, “The continuous ISASMELT^{TM} lead process”, (El proceso de continuo de plomo de ISASMELT^{TM}) Lead-Zinc ’90, T.S. Mackey y R.D. Prengaman, Eds., TMS,

Warrendale, 1990.

6. R.L. Player, C.R. Fountain y J.M.I. Tuppurainen, “ISASMELT^{TM} developments”, (Desarrollos de ISASMELT^{TM}) Simposio Mervyn Willis, University of Melbourne, Melbourne, 1992.
7. C.R. Fountain, M.D. Coulter y J.S. Edwards, “Minor element distribution in the copper ISASMELT^{TM} process” (Distribución menor de elemento en el proceso ISASMELT^{TM} de cobre), Copper ’91, Volumen IV, C. Landolt, A. Luraschi y C.J. Newman, Eds., Pergamon Press, Nueva York, 1991.
8. W.J. Errington, J.S. Edwards y P. Hawkins, “ISASMELT^{TM} technology – current status and future development” (Tecnología ISASMELT^{TM}: estado actual y desarrollo futuro), South African Institute of Mining and Metallurgy Colloquium, 1997.
9. R.B.M. Brew, C.R. Fountain y J. Pritchard, “ISASMELT^{TM} for secondary lead smelting” (ISASMELT^{TM} para la fundición de plomo secundaria), Lead 90: 10° conferencia internacional del plomo, Nice, The Lead Development Association, 1991.
10. S. Jahanshahi y R.L. Player, “Process chemistry studies of the ISASMELT^{TM} lead reduction process” (Estudios de química del proceso del proceso de reducción de plomo ISASMELT^{TM}), Simposio de fundición no ferrosa, Port Pirie, 1989.
11. P. Bartsch, B. Anselmi y C.R. Fountain, “The Radio Hill Project” (El proyecto de Radio Hill), Pryosem WA, E.J. Grimsey and N.D. Stockton, Eds., Murdoch University Press, Perth, 1990.
12. C.R. Fountain, J.M.I. Tuppurainen, N.R. Whitworth ay J.K. Wright, “New developments for the copper ISASMELT^{TM} process” (Nuevos desarrollos para el proceso ISASMELT^{TM} de cobre), Extractive Metallurgy of Copper, Nickel and Cobalt (Metalurgia extractiva de cobre, níquel y cobalto), Volumen II, C.A. Landolt, Ed., TMS, Warrendale,1993.
13. J.S. Edwards, “ISASMELT^{TM}– a 250,000 tpa copper smelting furnace” (ISASMELT^{TM}: un horno de fundición de cobre de 250.000 tpa), AusIMM ’98 – The Mining Cycle, AusIMM, Melbourne, 1998.
14. R.R. Bhappu, K.H. Larson, R.D. Tunis, “Cyprus Miami Mining Corporation smelter modernization Project Summary and Status” (Resumen y estado del proyecto de modernización de la fundición de Cyprus Miami Mining Corporation), Congreso EPD 1994, G. Warren, Ed., TMS, Warrendale, 1993
15. J.E. Salle y V. Ushakov, “Electric settling furnace operations at the Cyprus Miami Mining Corporation copper smelter” (Operaciones de hornos de retención eléctricos en la fundición de cobre de Cyprus Miami Mining Corporation), Conferencia internacional Copper 99-Cobre 99, Volumen V – Smelting Operations and Advances (Operaciones y avances de la fundición), D.B. George, W.J. Chen, P.J. Mackey y A.J. Weddick, Eds., TMS Warrendale, 1999.
16. Y. Li, P. Arthur, “Yunnan Copper Corporation’s new smelter – China’s first ISASMELT^{TM}” (Nueva fundición de Yunnan Copper Corporation, el primer ISASMELT^{TM} de China), Simposio internacional de Yazawa sobre procesamiento metalúrgico y de materiales, Volumen II – High Temperature Metal Production (Producción de metal a altas temperaturas), F. Kongoli, K. Itagaki, C. Yamauchi, H.Y. Sohn, Eds., TMS, Warrendale, 2003
17. P. Arthur, P. Partington, W. Fan, Y. Li, “ISASMELT^{TM} – Not just a flash in the pan”,

Copper 2003-Cobre 2003, Volumen IV – Pyrometallurgy of Copper (Pirometalúrgia del cobre) (Libro 1), C. Diaz, J. Kapusta, C. Newman, Eds., CIMM, 2003.

18. J. Ross y D. de Vries, "Mufulira smelter upgrade project - 'Industry' Smelting on the Zambian Copperbelt" (Proyecto de mejora de la fundición Mufulira - "Industria" fundición en Copperbelt de Zambia), Pirometalúrgica 05, Capetown, Minerals Engineering International, 2005.

19. P.Arthur, B.Butler, J.Edwards, C.Fountain, S.Hunt y J.Tuppurainen, "The ISASMELT^{TM} process – an example of successful industrial R&D" (El proceso ISASMELT^{TM}, un ejemplo de investigación y desarrollo industrial exitosa), Simposio internacional de Yazawa sobre procesamiento metalúrgico y de materiales, Volumen II – High Temperature Metal Production (Producción de metal a altas temperaturas), F. Kongoli, K. Itagaki, C. Yamauchi, H.Y. Sohn, Eds., TMS, Warrendale, 2003

20. A. H. Binegar, "Cyprus ISASMELT^{TM} startup and operating experience" (Puesta en marcha y experiencia operativa de ISASMELT^{TM} de Cyprus), Copper 95 – Cobre 95, Volumen IV – Pyrometallurgy of Copper (Pirometalúrgica del cobre), W.J. Chen, C. Díaz, A. Luraschi y P.J. Mackey, Eds., TMS, Warrendale, 1995.

21. R. Player, "Copper ISASMELT^{TM} – process investigations" (ISASMELT^{TM} de cobre, investigaciones de proceso), Simposio internacional Howard Worner sobre la inyección en la pirometalurgia, M. Nilmani y T. Lehner, Eds.,TMS, Warrendale, 1996.

22. J. Agarwal y F.Katrak, "Economic impact of startup experiences of smelters" (Impacto económico de las experiencias de la puesta en marcha de las fundiciones), Charles River and Associates, documento basado en "Startup of New Mine, Mill/Concentrator, and Processing Plants for Copper, Lead, Zinc and Nickel: Survey and Analysis" (Puesta en marcha de una nueva mina, molino/concentrador y plantas de procesamiento para el cobre, plomo, zinc y níquel: encuesta y análisis), Banco Mundial, 1979

23. T.McNulty "Developing innovative technology" (Desarrollo de tecnología innovadora", Mining Engineering Magazine, Octubre 1998.

ХАРАКТЕРИСТИКИ ШЛАКА ISASMELT И АГЛОМАТЕРИАЛОВ СВИНЦОВЫХ ШАХТНЫХ ПЕЧЕЙ

Баоджун Жао, Евгений Жак и Питер Хайес
Исследовательский центр по пирометаллургии, Школа инженерного искусства,
Университет Квинсленда, Qld 4072, Австралия
P.Hayes@minmet.uq.edu.au

Ганг Янг, Джикун Ванг и Йинг Донг
Yunnan Metallurgical Group, Кунминг. Юннан. КНР

Билл Ерингтон
Xstrata Technology, Брисбен, Австралия

АННОТАЦИЯ

Yunnan Metallurgical Group (YMG) собирается использовать свинцовую печь ISASMELT принципиально новым способом, для замены свинцовой агломашины в ее основной работе по производству свинца. Затвердевший шлак должен будет загружаться в шахтную печь.

При изучении фазы равновесия этих окисленных шлаков ISASMELT выявили влияние химического состава шлака на ликвидус, его влияние на защиту фурмы и целостность футеровки печи. Описание характера типов шлака играет важную роль в определении эксплуатационных пределов процесса.

Сравнили относительную производительность агло- и ISASMELT шлака в свинцовой шахтной печи посредством определения высокой температуры размягчения и характеристик восстановления данных материалов.

Исследования по экспериментальному и термодинамическому моделированию, таким образом, позволили выявить химические характеристики шлака, соответствующие работе усовершенствованного завода, и обеспечили научную поддержку для технологических изменений.

Ключевые понятия: процесс ISASMELT, свинцовый шлак, свинцовая шахтная печь.

ВВЕДЕНИЕ

Традиционно при производстве первичного свинца с применением свинцовой шахтной печи необходимо проводить спекание свинцово-сульфидных концентратов в процессе агломерирования Dwight Lloyd для удаления серы из материала. Агломатериал в форме больших кусков без серы затем загружают в шахтную печь.

В рамках программы по уменьшению влияния на окружающую среду Yunnan Metallurgical Group (YMG) будет использовать свинцовую печь ISASMELT с целью замены свинцовой агломашины в ее основной работе по производству свинца.

Применение печи ISASMELT сильно упростит работу и позволит более эффективно отводить и использовать SO2 и сдерживать неконтролируемый выброс в атмосферу. Шлаковый продукт с высоким содержанием свинца имеет низкое содержание серы ($<0{,}5\%$) по сравнению с агломатериалом, при этом уменьшается число экологических проблем при работе шахтной печи.

YMG провели полномасштабное испытание с целью оценить эффективность шахтной печи для восстановления богатого свинцом шлака. Одна из существующих свинцовых шахтных печей (3м^2) использовалась для переработки 3000 тон шлака богатого свинцом. Результаты показали, что когда агломерат замещается крупными кусками богатого свинцом шлака при подаче в шахтную печь, производство пыли сокращается на 40%, а общие затраты на производство первичного свинца сокращаются на 8%. Другие параметры, такие как производительность, скорость извлечения свинца, скорость кокса и свинец, оставшийся в шлаке – одинаковы для обоих процессов, как для технологии с агломератом, так и для технологии со шлаком, богатым свинцом.

Эти промышленные разработки поддерживаются усовершенствованными знаниями химии процесса. На основе промышленного агломерата и синтетического шлака было проведено всестороннее комплексное исследование для сравнения характеристик микроструктуры и характеристик размягчения данных материалов. Результаты данных исследований приводятся в настоящей статье.

РЕЗУЛЬТАТЫ

Характеристики размягчения и микроструктуры

В рамках широкомасштабного исследования характеристик агломерата свинцовых шахтных печей, были получены материалы агломерата из ряда разных промышленных аломерационных заводов. Нормализованный состав сыпучей смеси этих агломератов, включая основные компоненты, приводятся в таблице 1. Средняя пористость образцов была определена между 7 и 20 %. Образцы кусков этого агломерата были сложены и отполированы для металлографического исследования. Стандартные микроструктуры при получении образцов агломерата представлены на рисунках 1a и 1d.

Основные фазы, обнаруженные при получении образцов свинцового агломерата, включают мелилит $[(Ca,Pb)_2\bullet(Al,Fe^2,Fe^{3+},Mg,Zn,Si)_3O_7)]$, шпинель $[(Zn,Fe^{2+},Mg)O\bullet(Fe^{3+},Al)_2O_3]$, оксид свинца (PbO), сульфат силиката кальция $[Ca_3(Si04)^2S04]$ и стекло. Кроме того, некоторые другие фазы, такие как нерастворенный шлак шахтной печи, двухкальциевый силикат $[2(Ca,Pb)O\cdot SiO_4]$. Силикат Pb-Zn (ларсенит, $PbZnSiO_4$), силикат Pb-Ca (Pb_3Ca_2Si,O), свинцовый сульфид(PbS), медный сульфид (CuS), сульфаты кальция Ca ($CaO-CaSO_4$ и $2CaO\cdot CaSO4$), кремний (SiO_2), металлический Pb и Cu также наблюдались в свинцовых агломератах.

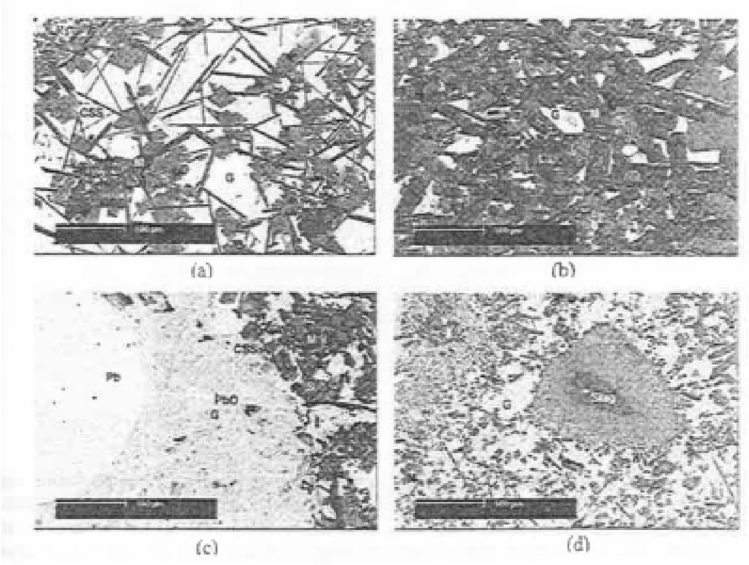


Рисунок 1. Примеры микроструктур наблюдаемых при получении агломератов:

a) Структура мелилита.

b) В основном мелилит и шпинель

c) Металлический свинец и свинцовый оксид

d) Нерастворенный шлак шахтной печи, где CSS = $Ca_5(Si04)_2SO_4$; G = стекло; M = мелилит; Pb = металлический свинец; S = шпинель

Шпинель присутствует в виде угловатых, равноосных кристаллов во всех образцах агломерата. Кристаллы мелилита представлены в виде соотношений сторон (длина к толщине) (см. рис. 1). Нижнее соотношение сторон мелилита постоянно формирует отдельные кристаллы, окруженные силикатным связующим материалом. В настоящей статье его назвали «изолированным» мелилитом, поскольку все кристаллы милилита изолированы и отделены друг от друга. Кристаллы мелилита при верхнем соотношении сторон могут быть изолированы или, при верхней концентрации, взаимодействовать, формируя структуру, подобную сетке. Исследование микроструктур свинцового агломерата непосредственно после получения показало, что структуры свинцовых агломератов не однообразные; даже в рамках исследования гранул отдельного агломерата (5-20 мм в диаметре) структура свинцового агломерата значительно отличается.

Таблица 1. Состав кусков нормализованного промышленного агломерата

Образец	Состав					
	Fe_2O_3	ZnO	CaO	SiO_2	PbO	CaO/SiO_2
11	13.6	7.3	9.7	8.5	60.9	1.14
12	14.1	7.7	10.8	9.2	58.2	1.18
13	13.3	7.1	10.2	8.9	60.6	1.15
18	16.1	9.1	12.0	9.3	53.6	1.30
111	15.6	9.5	12.3	9.6	53.0	1.28

114	14.8	9.4	11.0	9.1	55.7	1.20
117	16.4	10.3	12.9	9.9	50.6	1.30
120	14.3	8.9	10.8	8.9	57.0	1.21
123	17.7	9.6	8.5	8.6	55.6	0.99
MP1	15.2	7.0	10.0	10.7	57.1	0.94
MP2	16.2	7.4	10.9	11.2	54.3	0.97
NP1	21.7	6.6	9.0	9.4	53.3	0.95
NP4	23.4	6.3	10.8	12.3	47.2	0.88
PP1	14.5	9.9	7.4	10.3	58.0	0.72
PP2	14.3	10.1	7.0	10.4	58.2	0.67
PP5	13.5	11.1	7.2	9.1	59.1	0.79
PP8	13.4	11.1	6.7	8.5	60.3	0.78

В процессе производства промышленного агломерата, материал медленно охлаждается в пределах агломашины. Замеры установки показывают, что пиковая температура штабеля обычно достигается в диапазоне 900-1150^{o}C, а время охлаждения до температуры 800^{o}C от 10 до 30 минут. В процессе этого медленного охлаждения имеет место кристаллизация твердой фазы из жидкой. Объединение фаз представленных в образцах агломерата при температуре может быть выявлено посредством повторного нагрева в воздухе при выбранных температурах в течение 10 минут, быстрого охлаждения непосредственно в воде и затем повторной проверки структур. Стандартные микроструктуры повторно нагретого и быстро охлажденного агломерата показаны на рисунке 2.

После обработки при выбранных температурах между 900 и 1100^{o}C было обнаружено, что структуры агломерата становятся более однородными, чем в агломератах непосредственно после получения. Весь шлак шахтной печи и флюсовые материалы, присутствующие в материалах, которые поставляются на переработку, встраиваются в агломераты. Это указывает на то, что не все материалы агломерата подвергались воздействию данных температур в процессе агломерации. Очевидно, что соотношение стеклофазы в охлажденных образцах намного выше, чем в агломератах непосредственно после получения, поскольку не допускается кристаллизация. Было обнаружено, что жидкая фаза и фаза шпинель были обнаружены во всех повторно нагретых и охлажденных образцах. Кроме того, фазы мелилита и двухкальциевого силиката также присутствуют в некоторых образцах, в зависимости от соотношения CaO/SiO_2. В агломератах имеющих более низкое соотношение CaO/SiO_2 (менее чем приблизительно 0,8 по весу) двухкальциевый силикат не присутствует в исследованном температурном диапазоне. С увеличением соотношения CaO/SiO_2 могут присутствовать как двухкальциевый силикат, так и мелилит. При высоком соотношении CaO/SiO_2 двухкальциевый силикат превалирует. Для эквимолярного CaO и SiO_2 (соотношение CaO/SiO_2 приблизительно 0,93 по весу), формируется мелилит с высоким отношением длины к высоте, однако распространенные каркасные структуры не наблюдаются в этих образцах при температуре выше 1000^{o}C.

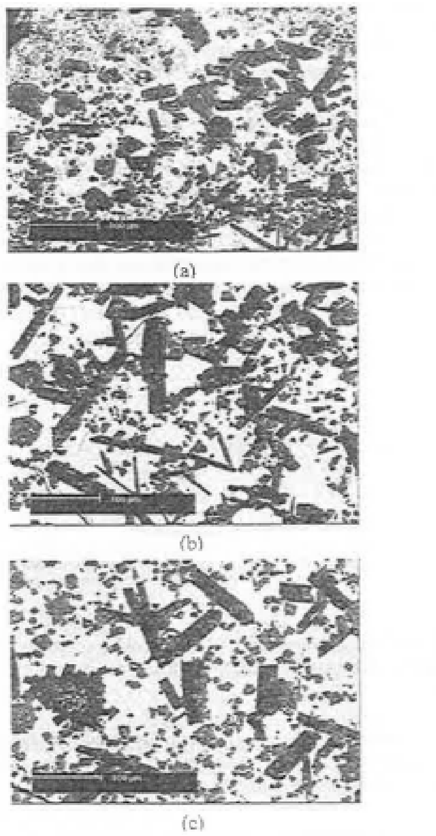

(a)
(b)
(c)

Рисунок 2 – Стандартные микроструктуры Свинцовых Агломератов охлажденные в воздухе при а) 900 ^{o}C; b) 1000 ^{o}C, c) 1100 ^{o}C

Были проведены замеры температуры размягчения образцов кускового свинцового агломерата в воздухе с применением щупа для измерения взвешенного смещения. Из собранных данных была построена температурная кривая на графике с двумя плоскостями, смещение в зависимости температуры. Стандартные примеры результатов испытания по размягчению агломерата, проведенного на образцах, полученных из данной партии агломерата, представлены на рисунке 3. Можно видеть, что в каждом испытании образец деформируется в диапазоне температур. Обнаружено, что в некоторых образцах первоначальное размягчение происходит уже при температуре 700 °C. Конечные температуры размягчения, температуры при которых имеет место полное разрушение образца под нагрузкой, были установлены в диапазоне от 800-1100 ^{o}C. Было проведено несколько отдельных замеров для каждого агломерационного материла, и совершенно очевидно, что имеются некоторые характерные изменения в поведении образцов из одного и того же материала. Можно увидеть, что стремительное частичное разрушение агломерата происходит в некоторых образцах по мере того как зонд проникает в образец. Исследование этих образцов показало, что образцы, имеющие небольшие колебания температуры размягчения, имеют более однородную микроструктуру. Образцы, имеющие большие колебания температуры размягчения, и которые подвергаются частому и беспорядочному разрушению из-за температуры, имеют неоднородную микроструктуру. Такие локальные различия в структуре агломерата в форме изменений в представленных фазах, а также соотношениях фаз, являются результатом локальных вариаций состава. Эти вариации в свою очередь приводят к разницам температур размягчения и плавления, а также вязкости материалов.

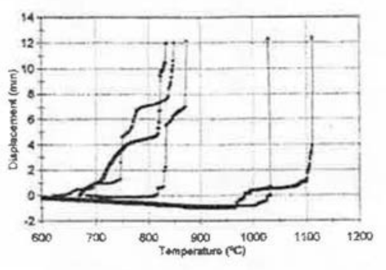


Рисунок 3. Кривые температур размягчения стандартного свинцового агломерата в воздухе.

(Скорость нагревания 400°С/минуту; щуп – 8 мм в диаметре, нагрузка 235 г.)

Хотя в настоящей статье не сообщается, обширные исследования, проведенные в рамках данной исследовательской программы, показывают, что следующие условия благоприятно влияют на формирование однородной структуры агломерата и высокий процент отведения серы:

- Высокая максимальная температура штабеля.
- Небольшой размер и ограниченный гранулометрический состав загрузки агломашины.
- Хорошо контролируемое содержание воды в агломерированной шихте.
- Равномерно высокая пористость штабеля и равномерное распределение подачи агломерата на слой агломерата
- Хорошее возгорание слоя агломерата

Эти проблемы в целом хорошо известны практикующим специалистам, но совсем не обязательно, что их можно легко контролировать на практике на заводе. Такие характеристики как размер оборотного агломерата и гранулометрический состав важны для определения характеристик гранулирования загрузки агломашины. Свойства оборотного агломерата в свою очередь зависят от предыдущей работы аглоустановки; временные константы устраняют тот факт, что эти вариации могут быть во много раз дольше, чем однократное время цикла аглопроцесса, в зависимости от буферной емкости, оборота агломерата и рабочих скоростей на конкретных заводах.

Характеристики микроструктур/размягчения синтетического шлака

Для сравнения характера промышленных агломератов с возможными шлаками, которые могли бы быть использованы в качестве сырья для свинцовой шахтной печи, в лаборатории подготовили несколько типов синтетического шлака. Образцы были подготовлены путем плавки приблизительно 1 кг аналитических химических реагентов в воздухе в алюмосиликатном тигеле. Состав этих шлаков приводится в таблице II. В образцах MIMPTS-8 соотношение $PbO/ZnO/Fe_2O_3$ и содержание SiO_2 поддерживаются приблизительно постоянными, а CaO прогрессивно увеличивается для производства шлаков с соотношением CaO/SiO_2 между 0,19 и 0,7. Эти синтетические шлаки обычно имеют более высокие Fe_2O_3 и ZnO и более низкая концентрация CaO по сравнению с традиционными агломератами свинцовых шахтных печей (см. Таблицу I)

Таблица II – Состав валового синтетического шлака

Образец	Состав по весу					
	Fe_2O_3	ZnO	CaO	SiO_2	PbO	CaO/SiO_2
MIMPT5	20.7	11.3	2.2	11.5	53.2	0.19
MIMPT5	19.9	11.2	3.7	10.8	54.0	0.34
MIMPT7	16.9	11.4	6.7	11.9	51.8	0.56
MIMPT8	17.3	10.8	7.6	11.0	52.1	0.69

Стандартные микроструктуры синтетических шлаков, охлаждаемых при разных температурах, приводятся на рисунках 4 и 5 для MIMPT5 и MIMPT6 соответственно. Для образцов MIMPT5 и MIMPT6, которые имеют более низкие соотношения CaO/SiO2, мелилит не формировался при температуре плавления (1100-1200oC). Мелилит, обнаруженный в медленно охлаждаемых образцах, указывает на то, что данная фаза формируется при более низких температурах в процессе охлаждения. Однако, в шлаках MIMPT7 и MIMPT8, в которых соотношение CaO/SiO2 выше, мелилит формировался при температуре плавления.

Состав фаз, представленный в образцах шлака, был проанализирован с применением электронного зонда рентгенмикроанализа (EPMA). Состав фазы шпинель во всех образцах был очень близок к стехиометрическому $ZnO \cdot Fe_1O_3$. Было обнаружено, что состав жидкой фазы, представленный в быстро охлаждаемых шлаках, не менялся значительно с увеличением температуры между 1050 и 1150 ^{o}C, а содержание PbO между 65-75 % по весу.

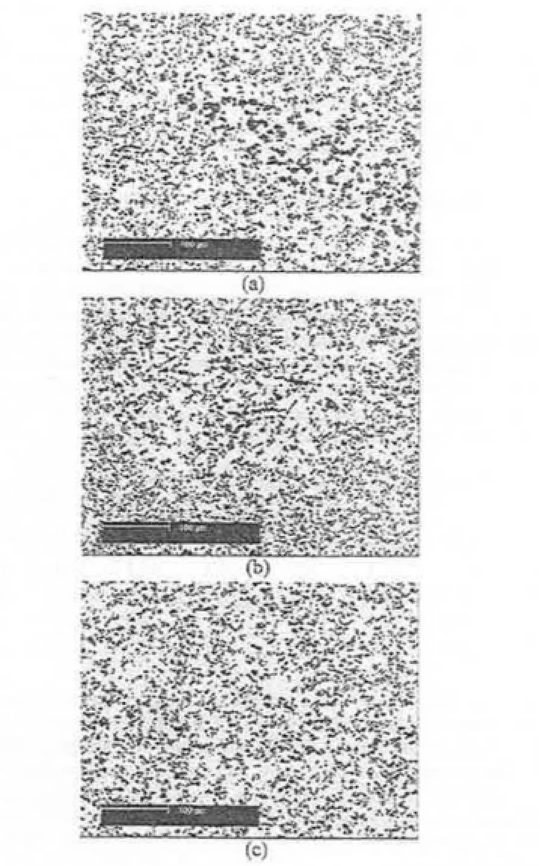
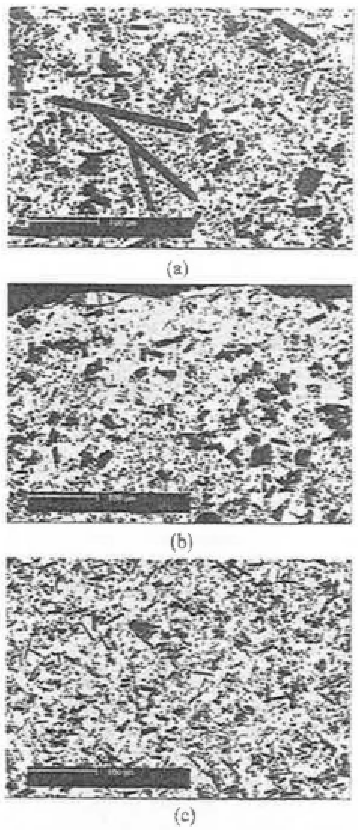


Рисунок 4. Стандартные микроструктуры MIMPT5 быстро охлажденные на воздухе с

a) 1050 °C; b) 11 00 °C; c) 11 50 °C



(a)

(b)

(c)

Рисунок 5. Стандартные микроструктуры MIMPT7 быстро охлажденные на воздухе с

a) 1050 °C; b) 1100 °C; c) 1150 "C

Были определены температуры размягчения на воздухе для медленно охлаждаемых шлаков MIMPT5 и MIMPT8. Было проведено несколько отдельных измерений для каждого образца шлака. Было обнаружено, что для одного и того же типа шлака, температура окончательного размягчения была в основном в пределах относительно узкого диапазона ($30^{o}C$). Это происходит в результате относительно однородных структур шлака по сравнению с агломерационным материалом. Если среднее значение принять за температуру размягчения для каждого шлака, влияние соотношения CaO/SiO2 на температуру конечного размягчения T_F свинцовых шлаков показано на рисунке 6. Можно увидеть, что если соотношение $PbO/ZnO/Fe_2O_3$ поддерживается постоянным, температура размягчения свинцовых шлаков увеличивается с увеличением соотношения CaO/SiO_2.

Это интересное открытие, поскольку можно было бы предположить, что добавление CaO в шлак приведет к снижению вязкости жидкой фазы, и, следовательно, к понижению температуры размягчения. Данные результаты указывают на то, что другие факторы, такие как присутствующее фазовое соединение, могут нести ответственность за данное направление.

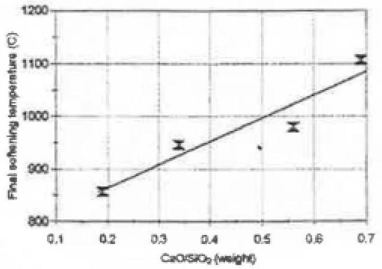


Рисунок 6. Отношения между средними температурами конечного размягчения T_F на воздухе и соотношение валового CaO/SiO2 для состава шлаков показаны в таблице II.

Как указано выше, существуют значительные вариации температур размягчения промышленного свинцового агломерата непосредственно после получения, эти вариации в основном связаны с местными различиями состава и структуры агломатериала. Эти различия возникают из-за локальных изменений физической и химической формы материалов подаваемых в процесс, и термических данных и данных окружающей среды, которые воздействовали на материалы внутри аглопроцесса. Поэтому очень сложно точно рассчитать характер размягчения агломатериалов. Однако, свинцовый шлак имеет более однородный состав и структуру и его характер размягчения можно рассчитать более точно.

Расчет фазового равновесия

С применением оптимизированной термодинамической базы данных для данной системы ZnO-Fc_2O_3-FeO-PbO-CaO-SiO_2 (4) представленные фазы и их пропорции в равновесии в свинцовых шлаках были рассчитаны при помощи FactSage 5.1. Весовая доля, полученная из расчетов была конвертирована в функцию объема данных фаз с допущением, что: a) частичный молекулярный объем каждого компонента является постоянным и одним и тем же в жидкой и твердой фазах, и b) в диапазоне рассмотренных температур нет температурного расширения. Неопределенности, связанные с данными допущениями незначительны и находятся в пределах неопределенности этих данных.

Расчетные функции объема основных фаз, представленных в синтетических свинцовых шлаках как функция температуры, представлены на рисунках 7a до 7d. Примеры кривой размягчения, определенной для каждого из этих синтетических шлаков, также обозначены на этих рисунках.

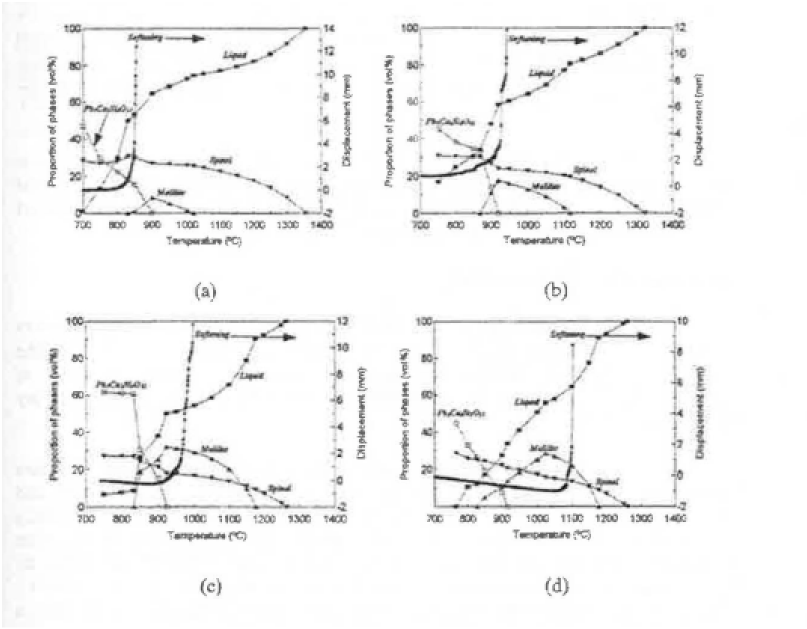


Рисунок 7 ·Соотношения фаз при разных температурах в воздухе.

(Рассчитанные при помощи FactSage) для

a) MIMPT5; b) MIMPT6; c) MIMPT7; d) MIMPT8

Из рисунков 7a - 7d можно увидеть, что для всех рассмотренных типов шлака, шпинель постоянно присутствует. Кроме того, мелилит и Pb-Ca силикат ($Pb_3Ca_2Si_3O_{11}$) также являются распространенными фазами в этих шлаках. С увеличением температуры доля жидкости постепенно увеличивается, а доля шпинеля уменьшается. Отмечается, что существует специфический температурный диапазон для стабильного формирования фазы мелилита. Например, из рисунка 7a видно, что когда MIMPT5 нагревают в воздухе, мелилит начинает формироваться при 830 ^{o}C и достигает максимума (7% об.) при 900^{o}C. Свыше 900 ^{o}C доля мелилита уменьшается с увеличением температуры и исчезает при 1020 ^{o}C. Также предполагается, что при температуре ниже 900 ^{o}C $Pb_3Ca_2Si_3O_{11}$ будет формироваться в этих образцах, а доля будет увеличиваться с уменьшением температуры.

Эксперименты с быстрым охлаждением для этих шлаков показывают, что мелилит присутствует в MIMPT5 только при температурах ниже 1050^{o}C. В MIMPT7 мелилит все также стабильный при 1150^{o}C. Эти наблюдения хорошо коррелируют с расчетами FactSage, которые показаны на рисунках 7a и 7c. Для измерения объемного содержания жидкой и твердой фаз для охлажденных шлаков был применен количественный анализ и эти результаты сравнили с расчетами FactSage. И снова они хорошо соотносились с расчётами FactSage и экспериментами.

Можно увидеть из рисунков 7a и 7d, что для синтетических шлаков MlMPT5 до MIMPT8, температуры окончательного размягчения соответствуют наличию 50-65 % об. жидкой фазы. Это значительный результат, поскольку он указывает еще раз на значимость расчетов фазового равновесия в технологии и выборе шлаковых материалов.

Восстановление свинцового агломерата и шлака

Были проведены предварительные эксперименты по восстановлению свинцового агломерата и синтетических шлаков в углеродных тигелях при температурах в диапазоне от 800 °С до 1200°С. Метод эксперимента был описан в другой работе (5). Схематически, степень завершенности реакции для каждого образца сопровождается измерением объема газа, производимого при любой температуре и за любой промежуток времени.

Агломерат I8 и шлак MIMPT7 были отобраны для предварительного эксперимента по восстановлению. MIMPT7 и I8 имеют сходный состав в плане содержания железа, цинка и свинца, если не учитывать, что I8 имеет коэффициент CaO/SiO2 = 1.3, MTMPT7 имеет коэффициент CaO/SiO2 = 0,56. Для эксперимента по восстановлению использовали одинаковое количество шлака и агломерата (5,00 кг). Размер частиц, использованных для эксперимента, был одинаковым для шлака MIMPT7 и агломерата I8. Изотермические эксперименты по восстановлению проводились при 800, 900, 1000, 1100 и 1200 °С. Обратите внимание, что газы, получаемые из реакции между агломератом/шлаком и углеродом, фактически представляют собой смесь CO и CO_2. Соотношение CO к CO_2 не было определено в настоящем исследовании. Однако общий объем CO и CO_2 полученный из данной реакции указывает на относительную степень завершенности реакций по восстановлению.

На рисунках 8 и 9 показаны результаты экспериментов по восстановлению I8 и MIMPT7 соответственно. Для каждого образца результаты представлены в виде объема выделения газа как функции от времени реакции и температуры соответственно. Из рисунков 8 и 9 можно увидеть, что

1) Как для шлака MlMPT7 , так и для агломерата I8 при постоянной температуре, объем первоначально выделяющегося газа, быстро увеличивается, а затем замедляется по мере продолжения реакции.
2) При температуре ниже 900 °С между агломератом/шлаком и углеродом происходит только ограниченная реакция. При температуре выше 900 °С объем выделяемого газа значительно увеличивается с увеличением температуры.
3) В рамках всего исследованного диапазона температур агломерат I8 имеет больший объем выделяющегося газа, чем шлак MIMPT7, но данная разница варьируется в зависимости от температур.

Можно увидеть, что за первые 20 минут данной реакции между I8 и углеродом, объем выделенного газа увеличивается с увеличением температуры. Однако, после 20 минут реакции, объем выделяемого газа при 1200°С становится ниже, чем при равных условиях при 1100 °С. При рассмотрении примера при 1200°С было обнаружено, что слой металлического свинца формировался на поверхности между образцом агломерата и углерода. Оказалось, что в данном случае слой металлического свинца сокращает контактную площадь между агломератом и углеродом, при этом сокращая общую скорость реакции.

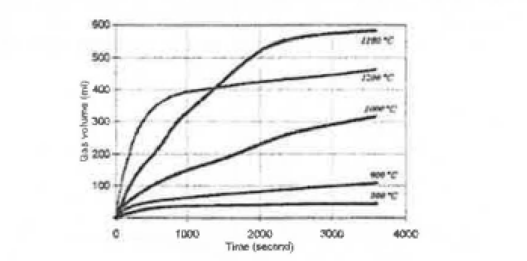


Рисунок 8. Выделение газа из свинцового агломерата I8 в графитовом тигеле как функция от времени реакции.

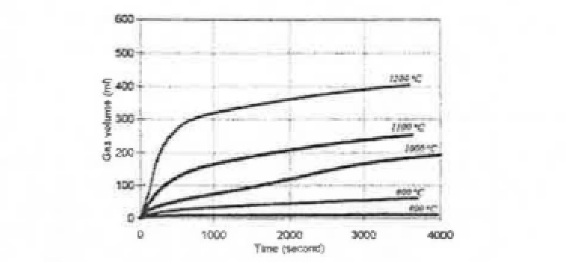


Рисунок 9. Выделение газа из шлака MIMPT7 в графитовом тигеле как функция от времени реакции.

На рисунке 10 показано влияние соотношения CaO/SiO_2 на скорость восстановления шлака при 1000 °C. Как видно при данной температуре увеличивающееся соотношение CaO/SiO_2 приводит к увеличению скорости восстановления. Однако последующие эксперименты показали, что при температуре свыше 1100°C скорость восстановления всех типов шлака, которые подвергались испытанию, оказалась почти одинаковой. Эксперименты также проводились на различных типах шлака для оценки влияния PbO, Al_2O_3 или MgO на скорость восстановления. Было обнаружено, что при низкой температуре (800-1000 °C) скорость восстановления увеличивается с увеличением концентрации PbO в шлаке. При более высокой температуре (1100 – 1200 °C) концентрация PbO в шлаке не имеет значительного влияния на скорость восстановления. Было обнаружено на основании экспериментов, что добавление 2% по весу MgO может немного увеличить скорость восстановления, а добавление 4% по весу Al_2O_3 уменьшает скорость восстановления.

В то же время некоторое косвенное восстановление агломерата за счет газа CO происходит при температуре ниже 900 °C, было показано экспериментально, что степень восстановления кускового агломерата относительно небольшая. Начальная скорость реакции шлаков с углеродом быстро увеличивается с увеличением температуры в диапазоне 900-1000 °C как показано на примере результатов, продемонстрированных на рисунке 9.

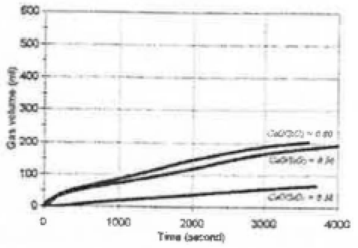


Рисунок 10. Влияние весового соотношения CaO/SiO2 на выделение газа из шлаков богатых свинцом при температуре 1000 ^{o}C в графитовом тигеле.

Значение для промышленной практики

В процессе работы свинцового завода ISASMELT в Маунт Айза было замечено, что присутствие кристаллов цинкового феррита в расплавленном шлаке было связано с формированием стабильного слоя на огнеупорных кирпичах, что обеспечило эффективную защиту кирпичей от износа и эррозии. Наличие этих равноосных кристаллов, таким образом, является благоприятным при условии, что они не значительно увеличивают вязкость шлаков с высоким содержанием свинца до такой степени, когда шлак невозможно выпустить из печи и при условии, что это не приводит к восстановлению в рабочем пространстве печи.

Чтобы свести к минимуму образование возгонов в процессе окислительной плавки ISASMELT свинцовых концентратов желательно поддерживать рабочую температуру ванны как можно более низкой. Расчеты по фазовому равновесию показывают, что температура ликвидуса этих шлаков увеличивается с увеличением CaO/SiO_2. В традиционной практике свинцовых шахтных печей используют коэффициент CaO/SiO_2 между 0,7 и 1,2 с целью сведения к минимуму концентрации растворенного и захваченного свинца в шлаке (6). В новом процессе ISASMELT/YMG необходимо добавлять известняк непосредственно в загрузку шахтной печи, а не в шлак для плавки.

ЗАКЛЮЧЕНИЕ

Была проведена обширная программа исследования с целью изучения характеристик агломератов свинцовой шахтной печи и окисленных шлаков ISASMELT,а также факторов влияющих на микроструктуру и свойства агломерата. Были исследованы характеристики размягчения свинцового агломерата при высокой температуре под нагрузкой. Существует явная корреляция между изменениями температуры размягчения агломерата и изменением агломерата на уровне макро-, мезо- и микроструктур; незавершенная реакция и большие различия в составе приводят к изменчивости свойств кускового агломерата.

В отличие от этого, синтетические шлаки имеют относительно однородную структуру и объемные свойства. Температуры размягчения синтетических шлаков аналогичны температурам агломератов того же состава, и увеличиваются при увеличении весового соотношения CaO/SiO_2 . Фазовое соединение и фазовый состав в синтетических шлаках соответствуют расчетам FactSage с применением оптимизированной базы данных для системы $ZnO\text{-}Fe_2O_3\text{-}FeO\text{-}PbO\text{-}CaO\text{-}SiO_2$, с указанием, что данные материалы могут быть описаны, если предположить, что химическое равновесие достигается при температуре. Характеристики размягчения шлаков, изученные в данном исследовании, хорошо коррелируют с фракцией жидкой фазы, присутствующей в данном материале.

Первоначальные лабораторные тесты по восстановлению, проведенные в графитовом тигеле, указывают на то, что при температуре ниже 900 оС происходит незначительное восстановление свинцового оксида из шлака и агломертов. Скорость восстановления шлака при температуре 1000°С увеличивается при возрастании коэффициента CaO/SiO_2 для исследованного состава шлаков. Однако, не существует значительной разницы в скорости восстановления этих агломерационных и шлаковых материалов при температуре свыше 1100°С. Эти данные указывают на то, что есть небольшая разница между свойствами свинцового агломерата и шлака в работе шахтной печи.

СПРАВОЧНАЯ ЛИТЕРАТУРА

1) B. Errington, J. Wang, P. Arthur a11d Y. Dong, «The lsasmelt/YMG Lead Smelting Process", Lead & Zinc'Q5, Kyoto, Japan. 2005.
2) E. Jak, B. Zhao, L Harvey and P.C. Hayes. "Experimental Study of Phase Equilibria in the PbO-ZnO-'Fe_2O_3 '-(Cao+SiO_2) System in Air for the Lead and Zinc Blast Furnace Sinters (CaO/SiO_2 Weight Ratio of 0.933 and Pb/(CaO+SiO_2) Ratios of 2.0 and 3.2)", Metall. Trans B, Vol.34B, 1003, 383.-397
3) E. Jak and P.C. Hayes, "Experimentai Liquidus in the PbO - ZnO -"Fe_2O;"-(CaO+SiO_2) System in Air with CaO/SiO_2 Weight Ratio of 0.35 and PbO/(CaO+SiO_2) = 3.2", Metall Trans B, Vol. 33B, 2002, 851-863.
4) E. Jak. S. Degterov, P.C. Hayes and A.D. Pelton, "Theanodynamic Modeliing of the System PbO-ZnO-FeO-Fe_2O_3-CaO-SiO_2 in Zinc Lead Smelting Conf. on Slags and Fluxes, Sydney, ISS, Warrendale. 1997, 621-628.
5) N.N. Kinaev, E. Jak and P.C. Hayes, "Kinetics of Reduction of Lead-Silicate Slags with Solid Carbon", Scand. J. Met. 2005, accepted.
6) A. H.-J. Siegmund, "Primary Lead Production - a Survey of Existing Smelters and Refineries", Lead-Zinc 2000, J.E. Dutrizac, I.A. Gonzalez, D.M. Henke, S.E. James and A.H.-J. Siegmund, Eds., Pittsburgh, Pa, TMS, Warrendale, 2000, 55-116.

БЛАГОДАРНОСТИ

Авторы желают выразить благодарность Австралийскому Научному Совету (ARC-SPIRT) за оказание финансовой поддержки, которая позволила провести данное исследование. Авторы также желают выразить благодарность руднику Mount Isa , компании Xstrata Technology, Pasminco Sulphide Corporation (сейчас Zinifex), Nornnda Ltd. и Yunnan Metallurgy Group за оказание финансовой помощи, предоставление полевой информации и образцов агломерата для данного проекта.

Г-ну Дугласу Прайсу и г-же Йинг Ю, которые предоставили основную помощь при лабораторной работе и проделали основную работу по подготовке. Г-ну Рону Рашу из Центра микроскопии и микроанализа при университете Квинсленда, который обеспечил техническую поддержку по применению рентгеновского микроанализа (EPMA).

云南铜业股份公司的新熔炼厂

——中国第一座艾萨炉

李 云[1]　 菲力普·阿瑟[2]

[1]: 中国云南昆明 西山区 王家桥（650102）云南铜业股份有限公司

[2]: 澳大利亚布里斯班 Wickham Terrace 87号（4000） MIM工艺技术公司

摘 要

1999年，云南铜业股份有限公司（YCC），决定对原有铜冶炼流程进行改造，即从原有的鼓风烧结机-矿热电炉联合流程改造为艾萨熔炼炉-贫化电炉联合流程。改造的主要目的是：节能降耗，改善环保。

总的来说，这个项目是成功的，且部分归功于选择了一个可信赖的、已取得成功实践的技术——艾萨熔炼。云铜艾萨炉的设计能力为60万吨干精矿/年。自2002年5月以来，该炉子已平稳运行了一年多。YCC在工厂操作的准备上花费了相当的精力。开炉前，一个YCC骨干人员组成的培训队伍被派到芒特艾萨，使他们对艾萨工艺的理解取得显著提高，并确保该项技术的顺利移交。

这片文章总结了云铜艾萨工程的历史和开炉初期的操作情况。

介 绍

中国的冶炼工业在迈入二十一世纪后，面临一系列的问题。首先，基于对提高生产效率的要求，促使一些国有企业开始进行股份制改革。第二，政府开始致力于降低重工业对环保的压力。这就极大地鼓励了那些进行股份制集资以实现其冶炼流程现代化的公司，以国内外现代化的技术更新其过时的工艺。云铜选择了引进国外新技术。

云南铜业股份有限公司

云南铜业股份公司位于中国云南昆明西郊区，是一个具有45年炼铜历史的企业，目前年产高纯阴极铜20万吨、硫酸45万吨、白银200吨、黄金2吨、电工用铜线杆6万吨。股票“云铜铜业”1998年6月2日上市交易，1999年4月1日入选美国道琼斯综合指数股，1999年10月8日深圳成份指数股。

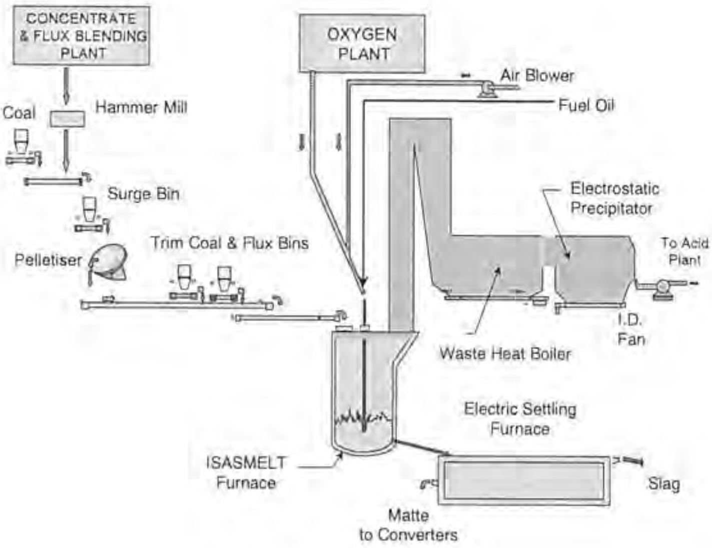
云铜的改造项目早在公司上市之前已经开始。以节能降耗改善环保为目标，1993—1996年公司对工艺流程的改造进行选型、论证。1997年—1998年针对艾萨炉炉型开展工作，1999年3月21日与MIMPT签订艾萨法铜熔炼技术合同，1999年9月28日—10月8日召开了项目分工会。开始破土动工，2001年设备开始安装，2002年5月9日炉子点火，15日投料生产。从开始生产至2002年9月30日，总日历作业率达到75.8%，扣除计划停产和外因停产后的艾萨炉作业率达87.1%。实现了一次投产成功。

1、　艾萨炉生产的基本工艺流程

艾萨铜冶炼技术，是由澳大利亚芒特艾萨矿业公司在约20年时间里发展一种熔池熔炼技术。该项技术已在一系列的发表文章[1~7]中详细阐述。通过喷枪把富氧空气强制鼓入熔池，使熔池产生强烈搅动状态加快了化学反应的速度，充分利用了精矿中的硫、铁氧化放出的热量进行熔炼，同时产出高品位冰铜。熔炼过程中不足的热量由燃煤提供。

云铜艾萨炉的设计能力为60万吨干精矿/年。工艺流程图见附图1。主要来源于国内矿山的各种铜精矿，在备料工序与熔剂进行混合。燃煤的主要部分在

精矿途经“鼠笼破碎机”后加入到精矿中，混有大部分燃煤的混合铜精矿经过4台制粒机中的几台后，补充进剩余的燃煤和熔剂，然后被运送进入艾萨炉中熔炼。熔炼需要的富氧空气通过喷枪鼓入熔池。为了便于生产期间的温度控制，还从喷枪加入燃油对炉温进行微调，熔炼产生的冰铜和炉渣混熔体进入贫化电炉进行澄清分离。冰铜被送到P-S转炉吹炼，炉渣水淬后抛弃。烟气经余热锅炉和电收尘器降温除尘后进入硫酸厂制酸，余热锅炉收下的烟尘经过破碎后返回电炉，电收尘器收下的烟尘送入电炉。



（*图1：工艺流程图*）

2、 工程设计、施工

艾萨工程建设也早在1999年初便开始动作,云铜围绕艾萨熔炼技术改造，开始了一系列的配套工程建设。为了便于指挥协调、加快工程建设，云铜将这些工程按区域和性质划分为四个大区：熔炼区、硫酸区、制氧区和动力区。

熔炼区即艾萨炉主体工程。艾萨炉的主体工程基础设计及相关接口设计由MIMPT负责，与艾萨炉相配套的余热锅炉的设计由德国OSCHAZS负责，其它的工程设计由中国北京有色冶金设计研究总院（ENFI）负责。YCC负责对各设计方工作的确认和协调。经过2000年4月在布里斯班的初步设计审查会、2000年8月在昆明的详细设计审查会，设计工作已基本完成。

工程施工由中国的本地施工单位负责完成。艾萨炉主厂房桩基的施工单位为有色昆明基础工程公司，试桩单位为昆明勘察院科技开发公司，艾萨炉基础承台及艾萨炉系统土建工程的承建单位是十四冶金建设公司，艾萨炉安装工程的承包方为云南省第一安装工程公司，工程内容：炉壳制造、厂房钢结构、艾萨炉系统设备、管道、电气、仪表，到2002年3月，绝大部分工作已完成。艾萨炉的筑炉施工由十四冶筑炉公司，在奥镁公司现场服务人员的指导下于2002年3月22日全部结束。艾萨炉钢结构厂房防腐工程、喷砂除锈、喷氯化橡胶底漆、面漆工程是由云冶建设公司负责，也在2002年4月全部完成。

艾萨炉各系统子项工程包括：熔炼炉、加料、控制室、应急熔剂及煤、喷枪、加热烧嘴、保温烧嘴、测量棒、泥炮开口机、熔体放出区域、新鲜空气（微正压送风）、通风（环保排烟）、返料、煤粉、卸砖、冷料、二次燃烧风、供风（氧）管道、燃油管道、冷却循环水、生产及生活给排水、吊车、电梯等

。

余热锅炉工程包含的各系统有：余热锅炉、锅炉给水、强制循环、自然循环、蒸汽管道、紧急冷却、刮板输送机（除灰）、破碎、烟尘输送、烟气排放、电动葫芦、保温等。

其它还有电力、仪表以及熔炼DCS和余热锅炉DCS、总图、厂房建筑结构等。

3、 工厂设计

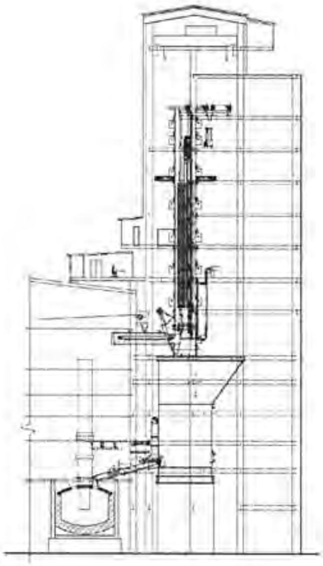
就象其它任何一个老厂改造的项目一样，云铜的工厂布置和工程设计面临着场地限制的挑战。艾萨熔炼炉及其余热锅炉，必须布置靠近原有的2#电炉，并在一个由原有2#电炉、转炉工段、电收尘器所构成的“包围圈”内的狭小场地上展开施工但设计者和施工人员最终还是解决了这一困难，同时也体现出艾萨熔炼炉占地小、适合旧址改造的优势。图2所示了拆毁掉不需要的建筑和设备后留下的施工现场。

（图2：施工现场）

由于昆明地处地震多发区，使施工的要求变得更加复杂。所以，厂房基础的设计须格外地小心。MIMPT提供了静态负荷和预期生产时动态负荷的相关数据，以进行基础和炉子支撑的设计。

通过新增的一条皮带廊，物料从原来烧结机使用的物料准备和运输线路运送到艾萨炉。艾萨熔炼炉被设计地尽可能靠近贫化电炉，熔体通过两个排放口和两个排放溜槽输送进入贫化电炉。余热锅炉安装在艾萨熔炼炉的对面。余热锅炉包括：一个辐射段、炉子烟气出口上面的上升烟道和下降烟道、最后是对流段。新增一段烟道，以将冷却后的烟气输送到一对改造过的电收尘器。

艾萨熔炼炉内径为4.4米，由高品质的耐火镁铬砖砌筑炉衬。炉顶由锅炉管束制成而成为余热锅炉的一部分。艾萨喷枪直径400mm，由特殊设计的喷枪架负责实现喷枪的下降和提升。图3显示了艾萨厂房的立面图。



（图3：艾萨厂房立面图）

4、　生产准备

生产准备工作早在1998年组成艾萨工艺小组，参与了技术方案的讨论、选择工作，以后又补充了其它专业人员参与了设备选购、安装调试、试车等工作。公司专门成立了艾萨炉生产准备领导小组，对艾萨炉的投产准备工作做系统的安排。总的认识是：扎实的培训和有效的准备工作，是评估和规避工程、生产风险的重要手段。所以公司决定采纳MIMPT建议的在芒特艾萨进行全面培训。

1、　人员培训：

- 1998年成立的工艺小组由冶炼工艺技术人员组成，针对艾萨工艺进行学习和论证。1999年至2001年，对熔炼分厂员工进行艾萨工艺知识的培训，共计向200名职工授课430个小时，通过考试选拔了一批预备开炉人员
- 2001年，在MIMPT的安排和管理下，YCC通过严格的考试选拔出的各专业（冶炼、电器、仪表、机械）工程师、CRO和排放工，赴芒特艾萨进行总时间为7个月的培训。受训者通过在芒特艾萨的培训，对艾萨工艺、设备、安全和配套管理，有了更深刻的认识。在从芒特艾萨回来前，YCC的培训人员已能在芒特艾萨操作工的监护下进行艾萨炉的操作，并准备好将所学到的理论和经验运用于而后的工作实践中。

（图4：YCC的培训人员在芒特艾萨）

·从芒特艾萨回到YCC后，根据开炉和生产的需要，从预备开炉人员中向各岗位补充了操作人员，再次进行培训，并组织所有预备操作人员参与了设备的安装和调试过程，使他们在生产前对设备有尽可能多的了解。

·余热锅炉的人员培训：2000年安排了13名操作工进行培训，工厂培训在国内企业进行，工程师和操作者同样参与了设备的安装、调试、检查过程。

实践证明，YCC从这样的培训准备工作中受益非浅。

2、 *操作规程及相关制度的编写*：2001年5月开始编写艾萨炉操作规程，7月编写艾萨区域的相关安全、管理制度。操作规程译成英文传给MIMPT修改，经修改后，10月按操作规程对所有操作人员进行再次培训、统一对规程的认识。

3、 *风险方案的制定及培训*：2002年1~2月，针对生产准备中发现的问题，提出艾萨炉和余热锅炉开炉、生产中可能的风险预测清单。4月，在MIMPT现场服务人员的帮助下完成最后的风险规避方案，并确保每一名操作工都得到一份复印件。

4、 *设备检查表*：为设备的使用及日常维护制定检查表格，明确设备检查的要求。包括：

- 检查内容、范围、方法的要求。
- 检查人员及培训的要求。
- 检查时间的要求。
- 检查记录和处理的要求。

5、 *参与安装调试*：成立工艺、机械、仪控、电器、热工、水分析等专业组，对现场的安装过程进行检查程序，包括：学习、检查、反馈信息、提出处理意见、记录等各步骤。

分工上进行两级检查，各专业组员执行第一级检查，组长执行第二级检查即全面检查。若有问题出现专业跨越，专业组之间进行合作，解决综合性问题。

为解决语言问题，公司组织了专门的翻译组配合工程进行。初专职翻译外，公司早在2000~2001年，就先后派出两批、共十余名公司年轻职工外出进行正规的英语培训，最后他们成为了具有工程师和翻译双重身份的现场翻译人员，减少了语言障碍带来的问题。

6、 *冷负荷试车及试车发现问题修改*：公司制定的试车工作方针是：单机试车要早、联动试车要全、负荷试车要稳。通过以上三级试车，为艾萨炉的投产铺平了最后一段道路：

•发现并解决了上料系统中的一系列问题。如：皮带跑偏、下料口易堵塞、最终皮带下料口改造等工作。

•解决了部分进口设备本身或安装中发现的问题。如：泥炮开口机钻杆进给不到位、喷枪旋转节不能灵活转动、负载栓不能工作、喷枪膨胀节尺寸不合导致喷枪更换困难等。

•解决了部分国产设备本身或安装中发现的问题。如：油泵不能满足要求供油流量和压力、部分国产皮带支撑结构不符合使用要求等。

•对部分不适应生产或安全要求的设计内容进行了修改。如：防爆片位置重新选择、熔池热电偶位置重新选择、阻溅板下捣打料掏出、

•增加炉底炉壳上排气孔使炉子加热过程产生的湿气能从炉砖中散发出去。

7、 *生产资料准备：*从2001年底就开始进行各种艾萨炉生产资料的准备，包括：劳保用品、排放层材料和工具、其它工种使用工具、生产用材料、备品备件、办公用品、防雨用品、防尘设备等。

（图5：YCC艾萨熔炼厂）

5、 艾萨炉投产和试生产：

1、 *投产最后的组织和准备：*

⑴MIM现场服务人员到厂。

开炉前芒特艾萨经验丰富的操作人员抵达YCC。这些现场人员协助最初几周的轮班，并根据YCC的要求，作为观察员身份，让YCC人员主持试生产，只有在需要时提供协助。

制定开炉计划。2002年3月将最初的YCC开炉、投产计划发送给MIMPT进行审核，根据MIMPT的意见进行修改。4月在MIMPT现场服务人员的协助下，由双方冶金工程师根据现场情况进一步讨论和修改开炉程序。最后形成以炉底铺渣、木材点火、加热烧嘴初期升温、保温烧嘴完成升温为框架的详细开炉计划，最终决定升温曲线和升温速率。并编制EXCEL升温记录及曲线显示。

⑵组建班组，实行12小时工作制，与芒特艾萨厂相似。任命班长和主CRO。各班班长、主CRO和排放层主操作工，都曾到芒特艾萨接受过培训。班组提

前一个月进入生产运转状态以度过磨合期。

⑶颁布实施艾萨区域安全规程。参考芒特艾萨的安全规范，制定了专门的艾萨区域安全规程，开始在艾萨区域个实施包括对个人劳动保护用品穿戴的要求、安全标识牌、标签使用、现场隔离制度等安全规范，使区域安全受控。

⑷公司组建艾萨炉开炉领导小组。根据开炉计划检查各项工作的进行、完成情况。每人每日向各自主管报告完成情况，领导小组对工作进行统一的协调和控制。

2、 试生产期间生产数据:

炉子升温结束后，工厂紧接着进入投料生产并将产量提升到满意的程度。在投料的初期，艾萨炉和原鼓风烧结机并行生产，冰铜品位保持在较低的范围，以使转炉逐步适应高品位冰铜的吹炼。当YCC认为艾萨炉的操作已足够稳定，就停止了鼓风烧结的生产，新熔炼炉逐步将产量提高设计能力。

表1总结了开始几个月的主要生产数据。

表1-主要生产数据

月 份(2002年)	5	6	7	8	9
平均料量 (干，t/h)	61.8	70.6	70.3	61.9	58.8
平均油耗 (L/d)	7,184	4,285	2,016	2,078	1,846
平均冰铜品位（%）	48.4	55.7	64.3	60.1	58.6
平均 SiO_2/Fe比	0.98	0.84	0.88	0.88	0.92
平均喷枪寿命 (天)	1.4	2.0	3.7	6.6	10.5

3、 投产以来主要问题简述

就象任何一个冶炼厂，YCC在试生产期间发生了一些问题。主要的问题和解决方法描述如下：

（1）加热烧嘴

加热烧嘴用于炉子升温，帮助炉子从环境温度升到操作温度。这种燃油喷枪可以精确的控制实际温度以目标升温曲线的要求完成升温过程。在加热烧嘴升温的后期，烧嘴的端头突然脱落，YCC操作工把加热烧嘴从炉子中提出，并使用保温烧嘴继续升温。YCC和MIMPT一致认为，将采取手段以保证此类事件不再发生。

（2）喷枪操作

在开始的几个星期里，喷枪使用寿命较短，主要是端部蚀损和弯曲。经过一段时间后，YCC生产操作人员积累了经验，炉温变得稳定了。稳定的操作大大减少了喷枪的弯曲，延长了喷枪使用寿命。

另外还发生了几次端部油嘴脱落的事件，但没有造成长时间停产。MIMPT的现场服务人员经计算后发现，是由于一个设计中的细小错误造成，并立即对现有的喷枪进行改造，之后没有再发生此类问题。

（3）给料量波动

最初的问题还包括艾萨炉的入炉料量不稳定，导致操作的不稳定：空气、氧气流量、熔池温度大幅波动。通过加强监控和对给料系统DCS逻辑进行改造，使问题得到控制。

（4）耐火砖剥落

在生产的头几周内，有耐火材料剥落并从排放口排出。尽管这对新炉子属于正常情况，因为暴露的耐火砖和边角的耐火材料会在这一时期剥落，YCC的操作工仍运用了他们在芒特艾萨学习到的知识，降低操作温度和温度波动，以控制耐

火砖的损失。同时，他们非常注意停产时间炉子的保温。

结 论

中国的第一台艾萨熔炼炉的设计和建设，通过YCC与MIMPT、欧萨斯、ENFI等国内外多家公司的成功合作，得以顺利实施、完成，说明YCC的选择是明智的。选择具有丰富实践经验的技术供应商，可以充分利用MIM的生产工厂的培训程序获得操作经验，其良好收益已显示了出来。

通过5个月的稳定生产实践证明，这一新兴熔池熔炼技术在云铜嫁接成功。YCC对未来充满信心，因为这种高效现代技术将给它带来巨大技术优势。

References

(1) **Bill, J.L., Briffa, T.E., Burrows, A.S., Fountain, C.R., Retallick, D., Tuppurainen, J.M.I., Edwards, J.S., and Partington P. " ISASMELT - Mount Isa Copper Smelter Progress Update", Sulfide Smelting 2002, R. L. Stephens and H. Y. Sohn, TMS, Warrendale PA, 2002, 181-193**

(2) **Edwards, J.S., Fountain, C.R., and Morland, R.L., "ISASMELT - Extending the Envelope," Proceedings of the Brimacombe Memorial Symposium, October 1 to 4, 2000: Poster Session Proceedings, ed. W. Poole, CIM, Montreal, QC, October 2000.**

(3) **Errington, W.J., Arthur, P.S., Fountain, C.R., "ISASMELT - Clean Efficient Smelting", GME'99 Global Metals Environment Conference, Beijing, 24-27 May 1999, Nonferrous Metals Society of China, Beijing, 1999, 164-172**

(4) **Player, R., "Copper ISASMELT - Process Investigations", The Howard Worner International Symposium on Injection in Pyrometallurgy, ed. M. Nilmani and T. Lehner, Melbourne, July 1996, TMS, Warrendale, PA, 1996, 439-446.**

(5) **Cribb, J.L., Edwards, J.S., Fountain, C.R., and Matthew, S.P., "ISASMELT Technology for the Smelting of Copper," 15th CMMI Congress, Johannesburg, SAIMM, Vol. 2., 1994, 99-103.**

(6) **Player, R., Fountain, C.R., Nguyen, T.V., and Jorgensen, F.R., "Top-entry Submerged Injection and the ISASMELT Technology," Proceedings of the Savard/Lee International Symposium on Bath Smelting, ed. J.K. Brimacombe, P.J. Mackey, G.J.W. Kor, C. Bickert and M.G. Ranade, TMS, Warrendale, PA, 1992, 215-229.**

(7) **Fountain, C.R., Coulter, M.D., and Edwards, J.S., "Minor Element Distribution in the Copper ISASMELT process," Copper '91, Volume IV, C. Diaz, C. Landolt, A Luraschi, C.J. Newman, Eds., Pergamon Press, New York, NY, 1991, 359-373.**

艾莎熔炼技术 - 蒙特艾莎公司铜冶炼厂的近况

J.L. Bill[1], T.E. Briffa[1], A.S. Burrows[1], C.R. Fountain[1], D. Retallick[1],
J.M.I. Tuppurainen[1], J.S. Edwards[2], and P. Partington[2]
[1]蒙特艾莎矿山有限公司，蒙特艾莎，昆士兰州，澳大利亚
[2]蒙特艾莎矿冶集团公司工艺技术部， 布理斯本， 昆士兰州，澳大利亚

摘要

蒙特艾莎矿山有限公司铜冶炼厂阳极铜的处理能力在一九九八年从十八万吨增加到二十六万五千吨。这一成就是以具有独创性的铜艾莎熔炼炉为基础取得的，其原料处理量已达每小时一百八十四吨。艾莎炉的进料量从每小时一百零二吨到每小时一百八十四吨的大幅度提高只需增加喷枪空气中的富氧量，无须对艾莎炉做任何变动。

扩建后耐火材料连续使用寿命已达两年。未来的耐火材料寿命在不使用冷却水的条件下预计可达四年。

目前艾莎铜熔炼炉已经在美国，比利时和印度获得应用。中国和德国的铜艾莎熔炼炉正在建造中。除此之外，在英国和马来西亚还投产了铅艾莎熔炼炉。

本文讨论了蒙特艾莎公司的铜艾莎熔炼炉及铜冶炼厂在扩建后的作业情况，作业数据以及耐火材料的使用情况。

1.0 引言

哈佛大学商校的 Clayton Christensend 教授的名著〞发明者的困境〞是一九九七年最畅销的管理书之一(1)。书中讲到新技术有时会使某一工业中的大企业被能够更快采用新技术的后起之秀所取代。书中列举的例子包括计算机硬盘产家的不断更换，缆绳操作的挖掘机被液压操作机器的取代，联合作业的钢铁公司受到来自小型工厂的攻击。

Christensen 教授称这类新技术为“破坏性”技术，因为他们具有打破某一工业现有次序的潜力。这类技术初期大都是从被主导厂家忽略了的边缘市场发展起来的。它们的存在虽然引起了大企业的注意，但同其它更稳妥的投资方案相比，发展这些技术的投资利润小。不过，小型企业认为边缘市场有利可图，而且它们有动力把新技术发展下去以推广到更广阔的市场上去。在这点上，小型企业的导向作用是那些已经确立了它们市场地位的大企业所难取代的。小企业变成主导者，而大企业慢慢从舞台上消失。

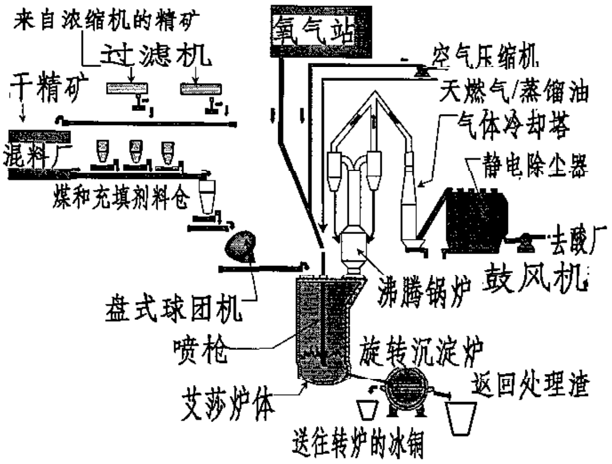
我们认为艾莎熔炼工艺与Christensen 教授所说的“破坏性”技术相似。艾莎熔炼技术是由蒙特艾莎矿山公司(MIM)发展起来的。 这项技术的前身是澳大利亚政府研究机构联邦科学与工业研究组织(CSIRO)发明的晒罗熔炼法(SIROSMELT)。艾莎熔炼工艺已经在澳大利亚，美国，英国，比利时，德国，印度，马来西亚和中国在取代传统熔炼技术的过程中得到了商业应用。对艾莎熔炼工艺的进一步改善使得三家采用艾莎法炼铜公司的成本降到世界最低。艾莎熔炼工艺处理铜精矿量已高达每年二百万吨。

艾莎熔炼技术的全套服务是由蒙特艾莎矿冶集团公司工艺技术部提供的。全套服务内容包括艾莎炉设计和组建，发放使用艾莎熔炼工艺的执照，提供关键设备，在蒙特艾莎公司的冶炼厂中进行现场培训，新炉调试以及对客户提供持续性的技术支持。

本文重点讨论艾莎炼铜法在蒙特艾莎公司的最近发展以及最新的生产指标。

2.0 艾莎熔炼工艺简介

艾莎熔炼炉是内衬耐火材料的竖炉，其内装有特殊设计的插入熔融物体池中的浸没式燃烧喷枪。空气或者富氧气通过喷枪射入融池中造成剧烈搅拌，产生快速反应。同早期的晒罗熔炼工艺比较艾莎熔炼工艺的诱人特点在于其独特的低压旋涡器的设计。艾莎炉喷枪典型的作业压力为八十千帕左右，采用单段吹风机即可提供燃烧气体，避免了昂贵压缩机的使用。艾莎炉已被用于熔炼铜精矿，镍铜精矿，铅精矿，铅和铜的二次材料。图一是蒙特艾莎铜冶炼厂艾莎熔炼工艺的流程。



图一　蒙特艾莎铜艾莎熔炼工艺流程

3.0 一九九七年至今艾莎熔炼工艺在世界范围的发展

艾莎熔炼工艺的发展史早期已有文章介绍(2-6)。本文重点讨论一九九七年以来的进展。表一列举了一九九七年以来新建的艾莎炉和正在建设中的项目。

表一　一九九七年以来新建的艾莎炉

日期	公司	厂址	厂型	处理量
1997	UNION MINIRE	HOBOKEN，比利时	二次铜熔炼	每年二十万吨原料
2000	METAL RECLAMATION INDUSTRIES SDN。BHD。	PULAU，印度	二次铅熔炼	每年四万吨铅金属
2002	云南铜业	昆明，中国	铜熔炼	年产冰铜中含十二万五千吨铜
2002	HOTTENWEREKE	LONEN，德国	二次铜熔炼	每年十五万吨二次铜料

4.0 一九九七年以来蒙特艾莎公司冶炼厂的发展

4.1 关键决定

一九九七年是铜艾莎炉发展的分水岭。蒙特艾莎公司决定关闭其焙烧炉和反射炉，使用铜艾莎炉熔炼全部铜精矿。蒙特艾莎公司同时开始扩建铜冶炼厂将其处理量提高到每年二十六万五千吨。澳大利亚西部矿山公司所属的化肥子公司也决定开发其在昆士兰省磷山附近的磷矿并建造一座酸厂将铜熔炼厂排放的二氧化硫转变为硫酸用于化肥生产。

下面列举的现场改造都促进了铜熔炼炉产量的提高
- 铜精矿品位从25%提高到27%
- 新艾莎炉进料系统的安装，包括混料设施
- 第二台日产氧气525吨氧气站的安装
- 艾莎炉废气处理系统中鼓风机处理量的提高
- 艾莎熔炼厂中转动沉淀炉长度加长
- KRESS型渣车取代了转动沉淀炉水淬渣系统
- 艾莎炉冰铜品位提高到62%
- 第四座 Peirce-Smith 转炉的安装
- 转炉车间安装了一台新的鼓风机
- 转炉车间两台吊车的更换，提高了机械可靠性
- 关闭焙烧炉和反射炉
- 阳极铜炉处理量增加
- 阳极铜浇机更新

艾莎炉本身没做任何改动。以上冶炼厂各项改造的调试另有文章介绍(7)，本文不加讨论。艾莎炉最近的详细作业指标请见表二。

4.2 艾莎炉耐火材料寿命

一九八七年到一九九一年，蒙特艾莎矿山公司投产运行了每小时处理原料十五吨的中间型艾莎炉试验厂，其耐火材料寿命为八十周。一九九二年工业规模铜艾莎炉建成时，耐火材料寿命是根据中间试验厂的八十九周耐火材料寿命估算的，预计寿命为两年。不过最初的结果令人失望。艾莎炉投产后最好的耐火材料寿命为六十五周。为了解决这一问题，蒙特艾莎公司对不同的耐火材料和砌炉技术做了大量的试验。同其它火法冶金工业上的耐火材料难题一样，确定最佳材料和操作方法化了多年时间。

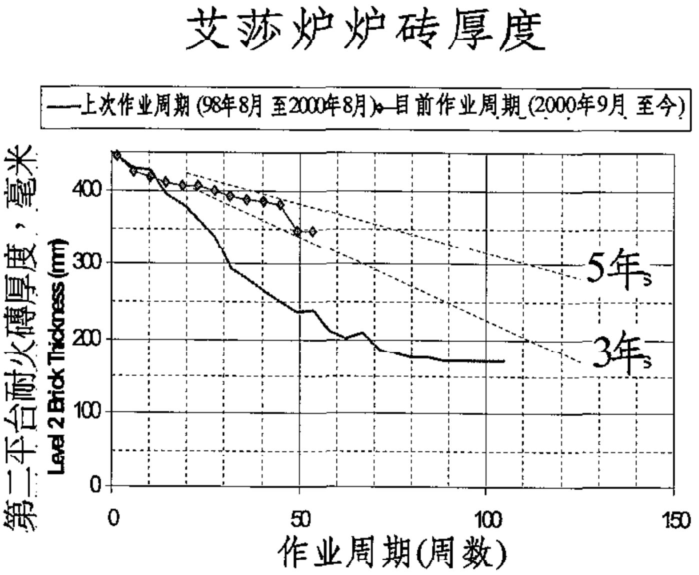
表二　　蒙特艾莎公司铜艾莎炉作业参数

参数	数值
设计每年处理铜精矿量	一百万吨
耐火砖内部炉体设计	
直径	3.75米
池深	1-2米
排料口数目	一个
排料口设计	水冷铜块

喷枪设计	
枪外部直径	0.5米
材料	不锈钢
枪头位置	入渣层 0.3米
旋涡器两端压差	80 千帕左右
吹风流量	每分钟 750 标米
富氧量	平均 60%
碳氢燃料	每小时 860 标米天然气
给料参数	
给料制备	潮湿球团料
处理精矿量(干量)	每天 2850 吨
精矿含铜量	25-27%
精矿水份	9%
加煤量	每天十吨
二氧化硅添加剂	每天 78 吨
石灰添加量	无
铜渣回炉	每天 100 吨
其它	每天 50 吨
产品	
冰铜-渣混合物	
下段作业	转动沉降炉
温度	1220 ℃
废气	
SO_2体积含量	锅炉入口处浓度 27%
温度	1220 ℃
灰尘量	每天 25 吨
沉降后的熔融产品	
冰铜	每天 1300 吨
冰铜品位	62% 铜
下段作业	Peirce-Smith转炉
出渣量	每天1270吨
硅铁比	0.78
渣处理	返回熔炼炉

对理解艾莎熔炼工艺耐火材料磨损机理的突破是在一九九八年取得的。在对作业方式做了相应改进后，效果明显，耐火材料磨损率大幅度下降。最近的一批耐火材料是在使用两年后由于锅炉的定期检查不得不报废的。在拆炉过程中发现，耐火砖的剩余厚度还可再坚持六到十二个月。

最近的运转周期中对耐火材料磨损控制作了进一步改善，损耗又有下降。图二给出了现在和上一个运转周期中耐火材料磨损的月指标。



图二　现在和上一个运转周期中耐火材料磨损的月指标。

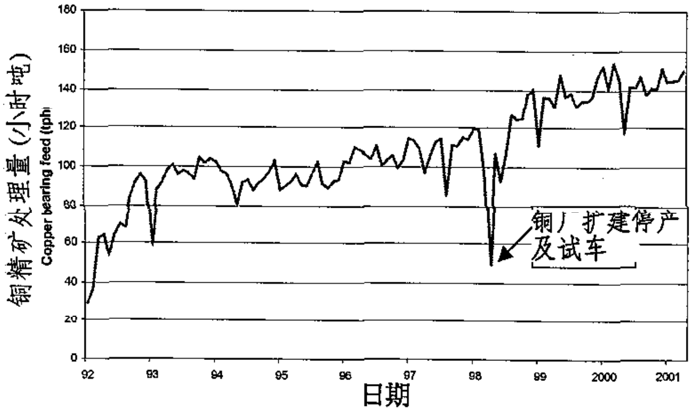
炉子运转四十五周后，磨损最严重处的砖厚度损失为七十毫米，而上次运转周期相应作业时间的砖厚度损失为二百毫米。图二中的另外两条线为三年和五年耐火材料寿命的预报线。

根据目前测定的磨损率，现在在炉中的这批耐火砖预计寿命可达四年之久。耐火材料寿命的延长没有采用水冷措施。水冷费用昂贵，作业复杂，在冶金炉上使用还有暴裂的危险。

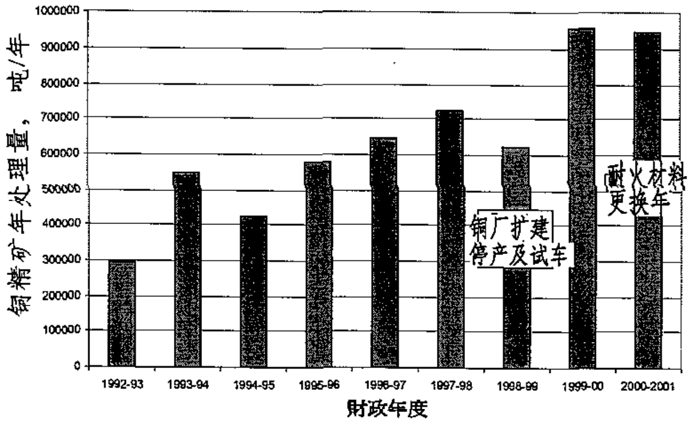
4.3　艾莎炉原料处理量

一九九七年以前，艾莎铜炉给料量的提高受到后续作业的限制。原因是Peirce-Smith转炉要同时处理艾莎炉和反射炉的冰铜而使其成为冶炼厂的颈口。一九九七年底反射炉关闭和一九九八年第二座氧气站投产后，艾莎炉的给料量有了明显的增加。最高小时处理能力达一百八十四吨。不过平均处理量仍然受到旋转沉降炉无备用处理能力的限制。当转炉，阳极炉或者阳极铜浇铸机出故障时，艾莎炉不得不降低处理量或停产。图三中给出的是艾莎炉的月平均处理量。

图四给出的是从一九九二年艾莎炉投产后每年铜精矿处理量。由图可知，处理量是每年增加的，最大产量是在一九九O到二OOO财政年度(七月一日到六月三十日)取得的，年产量为九十五万七千七百七十三吨。二OOO到二OO一财政年度的总产量由于换耐火材料停产的影响而有所下降。不过在目前的财政年度中，铜精矿处理量可期望超过一百万吨。处理量的增加在很大程度上是提高喷枪空气中富氧率取得的。富氧率目前通常高达67%。



图三 艾莎炉一九九二年投产以来月平均原料处理量



图四 一九九二年以来每财政年度处理的铜精矿总量

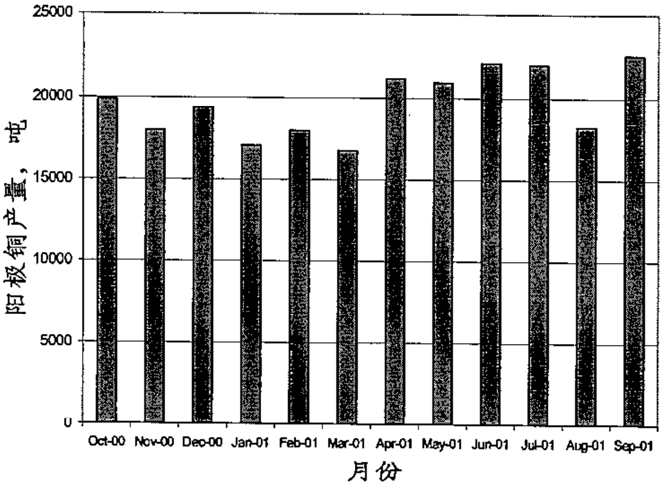
4.4 铜冶炼厂生产指标

图五给出了在最近的耐火材料使用周期间，阳极铜月生产指标。铜冶炼厂生产指标在持续改善，连续四个月生产阳极铜二十万吨。

4.5 杂质分布

表三列举了艾莎炉的杂质分布。PLAYER曾在文献(8) 中介绍过这方面的数据。杂质分布是对每个月生产混合样化验分析后，取二到十二个月的数据计算得来的。

从表三中的数据可知，与铜精矿相比，冰铜中只含大约 91%的砷，85%的镉，76%的铋，68%的锌，60% 的锑和铊，44%的铅和29%的碲。



图五 二〇〇二年到二〇〇一年蒙特艾莎冶炼厂月生产阳极铜指标

表三 蒙特艾莎公司铜艾莎熔炼厂杂质分布

元素	冰銅(%)	渣(%)	生产用气(%)*
Ag	98.5	1.0	0.5
Au	98.5	1.0	0.5
Pb	56.4	13.7	29.8
Zn	32.0	60.0	8.0
As	9.3	17.9	72.8
Co	63.0	36.5	0.5
Ni	97.3	2.2	0.5
Sb	39.9	51.9	8.1
Bi	24.0	1.0	75.0
Cd	14.9	5.0	80.1
Tl	39.9	43.7	16.3
Te	70.8	18.4	10.8

*在计算了冰铜和渣的元素之后平衡的结果

5.0 铜吹炼

艾莎熔炼工艺正在被试验用于铜的连续吹炼，有希望取代铜的间断吹炼作业，比如Peirce-Smith 吹炼法。它的诱人之处在于艾莎工艺可以用来灵活地处理固体物料，有效的扑收二氧化硫，并减少车间中的尘烟漏放。

澳大利亚联邦科学与工业研究组织(CSIRO)与蒙特艾莎公司合作对用艾莎炉连续吹炼铜精矿或冰铜生成粗铜的工艺进行了模拟和试验研究。蒙特艾莎公司还单独进行了大量的半工业试验，吹炼由艾莎炉生产的冰铜，证明该技术方向可行。两段熔炼和吹炼的艾莎炉工艺

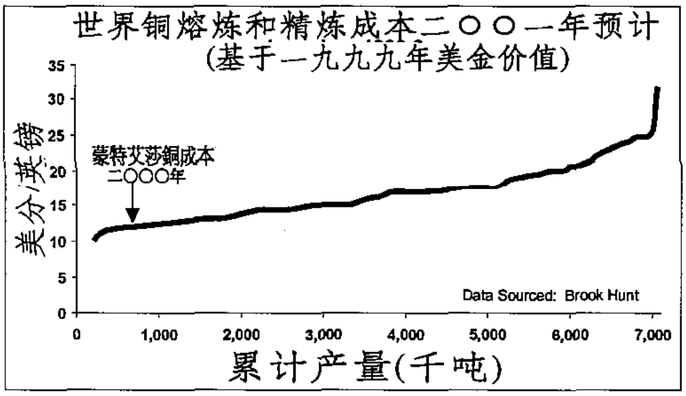
已被考虑将来在蒙特艾莎矿山实际应用。该工艺便于操作，而且在助燃气体的富氧量达30%时便可形成自燃。吹炼固体物料加上高富氧吹风生成的二氧化硫废气浓度高，体积小。

此工艺已生产出含硫量低达0.07%的粗铜，与Peirce-Smith 转炉生产的粗铜含硫量相同。
艾莎工艺用于连续铜吹炼的优点有，
1）投资比Peirce-Smith 转炉低10-20%。投资节省的原因是吹炼炉的体积和数目减少，小厂的卫生通风简单，无需昂贵的转炉车间吊车系统，艾莎炉可提供连续吹炼废气从而减小了所需酸厂的规模。
2）在艾莎炉多年积累起来的操作经验基础上，吹炼炉耐火材料的寿命可望延长。
3）如果在熔炼和吹炼之间设置冰铜料堆，可省去冰铜沉降炉，使两段作业分别操作，互不干扰。这种工艺设置解除了生产上的一个瓶颈，可明显增加工厂的利用率。
4）同现有的其它吹炼技术相比，艾莎工艺产生的灰尘率低，废气量小，二氧化硫浓度高，这些特点带来巨大的环保优势。

6.0 艾莎熔炼工艺的优点

下面介绍艾莎熔炼工艺的部份优点。
1）基建和作业成本低
艾莎炉构造简单体积小，废气排放量低，废气处理系统规模小。这些特点降低了基建费用和作业成本。蒙特艾莎公司铜艾莎炉于一九九二年建成，造价一亿两千万澳元，包括精矿过滤机，干料仓，废热锅炉，静电沉淀池，渣团粒系统，氧站，喷枪空气吹风机以及用于分离冰铜和渣的旋转沉降炉。

维修成本低，耐火材料寿命长，能源效率高和人工要求少导致了艾莎炉总作业成本大幅度下降。三座使用艾莎炉处理铜精矿的铜熔炼厂是世界上成本最低的。如图六所示，蒙特艾莎的艾莎炉与蒙特艾莎公司Townsville 铜精炼厂的艾莎工艺炼铜法(另一项蒙特艾莎公司专利技术)相结合使得蒙特艾莎公司的铜熔炼和精炼总成本位于世界产铜成本曲线的最低部位。最近，由于艾莎炉的处理量和利用率增加，作业成本又有大幅度下降。在二〇〇〇到二〇〇一的财政年度里，熔炼和精练的综合成本为每磅铜12.16美分。在二〇〇一到二〇〇二财政年度的头三个月中，熔炼和精练的综合成本为每磅铜8.64美分。



图六 铜熔炼厂和精炼厂成本曲线

艾莎炉充分利用了精矿本身含有的化学能，从而降低了冶炼能耗。表四列出了在铜艾莎炉建成前焙烧炉和反射炉处理一吨铜精矿的耗能。表中还给出了在铜艾莎炉的产量不断增加的过程中，每吨精矿所需能耗降低的数据。能耗的计算包括排料槽加热喷嘴和旋转沉降炉用油，但不包括熔炼厂用电。电耗对焙烧炉，反射炉和艾莎炉都很难估算。不过，根据二〇〇〇年三月数据估计的喷枪和氧站净耗电约为每吨精矿0.25 兆焦。

表四中的数据表明艾莎炉熔炼铜精矿可节能百分之八十。实际的节能还要高些，因为艾莎炉的原料中包括含自燃能量低的铜渣，渣的再选铜精矿和其它再生料。焙烧炉和反射炉是处理不了这类物质的。

表四　熔炼一吨铜精矿所需能源比较

	焙烧-反射炉 (1991.7 到 1992.6)	铜艾莎炉 (1993.7到 1994.6)	铜艾莎炉 (2000.3)	铜艾莎炉 (2001.4)
精矿处理量 (吨)	650,254	548,987	89,781	97,344
耗煤 (吨)	84,658	14,991	1,437	636
耗天然气(标米)	0	0	0	602,790
用油 (千升)	2,407	2,730	227.2	100.9
吨能耗 (兆焦/吨)	4.12	1.02	0.58	0.47

2)　原料和燃料的灵活性

艾莎熔炼工艺可用来处理范围很广的物料，包括铜和铜镍精矿，二次铜材料，铅精矿和二次铅材料。艾莎炉可使用的燃料包括各种等级的煤，焦粉，各种类型的油，天然气，或这些燃料的混合。设计燃料可采用价格最低的。如果燃料价格有变化，可再更换。蒙特艾莎公司艾莎炉的设计燃料为煤和油，现在已改用煤和天然气。美国Phelps Dodge Miami 的艾莎炉一直在使用天然气。印度 Sterlite Industries 的艾莎炉用煤和油，取决于其它燃料的价格。

艾莎炉最明显的操作优点之一是当辅助设备出现问题需要修理时可方便的将其停炉和启动。一般来讲，停炉需要二十秒钟，启动需要五分钟。由于停炉时间过长炉子需要冷态启动时，程序也不复杂。一小时之内熔炼生产便可继续进行。

3)　粉尘低

艾莎炉原料被带入废气系统的量比其它熔炼方法低得多，一般不超过原料重量的1%。

4)　无需燃烧喷嘴

艾莎炉使用喷枪而不是燃烧喷嘴，从而免除了喷嘴堵塞和喷嘴管道上的耐火材料损耗。喷枪更换是十分简单的，有些艾莎炉已做到了完全机械化。蒙特艾莎公司的铜艾莎炉在过去九年中对喷枪设计和控制方法做了改进，大大提高了枪的寿命。二〇〇一年四月和五月铜艾莎炉喷枪的平均寿命为十四天，每次换枪的大致时间为三十到四十分钟。至今为止取得的最长喷枪寿命为二十四天。枪的维修很简单，而且费用低，只需更换枪头部份即可。

7.0　结论

由蒙特艾莎公司发展的艾莎熔炼工艺可以说是当今最先进的铜熔炼技术。它的投资和作业成本都比其它铜熔炼技术低，而且有着多年的被工业实践证明了的性能指标。

艾莎炉目前在世界上许多国家都有应用，正在迈入主流技术行列，尤其是处理铜精矿目前已经有三家用户，第四家正在建造中。三家采用艾莎熔炼工艺公司的产铜成本是世界上最低的。除了熔炼作业外，环境保护法的日益加强使得艾莎熔炼工艺成为代替Peirce-Smith转炉进行铜吹炼的有力竞争对手。艾莎工艺吹炼所产生的粉尘少，作业简单，废气量低。

一九九九到二〇〇〇财政年度中，蒙特艾莎公司的铜艾莎炉熔炼了大约九十五万八千吨铜精矿。二〇〇一到二〇〇二政年度的处理量预计为一百万吨。

蒙特艾莎公司和世界上其它的艾莎炉作业成本和生产指标表明艾莎熔炼工艺已经远远跨越了其发展阶段，目前已经成为熔炼的主要竞争技术之一。

参考文献

(1) C.M. Christensen, The Innovator's Dilemma, Harvard Business School Press, Boston, Massachusetts, 1997.
(2) Errington, W.J., Fewings, J.H., Keran, V.P., and Denholm, W.T. (1987). "The ISASMELT lead smelting process." Transactions of the Institution of Mining and Metallurgy, Section C, 96, 1-6.
(3) M.D. Coulter and C.R. Fountain, "The ISASMELT process for copper smelting," Non-ferrous Smelting Symposium, Port Pirie, South Australia, 17-21 September 1989, Aus.I.M.M., Melbourne, 1989, 237-240.
(4) C.R. Fountain, M.D. Coulter, and J.S. Edwards, "Minor element distribution in the Copper ISASMELT process," Copper '91, Volume IV, C. Diaz, C. Landolt, A Luraschi, C.J. Newman, Eds., Pergamon Press, New York, NY, 1991, 359-373.
(5) R. Player, C.R. Fountain, T.V. Nguyen, and F.R. Jorgensen, "Top-entry submerged injection and the ISASMELT technology," Proceedings of the Savard/Lee International Symposium on Bath Smelting, ed. J.K. Brimacombe, P.J. Mackey, G.J.W. Kor, C. Bickert and M.G. Ranade, TMS, Warrendale, PA, 1992, 215-229.
(6) J.L. Cribb, J.S. Edwards, C.R. Fountain, and S.P. Matthew, "ISASMELT technology for the smelting of copper," 15th CMMI Congress, Johannesburg, SAIMM, Vol. 2., 1994, 99-103.
(7) J.S. Edwards, C.R. Fountain, and R.L. Morland, "ISASMELT-extending the envelope," W. Poole (Ed.), Proceedings of the Brimacombe Memorial Symposium, October 1 to 4, 2000: Poster Session Proceedings, CIM, Montreal, QC, October 2000.
(8) R. Player, "Copper ISASMELT-Process Investigations", Madhu Nilmani and Theo Lehner (Eds), The Howard Worner International Symposium on Injection in Pyrometallurgy, Melbourne, July 1996, TMS, Warrendale, PA, 1996, 439-446.

（高瑞敏 翻译， 高明炜 校对， 二〇〇二年十二月 于布理斯本，澳大利亚）

艾萨熔炼技术

----增产有道

J. S. 爱德华兹

MIMPT 工艺技术公司

C. R. 枫丹及 R. I. 摩尔兰德

芒特艾萨矿业公司

摘　要

芒特艾萨矿业公司（MIM）与联邦科工研究组织（CSIRO）联合开发了铜的艾萨法熔炼工艺。自 1992 年 8 月以来 MIM 就在芒特艾萨开始运行正规的艾萨炉。1997 年 7 月老式的沸腾焙烧炉及反射炉停产。目前在芒特艾萨熔炼的全部铜精矿均由艾萨炉处理。

MIM 的铜冶炼厂在 1998 年 8 月进行了改造，将阳极铜的年产量 提高至 27 万吨。此项改造包括安装日产 525 吨的第二套制氧设备、第四座皮氏卧式转炉、制酸厂与炼炉的连接装置，一台新的阳极铜浇铸机、一台艾萨炉炉料的堆取机，一台大型的与艾萨炉配套的排风机，并更换了老的转炉跨间天车。其他的改造包括延伸一台阳极炉的长度和艾萨炉旋转保温炉的长度，并以克力斯拖曳机取代了炉渣的水淬系统。

截至 2000 年 3 月底，十二个月来该冶炼厂生产了 21.9 万吨阳极铜。生产记

录月月更新。在西方矿业公司的肥料酸厂竣工投产后，这里的年产量还将进一步扩大。

前　言

近三十年来，全世界各矿业公司和冶金公司都面临两项无情的压力----它们的产品实价日趋下降和有关降低排放物的社会要求不断增强。芒特艾萨矿业公司（MIM）用发展艾萨熔炼工艺来回答这些挑战。这些工艺是与澳大利亚政府的研究机构------联邦科学与工业研究组织（CSIRO）共同开发的。

有关开发铜艾萨熔炼工艺的实际情况和历史已在别处予以详细叙述。铜的艾萨炉现在是 MIM 芒特艾萨铜冶炼厂的主要熔炼设备，年生产规模为 27 万吨阳极铜。该艾萨炉铜精矿的处理量高达 178 吨/小时。截至 2000 年 3 月底，十二个月来已处理含铜炉料（精矿加返料）91.6 万吨。

铜艾萨炉也是美国亚利桑那州迈阿密（Phelps Dodge）铜冶炼厂及印度南部土提科林（Sterlite）铜冶炼厂的主要熔炼设备。在中国云南昆明云铜公司正在建设第四座铜艾萨炉，在英国和比利时有其它的艾萨炉在运转，在马来西亚和德国新增加的艾萨炉也在建设之中。

芒特艾萨铜冶炼厂在 1997 年和 1998 年期间完成了升级，把年产约 18 万吨阳极铜的生产能力扩大为 27 万吨/年、从而能熔炼欧内斯特亨利矿山生产的精矿。该矿山距芒特艾萨约 200 公里，芒特艾萨冶炼厂占有 51%的股权。

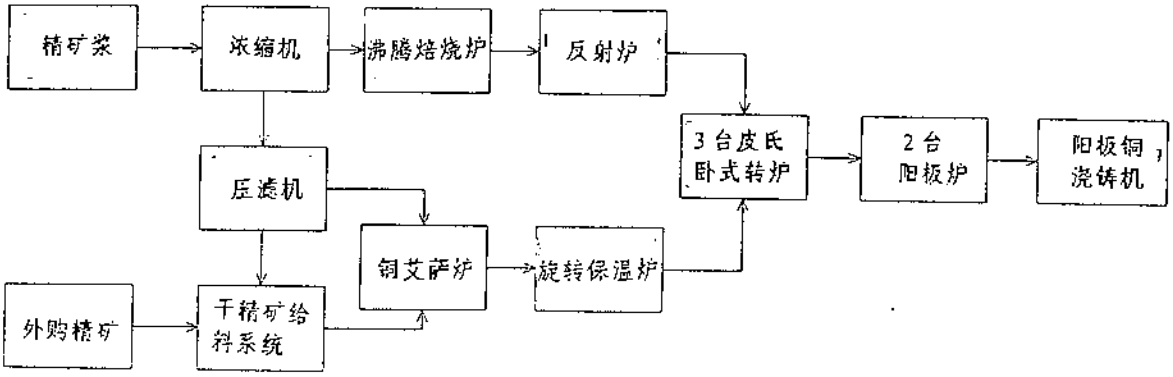
本文将叙述铜冶炼厂在升级前的情况，升级的原因，做出的变更，以及自此以后铜冶炼厂艾萨炉的生产情况。

铜冶炼厂原先的工艺流程

芒特艾萨铜冶炼厂铜艾萨炉的设计平均给料量为 104 吨/小时，于 1992 年 8 月开始生产（4）。原先的打算是以此艾萨炉取代 MIM 老的焙烧炉和反射炉，芒特艾萨冶炼厂的阳极铜生产能力不变，仍为 18 万吨/年。然而，人们发觉，如果焙烧炉和两台反射炉中较大的一台继续维持生产，只要增加少量的基建投资，从理论上讲就有可能将冶炼厂的生产能力提高到 26.5 万吨/年。转炉将有足够的能力处理生产出来的含铜 60%的冰铜。随后决定对此方案进行试验。

图一为该炉采取的工艺流程。来自 MIM 铜选矿厂的精矿浆经浓缩后分别送往焙烧炉和艾萨炉。送往艾萨炉的精矿先进行过滤，然后，或直接投入艾萨炉，或送往干精矿给料系统与外购精矿混合。两种炉子的产品最后在转炉汇合。

图一：1992 年铜冶炼厂包括艾萨炉的在内的工艺流程



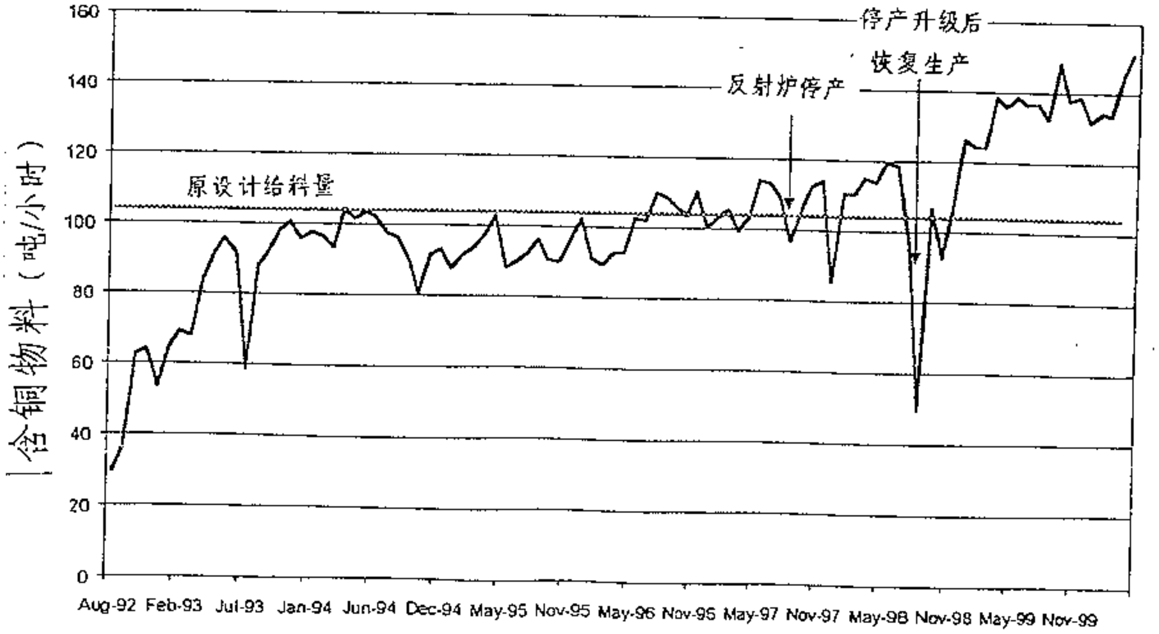
升级的理由

焙烧炉和反射炉的连续生产证明与冶炼厂的生产要求相悖。在这两台炉子运行的任何一年都未能生产出超过19.6万吨的铜（见图四）。不可能按冰铜品位为60%来使反射炉满意地运行。冰铜的熔点升高到了会使其在炉内结块的危险程度，而实际上它们在冰铜出口流槽和冰铜包中就已有结块 。这样就生成了大量返料。

来自反射炉的冰铜品位不得不下调，转炉就成了一个瓶颈。吹炼能力的制约导致艾萨炉明显地不能发挥其性能，因为它的处理量不得不降低以使通过反射炉的冰铜得以连续运动。如果做不到这一点就将造成反射炉结块的后果。

图二所示为艾萨炉自投产以来含铜炉料（精矿及返料）的处理率。该炉在1994年4月达到了设计的平均处理量，然而随后就受到反射炉需要量及艾萨炉炉渣水淬系统操作上的困难所制约。

有一点是显而易见的，即铜冶炼厂要达到处理欧内斯特.亨利精矿所需要的生产能力，就要进一步投资。



图二：艾萨炉自投产以来各月平均给料量

铜冶炼厂升级所需要的变更

已验证的变更

为提高铜冶炼厂生产能力所做出的变更如下：

*将精矿的含铜品位提高至27%；
*安装新的艾萨炉给料系统，包括配料设备；
*安装第二台日产525吨（吨/日）的制氧设备；
*使艾萨炉废气排风扇升级；
*加大艾萨炉旋转保温炉（RHF）的长度；
*把旋转保温炉的炉渣水淬系统更换为克力斯拖曳机；
*将艾萨炉冰铜品位提高至62%；
*安装第四座皮氏卧式转炉；
*在转炉上安装一台排风机；
*更换两台老的转炉跨间吊车以改善设备可靠性；
*停止运转焙烧炉和反射炉；
*增大阳极炉的生产能力；并
*增大阳极铜浇铸机的生产能力；

对艾萨炉本身未做变更。

预计这些变更将使铜冶炼厂得以生产出25.5万吨/年的阳极铜。此外，西方矿业肥料有限公司决定建立一座硫酸厂，配合磷灰岩山肥料项目，利用铜冶炼厂的废气生产硫酸。这一情况使铜冶炼厂得以将铜的生产能力提高到约27万吨/年。

以下将简单地探讨其主要变更。

配有堆取料机的新的给料系统

起初的意图是由压滤机直接给艾萨炉送料。然而，压滤机给料的不规则波动却无法消除。于是，决定由一个中间料堆给艾萨炉供料。

一种带棚的堆取料机被选用来混合芒特艾萨的精矿和欧内斯特亨利的精矿。原先的混料系统采用前端装载机混含精矿，不能充分地提供稳定的混合料以满足艾萨炉最大给料量的需要。此外，为应付更高的给料量，也需要增大给料系统的能力。

第二台制氧设备

为提供补充的氧气以满足提高艾萨炉给料量达到新目标的要求，需要有第二台制氧设备。喷枪空气中的含氧量预计为 60%。此外，氧气也将用以提高转炉的生产能力。

第四座转炉和高生产能力的圆盘浇铸机

先已提及，转炉被证明是冶炼厂的瓶颈。安装了第四座转炉。模型试验也表明，老的 45 吨/小时的阳极浇铸机太小而不能匹配。1997 年 8 月安装了一台新的 80 吨/小时维美克浇铸机。

制酸厂

铜冶炼厂至今仍将其二氧化硫排入大气。MIM 希望降低这些排放物的意向是其发展铜艾萨法工艺的动力之一。艾萨炉高浓度二氧化硫的废气使酸厂的生产成为可行之举。

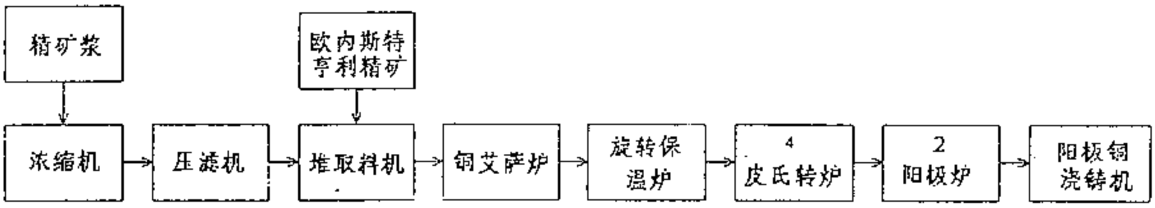
MIM 在芒特艾萨将地表二氧化硫浓度控制在标准以内，即相当于美国 EPA 环境二氧化硫的标准以内。当羽状烟柱似要增大地表二氧化硫浓度时，就让熔炼炉停止运转。其结果是，由于“空气质量控制”而停车，使铜生产时间的损失平

均约占年生产时间的10%。预计建成制酸厂后这种时间损失将减少一半。

冶炼厂升级后的运行情况

铜冶炼厂运行情况

铜冶炼厂于1998年7-8月停产以便安装新的设备，更换跨间吊车，安装艾萨炉用的排风机和加长RHF旋转保温炉。



图三：铜冶炼厂新的工艺流程

图二所示为自1998年8月以来艾萨炉平均给料量的阶段变化。在新设备安装后，试车曾出现某些困难。其中最值得注意是新吊车的电子设备受到意外的腐蚀（由于机键室密封不当及空调过度致使电路板冷凝所致），而主要困难发生在新混料机的取料部件上。该炉目前经常处理的含铜炉料量超过8万吨/月，而2000年3月的最新记录为8.9781万吨。

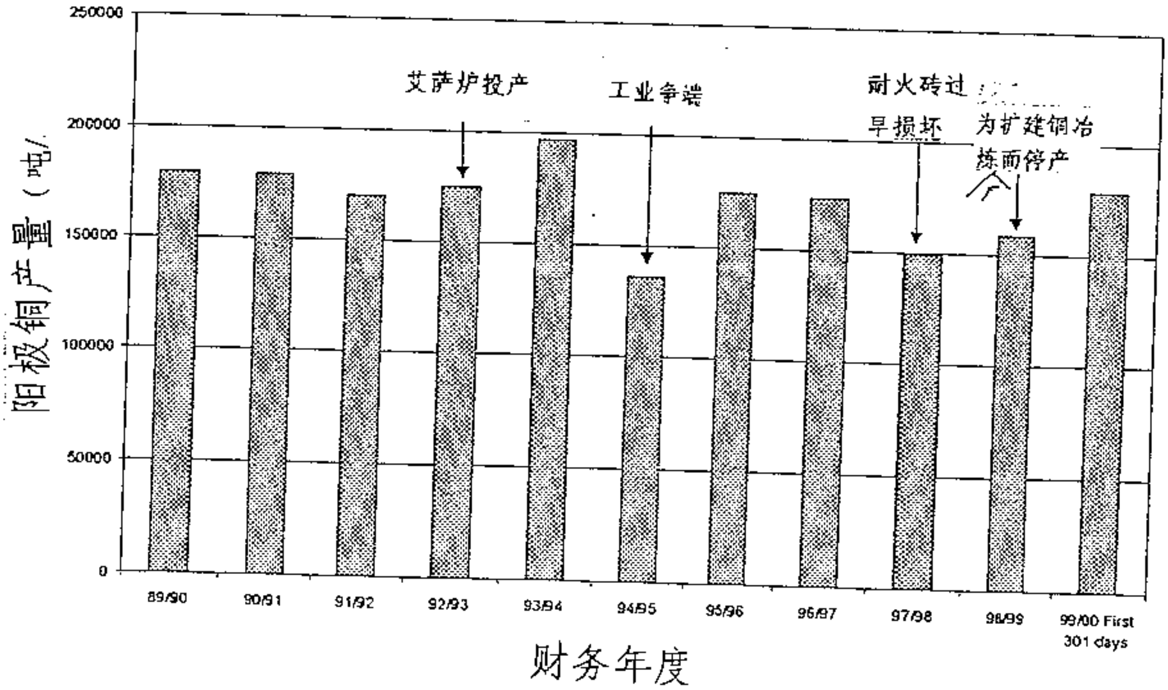
图四为当今铜冶炼厂的生产情况与前些年生产情况的比较（请注意：澳大利

亚的财务年度是7月1日至来年6月30日)。1994-1995年由于某工业争端而减产。1997-1998年的减产则是由于更换阳极浇铸机以及艾萨炉耐火材料炉衬过早损坏所致(详情稍后再述)。1998-1999年是因为铜冶炼厂扩建工程而停产。

至2000年3月底铜冶炼厂在12个月内生产了21.9万吨阳极铜。由于制酸厂仍处于试车过程中，因而在此铜产量数字中没有酸厂的贡献。

艾萨炉的运行情况

艾萨炉是铜冶炼厂的心脏。它已超过原设计的平均给料量162吨/小时，但某些辅助设备和铜冶炼厂其它部分的有关困难约束了它的处理量。尽管如此，铜冶炼厂已生产出的铜的成本仍低于扩建项目可行性研究中的假定成本。



图四-----铜冶炼厂1989-2000阳极铜生产情况

下列各点将论述某些主要的运行问题。

给料量

艾萨炉的铜精矿处理量可达 178 吨/小时。由于处理大量来自欧内斯特亨利矿山表生带的低黄铁矿的精矿，1999 年 9 月至今最高的月平均处理量为 148 吨/小时。2000 年 4 月份头 24 天的平均给料量是 151 吨/小时，包括欧内斯特亨利矿体原生矿带的精矿。

直到现在对给料量的主要制约是取料机的运行不适当，以及该炉通风不充分。

表一所示为艾萨炉在 2000 年 4 月 24 日晨 6 时为止的 24 小时的工作参数。在此期间制氧设备中的一台发生了问题，使当日艾萨炉的给料量下降。

取料机

堆取料机的取料机包括许多刮板，它们通过链条刨取料堆的表层。这一情况证明设备的不可靠性，磨损率很高。链条拉的很紧，经常折断。此外，该系统不能按更高于 120 吨/小时的给料量稳定地取料。在力图解决这些问题的近 12 个月以来，频繁的故障和给料量的上限限制了艾萨炉的作业率和处理量。该取料机现已不使用，改用前端装载机取精矿。

表一：2000 年 4 月 24 日早 6 小时为止的目的主要工作参数

参　　数	平均	最高	最低
含铜给料量（吨/小时）	151.1	165	120
喷枪空气流量（Nm3/秒）	5.91	6.9	5.1
喷枪氧流量（Nm3/秒）	7.70	8.2	6.2
喷枪空气中氧含量（%）	62.9	66	60
煤（吨/小时）	2.17	2.2	1.9
硅石熔剂（吨/小时）	6.25	6.8	5.0
石灰岩熔剂（吨/小时）	7.08	7.7	5.6
油（立升/小时）	76	1400	0
冰铜品位（%Cu）	57.4	59.3	55.9
工作时间	24		

通风问题

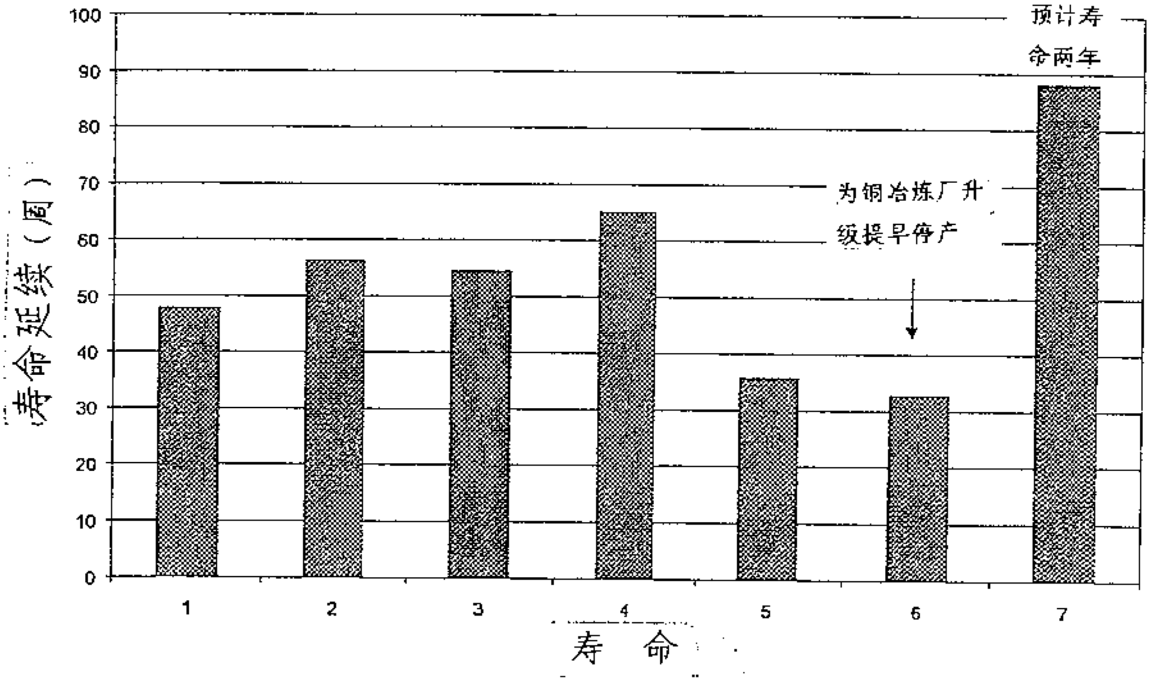
在最高给料量时艾萨炉遇到通风不充分的问题。该炉使用的是阿尔斯特隆（Ahlstrom）通流量（Flux-Flow）废热锅炉。它安装在炉顶，用循环流化床冷却废气。通过流化床的压力降较设计的炉顶负压相差 30 至 60 倍。当锅炉的热负载增大（例如，给料量增大）时，流化床中的粉尘量及压力降就增大。目前正以微调锅炉的运转和加大炉顶附近的二次通风来解决这一问题。

耐火砖寿命

图四为艾萨炉耐火砖在 1997 年 12 月发生过早损坏时的参数情况。其原因查明为多年来的不当操作实践所致。这一实践被纠正后，耐火砖的腐蚀率就降到了最低点，以致于在四个星期以后都难以测出其厚度的任何变化。据估计，一年来耐火砖的腐蚀率约为每周 2.2 毫米，过去的三个月中其腐蚀率较之更低。

达到这样低的腐蚀率并没有增加投入并使耐火炉衬水冷的复杂化。

艾萨炉原定于 1999 年 11 月更换炉衬。然而，由于稳定的低腐蚀率更换炉衬遂改期于 2000 年 8 月进行，其寿命为两年。表五说明当前耐火砖的寿命与以前六次耐火砖的比较。其中第五次是出乎意外的短。第六次是由于艾萨炉因铜冶炼厂扩建而停产。此时，炉衬在高腐蚀区还剩有可用厚度的一半。



图五：艾萨炉耐火砖的寿命

能效

在芒特艾萨建成艾萨炉以后大大地降低了炼铜的能耗。给料量虽增大，而能耗却降低。表二比较了在使用老的焙烧炉和反射炉期间、在铜冶炼厂扩建前使用艾萨炉期间和在 2000 年 3 月扩建后使用艾萨炉期间熔炼一吨精矿的能耗。

表二：熔炼一吨精矿的能耗

	1991年7月至1992年6月焙烧-反射炉	1993年7月至1994年6月铜艾萨炉	2000年3月铜艾萨炉
处理精矿量（吨）	650,254	548,987	89,781
煤耗（吨）	84，658	14，991	1，437
油耗（千立升）	2,407	2,730	227,2
每吨精矿能耗（Gj/t）	4,12	1,02	0,58

(YZL, 2000.10)

艾萨法—清洁、高效的熔炼技术

W．J．艾林顿，P．S．阿瑟
（澳大利亚，4000，布里斯班，MIM 工艺技术部）

C．R．枫丹
（澳大利亚，4825，芒特艾萨，芒特艾萨矿业有限公司）

摘　　要

MIM 控股有限公司工艺技术部(MIMPT)在全世界销售艾萨熔炼技术(ISASMELT)。艾萨炉在处理原生铜、铅精矿及铜、铅再循环物料(包括铅蓄电池)方面已得到商业规模的验证。现在，澳大利亚、美国、英国、印度和比利时都有这样的炼厂在运营。MIM 在芒特艾萨的艾萨法炼铜厂和 MIMPT 在英国的艾萨法中间试验厂继续与现有商业规模炼厂的客户们通力合作从事着工艺开发。

本文描述了艾萨炉的主要特点及其发展历史，包括其在已有炼厂现代化及新建炼厂中的应用。本文也论述了在环保远景方面使艾萨熔炼技术更具吸引力的某些特点。

简　　介

MIM 控股有限公司(MIM)是位于澳大利亚的国际采矿和矿产加工公司，其核心产品为铜、金、锌、铅、银和煤。该公司于 1923 年在澳大利亚昆士兰州西部边远的北方发现了芒特艾萨矿床的基础上建立。继承先辈们在建立矿山时表现出的创造性和智慧，新的工艺技术被引进 MIM 的所有冶炼厂。艾萨炉就是最为成功的这些革新的一个实例。

艾萨熔炼技术的发展

艾萨工艺最初是为原生铅的熔炼而开发的. MIM 和联邦科学工业研究组织(CSIRO)通过CSIRO 墨尔本实验室和芒特艾萨冶炼厂的半工业试验厂及示范厂的试验工作共同开发了此项工艺.

艾萨法炼铜示范厂在五年运行期间熔炼了约 53 万吨精矿, 大大地提高了 MIM 的铜产量. 示范规模的运转使许多关键性的工艺问题和操作问题得以提出. 这一谋略保证了把与提高工艺规模和商业化有关的风险降到最低程度. 关于该工艺发展的进一步细节可查阅文献[1-10].

铅和铜示范厂的成功使 MIM 有信心在其芒特艾萨冶炼厂中建立商业规模的艾萨法炼铅、炼铜厂, 并向外部客户提供使用该技术的许可证. 表一罗列了已投产的艾萨法熔炼装置.

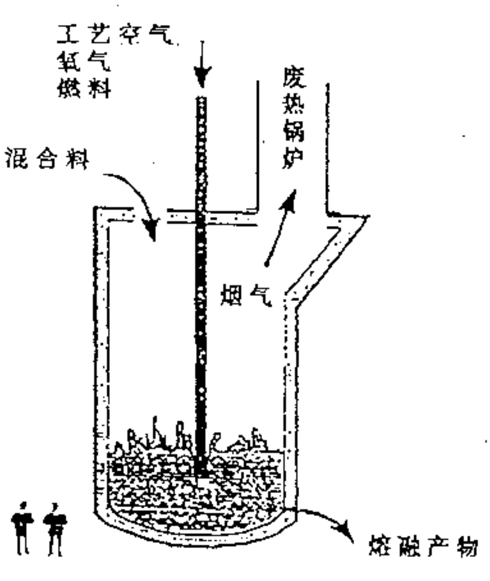
表一、截至 1998 年 12 月 31 日止的艾萨法熔炼装置

年 份	运 营 单 位	炼 厂 位 置	炼厂类型	规 模
1983	芒特艾萨矿业有限公司	澳大利亚芒特艾萨	铅炼厂	精矿 5-10 吨/小时
1985	芒特艾萨矿业有限公司	澳大利亚芒特艾萨	铅渣还原-浮渣处理	铅渣 5 吨/小时
1987	芒特艾萨矿业有限公司	澳大利亚芒特艾萨	铜炼厂	精矿 15-20 吨/小时
1991	芒特艾萨矿业有限公司	澳大利亚芒特艾萨	原生铅炼厂	铅金属 6 万吨/年
1991	不列丹尼亚精炼金属公司	英国 NORTHFLEET	二次铅炼厂	铅金属 3 万吨/年
1991	AGIP 澳大利亚有限公司	澳大利亚镭锭山	镍-铜炼厂	精矿 7.5 吨/小时
1992	塞浦路斯·迈阿密矿业公司	美国亚利桑那	铜炼厂	冰铜铜 16 万吨/年
1992	芒特艾萨矿业有限公司	澳大利亚芒特艾萨	铜炼厂	冰铜铜 18 万吨/年
1996	斯捷尔利特工业(印度)公司	印度土提科林	铜炼厂	冰铜铜 6-10 万吨/年
1997	联合矿业公司	比利时荷鲍肯	二次铜炼厂	进料 20 万吨/年

艾萨炉的特点

如图一所示，艾萨炉是一种简单的、用耐火材料衬里的熔炼炉．单一浸没式的燃烧喷枪经炉顶的入口插入熔池．用空气冷却的喷枪由于在其前端形成的凝固炼渣层的保护而免受炉内侵蚀环境的影响．如需维修喷枪，喷枪可轻易地从炉内取出．经由喷枪注入的工艺气体及燃料引起熔池剧烈湍动。由于熔池高度湍动，通过炉顶第二个入口投入的物料就十分迅速地溶解和反应，从而使设定的炉量达到高熔炼率。熔融产物从靠近炉体底部的水冷排放口排出．工艺烟气通过炉顶的出口排放，进入普通的废热回收和除尘系统．

图一、艾萨炉示意图



炉体结构简单．垂直的筒形炉体易于安装耐火炉衬．艾萨炉可采用煤、油或天然气等不同种类的燃料，可适应炼厂所在地区最为经济的燃料。艾萨法的进料置备是直通式的。物料仅需要混合或用水作粘结剂在制粒机上造粒。

炉耐火砖为高品级铬镁砖．艾萨炉耐火砖的寿命取决于被熔炼物料的性质，并对熔池温度非常敏感．炼铜艾萨炉的耐火材料炉衬在无水冷条件下通常可持续 12 至 18 个月．在示范炼厂其寿命曾达 21 个月．

艾萨法炼铜

自 1992 年以来，有四座炼铜的艾萨炉先后投产．表二所示为这些炼厂的某些细节．

表二、商业化的艾萨法炼铜厂

运营单位	设计规模	燃料	废气冷却	硫回收
芒特艾萨矿业有限公司 （澳大利亚）	原设计为 18 万吨/年，已提高到 25 万吨/年冰铜铜	块煤	废热锅炉 + 蒸发冷却器	空气中散逸(制酸厂计划于1999年投产)
塞浦路斯迈阿密公司 （美国）	16 万吨/年冰铜铜	天然气	传统的废热锅炉	制酸厂
斯捷尔利特工业公司 （印度）	6 万-10 万吨/年冰铜铜	块煤+油	蒸发冷却器	制酸厂
联合矿业公司 （比利时）	20 万吨/年二次铜料	块煤+油	辐射废热锅炉 + 蒸发冷却器	制酸厂

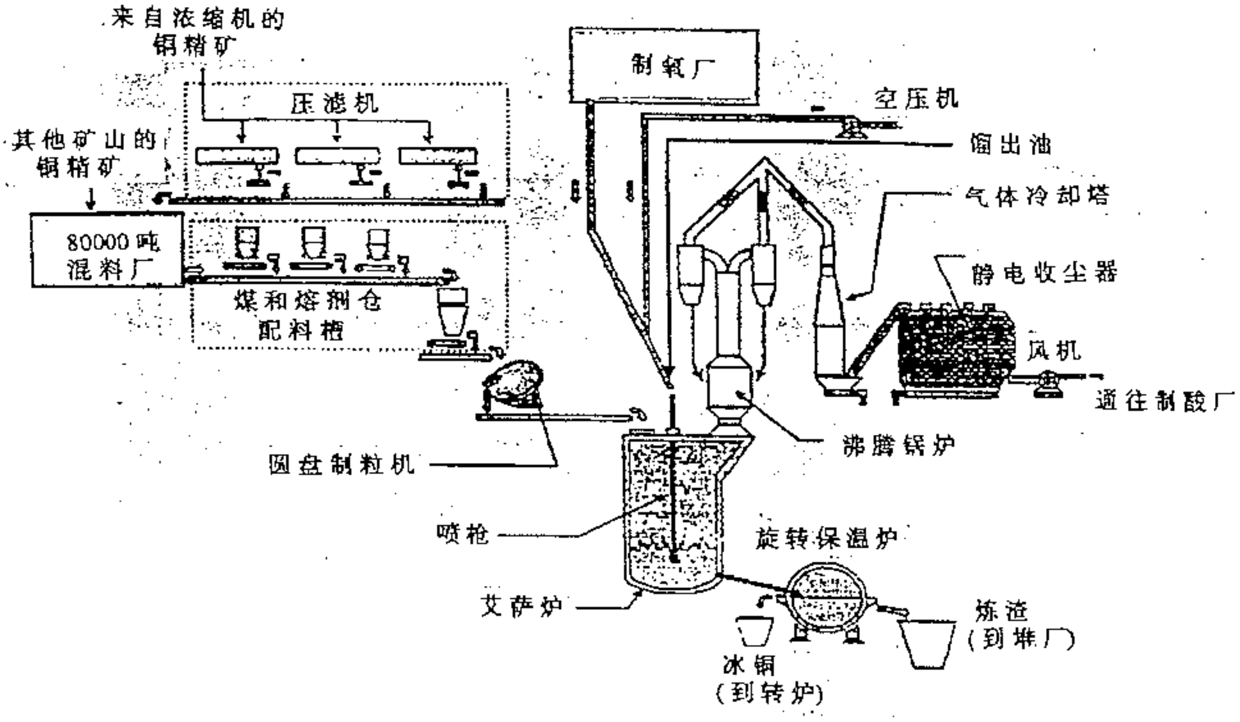
芒特艾萨艾萨法炼铜

MIM 公司以艾萨法工艺为基础改造了它的芒特艾萨炼铜厂．该炉熔炼铜精矿的生产率高达每小时 150 吨． MIM 公司选择艾萨法工艺的若干理由中包括：

- 在操作时很少有炉内气体泄漏工作场所；
- 所产生的烟气含有浓二氧化硫（ SO_2 ），可与较稀释的转炉烟气相混合，适于硫酸厂回收；
- 该工艺已得到示范规模炼厂五年运转的验证，进料率为每小时 15 － 50 吨；
- 与其他经过验证的铜熔炼工艺相比，其基建投资要低得多。

图二所示为芒特艾萨艾萨法炼铜厂的流程图．芒特艾萨的精矿与来自他处矿山的精矿在混料厂相混合．经过混合的精矿再与熔剂、返料和块煤混合，在园盘制粒机上制粒，然后送进艾萨炉．熔池温度是用块煤来控制的．注入喷枪的馏出油用于短时间的温度调节．该炉产出的冰铜及炼渣被一起排放入旋转保温炉．

图二、芒特艾萨艾萨法炼铜厂流程图



该炉的冰铜含量约为 60 %．在保温炉中的重力沉淀可产出含铜约 0.8 % 的废渣。冰铜经水冷流槽浇入铜水包，再由转炉跨间吊车送入转炉。炼渣由保温炉另一侧的第二个流槽直接注入 Kress Hauler 罐运往堆渣场。

1998 年的一个扩建项目取得了将艾萨炉铜精矿的平均处理量提高到每小时 150 吨。该艾萨炉上游和下游系统的改建包括安装了新的物料置备设备， 8 万吨规模的混合精矿贮料仓，一台四号 PS 转炉和新的阳极浇铸机．新的制氧厂加倍向艾萨炉供应氧气，并将注入喷枪的工艺空气的含氧量增至 60 %．新的制酸厂将于 1999 年底投产，以

处理艾萨炉的烟气并回收硫.

塞浦路斯·迈阿密矿业公司艾萨法炼铜

塞浦路斯·迈阿密艾萨法炼厂位于美国亚利桑那州，原设计年熔炼精矿 59 万吨. 先前是在电炉中进行熔炼的. 该电炉如作为艾萨法的冰铜和炼渣的沉淀炉来使用，则显不足. 电炉原功率为 51MW，平均耗电量约为 36MW. 改变了经济情况就意味着利用电炉熔炼已不再经济可行. 塞浦路斯选择了艾萨法来改造其炼铜厂.

该艾萨炉熔炼的精矿与一些外购的精矿经料场混合后送入精矿料仓. 精矿、熔剂和返料直接投入炉内. 熔池温度采用注入喷枪的天然气来控制. 塞浦路斯在工艺中不用任何煤. 工艺空气含氧 48 %. 该炉生产的冰铜含铜约 60 %. 冰铜和炼渣排放入电炉沉淀.

斯捷尔利特工业（印度）有限公司艾萨法炼铜

斯捷尔利特工业公司炼铜厂位于印度南端的土提科林，设计熔炼进口精矿. 该厂于 1996 年底投产，现正按设计年产 6 万吨冰铜铜，并计划通过增加工艺空气中的富氧扩大冰铜铜的年产量超过 10 万吨. 该厂系新建，包括了若干 PS 转炉、阳极炉和一座制酸厂.

精矿、块煤和熔剂相混合后经园盘制粒机制粒. 最终的团粒，含水份约 9 %，用传送带直接送入艾萨炉. 冰铜的目标含铜量为 50 %. 冰铜和炼渣在旋转保温炉中分离. 炼渣水碎粒化，冰铜用铜水包送入转炉.

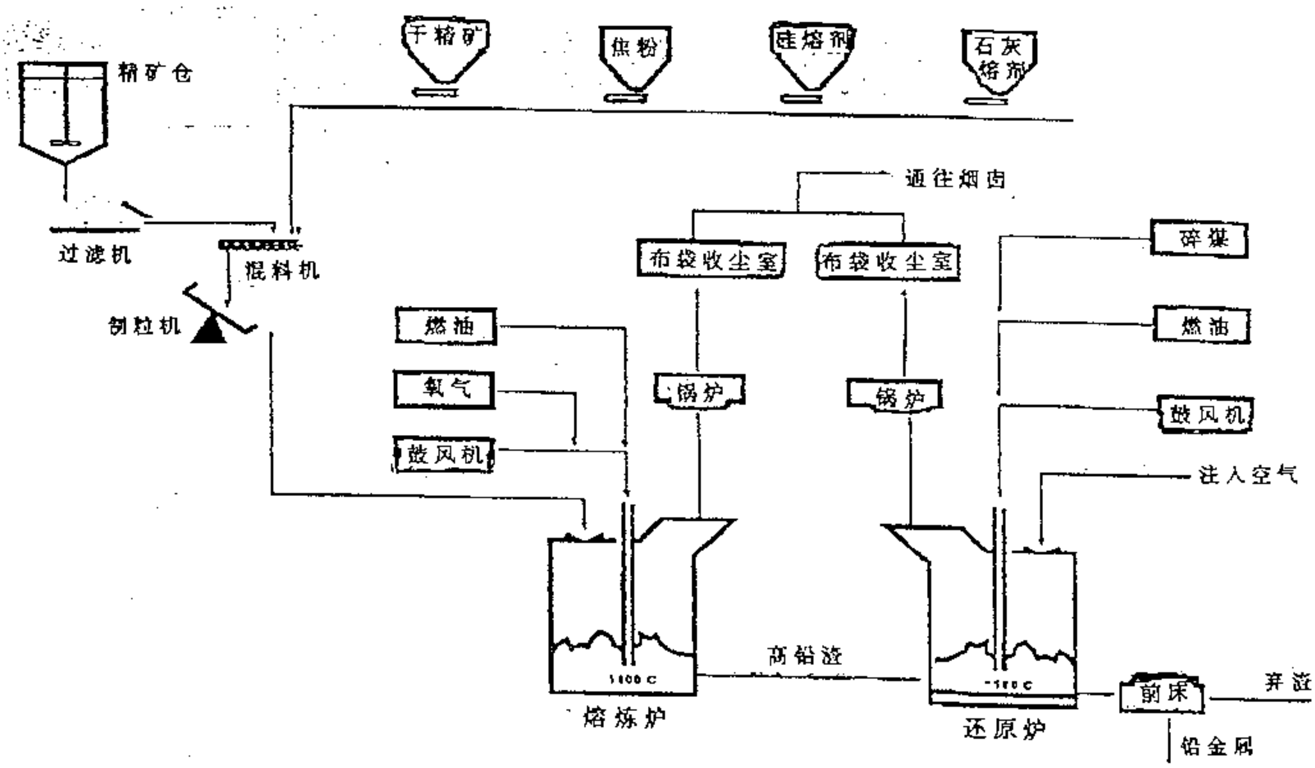
联合矿业公司艾萨法炼铜

位于比利时的联合矿业公司专营再循环有色金属废料. 联合矿业公司与 MIMPT 合作，开发了处理含有二次铜和铅的废料及残渣的工艺. 经 MIMPT 在英国 NORTHFLEET 的中间试验厂试验后，该工艺又经联合矿业公司荷鲍肯炼厂的示范厂验证. 随后，设计和建设了一座完全规模的商业化的艾萨法炼厂. 该新建的“2000 炼厂”于 1997 年第四季度投产，现正按设计能力在运转.

艾萨法原生铅熔炼

图三所示为安装在芒特艾萨的艾萨法两段炼铅工艺的示意图. 芒特艾萨炼厂设计年产量为 6 万吨金属铅, 每小时处理 20 吨含铅 47 % 的芒特艾萨铅精矿, 日用氧 70 吨. 在日用氧量 130 吨成为可能时, 该厂精矿处理量将增至每小时 36 吨, 而残渣中的含铅水平并不提高.

图三、芒特艾萨艾萨法炼铅流程图



两段工艺系采用一台氧化炉和一台还原炉. 精矿和熔剂在氧化炉中熔炼, 生成高铅渣(50 % 氧化铅). 高铅渣连续通过流槽进入还原炉, 被从喷枪注入的煤粉还原, 生成粗铅和废渣。

焦粉是熔炼炉的燃料。燃油是还原炉的主要燃料, 然而热需要量的重要部分是由于喷枪注入的煤的细粒组分提供的。两炉的烟气分别由单独的废热锅炉冷却至 200 ℃。每台废热锅炉都由竖式辐射室和相接的卧式对流室构成。冷却了的烟气在进入制酸厂或排放大气之前, 经由反向脉动袋式收尘室净化。

铅金属放入铅水包送往除渣锅，炼渣水碎粒化.

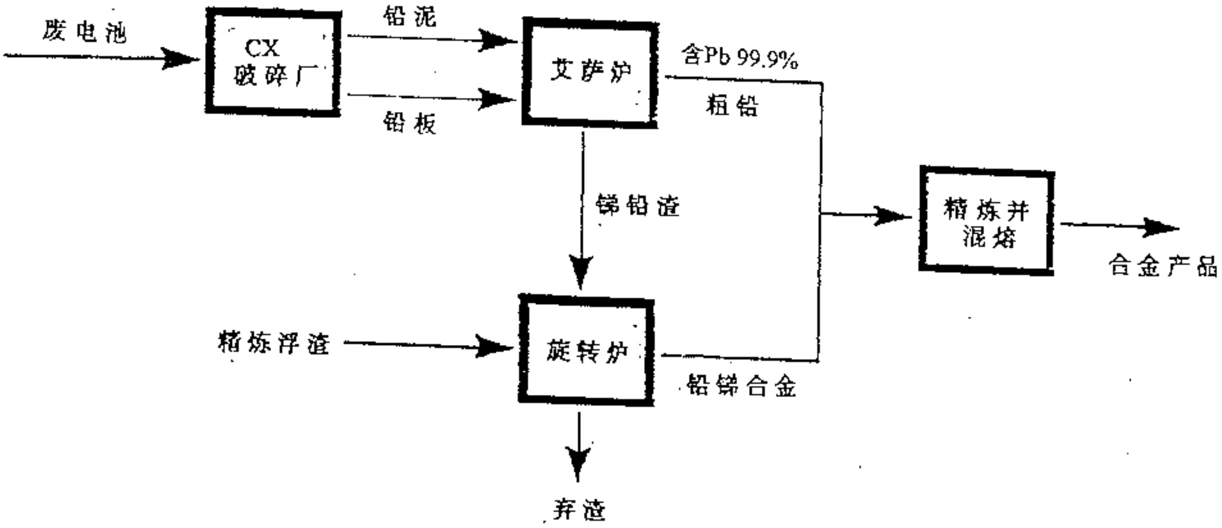
某次用 4000 吨含铅 67 % 的精矿所做的示范性试验表明，使用如此高品位的精矿可把炼厂能力提高到年产 13 万吨铅，且有约 50 % 的铅金属直接在熔炼炉中产出. 这一方案要求的日氧气量少于 50 吨.

目前芒特艾萨的铅艾萨炉没有使用，因为芒特艾萨生产的精矿不能满足他们及其它炼铅厂生产的需要. 在补充的精矿来源得到解决之前，正对它们进行保养和维修.

艾萨法二次铅熔炼

1991 年位于英国 NORTHFLEET 的不列丹尼亚精炼金属公司(BRM)以商业规模用艾萨法熔炼蓄电池铅泥和铅板. 该厂原先使用短型回转炉年产约 1 万吨铅. 该回转炉现仍保留，用以处理精炼浮渣和炼渣. 图四为该厂的简化流程图.

图四、 **BRM** 艾萨法二次铅熔炼流程图



蓄电池在破碎机内破碎并分成单一组分．其主要组分为金属铅板，氧化／硫化铅泥，聚丙烯，硬橡胶和聚氯乙烯．铅泥浆经与氢氧化钠反应脱硫后形成硫酸钠溶液和氧化铅泥．氧化铅泥再在压滤机上脱水．铅泥和金属铅板分别送入贮藏仓以便投入艾萨炉． BRM 的艾萨炉是以空气为基础，使用的燃料是回收油．

铅泥和铅板料一般是在分别的炉期中加以处理，以简化随后的精炼过程．在典型的铅泥炉期中，首先形成熔融铅泥的起始熔池．然后，铅泥和还原剂一焦炭一起投入炉中．大量的锑、硅、铁和铅泥的其它次要组分呈渣相．铅泥中的铅被还原成低锑(0.01 － 0.1%)软铅，定期地按一定间隔排放入罐，并以熔融状倒入精炼锅．

软铅，连续产出，直到 120 － 150 吨铅泥都入炉．在此之前，炼渣已高度含锑，并含有 55 － 65 % 的氧化铅．该炼渣可在艾萨炉中还原，以生成铅锑合金，但是用回转炉来还原炼渣，炼厂的处理可达最大限度．回转炉同样处理来自精炼锅的浮渣．

该炼厂设计为每小时熔炼 7.5 吨铅泥，但现在日常每小时熔炼 12 吨铅泥．铅板在单独的炉期中熔炼，速率为每小时 35 吨，生成的软铅含锑量要比铅泥周期的要高． BRM 艾萨法炼厂的生产目标是年产 3.5 － 4 万吨粗铅，满足环保标准并预期进一步降低单位生产成本．

继 BRM 炼厂成功之后， MIMRT 现正为马来西亚金属回收工业公司(MRI)设计一座新的艾萨炉． MRI 炼厂用二次铅废料年产 4 万吨粗铅．

艾萨法较传统工艺更具环保优点

炼厂改造项目的关键目的之一是减少排入环境的污染物的数量．艾萨法工艺在减少冶炼厂对环境的潜在影响方面具有许多引人注目的特点。

艾萨法工艺是以固定炉为基础．炉体的孔口局限于喷枪口、给料口和烟气通道．该设计容纳了炉与烟气系统之间的一个有效的定标系统．工艺气体及相关的尘埃被高效地吸收，很少泄漏入周边的大气空间，也很少有夹带空气进入烟气系统．在工艺过程中采用富氧，结合有效的定标系统，烟气中便产生高浓度 SO_2，因而要求较小的烟气处

理设备，简化了除尘和回收二氧化硫。

在芒特艾萨建立了炼铜艾萨炉，使修建硫酸厂以回收来自艾萨炉和炼厂 PS 转炉的 SO_2 成为可行．西方矿业公司（ WMC ）将利用该硫酸生产芒特艾萨以南杜切斯磷矿的化肥．

该工艺的热效率导致对燃料的要求比较低．芒特艾萨艾萨法炼铜厂在运转中没有添加煤或油．然而，在正常操作条件下，因为处理的精矿或返料中黄铁矿的含量较低，而仍需要若干煤．熔炼这种低黄铁矿的物料不能生成熔炼黄铜矿精矿时所能生成的那样多的热．芒特艾萨炼铜艾萨炉与传统的焙烧炉及反射炉的能量（煤 + 油）消耗对比见表三．在正常操作条件下，艾萨炉降低每吨精矿的煤与油的消耗量超过 70 %．尽管向混合进料添加低黄铁矿物料、转炉渣精矿、炼铜烟道炉灰和防污染池的泥渣，这种降低能耗的情况仍会存在．

表三、艾萨炉与焙烧炉及两台反射炉的煤、油消耗对比

	焙烧炉－反射炉 (1991 年 7 月 － 1992 年 6 月)	炼铜艾萨炉 (1997 年 7 月 － 1998 年 3 月)
每公吨精矿能耗(GJ/t)	4.12	1.14

该工艺所用燃料种类的灵活性意味着在各个现场都能采用最经济的燃料。艾萨法工艺能使用低质量煤作燃料的这一事实意味着可以省去鼓风炉工艺所要求的炼焦炉．取消了炼焦炉也就消除了诸如硫向大气散逸等的有关环保问题．

艾萨炉的设计容许混料、块状料和结饼直接入炉．这一特点使该炉可以处理来自其他工艺过程的残渣和废料，减少它们对环境的影响。循环物料，诸如来自炼铜转炉的返料等，可以容易地与进料混合，直接投入该炉．

艾萨炉会产生较少量的尘埃．通常，炼铜艾萨法工艺的烟气中所含尘埃量少于进料的 1 %．这一特点减少了炼厂循环物料的总量，从而提高了炼厂的效率．尘埃可以循环回入工艺过程或经处理后生产付产品，含锌尘埃的浸出处理即为一例．如果一座艾萨法炼铅厂位于锌

精炼厂附近，那么就有可能对还原炉的烟尘进行浸出处理，以产出硫化锌溶液和铅残渣．该残渣可以返回熔炼炉．

简单的艾萨法工艺可以减少所需操作人员的数量．炉体孔口少，较容易装配二次空气的风机系统．因此，熔炼会是非常清洁，有利于改善工人在炼厂的卫生条件．在这样的条件下，需要在炉边工作的操作人员接触到的是较少量的冶金烟尘．

未来的发展

MIMPT 继续为现有的和新的应用而发展艾萨法工艺，经常与本工艺的潜在用户保持联系．位于英国 NORTHFLEET 每小时处理量为 250 公斤的艾萨法中间试验厂用来试验客户的进料和检验工艺概念．

已经在商业规模中应用的铜的吹炼工艺是艾萨法工艺应用的一个实例．在中间试验厂已经展示了由冰铜生产粗铜．在现代化的炼铜厂中一座艾萨炉可以取代几台 PS 转炉．投入艾萨法转炉的进料可以是熔融状冰铜，也可以是粒状冰铜．艾萨法吹炼炉可以避免 PS 转炉的许多缺点．这里没有风咀，炉子将以连续方式，而不是间歇方式工作，收集烟气也较容易．

MIMPT 正在寻找感兴趣的伙伴将吹炼工艺发展为商业规模．MIMPT 也愿意讨论艾萨法工艺可能适用的其他应用．

结语

艾萨法工艺六年多来，以商业化规模在 MIM 集团的公司和外部客户中经受了检验，精矿进料率高达每小时 150 吨．该工艺在商业上用于熔炼铅精矿、铜精矿、铅电池泥以及铜的二次废料和残渣．

该工艺在现有的商业化炼厂及 MIMPT 的中间试验厂中的仍在继续发展． MIMPT 在艾萨法工艺基础上成功地试验了铜的吹炼工艺，并且寻找感兴趣的伙伴以求该工艺在商业上的发展．

艾萨法较传统工艺所具有的环保优点，包括减少向大气散逸、易于回收 SO_2 制成硫酸、低燃料消耗、燃料种类灵活、可循环废料及低尘埃量．在当今全世界大举实施现代环保规约之际，艾萨法的这些特点使其理想地适用于冶炼厂的革新提高．

参 考 文 献

1. W.J.Errington, J.S.Edwards and P. Hawkins, "ISASMELT Technology-Current Status and Future Development", Colloquium- Trends in Base Metals Smelting and Refining, The South African Institute of Mining and Metallurgy, Johannesburg, South Africa(1997)
2. R.L. Player, "Copper Isasmelt - Process investigations", Howard Worner Symposium on Injection in Pyrometallurgy, The Minerals,Metals and Materials Society of AIME, Warrendale, USA(1996)pp439-446
3. S. Jahanshahi, M. Somerville, and R. G. Hollis, "Direct converting of copper concentrate in Sirosmelt reactors", Copper 95 International Conference, Vol 4, Canadian Insitute of Mining, Metallurgy and Petroleum(1995)pp367-381
4. J. L. Cribb, J. S. Edwards. C. R. Fountain and S.P. Matthew, "Isasmelt technology for the smelting of copper",15th CMMI Congress Johannesburg, SAIMM Vol 2, (1994)pp99-103
5. R.R. Bhappu, K.H. Larson, R.D. Tunis, "Cyprus Miami Mining Corporation – smelter modernization project summary and status", EPD Congress 1994, The Minerals,Metals and Materials Society of AIME, Warrendale, USA(1993)pp555-570
6. C.R. Fountain, J.M.I.T. Tuppurainen, N.R. Whitworth and J.K. Wright, "New developments for the Copper Isasmelt process", Extractive Metallurgy of Copper Nickel and Cobalt, Vol 2, The Minerals and Materals Society of AIME, Warrendale, USA(1993)pp1461-1473
7. R.L. Player, C.R. Fountain, R.V.Nguyen, and F.R. Jorgensen, "Top-entry submerged injection and the Isasmelt technology", Savard-Lee International Symposium on Bath Smelting, The Minerals and Materals Society of AIME, Warrendalc, USA (1992) pp215-229
8. C.R. Fountain, M.D. Coulter and J.S. Edwards, "Minor element distribution in the Copper Isasmelt proccss", Copper '91, Vol 4 Pergammon Press, New York (1991) pp359-373
9. S.P. Matthew, G.R. McKean, R.L. Player, and K.E. Ramus, "The continuous Isasmelt lead process", Lead-Zinc'90, The Minerals, Metals and Materials Society of AIME, Warrendale, USA (1990) pp889-901
10. R.B.M. Brew, C.R. Fountain and J.Pritchard, "Isasmelt for secondary lead smelting", Lead 90: 10th International Lead Conference, The Lead Development Association, London (1990) pp170-181

(YZL 译校)

Paper Title:

LEAD / ACID BATTERY RECYCLING AND THE NEW ISASMELT PROCESS

Authors:
K. Ramus and P. Hawkins
Britannia Refined Metals Limited Northfleet, Kent (UK)

Date of Publication:
1992

Lead/acid battery recycling and the new Isasmelt process

K. Ramus and P. Hawkins
Britannia Refined Metals Limited Northfleet, Kent (UK)

Abstract

The recovery of lead/acid batteries has long been practised for economic reasons. More recently, battery recovery has also been influenced by environmental concerns, both in the general community and within the recycling plants. These influences will probably increase in the future. With these factors in mind, Britannia Refined Metals Ltd. introduced new technology for battery recycling at its Northfleet, UK operations in 1991.

A process description of the Britannia Refined Metals Secondary Lead Operation, the reasons for selecting an Engitec CX battery breaking plant in combination with an Isasmelt Paste Smelting Furnace, and commissioning and current operation of the plant are discussed.

Introduction

Britannia Refined Metals (BRM) opened its new recycling facility at Northfleet, UK, in September 1991. The plant is designed to treat traditional lead/acid battery scrap through the use of state-of-the-art technology.

The plant incorporates automated materials-handling systems, together with process technology, to achieve reduced emissions and effluent discharges of the lowest levels by practicable means.

Commercial aspects

During the last twenty years, the demand for lead on the part of the battery industry has grown by an average of 4% per annum from about 1.1 million tonnes to about 2.35 million tonnes at present (1). Here, the battery share of the total lead consumption rose from 38% in 1969 to 62% today (Fig. 1).

It is very likely that the battery industry will consume 75% of the lead produced by the year 2000. It can be assumed, for the foreseeable future, that lead still has a good growth potential in this area because of it: (i) metallurgical properties; (ii) reliable availability, and (iii) comparatively low price, which accounts for only about 25% of the value of a new battery.

In 1990, the Western world had a lead-refining capacity of around 5.3 million tonnes per annum. Of this capacity, about 3 million tonnes were attributed to primary lead smelters and 2.3 million tonnes to recycling, or secondary, lead smelters (Fig. 2).

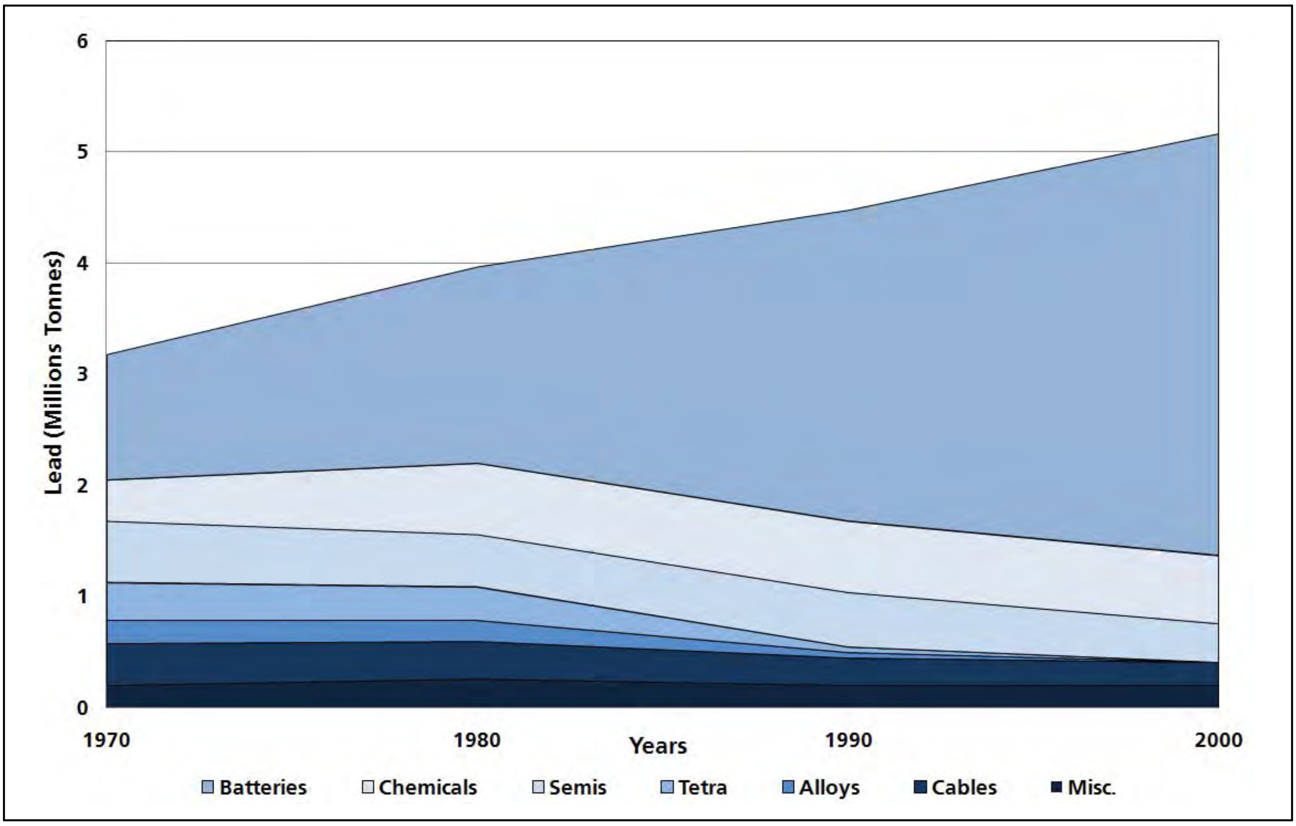


Figure 1 Recent and expected development of lead consumption in the Western world by end uses.

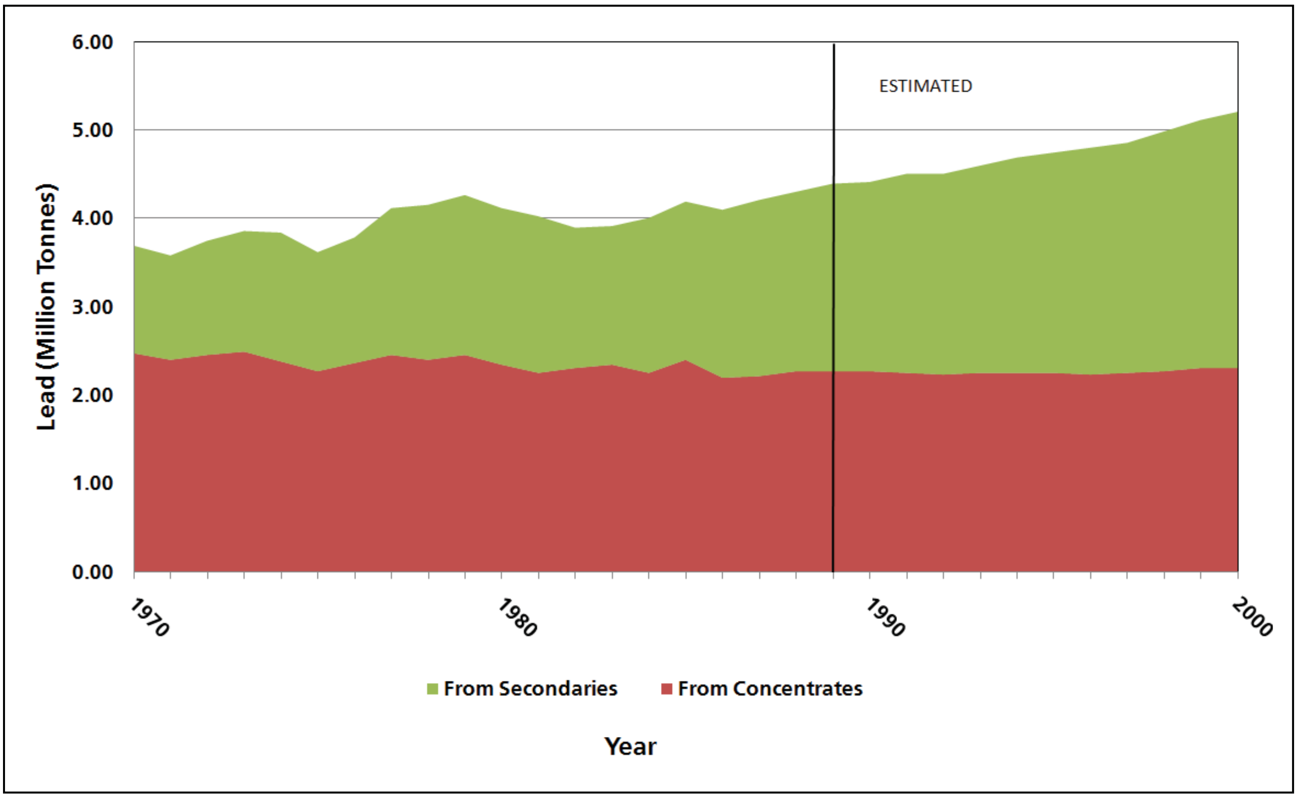


Figure 2 Development of lead production from primary and secondary raw material sources

In this decade there is a belief that secondary-lead production will have to increase by over 200 000 tonnes per annum, both in Europe and in the USA, because of continued growth in the generation of battery scrap and because governments are committed to increasing battery recycling rates to as close to 100% as possible.

In Europe and the USA, which together account for 75% of world secondary lead output, many producers are currently making operating losses due to low end prices. In addition, smelters are facing a rise in costs because of ever increasing environmental imposts. Additional costs could cause the closure of plants producing 200 000 tonnes per annum in the USA. A similar amount in Europe is vulnerable. The most vulnerable operation are small-to-medium sized plants, owned either by independent companies or by companies with no other lead interests. While smaller operations face problems, large greenfield secondary smelters are competitive. Upgrading larger operations and major expansions should also enable competitiveness to be maintained.

The initial objectives set up for the BRM project at its commencement in May 1989 were twofold: (i) to identify any opportunities in the lead business that would complement BRM's existing lead-refining business, and (ii) to identify opportunities to strengthen the economy and competitive position of BRM's existing lead-recycling business in the UK.

A broad spectrum of lead-based businesses were screened for industry attractiveness and financial returns. Expanded lead-recycling emerged as an attractive proposition. The installation of the CX battery breaking and paste desulfurization system, together with Lead Isasmelt, conforms with BRM's aims to be technically advanced and competitive by world standards. Further, approval for the project was based on the increasing need to demonstrate a commitment to improving the workplace and community environment.

Initially, BRM would invest £22.5 million in creating a £20-30 million per annum lead-recycling plant. This would exploit the advantages of the BRM site and Isasmelt technology and would also acquire the recycling business of Chloride Metals Limited.

Reasons for selection of technology

Based on an operating experience of the existing rotary-furnace-based plant and an awareness of the pressures facing the secondary-lead industry, a number of key areas were identified, where any new selected process had to offer improvements.

These were:
1) minimise sulphur emissions to atmosphere;
2) eliminate soda addition and produce a discard slag that met the TCLP leach test;
3) minimise low-level emissions of lead, through effective containment of lead oxide and paste during storage and handling and

4) select a process with low operating costs.

After careful consideration of the alternative technologies, BEM chose a process route based on the following principal process steps:

1) mechanical battery breaking and separation using the Engitec CX process for paste desulfurization;
2) Isasmelt to treat paste and grid metal to produce soft lead, antimonial alloy and an environmentally-acceptable discard slag; and
3) pyrometallurgical refining for final processing to alloys.

The capital costs of upgrading to CX type systems are more than twice that of less extensive changes to a conventional plant, but other savings offset the additional capital costs, particularly for larger plants. The cost of operating and slag-dumping are lower. Furthermore, there could be revenue from by-product sodium sulfate.

The CX System was developed by Engitec Impianti S.P.A., Milan, Italy. The Isasmelt technology was developed by Mount Isa Mines Limited, Australia, in association with CSIRO. The process was originally developed for the treatment of lead concentrates and the technology has now been extended to the treatment of copper and nickel concentrated and to battery-past smelting.

The innovative feature of the Isasmelt technology is the submerged top entry lance, inside of which are helical vanes to impart a swirling motion to the process gas stream. This design causes an increased rate of heat transfer from the lance to the process gas, as well as the formation of a frozen slag layer on the outer surface of the lance. This slag coating protects the lance so that it can be immersed in the bath for extended periods without excessive wear.

Submerged gas injection via the special lance provides an alternative to tuyeres in metallurgical processes. This technique greatly simplifies vessel design and eliminates refractory problems that are associated with tuyeres.

Process description

Materials' handling

The materials' handing system has been designed to minimise spillages around the plant. Deliveries of fluxes and reductant are discharged into covered storage bays from where they are moved to feed hoppers by a front-end loader. Subsequent transfer to the furnace is via a computer-controlled conveyer system with a local exhaust ventilation to all transfer points.

Battery storage

Incoming deliveries of batteries are discharged into a bay area after a thorough physical inspection. The batteries are stored ready for charging to the feed hopper that supplies the hammer mill. Acid drained from the batteries is collected in a sump at the rear of the bay; it is filtered and pumped to storage.

Battery breaking

Battery breaking technology is based on the Engitec CX process. The CX plant is designed to treat 16 tonnes per hour of scrap batteries. The hammer mill reduces the feed size to a maximum of 50 mm and deposits it onto a vibrating washing screen.

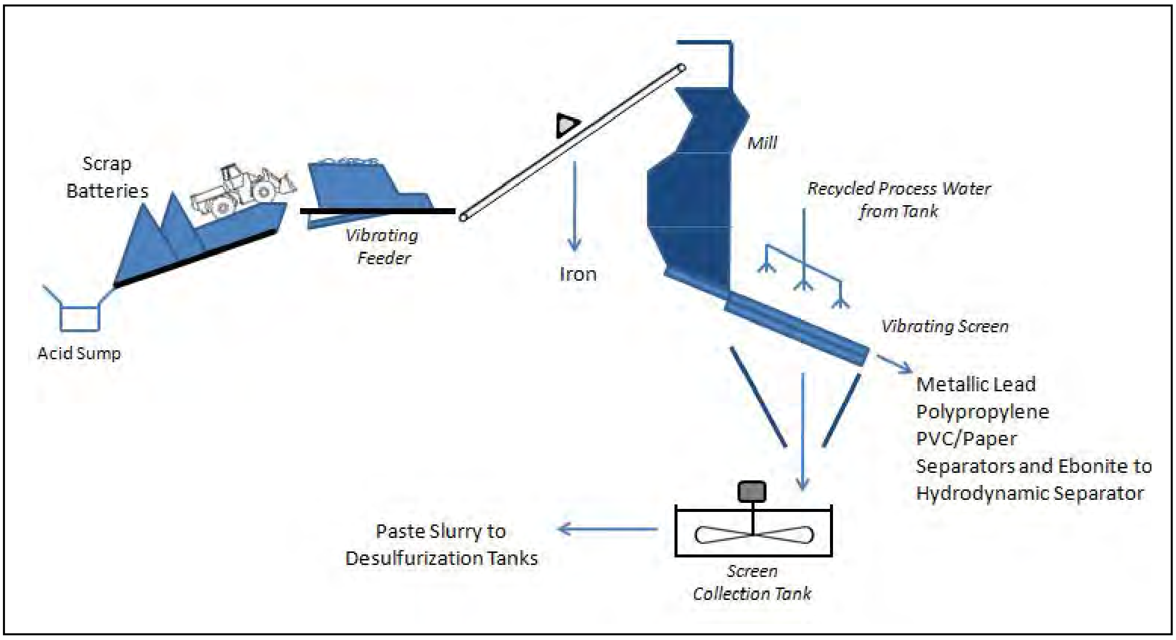


Figure 3 Battery crushing and screenings

The latter has a 0.8 mm mesh aperture that allows the fine lead paste to pass through and be collected in a stirred tank prior to desulfurization. Overside from the screen is fed to a hydrodynamic classification system that separates the component part, namely: polypropylene, metallic grids and poles, ebonite, and separator (see Figs. 3, 4).

Paste desulfurization

The paste slurry that contains lead oxide and lead sulfate is partially dewatered and pumped into the reaction vessels with collected battery acid and liquid caustic soda.

During this operation the following reactions take place:

$$H_2SO_4+2NaOH \rightarrow Na_2SO_4+2H_2O \tag{1}$$

$$PbSO_4+2NaOH \rightarrow Na_2SO_4+PbO+H_2O \tag{2}$$

Lead oxide is maintained in suspension and, upon completion of the reaction, is pumped to a pressure filter. The filter cake is washed before discharge (to reduce the sodium sulfate content) and then conveyed to storage ready for smelting. The sodium sulfate solution is further filtered and pumped to storage prior to further treatment in the effluent plant (Fig. 5).

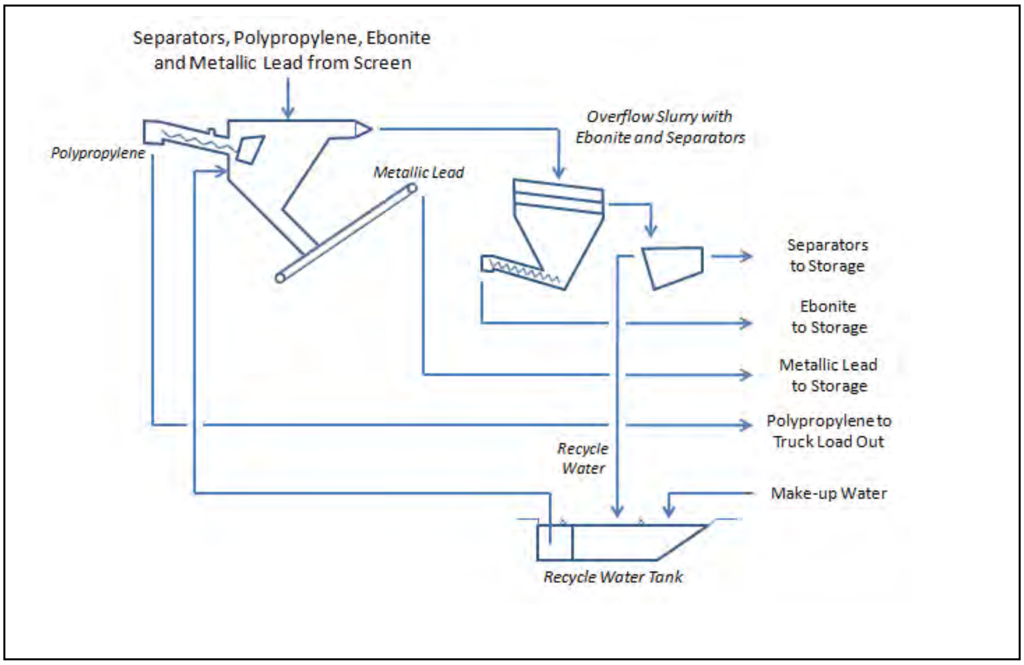


Figure 4 Hydrodynamic separation system

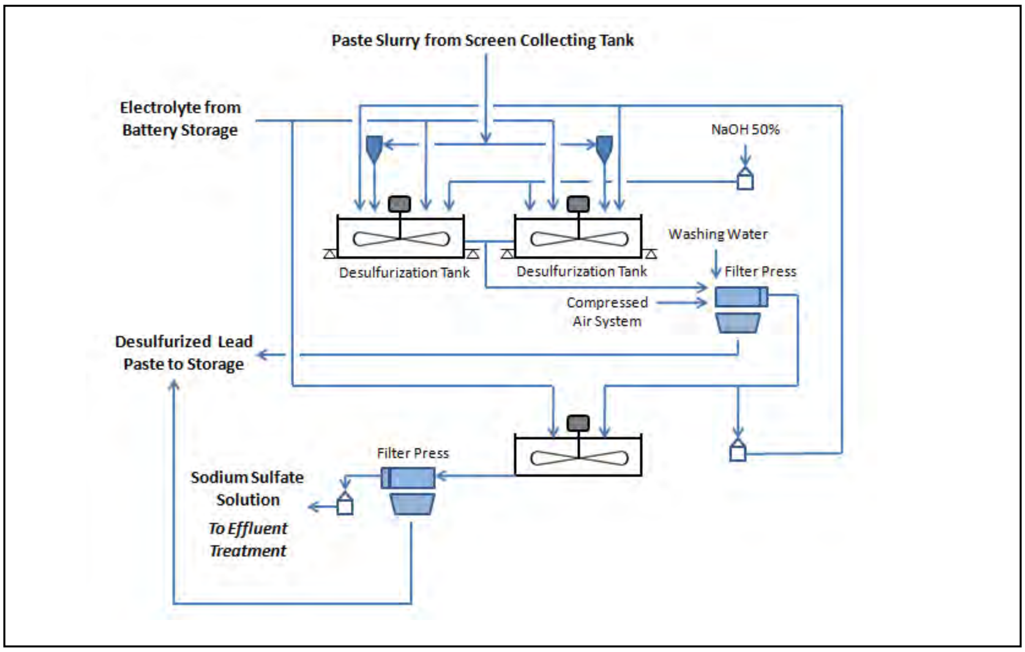


Figure 5 Paste desulfurization

Effluent-treatment plant

The effluent treatment plant has been designed to treat up to a maximum of 200 m³ per day of raw effluent from the CX plant. The raw effluent is pumped into the reception tank of the Exxflow unit. By careful control of the pH and pumping through a series of fabric membrane, the treated effluent meets quality standards of both current and pending legislation. The treated effluent is then pumped into storage prior to discharge (see Table 1 for discharge limits).

TABLE 1

Maximum allowable effluent discharge limits

pH value	6.0 to 9.0
Volume of discharge (m³/day)	200 m³/day
Biochemical oxygen demand	10mg/l in 5 days at 20°C
Dissolved antimony (mg/l)	5
Dissolved lead (mg/l)	1
Dissolved zinc (mg/l)	1
Dissolved arsenic (mg/l)	0.2
Dissolved silver (mg/l)	0.1
Dissolved copper (mg/l)	0.1
Dissolved nickel (mg/l)	0.1
Dissolved cadmium (mg/l)	0.05
Dissolved mercury (mg/l)	0.05

The Isasmelt furnace

The Isasmelt furnace consists of a refractory-lined cylindrical vessel (1.8m internal diameter) and is designed to produce 35 000 tonnes per annum of bullion (Fig. 6).

Desulfurized paste from the CX plant, together with reductant, is continuously fed into a bath of molten battery paste. The submerged combustion lance agitates the bath and produces a soft bullion that is intermittently tapped into a launder and pot system. At the end of each complete cycle, the furnace slag is conditioned and reduced to produce a high-antimony bullion and a discard slag [2]. The compositions of the furnace feed materials and the resulting products are given in Tables 2 and 3, respectively.

TABLE 2

Composition of furnace feed materials

Material	Battery Paste	Grid metal	Flue dust
Pb (wt.%)	66.20	90.8	60.00
Sb (wt.%)	0.45	2.5	2.00
S (wt.%)	1.00		

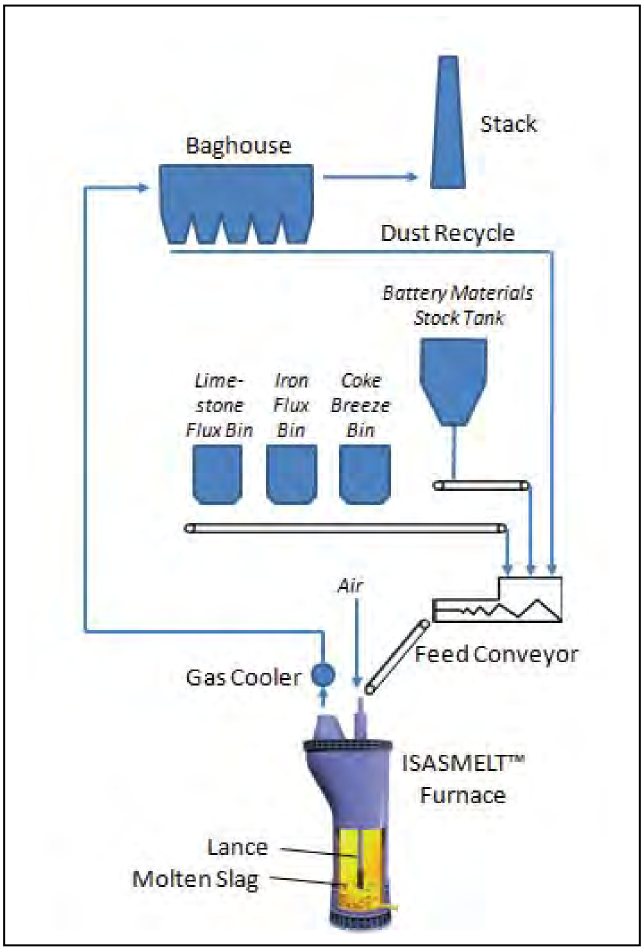


Figure 6 Schematic illustration of the Isasmelt plant

TABLE 3

Composition of products

Material	Soft Lead	Hard Lead	Slag	Flue Dust
Pb (wt.%)	99.8	79.2	0.5	60.0
Sb (wt.%)	<0.1	20.6		2.0
Cu (wt.%)	0.1	0.1		
FeO (wt.%)			50.0	
SiO_2 (wt.%)			25.0	
CaO (wt.%)			15.0	

The process is operated on a semicontinuous basis. When smelting paste, soft lead is tapped every 3 h. This allows slag to accumulate in the furnace. Antimony oxide becomes concentrated in the slag so that the metal tapped is relatively low in antimony. When about 250 tonnes of material has been charged, metal is again tapped and additions of lime and iron are made to give the required final slag composition. Similar procedures are adopted for grid metal smelting.

The reduction stage to produce the final slag is then commenced. Coal is added through the charge port and the oil flow on the lance is increased to raise a slag temperature in the final stages to between 1150 and 1250 °C. A ferrosilicate slag is produced with a low content of lead. The furnace is then tapped for metal and discard slag. The operational parameters of the furnace are summarised in Table 4.

The flow sheet for the smelting past and grid metal is shown in Figs. 6 and 7. This shows the two stages of smelting in which soft lead is tapped at intervals to allow antimony to accumulate in the slag, and then is reduced to produce a hard lead and low lead discard slag.

TABLE 4

Details of Isasmelt furnace operation

Operation parameter	Smelting		Reduction
	Paste	Grid metal	
Bath temperature (°C)	800-900	800-900	1200-1250
Feed rate (t/h)	10	18	
Air flow (N m^3/s)	1.4	0.9	2.1
Oil flow (kg/h)	500	390	570
Coal rate (t/h)	0.5.		0.5

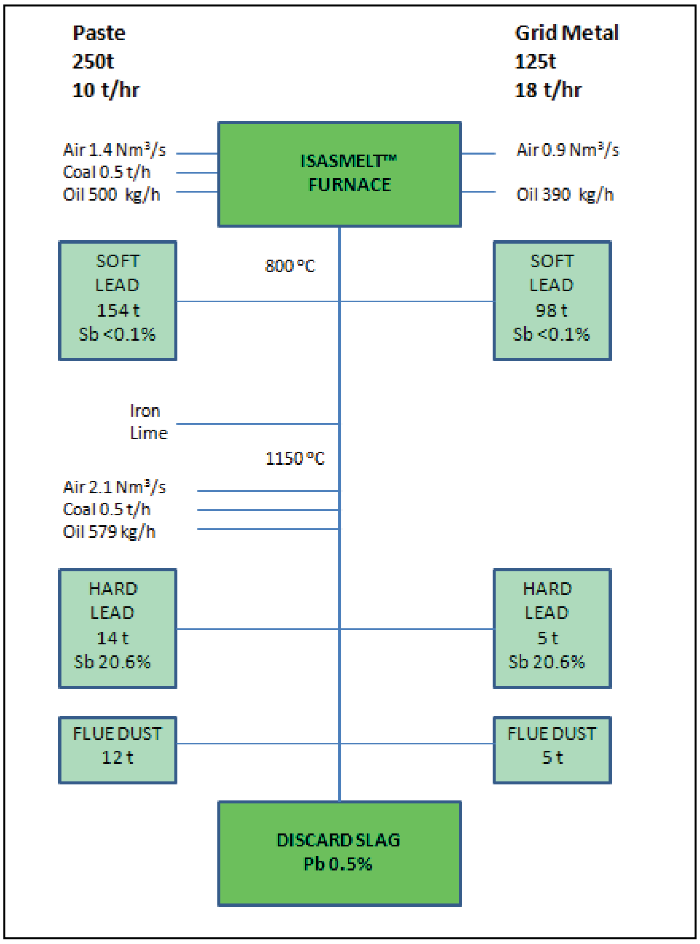


Figure 7 Flowsheet of Isasmelt process

Antimony distribution

The antimony distribution between the slag and lead metal is dependent upon the oxygen potential of the system or, in practical terms, upon the lead content of the slag. Thus: (i) under oxidizing conditions, i.e., high lead in the slag, the antimony is mainly in the oxide form Sb_2O_3 and therefore goes preferentially to the slag, and (ii) under reducing conditions, i.e., low lead in slag, the antimony is reduced to metallic form and dissolves in the lead.

During soft-lead production, oxidizing conditions are maintain, and results indicate that soft lead (<0.1wt.%Sb) will be produced down to 40to 50wt.% Pb in slag. This gives 10wt.% Sb in the slag generated per 200 tonnes cycle.

Slag reduction/hard lead production

During the soft-lead stage, the lead in slag is gradually reduced from 90% down to between 40 and 50% as the paste residuals and coal ash concentrate in the slag. As discussed above, this will be the limit of soft-lead production, and a full reduction and final slag stripping is required to produce a discard slag.

Final stripping of the slag to produce a low-lead discard slag will not be a problem as it is a batch process and, therefore, the reduction and fuming times can be extended. There is, however, one aspect of the reduction that requires special attention, namely fluxing/temperature control. As the lead is reduced, iron and lime are required to flux the slag. Optimum flux targets will require some plant testwork. As the lead contents drop, higher temperatures will be necessary to maintain slag fluidity, as discussed below.

Furnace Temperature

The furnace-operating temperature is dependent upon the lead in slag: low lead contents require high temperatures to maintain slag fluidity. There temperatures will also depend upon the soda content.

The upper operating temperature is limited in order to:
- enable the taphole to be closed on slag
- maintain a protective slag coating
- minimize fuel consumption

The lower operating temperature is limited in order to:
- prevent furnace accretions
- maintain slag with sufficient fluidity for high heat and mass transfer rates and for rapid settling of lead prills

During normal operations, the temperature is varied between 800 and 1200 °C. The temperature depends on the stage in the operating cycle.

Process-gas treatment

Process gad from the Isasmelt furnace is first rapidly cooled from 1000 to 450 °C by air-atomised water sprays in the furnace off-take. The gas can be further cooled by air admixture in the gas cooler, if required. The cooled gas is filtered in a baghouse that is fitted with 504 Goretex coated fibreglass bags to remove the dust and fume.

Isasmelt baghouse capacity is 80 000 N m^3 per hour of process gas. Filtration efficiency is 99.98%. The cleaned gas is discharged from a 90-m tall chimney.

The original intention to recycle the baghouse fume to the Isasmelt feed preparation system did not succeed due to the unsuitability of conveying equipment. A revised scheme to pug the fume with water and add to the Isasmelt past feed is currently being designed.

Kettle Floor

The kettle floor consists of 2X120-tonne hemispherical steel kettles each with an integral hood, mixer and dedrossing unit. Firing is by a tangential burner beneath each kettle. The burner runs on landfill gas.

Traditional refining techniques are used to produce a range of alloys to meet the specific requirements of customers. A dedicated filter unit provides ventilation during refining operations.

Moulding

The moulding machine is of the in-line design and can cast up to 23 tonnes per hour of finished product. The simple design allows for low manning and reduced maintenance requirements. Each 25 kg ingot is automatically stamped with a bath number then stacked and banded in 1 tonne bundles ready for final inspection prior to shipment.

Plant commissioning

The Isasmelt furnace was commissioned in June 1991 and temporarily fed lead oxide paste produced at the Britannia Recycling Limited operation, Wakefield, UK, until the CX plant came on line in September 1991.

CX plant

Basically, this is a well-engineered plant, capable of achieving production targets. There were problems that needed to be solved with equipment items such as pump-shaft seals, the operators' computer interface with the plant PLC control system, handling of polypropylene chips and separator fibre build-up on the recirculation water collection tank. Such problems have been largely addressed, or solutions have been developed that are awaiting resources to be installed.

The plant is capable of operating at a battery-breaking design rate of 16 tonnes per hour. Current availability is about 80% on a normal operating day, but averages 60% on a monthly basis. At present, the major reasons for lost time centre on the following:

- filter press paste collection, i.e., collection screw reliability
- paste-cake storage bin feeder restrictions
- pump-shaft seals
- effluent treatment plant capacity and operator control
- various control valves, electrical solenoids, limit switches, and tank level measurement probes
- reliability of the lubricating oil supply to the hammer-mill bearings

It is important to state that these problems are the collective responsibility of BRM, Engitec and other suppliers of specific items of equipment, and have been accepted as such. Non-familiarity with the process technology also caused a critical lack of spare parts for some pieces of equipment, which has now been remedied. Several hours of automation have also been unsatisfactory and are currently running in semi-auto or manual control mode. For example: the filler press cycle; neutralisation of effluent and the effluent treatment plant cycles and pH adjustment.

Effluent-treatment plant

An understanding of, and control over, the chemical reactions and pH of solutions at various stages of the effluent-treatment process took considerable time to develop. In the early stages of CX plant commissioning, considerable production delays were caused by the need to recycle effluent up to several times before it was within discharge limits. The addition of ferric sulfate, rigorous control over solution pH and rigorous control over the methodology for liquor sampling were necessary before this particular piece of plant was eliminated as a major bottleneck. Accurate and reproducible checking of pH control by plant operators was, and still is, a major factor in controlling this plant to obtain the consistent results achieved within the discharge consent limits.

Isasmelt furnace

The Isasmelt furnace has demonstrated that it is capable of achieving design capacity for both paste smelting and grid melting. Feed rates of up to 20 tonnes per hour of paste have been achieved on soft lead production. The soft lead process is extremely robust, and steady operation has been achieved.

Operator skills are steadily improving and all crews are capable of running the soft-lead and grid-melting stages with minimum supervision. More experience is required however, to successfully carry out slag reductions. Slag reduction to produce a final discard slag and achieve the full, five-stage, operating cycle as originally envisaged has still to be implemented. Averaged production rate and plant availability are not yet at the desired levels; current availability is 60%.

Until furnace operating hours have reached the desired level, soft-lead production will be maximised. This means that Isasmelt intermediate slag will continue to be recycled through the rotary furnace for final treatment. Lance tip lives of several days have been achieved, and it is expected that this will be extended. The major cause of lance failure is high-temperature excursions. A build-up of soda slag (the residue of paste desulfurization) also reduces the thickness of the protective slag coating on the lance and contributes to lower lance life.

Feed handling

Basically, the paste-feed system consists of a 200-tonne live bottom storage bin with an internal screw that delivers paste onto a series of standard conveyer belts.

Soon after start-up it was clear that handling non-free flowing and very sticky material such as wet-battery paste was a major problem. Belt cleaning, maintaining correct belt tracking, keeping tail and head drums clean and free running and transfer chutes clear of blockages all require a major effort on the part of the operators. Substantial redesign is required in the conveyor belt area to overcome all of these problems.

The live bottom Storall bin performed very well during the first six months of operation during which the past was not desulfurized. Delivery rates of 15 tonnes per hour were readily achievable. After paste desulfurization was introduced, the performance of the Storall decline markedly to about 5 to 7 tonnes per hour. The obvious change was a hard build-up of paste on the internal discharge screw so that, in effect, the screw resembled a solid shaft with little of the flights exposed. This build-up is exceptionally hard and a cementing reaction by residual sodium sulfate in the paste is thought to be the cause. Cake-washing tests in the filter press prior to release and experimentation with the Storall bin screw design and surface finish are currently in progress. Obviously, this problem has affected throughout of the Isasmelt furnace but is somewhat compensated by operating the rudimentary and temporary BRL paste-feed system in parallel. This means extra labour cost and lowers overall control of the plant.

Gas handling

The process-gas handling system is satisfactory when all operating parameters are functioning correctly, e.g., baghouse bag-cleaning system, cell isolation damper movements, inducted draught fan control, feed moisture content, flue-pressure measurement. Problems with any of these can cause reduced draught and the feed rate to the furnace must be reduced in order to maintain acceptable hygiene conditions.

Basically, the size of all of the baghouse devoted to the Isasmelt furnace is not large enough to cater for all that can go wrong in an operating place. This has been recognised and an extension of baghouse capacity is due to be commissioned in December 2002.

Plant Future

CX plant

The major mechanical items of plant that affect availability have been mentioned. In most cases, solutions have been identified and will continue to be implemented. It is expected that plant availability will be confidently at the 80% level by mid-1993. The effectiveness of paste washing to remove residual sodium sulfate will also be tested in the near future.

Other areas requiring longer-term investigation include the following:

- rapid wear on the rotating hammers of the breaking mill
- dealing with power-supply interruptions that result in bogging of the mixers in the paste-holding tanks
- dissolution of caustic dross (a by-product of the primary-lead refinery) to produce an essentially free supply of liquid caustic

Isasmelt furnace

Rates of paste smelting and grid melting for soft-lead production have been demonstrated to exceed production budget requirements. As with the CX plant, however m the availability must be further improved, to this end, the feed-handling and process-gas handling systems need to be modified and/or expanded. Both of these issues are currently being redressed.

In addition, longer-term strategies are being developed to overcome soda slag build-up in the Isasmelt furnace and the production of the final discard slag. With respect to the build-up of soda slag on the bath and consequent lance attack, two approaches to control are being investigated, namely:

effective washing of the sodium sulfate from the filter press paste cake, and

any residual soda slag in the furnace may have to be treated through a separate higher level furnace taphole for periodic removal; such a taphole eliminates the need to drain the entire bath through the normal taphole in order to remove the soda slag.

In respect of production to final discard slag, there are two basic types of slag reduction procedures that have been investigated on a pilot scale since November 1991. The first type is the semicontinuous treatment of slag as it is produced at the end of the soft-lead production cycle (as wet out in the original operating concept). The slag will need to be cluxed with millscale (iron oxide), limestone and silica to produce a slag composition, at the end of reduction, that has a melting point between 1200 and 1250 °C and a viscosity that is sufficiently low to enable tapping from the vessel.

The main problem experienced with the reduction of high-lead slag (i.e., above 50wt.% Pb) is that such a slag has a propensity to foam, especially at the start of reduction. The other difficulties with the semicontinuous treatment of high lead slags are:

- the additions of fluxes are made to an estimated weight of slag in the furnace; this can lead to frequent additions with a corresponding delay for slag analysis
- there is a need to run deep slag baths of high-lead slag to enable a lance to operate on the lower volume of discard slag
- the wear on bricks is expected to increase due to the frequent thermal cycling

Although there are solutions to some of these difficulties, on balance, the campaign basis retreatment of soft-lead slags after size reduction and analysis may be more attractive.

The slag treatment procedure is simply the charging of high-lead slag with fluxes and reductant to a low-lead (30wt.%) slag bath. The lead in slag is maintained at between 30 and 40% and metal is tapped periodically. When the depth of slag has reached a certain level, a batch reduction is performed and a hard metal and discard slag are tapped. If needed, the slag can be fumed to very low levels of lead i.e., 0.5wt.% PB. This procedure has been demonstrated many times in the pilot plant and once on the commercial scale. The discard slags so produced have passed the TCLP leach tests. The main attractions of this procedure are:

- minimal foaming of the lower –lead slags
- more control of flux additions – the weight and composition of slags in known before treatment
- frequency of plant dislocations for slag treatment is less by running campaigns of slag reduction
- less thermal cycling of the refractories

There are clearly time and fuel penalties in reheating slag, as well as more materials handling than in the semicontinuous procedure. The lack of slag foaming and simpler operations of the retreatment route, however, make it the currently favoured process option.

Conclusions

The major aspects of commissioning, performance to data and future activities of the Isasmelt process have been discussed. As is usual where new technology is being advance, changes would have been made given the benefit of hindsight. Fortunately, the fundamentals of this new technology for secondary-lead operation r sound in both process and engineering. By mid-1993, it is confidently expected that both lead alloy output and the cost of production will be within forecast.

References

1. R. Muller, Lead: Supply and Demand in the Western World, Recent Developments and Outlook into the 1990s. *Proc. 10th Int. Lead Conf. Nice, France,* Lead Development Association, London, 1991, pp. 30-45.
2. K.R. Barrett, The Isasmelt Process, *Conf. GDMC, Bleifachausschsuss,* Apr. 1992.

Text copied with permission from BRM, with updated diagrams by Xstrata Technology

Paper Title:
THE ISASMELT PROCESS – OPERATING EXPERIENCE OF A SECONDARY PLANT

Authors:
I. Mearns and P. Hawkins
Britannia Refined Metals Limited Northfleet, Kent (UK)

Date of Publication:
1992

The Isasmelt Process - Operating Experience of a Secondary Plant

I. Mearns and P. Hawkins
Britannia Refined Metals Ltd, Northfleet, Kent, UK.

Abstract

The emphasis on lead usage is moving steadily towards battery manufacture, with a projected rise to around 75% by the end of this decade and an increasing responsibility towards producers, consumers and recyclers. (Refer to Dia. 1) Following an in depth study, Britannia Refined Metals Ltd. embarked on the development of its secondary business at the Northfleet site. The construction of a new secondary refinery was completed and began commissioning in June 1991. The plant incorporates modern battery breaking technology and advanced smelting technique using the ISASMELT process for lead recovery from the scrap stream.

Plant and process descriptions are outlined together with experience gained during operation of the plant to date.

Introduction

BRM Secondary Operations

The BRM site occupies a prominent riverside position near to the river Thames main crossing point for the London peripheral motorway. The site incorporates both primary refinery and the new secondary smelter. Secondary operations also include BRL (Britannia Recycling Limited) which is situated at Wakefield in West Yorkshire. The two sites both recycle scrap batteries and water materials associated with the lead/acid battery industry.

Scrap Handling

In the U.K. during the past two decades, economic restraints at the smelters and increasing environmental concerns have seen the majority of dismantling operations transferring to the smelters. Collection and delivery of the cased battery scrap is now mainly handled through an existing network of scrap merchants. Automotive batteries are expected to last for approx. 4 years or longer before replacement is necessary.

In 1992 licences were issued to 24 850 000 vehicles to use the national road system with a further 25 500 licences for electric powered vehicles. New registrations are currently running at 1.9 million per annum. There are also large numbers of batteries used in unlicensed vehicles, standby and emergency power systems. For the secondary lead recycling efficiency to operate at high levels approaching 100%, the introduction of supportive collection schemes need to be implemented and if necessary legislated for. (e.g. Levy schemes or mandatory take-back, with increased consumer education.) As many smelters are operating at or near a loss during a period when lead use for battery production is increasing, smoothing the flow of battery scrap back to the secondary smelter is paramount for future economic and environmental demands. The BRM plant has been designed to meet these criteria. (Refer to Dia. 2.)

Plant Technology

Apart from the existing rotary furnace and pyrometallurgical refining methods, the battery breaking and furnace plants were chosen as:

CV phase 1 and 2.

The CX plant is well engineered and is capable of producing high quality desulphurised paste as furnace feed with reduced levels of entrained separator materials. This material extends the operating cycle of the Isasmelt furnace due to the low impurity levels, giving minimised sulphur emissions. The CX plant is manufactured in Milan, Italy, by Engitec Impianti.

Isasmelt Furnace

The furnace can efficiently produce high quality bullion from the CX paste product. Mount Isa Mines Ltd. in association with CSIRO (Commonwealth Scientific and Industrial Research Organisation) have developed the ISASMELT process. Application to secondary lead production follows successful demonstration with the treatment of lead, copper and nickel concentrates in Australia. The innovative feature of the technology is the use of a top entry Sirosmelt lance and static compact vessel. Helical vanes inside the lance impart a swirling motion to the process air which freezes a protective layer of slag onto the outside of the lance. This allows the tip to be immersed in a molten slag bath for long periods. A highly intensive reaction takes place as feed material is quickly absorbed into the turbulent slag bath.

Effluent plant

CV effluent is handled using cross microfiltration technology. Precipitation and flocculation of dissolved heavy metals is carried out prior to entry into the plant. Filtration through a series of porous fabric tubes takes place where the solid precipitates form the body of the filter medium on the inside of the tubes. Clean effluent passes through the tubular membranes, is collected and stored prior to discharge.

Process Description

Battery storage

Incoming battery scrap is stored in a covered, drained building with internal access to the CX crushing mill via vibrofeeder and conveyer belt. Incoming scrap quality is closely monitored. Free acid is filtered and stored for treatment in the CX plant.

Battery breaking

Batteries are crushed in a hammer mill at 16 tonnes per hour. The paste fraction is separated on exiting the mill through a vigorously washed vibrating screen and into a stirred tank. The remaining washed compartments pass over the screen and are conveyed to hydrodynamic classifiers. The classifiers separate the remaining components and deliver streams of grid/poles, polypropylene, ebonite and PVC separators to storage bays. (Refer to Dia. 3 & 4)

Paste Treatment

Thickened paste slurry from the bottom of the tank is pumped to one of two reactor vessels. Liquid caustic soda is added to react with sulphate ions present in the paste. Any lead oxides are unchanged.

Reactions: $H_2SO_4 + 2NaOH = Na_2S + 2H_2O$
$PbSO_4 + 2NaOH = Na_2SO4 + PbO + H_2O$

When the reactions are completed, the contents of the reactor are passed through a filter press. The filtrate is sent to the effluent plant for further treatment. Following washing, to remove residual sodium salts, the filter cake is discharged and conveyed to a storage bin.

Acid electrolyte is treated in the same way. (Refer to Dia. 5)

Materials handling

Storage of the paste feed (CX filter cake.), reductant, grids/poles, fluxes, stockpiled paste and mixed fume dust are contained in storage bins and hoppers. The furnace DCS system can control and feed these materials to the furnace singly or combined via weighbelts and conveyors. The main feed stream of CX paste is transferred from the storage bin over a weighbelt to the conveyor system. All other materials are kept in covered storage bays and transferred to the feed hoppers using loading vehicles, thus reducing spillage to a minimum.

Isasmelt furnace

The furnace consists of a static cylindrical vessel which is refractory lined to give an internal diameter of 1.8m. It is designed to produce 35 000 tonnes per annum of bullion. (Refer to Dia. 6)

Desulphurised paste from the CX plant storage bin is continuously fed into a bath of molten battery paste together with coke as a reductant. The submerged combustion lance produces a highly turbulent bath as the paste feed is reduced to a low antimony soft bullion. The metal is intermittently tapped into pots via a launder system. On completion of each soft lead cycle, the slag is conditioned and reduced to produce a high antimony bullion and discard slag.

TABLE 1

Composition of Furnace Feed Materials

Material	Battery Paste	Grid Metal	Flue Dust
Pb%	66.20	90.8	60.0
Sb%	0.45	2.5	2.0
S%	1.00		

TABLE 2

Composition of Products

Material	Soft Lead	Hard Lead	Slag	Flue Dust
Pb%	99.8	79.2	0.5	60
Sb%	<0.1	20.6		2
Cu%	0.1	0.1		
FeO%			50	
SiO2%			25	
CaO%			15	

The furnace is operated on a semi-continuous basis. When smelting paste, soft lead is tapped every 3 hours. Antimony from the feed concentrates in the slag as oxide which gives a consistent soft lead bullion. When approximately 250 tonnes of feed material have been charged, metal is again tapped and additions of lime and iron made to give the required final slag composition. Similar procedures are adopted for grids/poles.

The reduction stage is commenced at this point with increased coal additions at elevated temperature. (1150-1250C). A ferrosilicate slag is produced with low lead content. After settling, the furnace is tapped for metal and slag. The operational parameters are outlined in table 3.

TABLE 3

Isasmelt Furnace Operational Details

Operation	Smelting		Reduction
	Paste	Grid Metal	
Bath temperature (C)	800-900	800-900	1200-1250
Feed Rate t/hr	10.0	18.0	
Air flow Nm3/s	1.40	0.9	1.70
Oil flow kg/hr	500	390	600
Coal rate t/hr	0.5	-	0.5

Plant operation – first year

CX Plant

The CX plant achieved the designed crushing rate of 16 tonnes per hour. However, early problems with separator fibres, polypropylene conveying and the computer/plc interface reduced utilisation of the plant. The filter press was unreliable and improvements were made to the operating programme. Paste transfer proved to be at the root of many problems associated with plant availability. Paste transfer pipes to the reactors suffered blockages as did the filter cake discharge from beneath the press.

Modifications were made to the screw collection system under the filter press with limited success. A new design was initiated while the other problems were being addressed.

The CX plant is coupled to the Isasmelt furnace by the paste storage bin which has a capacity of 200 tonnes. The CX plant suffered limitations to its utilisation following difficulties with the bin discharge screw. It is thought that residual sodium sulphate solution in the paste caused cementing of the material around the screw conveyor flights. Discharge rate declined to 6 or 7 tonnes per hour. This change occurred after commissioning of the of the phase 2 desulphurisation. A modified screw was fitted and the problem resolved.

The first year of commissioning and operation placed high demands on the operators and engineers. Many failures of small items such as pump shaft seals and electrical components occurred as a result of the learning process involved for reliable plant operation.

Isasmelt Furnace

Until the CX plant began operation, manually fed battery paste via temporary conveyors allowed the furnace to operate and continue commissioning. The attention focused mainly on lance flow measurements and the process control programming for the DCS system. As the CX plant supplied more feed, output from the plant increased due to the greater utilisation of the developing control system, up to period 6-1991. (Refer to dia. 7) The main difficulty on the practical operation rested with tapping procedure. Increased wear on water cooled tapholes resulted from inexperience with the methods involved. Further reductions in output resulted following loss of lance airflow calibration which took longer than expected to remedy. Inaccurate lance air caused difficulties with control of the slag bath depth. Coupled with a refractory reline, output was substantially reduced during period 10-12. (Refer to dia. 7) Production returned to a slightly improved performance by the end of period 13. During the first year of operation, the soft lead stage emerged as a flexible and robust process capable of producing high quality bullion. Grid metal from the CX plant was commissioned into the process after 6 months to augment the operation.

However, reductions in output through isolated case of equipment failure could not mask the underlying restrictions to performance which were split into two areas:

1) Feed transfer
2) Gas handling

Feed handling

The feed handling system reduced plant availability and demanded more of the operators' time than expected. Battery paste transfer caused problems with conveyor belts, transfer chutes, drive systems and screws. Modifications were actioned following analysis of specific problems throughout the system. During this time a high degree of effort was maintained by the production operators to keep the plant running.

Paste discharge from the CX plant storage bin was restored to design rate after screw modifications, then upgrading of belt and chute cleaning implemented while further improvements were considered.

Also in progress was the design and construction of additional feed handling equipment, which would replace the temporary conveyors used initially, to produce a more flexible system.

Gas Handling

The limits of capacity in the system were easily reached when the furnace began to achieve above design feed rates. Also, any difficulties with mechanics or instruments resulted in feed rate reductions in order to maintain acceptable hygiene conditions. Again, improvements were made to all parts of the system, especially controls in the baghouse itself and I.D. fan damper controls. It became clear that the system was too highly stressed and plans were made to erect and commission additional filter capacity by Dec. 1992.

Plant performance from period 1 – 1992

CX Plant

The phase 1 section containing hydrodynamic separation and discharge has presented few problems to date. Attention has been mainly focused on the phase 2 section and effluent plant. Standards of paste quality are improved and with the engineering experience gained since start-up, the plant is capable of exceeding the crushing rate of 256 tonnes per day. There are some longer term tasks to complete regarding the computer interface and programme improvements together with objectives for mill hammer life, oil cooling to the bearings and rotor changing.

In January, various changes were made to pipework, control valves and the filter press.

The replacement paste transfer equipment was installed under the filter press. A further increase in output was achieved as a result. Operation to crush forecast tonnage of batteries continued throughout February.

Isasmelt furnace

Having attained a peak in output during period 4, (Refer to dia. 7), some restrictions to production were addressed. Levels of soda in the furnace increased as more CX paste became available. The soda forms a separate layer and floats on the slag bath. The soda layer at times proved difficult to tap if the slag bath was not completely drained. The soda remained trapped in the furnace where it interfered with the protective coating on the lance tip. A second taphole has been installed at a higher level which allows tapping of the soda slag without having to drain the furnace.

Closer control of the plant operation with detail on temperature and reduction rate also contributed to record production in period 4. Following refractory reline in period 5, production was subject to restrictions in feed handling. A planned stoppage of the furnace allowed further design improvements to be installed on the conveyor system and gas handling. The results showed a further substantial increase in production in period 9. With the additional feed handling equipment planned for April commissioning, furnace utilisation is expected to increase further.

Final slag reduction is still under development and application to the furnace is being progressed. However, with the production of soft lead being maximised for the foreseeable future, the final stage of the originally envisaged five stage process will be developed and trialled when furnace time is available.

The reduction route at present involves two possible methods:

1) In situ reduction of the remaining slag bath following completion of soft lead production. Fluxes are added to give the final slag composition. All remaining metals in the slag are then reduced to give a high Pb/Sb alloy and discard slag.
2) Slag retreatment method of feeding crushed high lead slag to a low lead (30%-40%) slag bath. Lead level in the slag bath is maintained at 30%-40%. Flux composition is monitored and trimmed on the run. Lead is tapped and the slag bath depth checked.

When the slag volume has increased sufficiently, a full reduction takes place to give a high Sb/Pb alloy metal and insert discard slag.

Option 1) has displayed a tendency to foam at the start of reduction when the lead is high. (i.e., above 50%).

Other difficulties are encountered regarding the calculation of slag volume in the furnace for flux addition and small residual volume of slag after reduction.

Option 2) is favoured as a simpler operation which can be used to campaign pre-crushed slag. There are additional advantages:

- More accurate control of flux additions.
- Less thermal cycling of the refractory lining.
- Minimal foaming of the lower lead slag.

The life of the refractory lining and not reached satisfactory lengths to date and a development program is in progress with refractories and vessel cooling.

Feed handling

The conveyor system received improvement to the transfer points, drives and operator access during a planned shutdowns for work on the main chimney. On start-up the system allowed improved furnace operation during period 9. (Refer to dia. 7) This allowed the CX plant to operate and match the increased feed demand with a corresponding rise in output.

The additional feed system was installed and commissioning during April. The plant consists of a similar storage bin to the CX paste bin and conveyors which are integrated into the materials handling area. The more flexible and controlled feed system will improve furnace control.

Gas handling

The increased baghouse facility completed commissioning at the beginning of 1993. Process and hygiene gas handling capacity has been doubled to the Isasmelt and Rotary furnaces. Much improved operations have resulted, especially with Isasmelt, where changes to the filter bag construction and reduction to spray water in the process gas have improved the performance of the filter plant.

Effluent plant

Following the addition of extra settlement tanks from the original treatment plant, the filter curtain duration has been extended and curtain cleaning reduced. This is coupled with in situ acid washing which reduces the frequency of removing the curtains for cleaning. Manual operation of the control system is still favoured due to the fine limits placed on pH control.

Regular inspection and monitoring by the CX plant operators ensures consistent effluent quality is maintained to comply with the consent for discharge.

Table 4 lists the maximum discharge limits for the plant effluent.

TABLE 4

Maximum allowable effluent discharge limits

pH value		6.0 – 9.0
Max. volume of discharge (m3/day)		200
BOD (Biochemical oxygen demand)		10mg/l in 5 days at 20C
Dissolved metals in mg/l		
	Antimony	5
	Lead	1
	Zinc	1
	Arsenic	0.2
	Silver	0.1
	Copper	0.1
	Nickel	0.1
	Cadmium	0.05
	Mercury	0.05

Metal refining and moulding

Preparation of the finished products has continued with few problems to date. The increasing flow of metal from both furnaces will allow optimisation of ingot production from the kettle floor. Subjects for investigation will cover metal transfer from the furnaces, blending efficiency and procedure, using the two 120 tonne capacity refining kettles. The blending operation arises from the combination of a high quality soft bullion source provided by the Isasmelt furnace and more selective antimonial bullion produced from the Rotary furnace.

Conclusions

The issues affecting commissioning and development of the secondary plant at BRM have been discussed. The Isasmelt process technology for battery recycling through the secondary plant will continue to advance, with commitment to both quality and innovation. From the present foundation of operating experience, metal production is expected to be within forecast by the middle of 1993. Production costs will still be subject to completion of developments actioned during the early part of the year and are expected to meet forecast during the latter part of 1993.

References

1. R. Muller, Lead: Supply and Demand in the Western World, Recent Developments and Outlook into the 1990s. *Proc. 10th Int. Lead Conf. Nice, France,* Lead Development Association, London, 1991, pp. 30-45.
2. K.R. Barrett, The Isasmelt Process, *Conf. GDMC, Bleifachausschsuss,* Apr. 1992.
3. K. Ramus and P. Hawkins, Lead/Acid Battery Recycling and The New Isasmelt Process. *3rd European Battery Conf., Munich,* Oct. 1992.
4. Dr D.N. Wilson, The Recycling of Lead. *Recycling '93 Conf., Geneva,* 1993.

Text copied with permission from BRM, with updated diagrams by Xstrata Technology

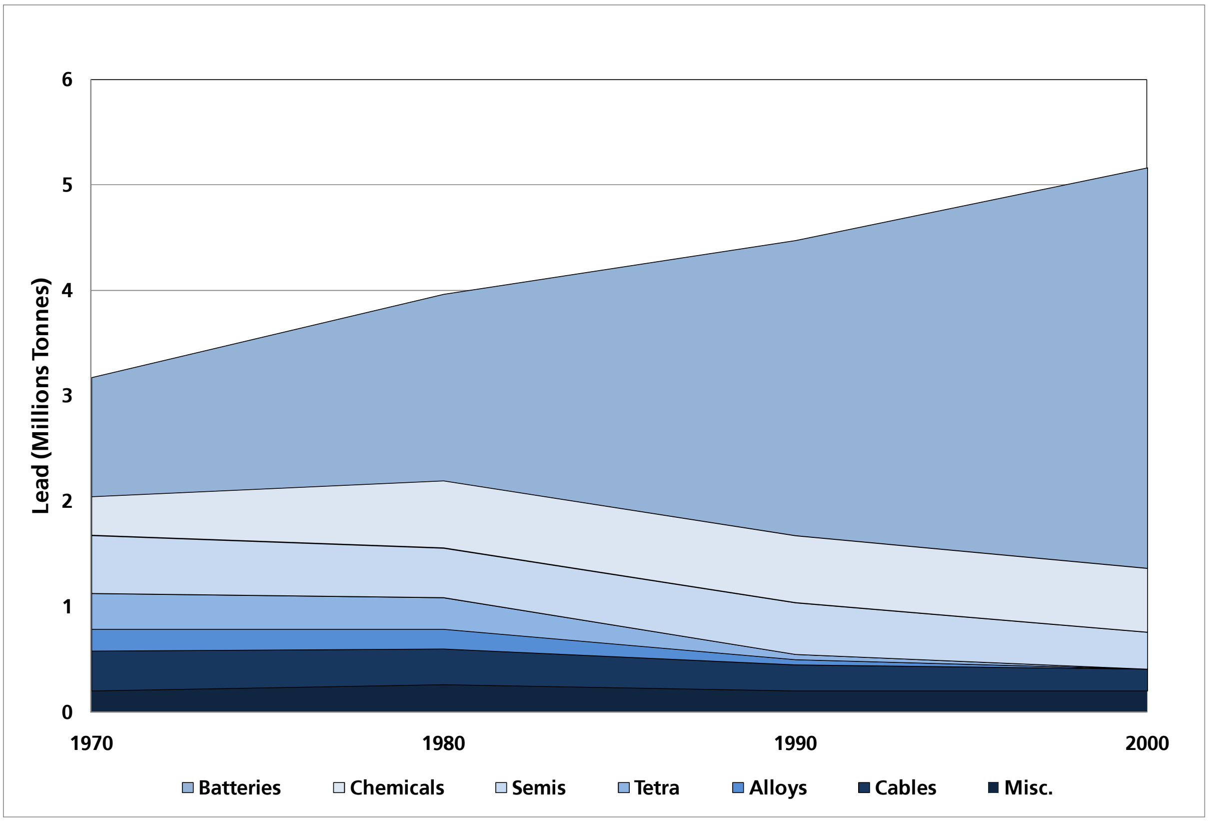


Diagram 1 Recent and Expected Development of Lead Consumption in the Western World by End Uses

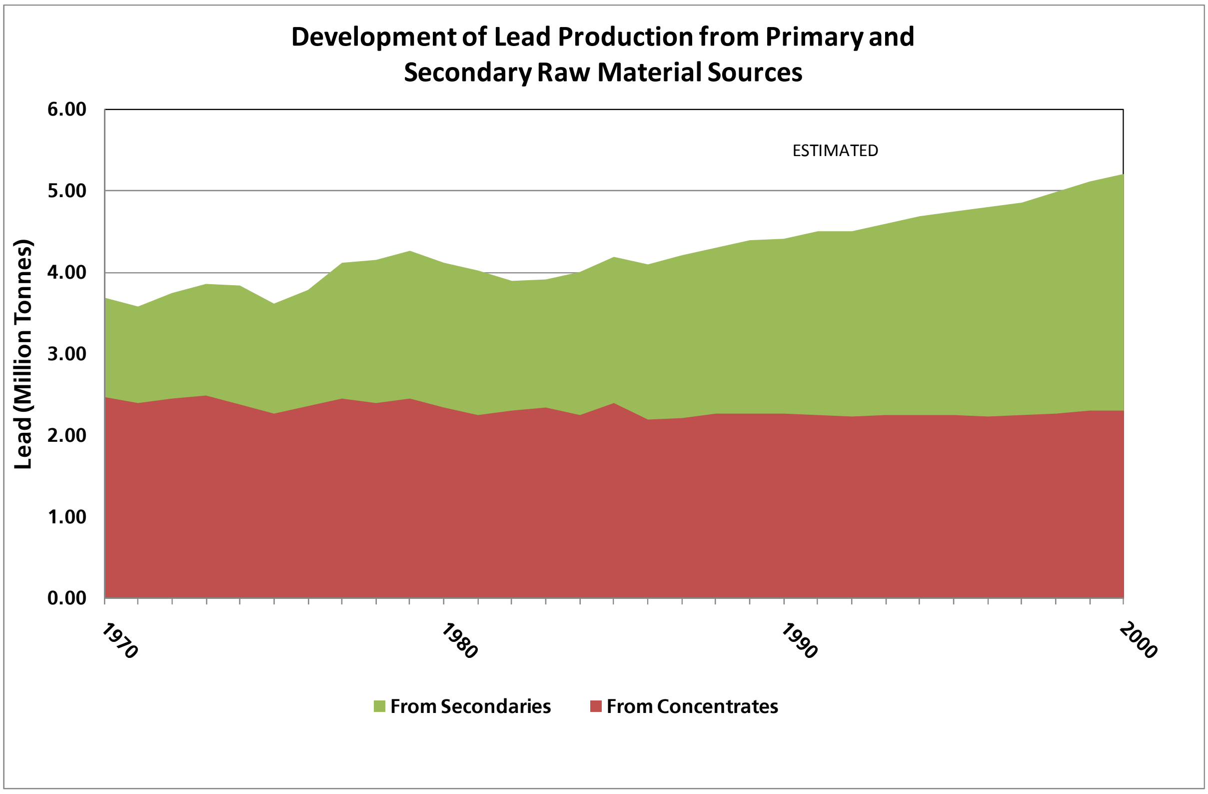


Diagram 2 Development of Lead Production from Primary and Secondary Raw Material Sources

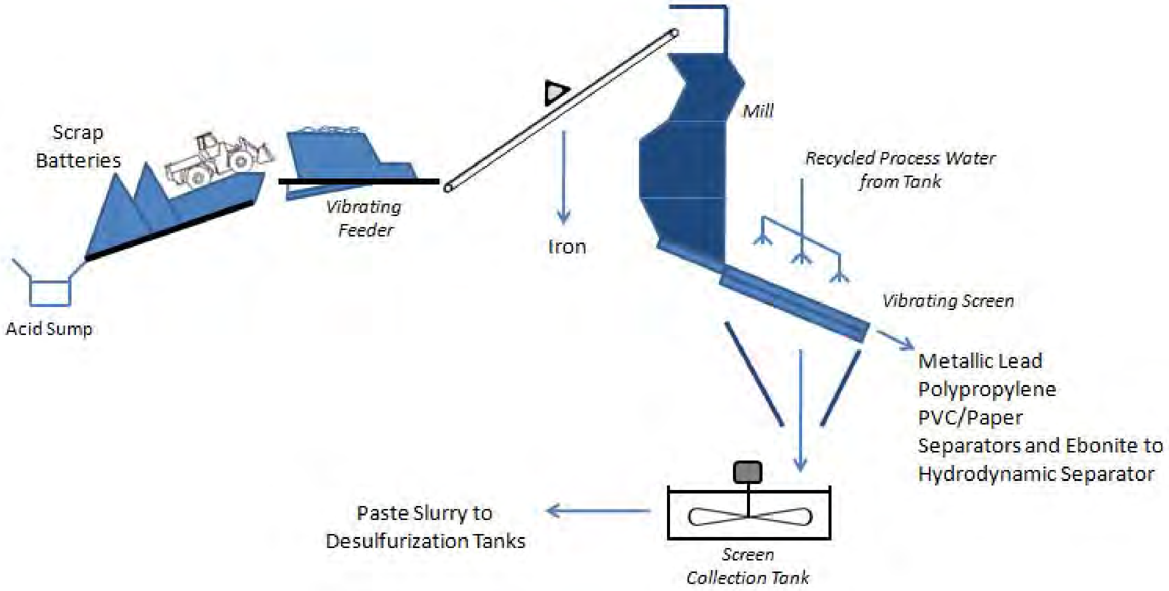


Diagram 3 Battery Crushing and Screening

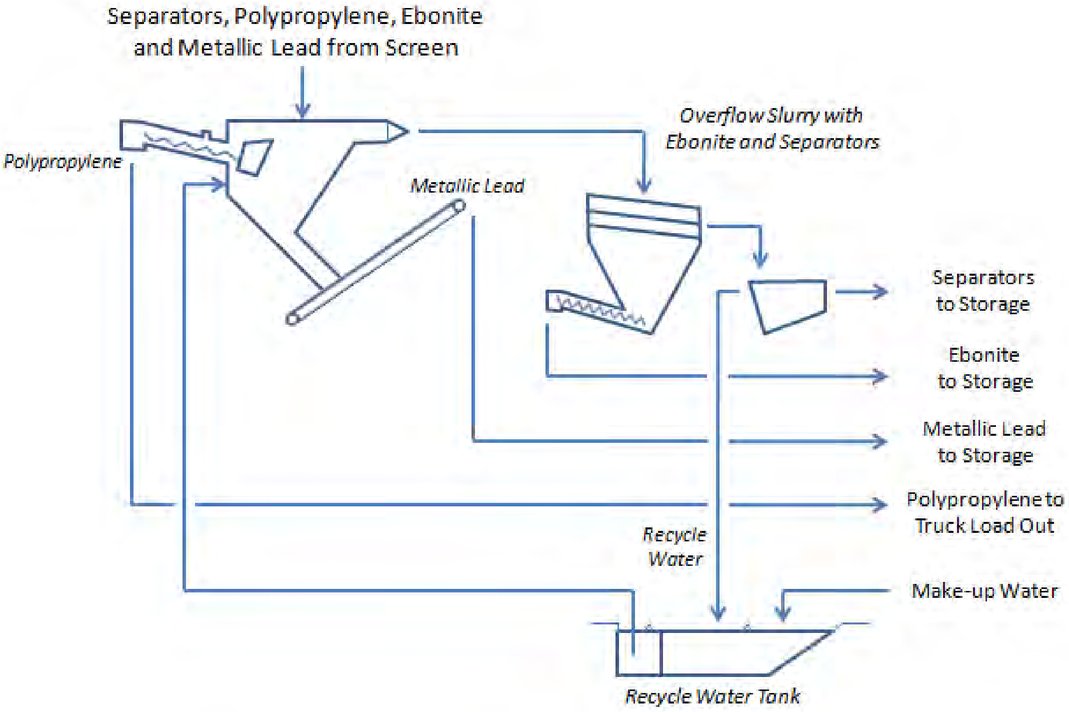


Diagram 4 Hydrodynamic Separation System

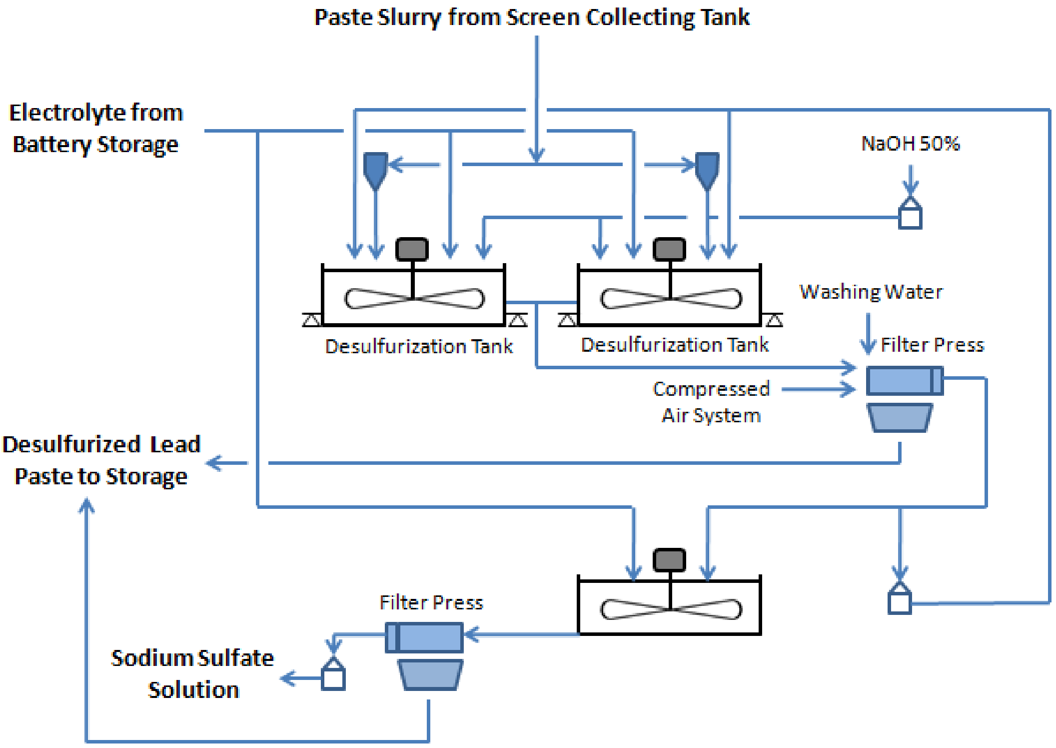


Diagram 5 Paste Desulfurization

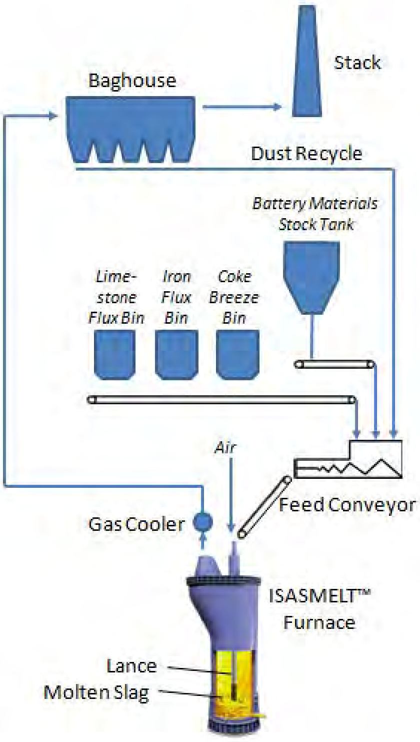


Diagram 6 Schematic Illustration of the Isasmelt Plant

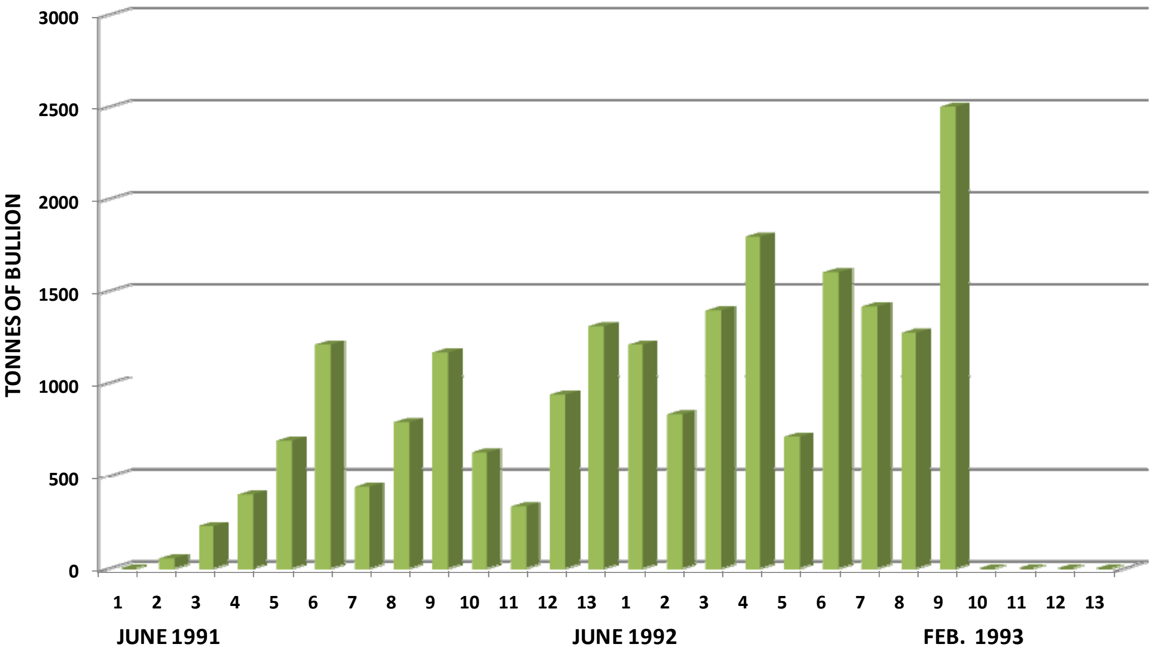


Diagram 7 Isasmelt Furnace Period Production

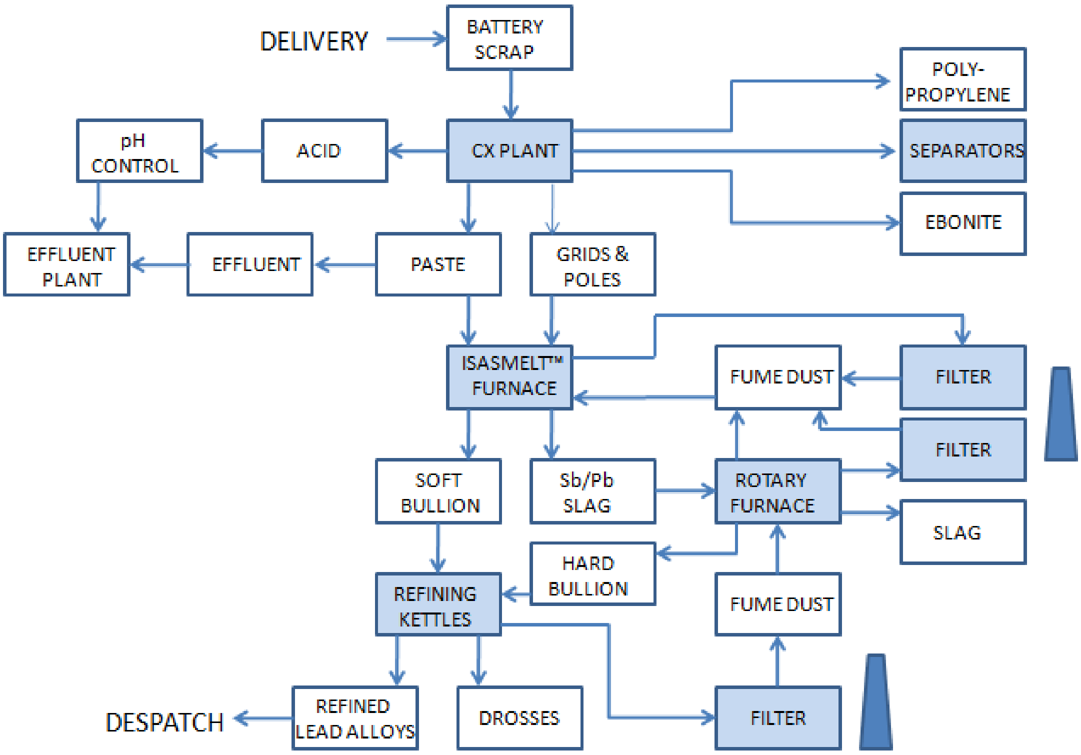


Diagram 8 Secondary Refinery Flow Sheet

ISAKIDD™

Copper Refinery Modernisation, Mopani Copper Mines Plc, Mufulira, Zambia

Maambo Chooye[1], Rakesh Patel[1], Addin Pranowo[2] and Brendan O'Rourke[3]*

[1]Mopani Copper Mines, Mufulira, Zambia
[2]Glencore Technology, Townsville, Australia
[3]Glencore Technology, Brisbane, Australia

**Corresponding author: brendan.orourke@glencore.com.au*

The electrolytic refining of copper has been conducted at Mufulira, Zambia, since 1939. The original refinery has been upgraded and expanded since then to match the output of anodes produced at the Mufulira, Luanshya, and Nkana smelters at various times; however, by the end of 2012, it was recognised that customer expectations regarding–cathode physical quality and chemical purity have increased due to the introduction of new copper refining technologies. In June 2013, Mopani Copper Mines Plc (MCM) committed to the modernisation of the copper refinery that included the conversion of the existing refinery to the ISAPROCESS™ technology, installation of electrode handling equipment, upgraded electrolyte circulation systems, and upgraded process control systems. This paper describes the equipment, technologies and new operational practices that constitute the Mufulira Refinery Upgrade Project. All activities have been progressively completed since late 2014 and have culminated with the registration of a new London Metals Exchange cathode brand, MCM2, in July 2017. The MCM Refinery is the first copper electrorefining plant in Africa to utilise the ISAPROCESS™ technology.

INTRODUCTION

Copper has been electrolytically refined in Mufulira, Zambia for over 75 years. The original refinery has been expanded several times and one part of the refinery was expanded to conduct commercial-scale electrowinning operations. These operations included in-situ leaching, vat leaching, and heap leaching of oxide ores. A dedicated solvent extraction plant was associated with each leach operation.

The primary objective of the Refinery Modernisation Project was to ensure that current and future customer expectations could be fully satisfied, and to implement productivity improvements, whilst allowing the refinery to have the flexibility to source anodes with higher impurity levels than currently received. This necessitated a change in the refining technology used by Mopani Copper Mines Plc (MCM).

MCM selected the ISAPROCESS™ technology due to its extensive record of converting starter-sheet refineries to permanent stainless steel cathode plates. At the same, MCM also took the opportunity to upgrade and modernise its electrode handling systems, electrolyte circulation systems and process control systems.

Owing to the requirement for the existing refinery to maintain copper production operations at full capacity, the upgrade activities were conducted progressively over a four-year period. The constructions activities are described below in more detail. The operational improvements since completion of the project are also described in detail.

PROCESS SELECTION

Layout

Various combinations of cell house layout and design have been implemented in ISAPROCESS™ copper refineries. MCM was constrained in some cases due to the requirement to maintain full copper cathode production capacity during the modernisation activities. In particular, this meant that the new electrode handling machines had to be located in areas that were not being used for existing production activities. The location of anode receipt from the smelter was fixed by the existing transport infrastructure.

Design Criteria

The principal design objectives for the project were:

- Long-term production forecasts;
- customer requirements;
- productivity improvement;
- upgrade of electrolyte circulation systems;
- future expansion options.

Long-term production forecast

MCM is upgrading the existing Mufulira Mine and expects to complete the new Synclinorium Mine and Concentrator Project in 2019. When combined with external concentrate feed treatment at the MCM smelter, this will enable the Mufulira Smelter to increase production to 225 000 tonnes per annum. The modernised refinery production capacity matches this requirement.

Customer requirements

MCM has been producing copper cathode marketed under the London Metal Exchange (LME) Registered Brand MCM. Cathode produced by the modernised refinery is now marketed under the LME Registered Brand MCM2. The cathode bundle being produced now has many advantages over the previous style. It is compact, easier to handle and transport and no longer has the protruding starter-sheet loops associated with starter-sheet cathode.

Productivity improvement

Using starter-sheet technology, MCM was operating four tank houses plus a starter-sheet production aisle with-in their refinery complex. Glencore Technologies experience with ISAPROCESS™ copper refinery conversions conducted over the past 30 years was that conversion to stainless steel cathode plate technology achieved direct productivity improvements of 25% to 30% when compared with starter-sheet technology. In the specific MCM case, this has allowed MCM to reduce the operational footprint to three tank houses, with an associated reduction in the number of operating overhead cranes, refining cells, and, importantly, copper inventory.

Further productivity improvements have been made in the electrode handling task. The elimination of starter sheets has removed all aspects of starter-sheet production – starter-sheet production cells, manual stripping, sorting, and preparation activities, loop production, hanger bar assembly, manual cell loading, alignment activities and replacement of falling cathode due to broken loops. The introduction of the ISAPROCESS™ permanent stainless steel cathode system requires a dedicated cathode stripping machine (CSM) to perform cathode stripping operations. The CSMs selected at MCM are capable of fully automatic operation and eliminate all steps that were performed manually. At the same time, MCM elected to introduce a fully automated anode preparation machine (APM), anode scrap washing machine (ASWM), and upgraded its overhead refinery cranes to suit the new electrode handling machines.

Upgrade of electrolyte circulation systems

With the expected increases in copper production intensity, it was determined that the existing circulation systems required additional circulation capacity and improved process control. At the same time the refinery circulation equipment configuration was changed so that each tank house now has its own, independent circulation system. All piping associated with the refining cells has been upgraded and the ability to collect decant solution from each refining cell during anode change operations has been implemented. The reagent dosing system was replaced for each tank house with a system that has the ability to prepare and dose thiourea, glue and salt. Automated process control has been implemented for all circulation system equipment via the introduction of a SCADA PLC-based system.

Future expansion options

The current MCM refinery building has five individual tank houses plus a dedicated starter-sheet section. The design, specification and placement of all electrode handling equipment will allow MCM to expand its refinery production to five tank houses without the need to purchase additional anode handling equipment.

Final Process Selection

The modernisation of the MCM copper refinery was implemented with the following items:

- introduction of the ISAPROCESS permanent cathode plate technology;
- replacement of starter-sheet preparation machines with automated cathode stripping machines;
- introduction of automated anode scrap washing machine and automated anode preparation machine;
- introduction of automated anode and anode scrap transfer system;
- anode mould design was upgraded to match new anode dimensions;
- total replacement of electrolyte circulation system piping;
- installation of dedicated cell decant electrolyte piping system;
- replacement of fixed-speed circulation pumps with variable-speed drives;
- replacement of all reagent dosing equipment;
- partial replacement of overhead cranes;
- upgrade of anode and cathode lifting bales on the refinery overhead crane;
- installation of a SCADA PLC-based process control;
- installation of cell voltage monitoring system;
- implementation of an intermediate bar-based cell top furniture system;
- installation of new electrolyte circulation system tanks in Tank house Five;
- replacement of 336 refining cells in Tank House Five.

PROCESS DESCRIPTION

Modernised Refinery Operating Statistics

Table I shows the key operational statistics for the modernised refinery.

Table I. Operating statistics for new refinery.

Capacity	225 000 t per annum
Cells	1320
Electrodes per cell	36 cathode / 37 anode
Electrode pitch	100 mm
Cathode deposition area, per side	0.82 m^2
Maximum anode mass	320 kg
Maximum cathode mass	60 kg
Anode cycle	18 days nominal
Cathode cycle	2 crop or 3 crop
Current density	305 A/m^2
Electrolyte circuits	3
Number of cranes	6

Refinery Layout

The existing MCM Refinery design is typical for the era in which it was built. Overhead refinery cranes from that era had very slow operating speeds hence the design objective was to minimise the length of crane runs. Crane rail heights were also set as low as possible to reduce lifting and lowering times for electrodes. This lead to refineries being built with multiple aisles, short crane runs with a low operating height over the electrolysis cells. Movement of material was best managed by having a common centrally located aisle running the length of the refinery building. Industrial standard forklifts were also not common at the time so rail was used as a method for moving anodes, anode scrap and copper cathodes.

Owing to the existing layout, all of the new electrode handling machines can only be located in the central common aisle. This led to the situation where anodes and anode scrap need to be conveyed up to 120 m if only one APM and one ASWM unit are installed. By working with the supplier of the APM and ASWM, an innovative ground-running transfer car system for carrying anodes and anode scrap was specially developed for each machine. All of the electrode handling machines are designed to be serviced by forklifts.

Figure 1. New overhead crane with drip tray carrying cathode plates.

Electrode Handling Equipment

Anode preparation machine
Anodes are received from MCM’s copper smelter via rail and forklift. Anodes are loaded to the APM by forklift for weighing, blade pressing, lug-offset pressing and contact milling. Contact milling is used to ensure that anodes hang vertically in the cell and ensure best possible current distribution in the refining cells. The anodes are then transported in loads of 36 anodes to the appropriate tank house using the anode transfer car system. The APM was supplied by MESCO, Japan.

Anode transfer cars
A transfer car system is used to move anodes and anode scrap with-in the refinery building. Each transfer car runs at ground level and carries 36 pieces of anode or anode scrap to or from the respective APM or ASWM. Each transfer car services nine fixed set-down or pick-up positions located across the length of the refinery building. The Anode transfer car system was supplied by MESCO, Japan.

Cathode stripping machines
Three medium speed CSMs have been installed at MCM. One CSM is installed in each operating tank house; however, each CSM has the capacity to service the electrodes from the two bays in each tank house. The CSMs have been specially designed to interact with manually operated and low clearance overhead cranes. This means that conveyors are used to transfer cathode plates instead of fast-moving trolleys. This eliminates the possibility of a crane bale to trolley collision. A forklift is used to remove the finished cathode bundles to the product despatch area. The wash water used to remove electrolyte

from the cathode deposits is heated using electric heaters. The wash water system is designed for future conversion to steam heating. The CSMs were supplied by MESCO, Japan.

Anode scrap washing machine

Anode scrap is moved from the refining cells by overhead crane and then conveyed to the ASWM via the transfer car system. In the ASWM, the scrap is washed to remove any adherent slimes and then formed into bundles for forklift removal. Anode scrap is returned to the copper smelter. The ASWM was supplied by MESCO, Japan.

Overhead crane upgrade

The change in refinery electrodes, electrode pitch and cell-top furniture necessitated that new anode and cathode plate bales were required for the overhead cranes. At the same time, MCM decided to replace four of the six existing refinery cranes. MCM utilises crane transfer cars located at the end of each tank house. This allows the overhead cranes to be moved to another aisle when needed, thus reducing the overall number of cranes required in the refinery. The new cranes and lifting bales were designed and supplied by DEMAG, South Africa.

Electrolyte Circulation

Many parts of the existing circulation systems were not suitable for the increased operating intensity associated with the ISAPROCESS™ technology. A higher and more consistent cell electrolyte flowrate was the major change required. This meant that all electrolyte piping was upgraded, together with the associated circulation pumps. The use of variable-speed pumps was required so that a responsive flow control scheme could be implemented. The electrolyte is distributed to the cells by direct pumping without an intermediate head tank. Owing to the introduction of the thiourea reagent to replace Tembind, the existing reagent dosing systems were also replaced.

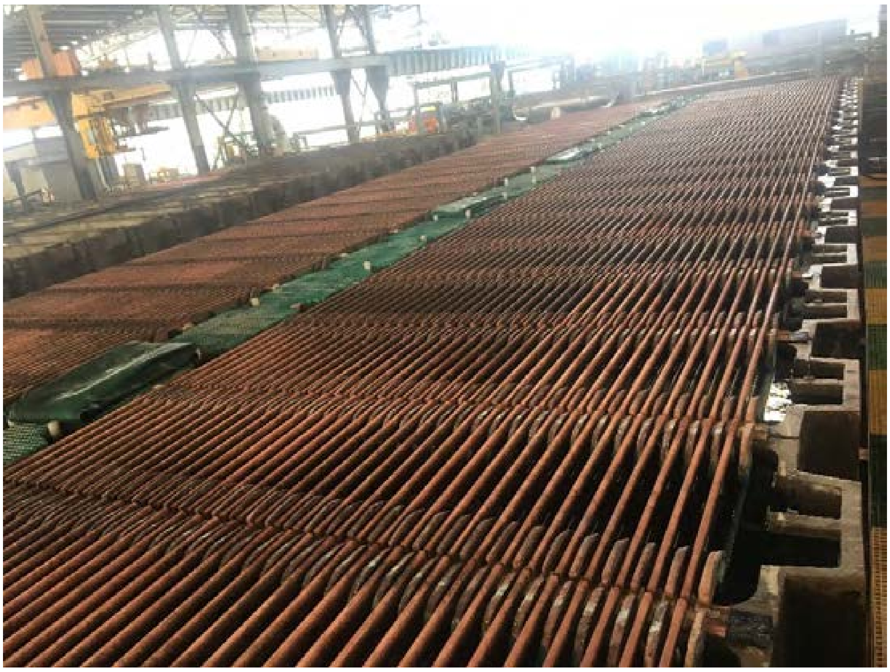


Figure 2. ISAPROCESS™ permanent stainless steel cathode plates in service.

Process Control and Monitoring

To ensure the correct electrolyte conditions for electrolysis at the higher operating intensities, improved process control of the operation was required. This has been achieved by implementation of a SCADA-based PLC system for the electrolyte circulation. The Siemens PLC-7 system was used for the SCADA configuration.

Refinery Cells
As most of the existing refining cells were retained, the dimensions of the new stainless steel cathode plates were matched to the internal cell dimensions. The dimensional design of the anode was changed to match the new cathode plate dimensions. Owing to the inherent straightness of the stainless steel cathodes, the electrode pitch in the cell was able to be reduced to 100 mm, which provides reduced power consumption when compared with the starter-sheet design. The cathode plates were designed and supplied by Glencore Technology, Australia.

The upgrade of the electrolyte circulation system required changes to the electrolyte feedpipe for each cell. At the same time, an individual cell flow-control device was included in the new piping design. New cells were installed in Tank house Five to allow for full implementation of the decant collection system.

It was necessary to change all cell-top furniture to an intermediate bar design. The end cell busbars on all sections required modification to suit the ISAPROCESS™ system. The additional busbar elements and intermediate bars were designed and supplied by Copalcor, South Africa.

CONSTRUCTION

In all cases, the scheduling of construction activities was dictated by cathode production requirements. With four tank houses still available for production, each of the three modernised tank ouses was consecutively taken off-line for refurbishment over a two-year period. At the same time, MCM was forced to curtail production during 2016 due to power supply restrictions in place in Zambia.

The major construction milestones are listed below:

Jan–May 2014	New end cell busbars designed and installed
July–Dec 2014	Preparation of electrode handling machine areas
Oct–Dec 2014	Installation of electrode handling machines – TH1 & TH4 CSM's
Feb–March 2015	Installation of electrode handling machines – APM, ATS
May 2015	Installation of electrode handling machines - TH5 CSM
June–Aug 2015	Replacement of reagent system, piping and pumps – TH4
May–Aug 2016	Installation of new overhead cranes
Feb- July 2016	SCADA and cell voltage monitoring installed
Mar–Dec 2016	New circulation system for TH5 installed
May–Oct 2017	Installation of new cells – TH5
May–Nov 2017	Replacement of piping and pumps – TH 1
November 2017	Replacement of piping TH5

COMMISSIONING

MCM formed a dedicated commissioning team to ensure that the transition between the two operating systems was performed safely and in step with the completion of construction/installation activities. A major emphasis was placed on training across all disciplines: metallurgy, operations and maintenance. At the same time, the operational requirements and testing to obtain LME registration for the new cathode were completed.

The key commissioning milestones are listed below:

Jan–Feb 2015	Operations and maintenance training program completed at CRL Australia and PASAR, Philippines
March 2015	Full commissioning of CSM TH4, APM, ATS, ASWM and TH4 circulation systems

Apr–July 2015	Full conversion of TH4 for first LME registration testing program
Jan–Dec 2016	Second LME registration testing period – TH4
June 2016	Full conversion of TH1 and commissioning of CSM TH1, CSM TH5
June–Nov 2016	Progressive commissioning of new refinery overhead cranes
June 2017	LME registration granted for MCM2 brand
October 2017	Central control room operational for SCADA and cell voltage monitoring systems
October 2017	Full conversion of TH5 and commissioning of TH5 circulation system
October 2017	Starter-sheet oOperations ceased

Commissioning engineers were supplied from MESCO, DEMAG and Glencore Technology for the electrode handling equipment. Specific ISAPROCESS™ training services and process engineers were also provided by Glencore Technology. MCM also retained the services of two expatriate electrode handling machine specialists to provide on-site training.

PRODUCTION AND OPERATIONAL IMPROVEMENTS

Table II details the improvements in several refinery production performance indicators since the conversion of the refinery to the ISAPROCESS™. The level of improvement achieved is consistent with the levels foreseen during the feasibility study phase of the project.

Table II. Cathode purity *improvement – ISAPROCESS vs conventional*.

Impurity (ppm)	Conventional			Full ISAPROCESS	Note
	2014 average	2015 average	2016 average	2017/8 average	
S	3.74	6.29	4.93	4.43	Similar
Pb	1.50	1.81	1.42	0.53	Improved
Ni	1.19	1.27	0.70	0.58	Improved
As	1.24	1.48	1.27	0.50	Improved
Bi	0.43	0.41	0.26	0.20	Improved
Sb	0.36	0.51	0.54	0.47	Similar
Se	1.07	1.24	0.84	0.28	Improved
Ag	6.48	6.20	5.35	5.95	Similar
Sn	0.28	0.24	0.31	0.25	Similar
Te	0.45	0.43	0.46	0.31	Improved
Zn	0.56	0.46	0.48	0.418	Similar
Fe	<2	<2	0.48	0.35	Similar

Table I shows the long-term improvement in copper cathode impurity levels for the common cathode copper impurities. These levels easily comply with the requirements for LME Grade A copper cathode. At least four different mechanisms for cathode impurity exist in an industrial-scale copper electrorefinery. Each mechanism and the associated impurities are briefly described below with reference to Mopani's specific anode impurities:

- inclusion of slimes in cathode deposit: indicators are lead, selenium, arsenic, bismuth, tellurium and antimony;
- occlusion of electrolyte in cathode deposit: indicators are nickel, iron and arsenic;
- electrolytic codeposition: indicator is silver;
- precipitation on cathode deposit surface: bismuth, antimony, tellurium and arsenic.

Impurities associated with slimes inclusion have shown the greatest improvement. The inherent straightness of the permanent stainless steel cathode plate means the probability of a slimes particle making contact with the cathode deposit during electrolysis is significantly lower than for a starter sheet. In addition, permanent stainless steel cathode plate technology allows for shorter crop ages, which reduces the probability of slimes inclusion occurring.

The level of impurity associated with electrolyte occlusion has also improved noticeably. This is attributed to the improved cathode deposit smoothness achieved by MCM since introducing the thiourea reagent. MCM upgraded its electrolyte circulation system to ensure that all refining cells are now receiving the correct flowrate of electrolyte, and thus reagents are always available at the cathode deposit surface during electrolysis operations. The inherent flatness of a permanent stainless steel cathode plate and the shorter crop ages also contributes to the cathode deposit smoothness.

No change in the silver assay at MCM indicates that the level of silver codeposition is not influenced by the conversion to permanent cathode plates. MCM does not have sufficient quantities of bismuth, antimony and tellurium in its anode and electrolyte for the surface precipitation impurity mechanism to be noticeable.

Customer Acceptance

Figure 3 shows the percentage of copper cathode production that was suitable for direct despatch to customers as LME Grade A cathode. The long-term trend for starter-sheet production was 86.6%, with significant variability in conformance. This indicator has improved to 97.7% since the conversion of the refinery to the ISAPROCESS™, with a vast reduction in variability.Figure 4 shows a photograph of typical cathodes.

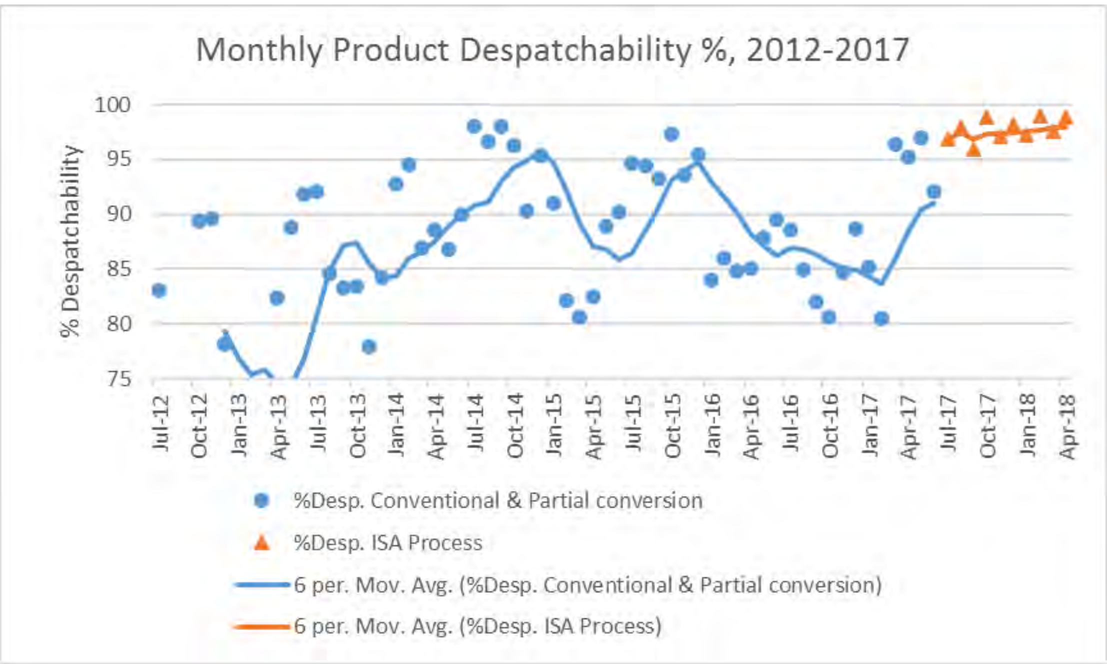


Figure 3. Cathode product dispatchability.

Figure 4. Copper cathode at CSM and existing trolley system.

Current Efficiency

Figure 5 shows the long-term trend for refinery current efficiency (CE). The long-term trend for CE was 85.9%. This indicator has improved to 97.5% since the conversion to the ISAPROCESS™, due to significant reduction in the number of short circuits. The variability in month-to-month performance has also reduced significantly.

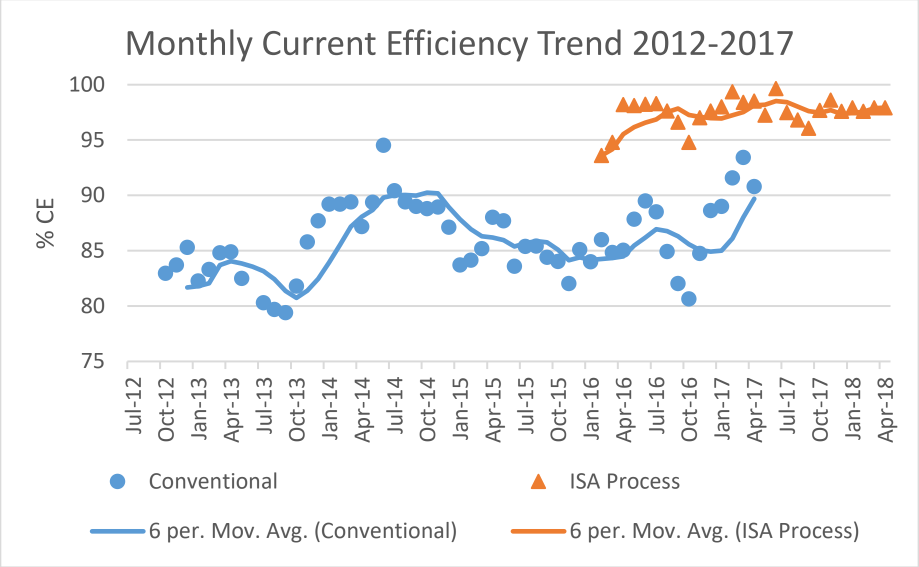


Figure 5. Current efficiency improvement after conversion.

Productivity

Figure 6 shows the long-term trend in labour productivity for the refinery. It is important to correlate this measure against the actual production level at the time. Refinery output was constrained significantly for two years between August 2015 and September 2017 due to the countrywide power shortages and MCM's business response to the crisis. Thus, to make a valid comparison, only months with production above 12 500 t are compared.

The long-term productivity trend for the starter-sheet refinery was 30 t cathode production per person-month. Since full conversion to the ISAPROCESS™, the productivity measure has averaged 41 t cathode production per person-month. This is a 37% improvement in labour productivity. As MCM returns to previous production levels, the productivity improvement achieved will continue to improve.

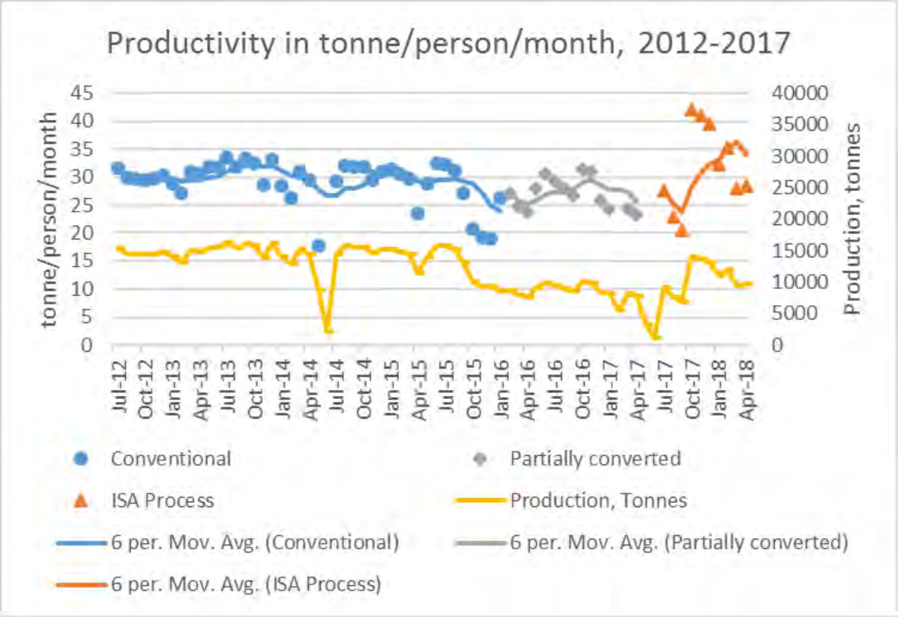


Figure 6. Productivity in tonne/person-month

REFERENCES

Aslin, N.J., Eriksson, O., Heferen, G.J., Sue Yek, G. (2010). Developments in cathode stripping machines-an integrated approach for improved efficiency. *Proceedings Copper 2010 International Conference, Volume IV Electrowinning and –refining*, GDMB. Germany. pp. 1253–1270.

Barrios, P., Alonso, A., Ortiz, C. (1999). Improvements in the operating practices at the Atlantic Copper Refinery. *Proceedings Copper 99–Cobre 99 International Conference, Volume III – Electrorefining and Electrowinning of Copp*er, Dutrizac, J.E., Ji, J., Ramachandran, V. (eds.). The Minerals, Metals and Materials Society, Warrendale, PA. pp. 291–299.

Dobner, R., Schwill, M. (1995). Modernisation and extension of a conventional tank house. *Erzmetall.,* 48, Jahrgang, Nr. 4/95, S. 272-276.

Jenkins, C. (1962). Plant designing for electrolytic copper refining. *Australasian Engineer*, 105, 42–45.

Moats, M., Robinson, T.G., Davenport, W.G., Karcas, G., Demetrio, S. (2007). Electrolytic copper refining – 2007 world tankhouse operating data. *Proceedings Copper–Cobre 2007 International Conference, Volume V-Copper Electrorefining and Electrowinning*. Houlachi, G.E., Edwards, J.D., Robinson, T.G. (eds.). The Metallurgical Society of Canada, Toronto .pp. 195-201

O'Kane, J. (1997). ISAPROCESS – Benchmarking copper refining electrolysis. *Proceedings GDMB 33rd Metallurgical Seminar*, Lunen, Germany. pp 67-82

O'Rourke B.J. (1999). Tank house expansion and modernisation of Copper Refineries Limited. *Proceedings Copper 99–Cobre 99 International Conference, Volume III – Electrorefining and Electrowinning of Copper*, Dutrizac, J.E., Ji, J., Ramachandran, V. (eds.). The Minerals, Metals and Materials Society, Warrendale, PA. pp. 195–205.

Robinson, T.G. (2002). Electrolytic refining. *Extractive Metallurgy of Copper*, 4th edn. Davenport, W.G., King, M., Schlesinger, M., Biswas, A.K. Elsevier Science, Oxford. pp. 265–288.

Wenzl, C., Filzwieser, A., Antrekowitsch, H. (2007). Review of anode casting – Part I: Chemical anode quality. *Erzmetall.,* 60 (2), 77–83.

MOUNT ISA MINES NECESSITY DRIVING INNOVATION

*V. Lawson, H. DeWaal, G. Heferen, N. Aslin, P. Voigt, and M. Hourn

Glencore Technology
Level 10, 160 Ann St
Brisbane, Australia, 4000
*(*Corresponding author: virginia.lawson@glencore.com.au)*

ABSTRACT

Mount Isa Mines (MIM) acquired a reputation for the successful application of R&D to develop break-through technologies for the mining industry starting in the 1978's through until the early 2000's. The ISAPROCESS^{TM} tank-house technology has been licensed to copper refineries throughout the world, and a significant per cent of the world's copper is refined using this technology. Since development in the late 1980's more than 20 ISASMELT^{TM} copper and lead smelting furnaces are now installed in countries around the world. Jameson Cell flotation technology developed jointly by Mount Isa Mines and Professor Graeme Jameson is widely used in the Australian coal mining industry and increasingly in the base-metal and gold industry. The IsaMill^{TM}'s developed at Mount Isa and McArthur River made it possible to develop the McArthur River and George Fisher orebodies and has been successfully implemented into base metal fine grinding applications around the world. The most recent commercialised innovation is the atmospheric leach Albion Process^{TM} with its supersonic HyperSparge^{TM} gas sparger, is being adopted as a solution to the increasing complexity of orebodies.

MIM's contribution to the industry was significant given the size and the remote location of its operations with Townsville Copper Refineries more than 1350 km and Mount Isa 1800 km from the nearest state capital of Brisbane. This paper will briefly discuss the development of each of these technologies and why MIM – now owned by Glencore - was so successful innovating and developing such technologies over a period of nearly 40 years.

KEYWORDS

Innovation, Mount Isa Mines, ISAPROCESS^{TM}, IsaKidd^{TM}, ISASMELT^{TM}, IsaMill^{TM}, Jameson cell, Albion Process^{TM}, HyperSparge^{TM}, ZipaTank^{TM}

INTRODUCTION

Mount Isa is located in the Gulf Country region of Queensland about 1800 kilometers North West of Brisbane (see Figure 1). It came into existence because of the world class mineral deposits found in the area. In 1923 the orebody containing lead, zinc and silver was discovered by the miner John Campbell Miles. Mount Isa Mines Limited (MIM) was founded in 1924 to develop the minerals discovered by Miles, but production did not begin until May 1931. It paid its first dividend in 1947 after 16 years of troubled production. In 1954 the 1100 copper orebody was discovered and with rapidly rising reserves during the 1950's and 1960's led to the construction of new concentrators to treat lead/zinc/silver ores in 1966 (#2 concentrator) and copper ore's in 1973 (#4 concentrator). The difficult nature of the Mount Isa lead-zinc orebodies has meant that the company had always needed to be at the forefront of mining technology. In the 1970's through to the 1990's, it became a world leader in developing new mining techniques and processing technologies as a response to declining metal prices and rising costs. Mount Isa has been smelting copper since 1953 and lead since the early 1930's. Copper Refining at Mount Isa's fully owned subsidiary of Copper Refineries Proprietary Limited (CRL) had commenced operations in 1959.

Figure 1 – Location of Mount Isa and Townsville relative to Brisbane the nearest Capital City

Technologies to come out of Mount Isa include the ISAPROCESS$^{\text{TM}}$ copper refining technology, the ISASMELT$^{\text{TM}}$, The Jameson Cell, the IsaMill$^{\text{TM}}$, the Albion Process$^{\text{TM}}$ and the Hypersparge$^{\text{TM}}$. Mount Isa Mines Ltd was acquired by Xstrata in 2003 and Xstrata was then merged with Glencore in 2015. The level of innovation achieved at Mount Isa Mines is unsurpassed and was the result of the difficult nature of the Mount Isa ore bodies and its response to declining metal prices and rising operational costs in the 1970's and 1980's. By the 1990's, Mount Isa had become a world leader in innovative mining techniques and state of the art processing technologies. The processing technologies are discussed below.

INNOVATIONS

Each of the innovations developed at Mount Isa Mines had a driver but the overarching desire was to make technology more efficient and cost effective. Each of these process developments will be discussed separately.

ISAPROCESS™

The development of the ISAPROCESS™ tank house technology had its beginning in the zinc industry. During the mid-1970s, MIM was considering building a zinc refinery in Townsville to treat the zinc concentrate produced by its Mount Isa operations. As a result, MIM staff visited the zinc smelters using the best-practice technology and found that modern electrolytic zinc smelters had adopted permanent cathode plate and mechanised stripping technology. MIM realised that the copper refineries performance was constrained by the conventional practice of copper starter sheets. The preparation of these copper starter sheets was labour intensive and the overall cycle was several weeks in duration.

MIM initiated a research program aimed at developing similar permanent cathode technology for copper refining. CRL, a subsidiary of MIM, had been operating in Townsville since 1959, using conventional starter-sheet technology and treating blister copper produced in the copper smelter at Mount Isa. Permanent cathode technology was developed and adapted over many years of in-plant experimental work and successfully introduced to the Townsville refinery in 1978. . The fundamental difference between the new ISAPROCESS™ and the conventional starter sheet technology is the use of a permanent reusable cathode blank instead of a non-reuseable copper starter sheet and the introduction of mechanised and automated electrode handling machines replacing labour-intensive manual operations. The vertical edges had plastic strips and the bottom cased in wax to prevent copper cathode from growing around the edges of the cathode plate during, stripping and allowing two separate copper sheets from each cathode plate. This technology led to major advances in the electrode handling systems and automation in copper tank houses. The improved geometry of the cathode plates and the significantly shorter cathode cycle times allowed for increased intensity and efficiency of the refining process. Introduction of permanent cathode technology resulted in higher capacity, better copper cathode quality with less defects, safer operation and a four-fold improvement in productivity. Considerable development work was required to modify the original stripping machines from their zinc cathode origins due to the heavier cathodes. The stripping capacity of the machines has increased from 250 plates per hour to 600 plates per hour in the latest designs. More recent developments include the elimination of wax masking from the cathode plate, robotic electrode handling machines, and the introduction Duplex Stainless Steel cathode plates giving greater durability and corrosion resistance. Through the use of ISAPROCESS™ user forums, to exchange ideas and developments in the technology and to share operational experiences, the technology has enjoyed continued improvement with higher productivity and improved quality at low cost.

Figure 2 – The IsaKidd process

In mid 1981 Falconbridge Limited commissioned a copper smelter near Timmins to treat concentrate from its Kidd Mine. The original copper cathode produced at Kidd suffered from the presence of higher concentrations of lead and selenium and could not meet customer specifications. It was determined that the use of copper starter sheets was preventing the Kidd refinery from meeting its cathode quality targets. Testwork began with the use of permanent stainless steel cathodes after preliminary tests showed a significant reduction in deleterious elements. The Kidd Process cathode used a solid copper header bar welded onto stainless steel resulting in a lower voltage drop than the ISAPROCESS™. Falconbridge began marketing the Kidd Process technology in 1992 providing competition between the two suppliers of permanent cathode technology. Between 1992 and 2006, 25 Kidd technology licenses were sold and 52 ISAPROCESS™ licenses.

The development of the ISAPROCESS™ and Kidd Process set the scene for a run of technology developments that continued until the mid 2000's. Xstrata took over MIM in 2003 and then Falconbridge in 2006. The Kidd Process technology consequently became part of the tank house package and together they have since been marketed as IsaKidd™ representing the dual heritage of the technology. The current robotic stripping machine (Figure 2) is based on over 30 years of copper refining and winning technology. Today over 100 licensees are using IsaKidd™ technology.

ISASMELT™

The sinter plant/blast furnace combination was the dominant technology for lead smelting throughout the 20th century. In the early 1970's companies using this technology came under sustained political and economic pressure as tighter environmental regulations were introduced, and energy costs increased, leading to higher capital and operating costs (Fewings 1988). It was in this environment that Mount Isa Mines sought a process that would improve the performance of the operations at their lead smelter in Mount Isa. After investigating the various processes under development, researchers turned their attention to the Sirosmelt lance. It had recently been developed on a laboratory scale at the Commonwealth Scientific and Industrial Research Organisation (CSIRO) in Melbourne. Following initial investigations Mount Isa Mines recognised the potential of the novel concept for smelting of lead concentrates and embarked on an extensive development program.

In 1978 a joint project was initiated between Mount Isa Mines and CSIRO to investigate the application of the Sirosmelt submerged-combustion technology to the smelting of Mount Isa lead concentrates. The ISASMELT™ process, as it became known, was developed to maturity for smelting copper, nickel, lead and zinc feeds by Mount Isa Mines through the 1980's and 1990's using incremental scale up. Commercialization only occurred once the process had been proven on laboratory, pilot and demonstration scale over many years. Approximately ten years were required for development of the lead and copper ISASMELT™ from crucible to demonstration scale (refer to Figure 4). During this decade the core know-how that was accumulated enabled the development team to reach the point where they were much better equipped to design and construct a full scale commercial plant – the final stage of the scale up process. Key aspects in this process were the selection of the scale up factors and the systematic design, development and re-engineering of several components of the technology. Figure 3 shows a comparison for the scale up stages for the lead and copper ISASMELT™ processes. Pilot scale was defined as unity for scale up comparison.

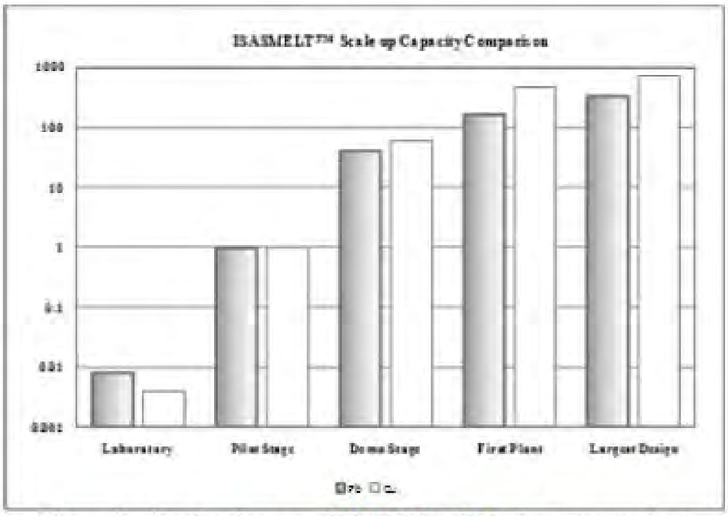


Figure 3 – Lead and copper ISASMELT™ Scale up comparison

During the scale up process, refer to Table 1, several aspects of the technology were developed to a high standard that allowed the ISASMELT™ technology to become a commercial success. As a result, ISASMELT™ technology now operates successfully at numerous plants around the world. The methodical approach to development of the technology has allowed owners to modernise their existing operations or create new businesses with significantly reduced technical risk.

An important parameter in the evolution of the ISASMELT™ technology has been the refractory campaign life. Figure 5 shows the history of the refractory campaigns at the commercial copper ISASMELT™ plant at Mount Isa since commissioning. At the time Mount Isa Mines management considered the installation of water cooling on the furnace refractories undesirable because of the potential for fatal incidents and increased operating costs. As a result the commercial scale furnaces were constructed with minimal water cooling. Although this led to shorter campaign lives initially, a development program was begun that focussed on optimising refractory materials selection and installation methodology. When coupled with process control strategies and continuous on-line monitoring of the bath temperature using systems developed over more than 10 years of operation, it allowed Mount Isa Mines to achieve campaign lives of more than 3 years without using any water cooling of the furnace refractories.

Figure 4 – Tapping matte from the copper ISASMELT at Kazzinc

Table 1 – Key Indicators of ISASMELT™ Plants from pilot to commercial scale

Topic	Unit	Pilot Scale		Demo Scale		First Full Scale		Current Design[1]	
		Pb	Cu	Pb	Cu	Pb[3]	Cu	Pb	Cu
Furnace ID	m	0.4	0.4	1.8	2.3	2.5	3.75	3.6	4.4
Lance Diameter	mm	38	38	150	250	250	350	250	500
Lance Control	-	Manual		Semi Automatic		Semi Automatic		Automatic	
Oxygen Enrichment	%	21	21	21	28	35	45	70	90
Nominal Feed Rate	tph	0.12	0.25	5	15	20	101	40	183
Offgas Treatment	-	Flue System / Baghouse		Gas cooler/ Baghouse		WHB		WHB[2]	

Notes:

ID: Internal Diameter; WHB: Waste Heat Boiler

[1] Refers to maximum throughput

[2] Some of the plants use a combination of radiation section and evaporative cooler for offgas treatment

[3] Refers to the smelting furnace from the two stage lead ISASMELT™ process

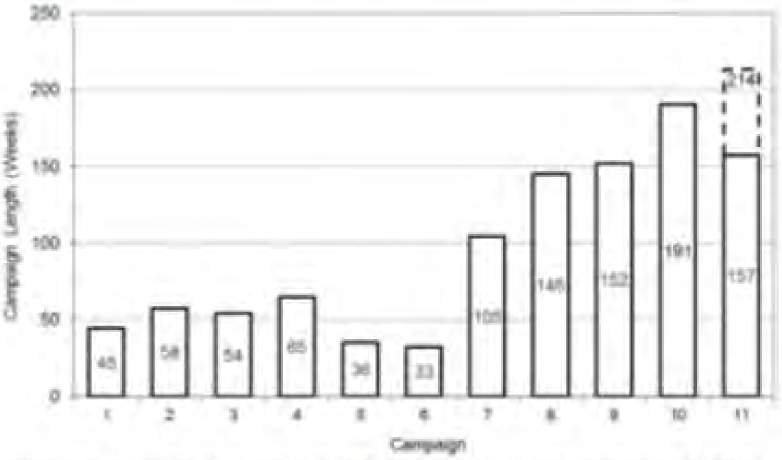


Figure 5 – Mount Isa copper ISASMELT™ plant campaigns (as of 2013)

Jameson Cell

The Jameson Cell (Figure 6) was jointly developed by Mount Isa Mines and Laureate Professor Graeme J Jameson (AO) of the University of Newcastle. Mt Isa had commenced operations with conventional flotation cells but was installing columns in cleaning duties in the mid 1980's. The columns had the benefit of froth washing that was likely to allow significant grade benefits in the very fine lead-zinc circuit. The first observations of the columns was that the collection process was slow necessitating long residence times and large volumes which remains a limitation of columns even today. In 1985 Professor Jameson was commissioned to undertake a project to improve the column sparger design.

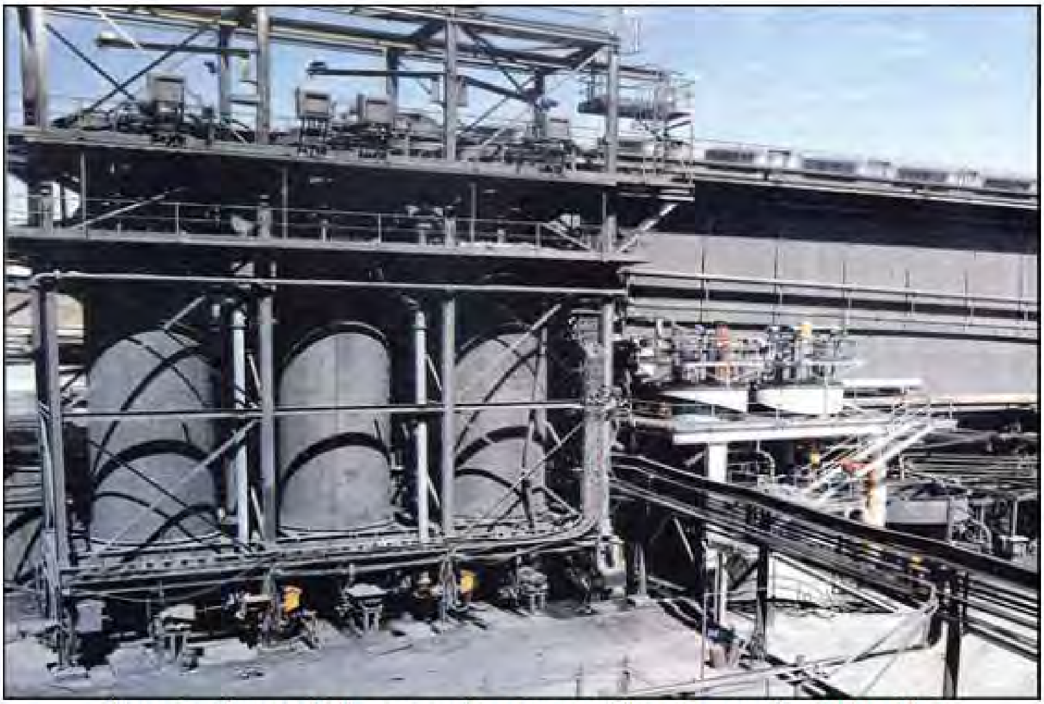


Figure 6 – Jameson Cells compared to columns of the same capacity at Mount Isa

Following initial work to provide an alternate method to bring together bubbles and particles, the downcomer was created. In the downcomer the air and the slurry are co-current with the air being entrained into the plunging jet under vacuum. Investigation showed that all of the bubble particle contact took place in the downcomer and thus the flotation tank could be much smaller. The first application at an industrial scale was in the lead zinc concentrator on the heavy media plant (HMP) lead slimes circuit. The initial improvement in performance were attributed to the very short residence time that allowed the minimisation of oxidation of galena fines. The cells were significantly smaller than the columns and there is no doubt the performance was superior as shown in Figure 3.

The testwork and trials in the early applications showed improved metallurgical performance when operated correctly. The challenge was operating them correctly. The technology hadn't been sufficiently developed to be successfully adopted into plant operations. The cell fell out of favour in base metals and in the 1990's was adopted into the Australia Coal industry and into niche SXEW applications where the main design challenges were resolved. The operability was improved by the introduction of a partial recycle to maintain constant flow and the maintainability of the cell was improved through various design modifications in operating plants. It was a period of continuous improvement. The result was a robust, low maintenance, easy to operate cell with the original features of excellent bubble particle contact.

The final obstacle was overcome when its adaption into the flowsheet was recognised to enable successful installations at the head of cleaner circuits and as low cost brownfield expansions. It is clear that the fast failures have had a significant effect on the success of the cell limiting its adoption into the industry. It is interesting that a significant proportion of sales are to return customers. Once you get over the hurdle of getting a Jameson Cell into your plant then seeing is believing. 2016 was the best year for Jameson cells into base metals and include the first sales back into South America where the cell had been abandoned after the difficulties of operations and maintenance of the Alumbrera installation. The metallurgical performance in Alumbrera was never the issue but the operators and maintainers hated the cells and they failed fast and hard.

The Jameson Cell celebrates its 30th birthday this year and has finally been adopted into mainstream base metals concentrators mainly as cleaner scalper at the head of the cleaner circuit. The cells generally recover up to 80% of the cleaner feed at high grades enabling much lower capital expenditure on the entire circuit. Process performance can be predicted from laboratory and pilot plant testing with demonstrated direct scale-up. It may have taken 30 years but the Jameson Cell is finally a success story. There are many lessons that can be learned from the implementation of innovation into industry from this case study.

IsaMill™

Unlike the developments of some of the other technologies at Mount Isa where efficiency was the main driver, the IsaMill was developed based on necessity. Figure 7 shows photomicrographs with the same scale of 40 micron demonstrating the increased complexity of Mount Isa ore over Broken Hill ore and the very difficult McArthur River ore. Although McArthur River was discovered in 1955 it was not able to be economically processed until the successful development of ultrafine grinding. McArthur River processing began in 1995 – 40 years after discovery when the IsaMill™ made it technically and economically feasible to grind all of the rougher concentrate to 7 micron to facilitate the rejection of non-sulphide gangue. Even at 7 micron galena liberation is not possible and a bulk zinc-lead concentrate is produced.

Figure 7 – Photomicrograph of a) Broken Hill ore b) Mount Isa ore c) McArthur River ore

Investigations into fine-grinding started at Mount Isa started in the 1970s using conventional grinding technology to increase mineral liberation by grinding to fine sizes. These technologies were not only found to have high power consumption but also proved to be detrimental to flotation performance as a result of pulp chemistry and iron contamination from steel media. These poor results were revisited during pilot plant and tower mill testwork in the 1980s which also showed an inability of tower mills to economically achieve the required sizes. When it became clear that the solution to efficient fine-grinding did not exist in the minerals industry, MIM looked for ideas to "crossover" from other industries that also ground fine particles – pigments, pharmaceuticals, foodstuffs (e.g. chocolate). While these mills operated at a much lower scale and treated high value products they demonstrated the principle that stirring fine media at high speed was highly efficient. The challenge was transferring this concept to continuous, high tonnage and lower-value streams in the minerals industry.

In 1991 the introduction of a Netzsch laboratory stirred mill to the Mount Isa site was a turning point in fine-grinding and ultrafine grinding. The ½ litre bench scale mill resembled a milk shake maker and used fine copper smelter slag as grinding media. Testwork on McArthur River ore started in 1991, and by January 1992, a small pilot scale mill, LME100, had been designed and installed at the Mount Isa pilot plant. The testwork showed that high speed, inert, horizontal mills could efficiently grind to 7 microns at laboratory scale providing major improvements in metallurgical performance. To make ultrafine grinding applicable to full-scale production a program of development was undertaken between Mount Isa Mines Limited and NETZSCH-Feinmahltechnik GmbH.

After 7 years of development and testing of prototypes in the Mount Isa operations, the IsaMill™ evolved. It was large scale, continuous, and most importantly robust because it was developed by operators. The crucial breakthrough was the perfection of the internal product separator – this allowed the mill to use cheap natural media (sand, smelter slag, ore particles) and to operate in open circuit. These are significant advantages for operating cost and circuit simplicity. Scale-up was tested using trial installations at the Hilton and Mount Isa lead/zinc concentrators. By the end of 1994, the first full scale IsaMill™ (1.1MW) was installed in the Mount Isa concentrator. Improvements to the technology were continually made by the operators, maintainers and engineers working with the technology.

In 1998 the rights for commercialisation of the IsaMill™ where transferred from Mount Isa Mines Limited to MIM Process Technologies (now Glencore Technology) and under an exclusive agreement with Netzsch. In December 1998, the IsaMill™ technology was launched to the metalliferous industry as a cost effective means of grinding down to and below 10 microns. The IsaMill™ is now a mainstream fine grinding machine with over 130 installations around the world.

The Albion Process™

In the 1990's, MIM were studying options for the development of the large Frieda River/Nena project in PNG through its subsidiary Highlands Pacific. The Nena ores were not amenable to smelting, due to the elevated arsenic content, and several hydrometallurgical options were examined. Out of this work, MIM developed the Albion Process™, named after the suburb in Brisbane where MIM's development laboratory was located. The Albion Process™ is a combination of ultrafine grinding using Glencore Technology's IsaMill™, followed by oxidative leaching at atmospheric pressure in a series of reactors designed to achieve high oxygen mass transfer efficiency. The HyperSparge™ was also developed to deliver oxygen to the reactors efficiently.

Various small scale continuous pilot plant campaigns were conducted in 1994 and 1995. A larger pilot plant (120kg zinc cathode/day) was constructed in 1997 to conduct testwork as part of a feasibility study on the zinc/gold resources of Pueblo Viejo in the Dominican Republic. Extensive piloting was also conducted on lower grade chalcopyrite concentrates for Cyprus Amax in 1998, and for Mount Isa Mines in 2000. Pre-feasibility and feasibility pilot testing was conducted on the zinc/lead bulk concentrates from McArthur River and Mount Isa in Australia between 2001 and 2005. During this time the Albion Process™ was successfully tested on over 70 different ores and concentrates. The process is designed to recover gold and base metals from refractory ores. The key to the process is the ultrafine grinding stage followed by a hot oxidative leach at atmospheric pressure.

In the period from 1994 until 2004, the Albion Process™ (see Figure 8) was seen as strategic to the MIM/Xstrata group, and was not marketed externally. In 2005, a decision was made to offer the technology to external clients under licence, and a marketing agent – Core Resources, was appointed to market the technology globally. Interest in the technology has been very strong in the subsequent period, with early licences signed in 2005 for the Las Lagunas Project, and 2006 for the Certej Project. The technology moved into commercial production in 2010 with the commissioning of Glencore's Albion Process™ plant in Spain (4,000 tpa zinc metal), followed in 2011 by the commissioning by Glencore of a second plant in Germany (16,000 tpa zinc metal). The Las Lagunas refractory gold project commissioned in 2012, and the GPM Gold refractory gold project commissioned in 2013.

Figure 8 – The Albion Process oxidative leach plant in Armenia

The major scale up risk with any oxidative leaching technology is oxygen mass transfer. High agitator power demands are common to achieve the shear rates in the vessel required for effective mass transfer at a commercial scale. A different approach was taken in the design of the Albion Leach Reactor to lower the agitator power demand. Glencore developed the HyperSparge supersonic gas injection lance to provide gas injection velocities of the order of 500 $m.s^{-1}$ within the leaching vessel, compared to the 4 – 8 $m.s^{-1}$ achieved with a typical agitator. Supersonic oxygen injection is a far more efficient method of generating shear than conventional agitation, allowing the total power input into the vessel to be significantly reduced, and greatly reducing the scale up risk for the oxidative leach.

The Albion Process™ was enabled by the fine grinding of the IsaMill™ and the process was designed to deliver a lower cost processing option for treating refractory mineral resources. There are now six operating Albion Process™ plants and the process has now an extensive database of potential applications.

CONCLUSIONS

MIM developed a significant number of processing innovations that are technical and economic successes. The ability to innovate at MIM was enabled by very challenging orebodies and the need to process efficiently to remain economically viable. The success has been attributed to the development of these technologies on an operating site with the R&D group solving the technical issues on small scale. Each subsequent scale up was completed in the operating plants where the operators, maintainers, engineers and metallurgists were required to achieve production goals at each step of the scale up to ensure funding for the next step.

The number of innovations, at MIM, was disproportionate to the scale of operations and may have been enabled by the remoteness of the site and the researchers and operators working collaboratively to solve economic and technical problems. The research group were not capital city based but worked on the same site and were required to assist with installation, commissioning and operation of the various stages. This co-operation led to adoption into the plant and a fast feedback loop for improvements. The ultimate success of the innovations has been their widespread adoption into the mainstream industry where feedback from operating sites based on a user group model has enabled continuous improvement of each of the technologies.

ACKNOWLEDGEMENTS

The authors would like to acknowledge Glencore Copper and Glencore Technology for permission to publish and to all the research and production personnel who enabled and improved the

technologies in their plants. The success of these developments continues with the input from end users in the ongoing development.

REFERENCES

Armstrong, W. "The Isa Process and its contribution to electrolytic copper," paper presented at the Rautomead Conference, Scotland, August 1999.

Fewings J.H., Management of Innovation – The IsaSmelt Process. Presented at AMIRA "Innovation in Metal Production" Technical Meeting at Mount Isa, October 3-4, 1988.

Fountain C. Isasmelt and IsaMills – Models of successful R&D – AusIMM Young Leaders conference, 19-21 March 2002, Kalgoorlie, W.A.

Nihill D.N., Stewart C M. and Bowen P. the McArthur River Mine – The First Years of Operation. The AusIMM '98 Mining Cycle. Mount Isa 19-23 April 1998.

Pease, Joe (August 2005), "Complex leaching becomes much simpler" (PDF), Australian Mining, pp. 26–32, archived from the original (PDF) on 12 September 2009, retrieved 6 January 2010

Pease, Joe September 2016, "Crossing the innovation valley of death" Presented to JKMRC.

Current distribution in modern copper refining

N.J. Aslin
Xstrata Technology
Hunter St
Stuart,QLD, Australia

D. Stone
PI International
3094 Emery Circle
Austell, GA 30168 USA

W. Webb
Xstrata Technology
Hunter St
Stuart,QLD, Australia

ABSTRACT

In today's modern copper electro-refineries, increasingly higher average current densities are being employed. With these increases many refineries are approaching their limiting current density. The nearness of the average operating current density to the limiting current density has placed increasing emphasis on the need to maintain an even current distribution. This paper explores the importance of maintaining even current density and discusses the factors, processes and practices that are necessary to achieve and maintain high quality production at high operational intensity.

INTRODUCTION

The copper industry was based essentially on the use of a copper starter sheet as the substrate for the refined copper deposition. In the 1970's operating current densities with this technology were typically around 220-250 amps per square metre.

There was a clear recognition that the maintenance of electrode spacing or geometry was crucial in minimising short circuits and rough growth within the cells. A number of systems aimed at rigidising the fragile copper starter sheets were introduced. These included a number of both pre and post-installation straightening systems including starter sheet embossing and restraightening systems such as the PD press.

The pursuit of the vertical electrode culminated in the introduction of permanent stainless steel technology by the ISA PROCESS^{TM} group in 1979 at MIM's Townsville Refinery. The introduction of this inherently straight permanent cathode technology led to its combined use with high quality anode straightening machinery and crane placement systems. These combined systems led to very predictable electrode geometry and inter-electrode gaps, resulting in superior cathode quality at high current density.

The industry now had a refining system, which had overcome much of the labour intensive tasks associated with maintaining correct and even electrode spacing.

Limiting Current Density

A key industry target has been to increase productivity and reduce costs while improving product quality. Increasing current density has been an important element of this aim, along with larger electrodes, closer spacing, larger numbers of plates per cell and higher time efficiency.

The maximum current density possible is related to the ability of cupric ions to migrate to the cathode surface as quickly as those ions can discharge from the anode. This is driven by the diffusion rate of cupric ions across the boundary layer at the cathode face. The thickness of the boundary layer depends on many factors including flow rate of the bulk electrolyte and the concentration gradient across the boundary layer. This process is described by Fick's Law (1), which can be written as;

$$\frac{-dQ}{dT} = \frac{D(C_b - C_e)}{d} \qquad (1)$$

where C_b is the concentration of the cation (cupric) in the bulk solution (mol/m^3), C_e is the concentration of the cation at the electrode surface, d is the distance over which the concentration change occurs (m), D is the diffusion coefficient (m^2/s). dQ/dT is the flux in mol/m^2/sec.

If the current density exceeds the ability of cupric ions to diffuse across the boundary layer the current will be carried by cations other than copper, and a reaction other than copper reduction at the cathode will occur. The limiting current density can be written as the equation;

$$i_{\mathrm{lim}} = \frac{nFDC_b}{d} \qquad (2)$$

where i is the current density (A/m^2), F is Faraday constant (C/mol) and n the number of moles of electrons in the electrochemical reaction.

This condition exists when the cupric ion concentration is zero at the electrode. If more current is driven through the electrode it will be carried by cations other than copper. In electrorefining, this would normally be Arsenic and possibly Bismuth or Antimony.

A key factor here is that the limiting current can be reached at any electrode, or part of an electrode in a cell, prior to the full set of electrodes reaching their limiting current. This results in the generation of a rough and open structure in the high current density regions. This cathode will not comply with the criteria specified by international standards. Rough growth in turn can result in the inclusion of slimes and electrolyte within the structure. Both occurrences will result in non-compliant product.

Quality Considerations

In a copper market where demand outstrips supply, the minimum standard is often sufficient. However in a less favourable market, only suppliers of the highest quality copper will maintain full sales of their product and achieve maximum premiums over standard product value. In the modern era, downstream fabricators are under constant pressure to reduce their costs. These companies are becoming less inclined to accept the need for rework due to poor raw material supply.

Table I – Cathode Quality Standards versus Typical ISA PROCESS^{TM}

Element	LME Limit ppm	ASTM B115 (COMEX) Limit ppm	Xstrata Refinery (12 month average) ppm
Pb	5	8	0.1
As	5	5	0.5
Sb	4	5	0.2
Bi	2	2	<0.1
Ag	25	25	12
S	15	25	5
Fe	10	12	<0.1
Ni(+other)	20	8	0.6
Se	2	4	<0.1
Te	2	2	<0.1

While the LME and Comex standards are recognised internationally as good supply, copper producers are now aware that simple compliance is not enough. Some of the world's key wire-rod producers will simply not accept sulphurs above 5 ppm. Lead concentrations should be maintained well below 3 ppm.

Current Distribution - Theory

Electrode pairs in a cell are arranged in parallel with the direction of current flow, such that total cell current divides between the electrode pairs in accordance with Ohms Law. The current passing through each electrode pair is inversely proportional to its component resistance.

Ideally, if all resistance paths are equal, the cell current will divide so that all the electrodes will operate at the mean current density over the entire surface of the cathode. In practice however, variations in ohmic resistance between electrode pairs leads to non-uniform current distribution. The range of current densities within each cell approximates a ‘normal’ distribution. Cathode plates at the extreme high end of the range are the first to exhibit rough growth and ultimately cause short circuits. These highs also restrict the ability to raise the mean current density because they impact on the current efficiency and cathode quality.

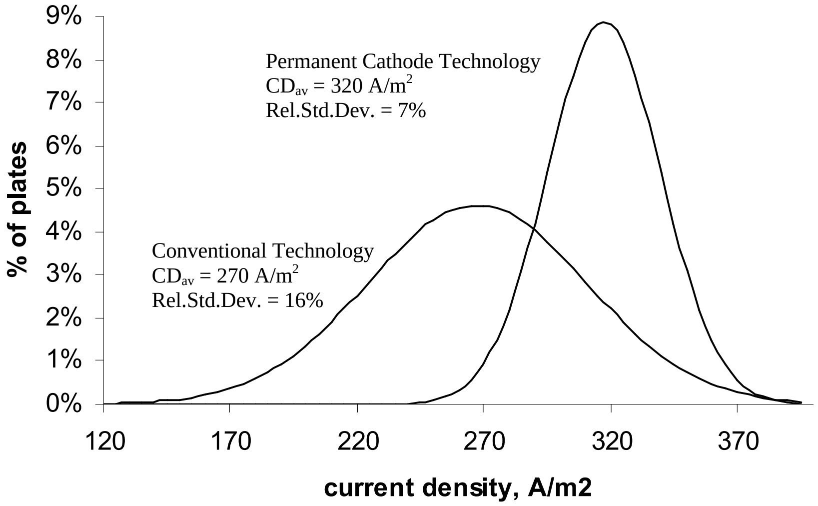


Figure 1 – Current Distribution Comparison - Conventional and Permanent Cathode

Factors affecting Current Distribution

The causes of non-uniform current distribution are simply those physical characteristics that affect the electrode pair resistance, namely;

- Electrode cell spacing
- Electrode alignment
- Electrode physical geometry
- Electrode contact resistance
- Electrode internal resistance

The factors that have greatest impact on current distribution will be those which contribute the greatest component voltage to the overall cell voltage.

Table II – Cell voltage Components, typical modern refinery*

	Components of Cell Voltage, mV		
	Crop 1	Crop 2	Crop 3
Anode contact voltage drop	10	10	15
Electrolyte voltage drop	220	270	320
Cathode plate internal resistance	25	25	25
Cathode plate contact	25	25	25
Anode overpotential	-340	-340	-340
Cathode overpotential	340	340	340
Total Cell Voltage	280	330	385

* permanent cathodes, 600 Amps/plate (300 A/m^2) electrode pitch 100mm

Electrolyte Resistance

Electrolyte resistance is by far the major component, representing 80-85% of total cell voltage. Therefore small changes electrode geometry that affect inter-electrode gap will have a major impact on the electrode pair resistance and current distribution. Electrode spacing and geometry are the key variables that must be controlled to optimise current distribution. This is particularly true in modern high-intensity refineries that use increasingly thicker anodes and closer electrode pitch. As anode thickness increases, the current distribution becomes increasingly sensitive to variations in inter-electrode gap.

Electrode Contact Resistance

In a modern refinery, the average cathode plate contact voltage accounts for 8-10% of the overall cell voltage. However contact voltage is often highly erratic, depending on the physical condition of the contact surfaces. Field measurements show that individual contact resistance typically ranges from 20-200 µO, equivalent to 5%-25% of total cell voltage. Cell contacts can therefore have a substantial effect on current distribution if not correctly managed.

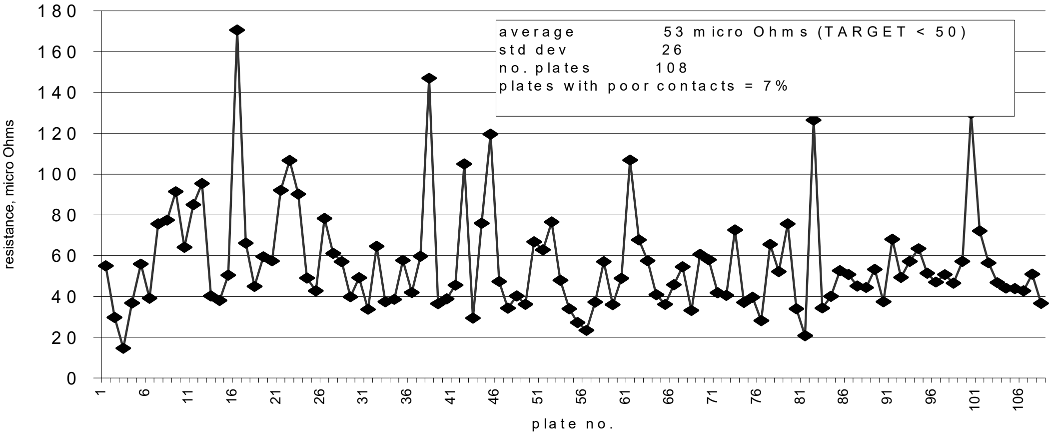


Figure 2 – Cathode Plate Contact Resistance

Electrode Internal Resistance

Cathode plate internal resistance depends upon the plate design characteristics. A typical ISA PROCESS^{TM} plate accounts for around 8% of total cell voltage. However more important is the ability of the cathode plate to maintain low resistance over the duration of its life. Inferior plate designs result in a marked deterioration of electrical properties over time. Therefore plate internal resistance becomes a significant component in the overall cell voltage, and variable plate resistance will impact current distribution.

Anode resistance (both internal and contact resistance) is typically less than 2% of the cell voltage and therefore has negligible effect on current distribution.

OPTIMISATION OF CURRENT DISTRIBUTION

Electrode Alignment

Electrode geometry and alignment have long been recognised as the essential requirements for producing high quality cathode at high current density. With the use of increasingly narrow inter-electrode gaps, small deviations in electrode spacing have a proportionately larger impact on the inter-electrode gap and therefore on current distribution.

The aim of alignment is simple in theory. Anodes are placed at a fixed and uniform pitch in the cells, using the mould face of the anode lugs as a reference. Plates are then interleaved so that the blade is equal-distant from each adjacent anode face.

Alignment practice is carried out either automatically with advanced crane systems, or manually by the tankhouse operators. Both methods are capable of good results when implemented correctly. The main benefit of crane alignment is consistency and repeatability.

Manual alignment techniques commonly employed include the following;

Torching in – The gap between anode and cathode blades is checked visually with the aid of a hand-held light during the anode change, without electrolyte in the cell. This time-consuming method is most useful when anode physical quality is poor.
Visual Spacing – Anodes and cathodes are positioned in relation to reference points on the cell-top furniture (insulators and / or contact bars).
Spacer tools – A hand-held spacer bar is used to re-position the cathode hanger bars to a set distance from the mould face of the anode lugs (equal to the theoretical spacing for nominal anode thickness).

Some modern operations have the capability of automatically aligning the electrodes, such that little or no manual adjustment is necessary. This requires integration of the anode preparation machine, cathode stripping machine and overhead cranes, such that;

- The anode preparation machine and cathode stripping machine deliver electrodes to the crane at precisely the correct pitch. Anodes are positioned via the ‘mould’ side, which has less physical variance than the ‘set’ side.
- The crane is capable of maintaining the electrode pitch during loading / unloading and during transit. The crane hooks must positively locate the cathode plate hanger bars and have minimal free tolerance. Hooks must be robust enough to resist bending.
- Final placement (fine-positioning) of the crane bale on the cells must be highly accurate. Positioning devices used include laser targets and the more positive mechanical systems (cone or pyramid). ‘Stiff-leg’ cranes facilitate location of the bale onto the positioning device.

- The position of the cathode hooks relative to the anode hooks must be adjusted to the correct spacing, and checked by actual observation of the cathode. This action must be precise and repeatable.
- To enable the crane system to function as designed, the cells and cell-top furniture must be positioned accurately and remain fixed in place.

Xstrata's refinery in Townsville has operated two fully automatic cranes since it underwent a major refurbishment in 1998. These cranes are highly reliable and consistently place the electrodes within 2mm of their intended target. The alignment capability of the anode preparation / cathode stripping machine / overhead crane system is checked daily, by placing one set of electrodes in a calibrated portable rack.

Anode Quality

Variable anode geometry has a significant impact on inter-electrode gap and therefore current distribution. To fully realise the benefits of permanent cathode technology, significant improvements to anode quality were needed. Anode geometry had become a limiting factor in refining performance, which led to improved casting practice and better anode preparation.

There is now greater onus on casting operators to deliver anodes of consistent weight that are free from bowing, taper, fins or wash. A five percent variation in anode weight can result in blade thickness variation of 2-3mm. This is significant in high intensity refineries where inter-electrode gap may be less than 20mm on crop 1[1]. Modern weight-controlled casting systems are capable of delivering weight control within 2% of the target.

Key improvements to the anode preparation machines include lug contact milling, face milling and lug centring. More sophisticated machines also measure lug and blade thickness at various points in the press, and reject / accept anodes based on thickness, taper and other dimensional criteria.

Lug face milling and lug centring reduces interference between lugs that would otherwise prevent proper alignment, particularly in high-intensity cells with narrow gaps. These features also facilitate crane handling by ensuring the anode lugs are compatible with the crane hooks. Lug centring also aids manual alignment by allowing operators to more easily judge by eye the correct position of the lug. This can be difficult with off-set lugs.

A further contribution to improved alignment comes from the introduction of narrower cathode plate hanger bars, made possible by the high strength of the stainless steel hanger bars system. ISA PROCESS^{TM} has supplied hanger bars to a width of 25mm in response to customer requirements. The strength of the stainless steel hanger bar

---

[1] Based on ISA PROCESS plant operating with 400kg anode and 95mm pitch

ensures the mechanical properties and geometry of the plates will not be compromised over time.

Operating Cycles

Permanent cathode technology offers greater flexibility with stripping schedules than can be achieved with conventional technology. Anode / cathode cycles can be varied to suit operational requirements. A practice commonly used in high intensity refining is to vary the anode / cathode cycle to optimise current distribution, including;

- Reducing the anode weight and cycle time, to increase average inter-electrode gap. Anode cycles from 14 to 21 days are used amongst ISA PROCESS^{TM} operators. Example – 5d / 5d / 6d (crop 1 / crop 2 / crop 3)
- Shortening crop 1 duration and extending crop 3 duration, to maximise current efficiency (inter-electrode gap increases with crop number). Example – 6d / 7d / 8d
- Shortening crop 3 duration, to minimise poor current distribution arising from light anode scrap / poor anode contacts in latter crop 3. Example – 6d / 8d / 7d
- Converting to a 2 crop operation instead of the traditional 3 crop operation. This reduces anode weight and therefore inter-electrode gap. Example 7d / 7d. (Often two crops are used to achieve other objectives such as increased cathode weight, or reduced workload on the machines – Example 10d / 10d).

Electrode Geometry

The single most significant property of permanent cathode technology is the vastly improved plate geometry. This is particularly well demonstrated by the benchmark ISA PROCESS^{TM} refineries around the world.

Performance of traditional refineries was constrained by the poor cathode geometry inherent with copper starter sheets. This was despite innovations such as embossing, rigidising and pressing of the starter sheets. Permanent cathode technology provided the step-change improvement in cathode geometry that was needed to make high intensity refining possible.

Cathode Plate Verticality

Verticality of the cathode plate is essential for achieving uniform current density over the face of the cathode plate. Non-vertical plates are subject to localised high current density in the bottom portion of the plate, leading to rough growth, increased entrapment of impurities, proximity shorting, and lowering of the effective limiting current density.

In modern refining with narrow inter-electrode gaps, small deviations in verticality can have significant impact on current distribution. A plate that is hanging 6mm off-plumb will raise the current density by a factor of 30% in the lower region of the plate[2]. As the intensity of refining increases, demands on plate verticality become greater.

Manufacturers of quality cathode plates should achieve verticality tolerance at least ± 5.5mm (centre-line deviation from vertical). Operators are demanding even stricter verticality tolerance in some operations.

While construction tolerances are important, the ability of the plate to maintain its geometry in service has a far greater impact on long-term plant performance. Plates must be robust enough to resist bending. The hanger bar system is a critical design feature that imparts overall strength to the plate, and provides rigidity to the blade. Hanger bar systems can be either copper or stainless steel. A copper-plated RHS stainless steel hanger bar, welded to a high chemical and physical quality stainless blade, has proven to produce the most consistent long-term performance.

Proper management of process parameters including electrolyte composition and reagent levels, will preserve the blade surface condition and maintain copper stripability. This in turn minimises mechanical damage during stripping.

Routine checking of plate verticality is also highly important. Non-vertical plates can generally be repaired on site using a simple peening technique.

Anode Verticality

Anode verticality is equally important as cathode verticality. Traditionally anode verticality was often achieved by inserting packing under the lugs to alter the hang of the anode, during torching-in.

Significant improvements were made with the introduction of pressing and contact milling in the anode preparation machines, as highlighted already. Lug pressing should incorporate re-setting of the lugs to the centre of the blade. Measurements have shown (2) that anodes with off-set lugs tend to hang 7-8mm off-plumb, and this can be overcome by centralising the lugs.

Contact milling improves verticality by providing a flat, regular contact surface. Correct maintenance and set up of the milling equipment and cutting heads is critical.

Cathode Plate Flatness

The inherent flatness of stainless steel cathode blades is a key factor in the success of permanent cathode technology. Today’s manufacturers can supply to a flatness tolerance of 3mm. However, the on-going flatness of the plate is more important

---

[2] assuming 17mm inter-electrode gap, crop 1

than the original flatness tolerance. This is a function of blade thickness, hanger bar type and plant operating conditions.

A blade thickness of 3.25mm for most applications has proven to be the most cost-effective, in terms of current efficiency, plate maintenance costs and ultimate service life.

The hanger bar system provides much of the overall strength and rigidity to the overall plate assembly. Alternative hanger bar systems provide varying degrees of strength, however it is generally held true that stainless steel hanger bars provide optimum strength and durability. While solid copper hanger bars are widely used, a long-term bond between the copper bar and the stainless steel blade continues to elude manufacturers.

Plates that are bent or bowed may be tolerated provided they hang within an acceptance envelope (eg 14mm for a 3.25mm plate). The allowable envelope becomes tighter as current density increases or electrode pitch decreases. The decision to straighten a bowed plate should be based on hang-test results rather than absolute blade flatness.

On-going management of cathode plates is the key to their long term performance and extended life. Mechanical damage to the cathode plates can occur during crane handling, in the stripping machines, or through manual handling during repair and manual stripping. Areas that are often problematic include;

- Plates can collide with anodes or cell walls during cell loading, or strike feed conveyors during machine loading. Cranes should have accurate bale positioning, and incorporate sensors, which stop the bale from lowering when ‘plate-high’ is detected.
- Stripping machines must be engineered and set up to eliminate impact points. Automatic hammering of cathode plates to remove difficult-to-strip copper is not recommended by the ISA PROCESS^{TM} as it can stretch and deform the plates.
- Incorrect manual stripping techniques have the potential to cause severe mechanical damage to the plate, which can affect its hanging geometry.

Cathode Plate Contact Resistance

Contact Maintenance

The contact resistance of individual cathode plates typically accounts for between 5% and 25% of the overall electrode pair resistance. The large variation is due to the high sensitivity of plate resistance to the condition of the contact surface.

Uniform contact voltages within a cell are far more important than the absolute value of contact voltages. Average contact voltage impacts power costs, while the variability determines current distribution. Uniform contact voltages are realised by having well maintained cathode plate hanger bar and intermediate busbar contacts.

The stripping machine washing system must incorporate contact cleaning for removal of organics, copper oxide and electrolyte salts. Modern machines have targeted contact cleaning systems using high-pressure hot water. Small quantities of sulphuric acid can be added to the wash water to aid removal of copper oxide and improve contact voltages.

Routine cleaning of the intermediate bars must also be undertaken. Contacts should be cleaned by scrubbing with dilute acid during anode changes, and by hosing during cathode harvests. More frequent wetting of the contact zones should be avoided because copper corrosion will result in variable contact voltages.

To avoid dripping electrolyte on contacts, the cranes can be fitted with drip trays. The trays should be fully engaged prior to travel.

Intermediate Bar Design

An important aspect of intermediate bar design their ease of cleaning. There should be no recesses or crevices that allow electrolyte to pool, or the resulting salts that accumulate will corrode cell-top furniture and hanger bars, leading to poor current distribution.

The conventional busbar system used is the Walker system (3), where electrode pairs within each cell are connected electrically in parallel. Intermediate bars between each cell allow equalization of current passing from one cell to the next. Intermediate bar contacts are generally of dog-bone or triangular profile, which provide a high-pressure point contact when used in conjunction with round-contact hanger bars.

Alternative designs have been developed that are aimed at overcoming the perceived shortfalls of the conventional busbars system.

Wet contact systems (developed Hibi Kyodo Smelting, Japan) was used for many years at Xstrata's refinery and gave clear benefits with regard to current distribution and power costs. However there was also a cost associated with increased corrosion of intermediate bars and hanger bar contacts, and high water inputs to the electrolyte (4).

Double contact systems have been promoted for improved current distribution. All like-electrodes (cathodes to cathodes and anodes to anodes) in each cell are connected via a secondary copper equalizer bar, providing an alternative electrical pathway between electrodes. This system has proven useful where the primary contact is compromised. While conceptually sound, there are issues relating to cleaning and maintaining currently offered systems.

Optibar have developed a Segmented Contact System (5), which is based on a similar principle to the original Whitehead system (6). Each cathode is electrically connected to an anode in the following cell, while each anode / cathode pair is insulated from the other electrodes (no intermediate distributor bar). It is claimed to improve current density dispersion, giving higher current efficiency, better cathode quality and less shorts.

Cathode Plate Internal Resistance

Internal plate resistance normally accounts for 8-10% of the overall cell voltage. The majority of cathode plate resistance occurs within the stainless steel ‘free-board’ zone between electrolyte solution line and the first copper plating. A much smaller resistance exists in the hanger bar itself.

The ISA PROCESS^{TM} electroplated hanger bar design makes use of this property to significantly reduce overall plate resistance, so lowering power consumption. The copper coating extends from the hanger bar, across the welded joint and partially down the blade. This minimises the high resistance path across the stainless steel. The latest ISA Cathode BR^{TM} plate extends the copper depth from 15mm to more than 50mm giving superior electrical performance.

Solid copper hanger bars incur much larger power losses through the high resistance path between solution line and hanger bar. This results in a significantly higher internal resistance than electro-plated designs.

Low internal resistance is important for minimising power consumption, however uniform current distribution requires uniform plate resistance from plate to plate within a cell. Therefore the electrical properties of the plates must be maintained over many years. The predominant cause of diminishing cathode plate electrical performance is the corrosion of the hanger bar to blade brazed joint that occurs in some plate designs. Since corrosion rates vary between plates, this results in variable current distribution within a cell, particularly when new plates are intermixed with older plates.

The electrical performance of electro-plated hanger bars is essentially unaffected by corrosion in refineries. This is evidenced by operations at Brixlegg, Olympic Dam and Copper Refineries, where plates have operated for 15 years to date without significant corrosion.

Measurement of Current Distribution

Routine measurement of current distribution and contact voltages should be undertaken. This provides a valuable measure of the capability of the overall system encompassing anode and cathode geometry, alignment accuracy and contact condition. Plate currents are measured by inserting a DC clamp meter through the lifting window closest to the contact. A scale-up factor is applied to individual measurements, to account for the portion of current not measured at the window (scale up factor equals rectifier current divided by the sum of all measured currents within in a cell).

Cathode plate contact voltages are measured using a millivolt meter, then resistance calculated by Ohm’s Law. Resistance is independent of current density so provides a better performance measure than contact voltage.

The measured plate currents should approximate a normal curve, with standard deviation ideally less than 10% of the average plate current, excluding shorts or open-circuits. Average plate contact resistance should ideally be less than 50 micro Ohms.

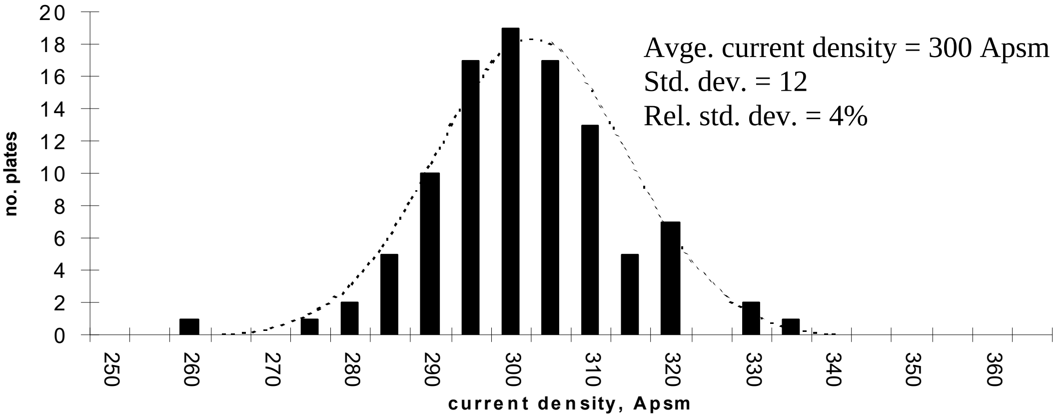


Figure 3 – Measured current distribution data, ISA PROCESS™ tankhouse (7)

Conclusion

The widespread introduction of the ISA PROCESS™ developed permanent stainless steel technology initiated a rapid increase in the intensity of the copper refining process. The superior and predictable verticality of permanent electrodes led to major improvements in current distribution and cathode quality, and increased intensity of operation.

Many refineries have benefited by achieving increased capacity of their plants at lower operating cost. The increased current densities being employed have required further improvements in the current distribution within cells and on cathode surfaces.

Improvements in anode preparation machinery and crane systems in conjunction with permanent stainless steel cathodes have further facilitated the improvement in electrode geometry. However the drive to reduce refining costs by increasing current density has caused greater emphasis on the ancillary components within the system. The need for a complete operating system with a key focus on the maintenance of current distribution is essential for production of high quality cathode demanded by the market place.

REFERENCES

1. J.O'M Bockris, D.M. Drazic, Electrochemical Science, Taylor and Francis Ltd, 1972

2. N.Kimlin, P.Hall, Anode Hanging Test, Internal Memorandum, Copper Refineries Pty Ltd, Box 5484 M.C, Townsville, QLD, Australia, Sept 2001

3. A.L Walker, Plant for the Electrodeposition of Metals, United States Patent No. 687,800, 3 December 1901

4. P.Hall, Dry Contact Trial, Internal Filenote, Copper Refineries Pty Ltd, Box 5484 M.C, Townsville, QLD, Australia, May 1994

5. G.A.Vidal, E.P.Wiechmann, A.J.Pagliero, Technological Improvement in Copper Electrometallurgy: Optibar Segmented Intercell Bars (Patent Pending), University of Concepcion, Faculty of Engineering, P.O.Box 160-C, Concepcion, 8 Region, Chile, Oct 2004, Canadian Metallurgical Quarterly, Vol 44, No.2

6. R.L Whitehead, Electrolytic Apparatus, United States Patent No. 1,206,965, 5 December 1916

7. N.Aslin, Commissioning Report, Copper Refineries Pty Ltd, Townsville, February 2003

Developments in Cathode Stripping Machines - An Integrated Approach for Improved Efficiency

N.J. Aslin, O. Eriksson, G.J. Heferen, G Sue Yek

Xstrata Technology
Hunter Street, Stuart
Townsville, Australia

Abstract

Copper cathode stripping machines and stainless steel cathode plates are critical elements determining the longevity, productivity and efficiency of modern tank houses. In the last decade notable innovations have been the ISA2000 and Kidd robotic cathode stripping machines, and enhanced integration with the overall electrode handling system. The acquisition of Falconbridge by Xstrata in 2005 provided a unique industry opportunity to bring two competing technologies together and has led to the development of a new stripping system which accesses the successful IP, experience and know-how of both ISA and KIDD systems of the past.

This paper presents a new copper cathode stripping system, developed and prototyped by Xstrata Technology in 2009, for the electro refining and electro winning industries. It features the use of robots for both the handling and stripping functions, providing a highly reliable, low-maintenance handling system. The flexibility that comes with robotic operation also offers shorter delivery time, simplified installation and high operational capacity.

The performance, safety and efficiency features of the new Xstrata Technology Robotic Cathode Stripping system are presented, along with an introduction to flow on impacts for other areas of tankhouse design and operation.

Introduction

The adoption of permanent stainless steel cathode electrodes in the electro-refining of copper was the most significant process development since the first commercial electro-refineries were built in the late 19th century. Invented and commercialised by Mount Isa Mines, the ISA PROCESS™ was first implemented into Copper Refineries Limited (CRL), Townsville, in the 1970's. The KIDD PROCESS by Falconbridge evolved in the 1980's and was first implemented into the Kidd Creek Tankhouse, both operations converting from traditional copper starter sheets.

Due to the permanent re-usable nature of the stainless steel cathodes, it has been a critical design requirement of electro-refineries to have an efficient and reliable integrated electrode handling and cathode stripping system. The acquisition of Falconbridge by Xstrata in 2005 has led to the development of a new cathode stripping system which accesses the successful IP, experience and know-how of both the ISA and KIDD systems.

Background

Permanent Cathode plate Technology

Numerous benchmarking studies [1,2] clearly demonstrate the widespread use and operational record of the ISAPROCESS™ and KIDD PROCESS for copper electrorefining and electrowinning. Today more than 11 million tonnes/annum of copper cathode is produced worldwide using ISAPROCESS™ and KIDD PROCESS. The significant features are summarised in terms of;

- Improved copper cathode quality
- Higher operating intensity and current efficiency, giving increased production rate per cell
- Longevity and Reliability
- Improved labour productivity
- Operational flexibility

Although stainless steel permanent cathode plates are critical components in maintaining a productive and efficient tankhouse, just as important are the stripping machines that process the Cathode plates on a daily basis.

Permanent Cathode Stripping Machines

The ISA and KIDD cathode stripping systems have both constantly improved and evolved as a result of the ongoing research and development carried out by MIM and Falconbridge respectively, resulting in several generations of CSM design over the past 30 years. The historical developments of the ISA and KIDD technologies have been well documented [3,4,5], and today there more than 100 installations worldwide.

The method of stripping copper from stainless steel and associated handling of the permanent cathode plates and the copper cathode product all have several options in their design and commercial application. Commercial applications can be grouped by the stripping mechanism, and material handling methods, as shown below in Table 1, summarizing the various stripping systems developed by Xstrata Technology.

Table 1 - Comparison of Layout, Materials Handling and Stripping Mechanism

	Low Capacity (<150 plates / hour)	**Medium-High Capacity (150–600 plates/ hour)**
ISA	➢ Flexor Stripper with Pivot Arm. ➢ ISA 2000 Flexor stripper	➢ Original ISA Machine, traverse conveyor, using wax bottom masking. ➢ ISA 2000
KIDD	➢ Standard Linear KIDD machine.	➢ Original KIDD carousel machine. ➢ KIDD multi-function stripping station using Robotic loading / unloading.

The key difference between the ISA and KIDD stripping mechanism is related to the physical form of the copper product. The ISAPROCESS™ stripping produces a split sheet product, 2 separate sheets of cathode from each Cathode plate, while the KIDD PROCESS stripping produces a V-sheet or “taco” style cathode copper which remains joined along the bottom edge.

Both types of copper product are widely accepted, and user preference often dictates design. For example, the ability to automatically reject only one side of the cathode plate using the ISA stripping system is preferred by many electrowinning operations, whereas the new KIDD stripping system is preferred by some refining operators.

Electrode Handling Systems

Efficient integrated electrode handling is paramount to a productively operating efficient tankhouse. The prime objective of the cathode stripping machine is to safely process the copper as quickly as possible, in order to minimise the downtime associated with cathode stripping operations. In business terms, asset utilisation and productivity comes through continuous plating of copper and this cannot occur unless blank SS Cathode plates are routinely returned to the electrolytic cells. Lost time efficiency from stripping operations generally accounts for around 4-6% in electrorefining operations.

The electrode handling system productivity depends on both mechanical and process factors. This section considers the main process related factors that affect overall cathode stripping machine throughput. Common causes of reject copper in the CSM are briefly given below;

- **Thin copper deposits** are caused by short circuits, limited plating time, poor alignment or poor electrical contacts. Thin copper is difficult to separate from the stainless steel plate due to its lack of rigidity, and generally requires rejecting and manual stripping
- **Sticky copper deposits** are generally related to poor surface condition on the cathode plate, such as corroded surface or improper mechanical treatment. These are also problematic to separate in the flexing station.

- **Heavily nodulated cathode** can often cause stoppages in the cathode stripping machine as the protruding nodules interfere with guides and other parts of the machine. It is generally more efficient to reject these plates than to attempt to strip them through the machine and risk long interruptions to the stripping.
- **Laminated Copper** is a particular issue that occurs with ISA 2000 cathode stripping machines. This results when DC power supply to the electrodes is stopped during the growth cycle, then resumed again, causing a lamination in the cathode deposit. While these events are rare, they can have a significant impact on cathode stripping rate for the affected cathodes.

  ISA 2000 stripping relies on the separation of the 'frangible' portion of copper inside the bottom edge v-groove, see Figure 1. This portion has a natural line of weakness starting from the void inside the v-notch and travelling downwards vertically through the deposit. Laminated cathode has a secondary line of weakness along the lamination which often affects the ease with which the two sheets can be separated. The ISA2000 cathode stripping machine successfully deals with this cathode by detecting splitting failure and automatically carrying out one or several repeat down-ending cycles (referred to as 'flapping'), however does lead to slower overall stripping rates as reported at the Hitachi refinery [6].

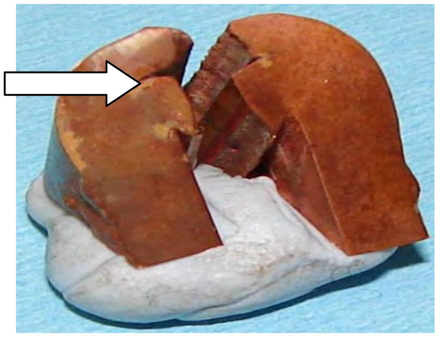


Figure 1 – Laminated Copper

Figure 2 – Magnified view of lamination

- **Strong Envelope c**aused by thick copper deposits or exceptionally hard copper. Causes curving above bottom of the cathode, difficult to separate, difficult to bundle neatly, aesthetics of final product lowered.

In summary, various operating factors can result in longer CSM cycle times, thereby lowering productivity and efficiency. Furthermore, manual handling by operators and other staff is generally required (increasing chances of injury) if separation can not be achieved. Therefore, it is apparent that a stripping machine needs to be designed to handle all types of copper deposit conditions as ideal conditions can not always be achieved in the real world.

Objectives and Methodology

The objective for Xstrata Technology in this development was to create a more accommodating and universal stripping machine, initially focussing on the production of spilt sheet cathode copper product. This new process was created with the goal of bringing the designed stripping rate, effective stripping rate and actual shift rate closer together. In order to achieve this, rather than designing to a specific and strict set of copper conditions, the new process and machine were designed to flex and strip copper regardless of how variable the deposit. Therefore, improving the reliability of the machine and meeting scheduled production rates independent of process variables.

The Xstrata Technology core designs principles were maintained, namely:

Safety - specifically the controlled handling of the copper cathode sheets during stripping operation, Reliability, Ease of Operation, Noise Reduction, Minimise potential for damage to the SS Cathode plates, Flexibility to handle all types of copper deposit and cathode plates, Maintainability and Sustainability across initial capital costs and investments to product and materials life cycle.

The design for the latest concept came as a result of a methodical and fresh approach adopted from the following:

1. Research and development conducted on several other successful Xstrata Technology designed, implemented and operating cathode stripping machines / processes.
2. Experience from numerous ISA/KIDD licensed refineries around the world.

Strengths and weakness from all machines were drawn upon when formulating the new Xstrata stripping concept. Standard design process and stages were then implemented.

Current CSM Technology

A number of machines were examined and tested in the creation of the latest robotic cathode stripping machine. A summary of operational backgrounds and key features from the existing designs that were critical to the development of the latest Xstrata Technology stripping system are presented below:

ISA 2000 CSM

The ISA 2000 Waxless machine uses hydraulics to drive a set of stripping knives downwards in between the copper cathode and stainless steel blade. This machine opens the copper envelope approximately 30 degrees each side (60 degrees total). The grippers then engage onto the copper sheets and the down enders rotate downwards about the bottom edge of the cathode until the copper sheets are taken just past horizontal position (approximately open to 195 degree angle). If separation does not occur, the downenders go up and down repeatedly (flapping) until the copper fatigues, cracks and separation occurs. The copper is then dropped onto a transfer conveyor where it is taken to be sampled, weighed, labelled, bundled and strapped.

The ISA 2000 CSM’s have been operating successfully in refineries and electrowinning operations since 2001. They provide full control of copper sheets, produce a split sheet copper cathode product bundle and require minimal operator intervention during stripping.

Figure 3 – ISA 2000 Waxless machine stripping station

The need for improvement on this machine comes as a result of being originally designed to a strict set of “normal operating condition” copper deposits. Where there are variations in the operating conditions, the affects may result in increased cycle times. In addition, the force application position of the down ender grippers affects the separation reliability, especially when the copper is laminated or has a strong envelope.

ISA 2000 Flexor Stripper

This stripping system is a development based on the proven low capacity flexor stripper system patented by Xstrata Technology for use in electrowinning operations to allow operation without bottom edge strips or wax. The ISA 2000 Flexor Stripper uses a similar stripping function to that of the high capacity ISA 2000 Waxless machine.

The hydraulically driven knives on this machine extended the full width of the plate. This allowed the support and application of the stripping force across the full width of the stainless steel blade and copper respectively. It also worked well in helping maintain a straight copper cathode in addition to aiding the separation of the cathode in the corners where it is generally more difficult to separate as shown in Figure 4and Figure 5 below.

Figure 4 – Thin Wedge, Curved copper due to separation issues on edges due to faulty edge strips

Figure 5 – Full Width Wedge, uniform copper separation

The first set of knives typically opened and split the copper. The secondary set of knives was used only when the bottom bond was strong or separation didn’t initially occur (by opening the copper to the horizontal position). If separation still did not occur, flapping arms raised the copper back up to approximately 45 degree opening (90 degree open envelope) and then the knives were used again to bring it back down. Both sets of knives were attached to a “cassette” allowing the first set of knives to get closer to the bottom edge of the cathode (pivot axis of the copper envelope). The cassette moved up and down vertically and also had the flexing station attached.

This design allowed a full function machine to be compressed into the size of a single station making it ideal for space or capital limited operations, although cycle time is extended.

Critical learning’s during testwork included the difficulty in flapping using the two separate mechanical devices (knives and flapping arms) as they had to be well co-ordinated. It was also observed that very large forces were required to strip the copper due to the linear movements and no mechanical advantage or leverage in the mechanical devices. However, this was easily overcome through the use of hydraulics.

Successful implementation and operation of this stripping system has been applied at the Tenke Fungurume electro-winning operation, with several more systems due for commissioning in developing projects. It is best suited for low production applications.

Figure 6 – ISA 2000 Flexor Stripper

Roller Stripper

The Roller Stripper CSM was developed and commissioned on site at Townsville copper refineries in 2008. This machine was primarily designed for research and test purposes with some potential for retrofitting to existing ISA 2000 machines and greenfield installations.

The prototype machine employed a new concept in copper cathode stripping (patented by Xstrata Technology). The objectives of this design and development were to:

1. Reduce friction
2. Eliminate Plate damage
3. Maintain a flatter copper cathode

A three roller combination was formulated and uses similar motions to that of a normal wedge. The rollers extend the full length of the plate providing maximum support to the plate and stripping force to the copper. As shown in Figure 7 below, the top roller supports the plate and provides a reaction force to help separate the copper from the stainless steel blade. The middle roller supports and applies the stripping force to the copper cathode (this is normal to the roller and perpendicular to the face of the copper cathode). As the copper rotates towards horizontal, the force applied by roller 2 becomes vertical. The bottom roller is used when separation does not occur first time and provides support to the copper when bringing the rollers back up to the start position. The cycle is then repeated.

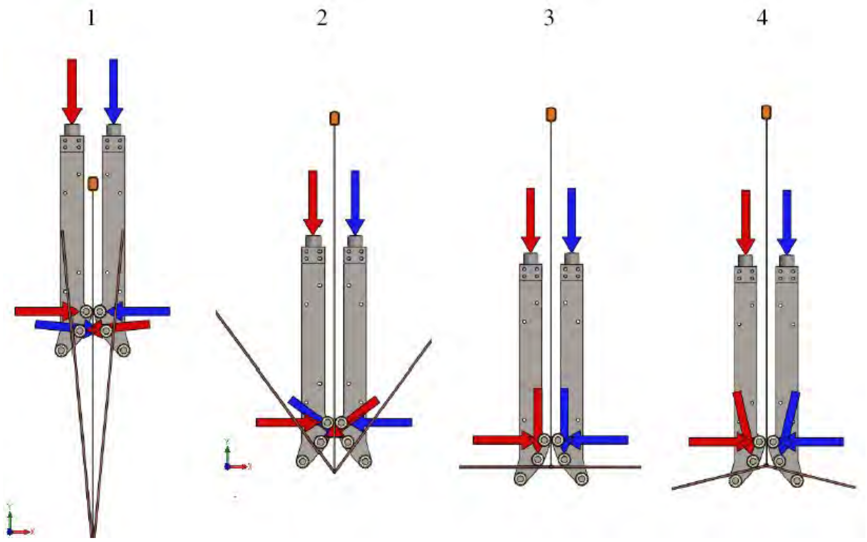


Figure 7 – ISA 2000 Roller Stripper

During testing it was observed that this machine was far superior in separating the copper sheets when compared to the previous two ISA 2000 machines. Test work demonstrated that in addition to supporting the copper across the full width, the rollers applied the force very close to the envelope pivot axis and concentrated all the fatiguing of the copper into the correct separation area.

This study gave an understanding of the critical force values which allowed the continuation and development of the new concept and confirming the possible use of robotics [7].

Other observations included:

- The ability to open the copper cathode past horizontal also aided in the fatiguing and crack propagation when trying to separate the copper. When the rollers moved upwards again, the force was still applied in the same position where as in other machines it can be dissipated and translates into bending of the copper cathode.
- The rollers were extremely quiet and left no plate damage.
- Maintains a very flat copper cathode sheet

The New Concept

It was apparent from the previous research and test work that the point of force application on the copper cathode had a substantial impact of the separation reliability. This was the main driving factor in the design and development of the new CSM process concept.

The following design parameters for the new concept were set based on the test work and operational experience from both the ISA and KIDD stripping systems:

- Small opening angle - Initially open the copper envelope as minimal as possible to prevent any bending of the copper in the sometimes weaker section just above the bottom edge of the envelope.
- Full width mechanism - Support the full width of plate (eliminate risk of plate damage) and copper (maintain straight flat copper cathode)
- Maintain or Increase Speed – Maximum speed to ensure fast cycle times
- Reliability – be able to strip the complete range of deposited copper from normal operating conditions to laminated
- Reduce Friction – to provide an efficient and lower maintenance machine

The following prototype was designed and constructed for initial testing on site at CRL. It comprised of four vertical members profiled in the shape of a wedge, it was designed to support the entire face of the copper cathode when rotating or stripping (Figure 8). The sharp angle allowed the wedge to reach the bottom whilst maintaining a minimal opening angle on the way down to prevent the copper bending as mentioned previously. The arms on either end locked in to a horizontal shaft to help keep the wedge in position during rotation, ensuring that the wedge made it to the bottom and guarantee that all forces were applied into the correct position on the copper cathode during separation.

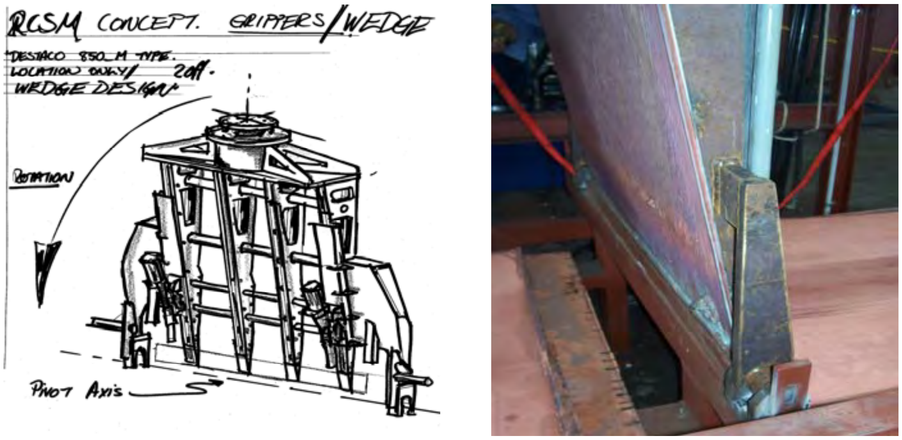


Figure 8 –Concept and prototype design

Some of the challenges in the design and development stages included:

- Forces involved with reaching the bottom

  - Initial test results from the roller stripper and prototype wedge suggested that the forces required to get to the bottom of the cathode plate might have been above the capabilities of the available robots. After further testing and an investigation by John Hart, it was found that with the recent improvements and advancements in robots, they would be able to handle to task with slight modifications being made to the process and program of the robots to aid and ensure they would reach the desired point.

- Process difficulties (the range of movements required)
  - The new concept could be applied to either hydraulically driven mechanical components or robotics. However, due to the complex range of movements required by the concept process, it was thought that it was not practical, economical or desirable to construct using hydraulics.

The New Xstrata Technology Robotic CSM

Xstrata Technology has now taken the new concept through a complete design process and obtained a patent on the new cathode stripping process and design. Mechanical components driven by electrical devices, pneumatics or hydraulics could be used, however XT's determined that robotics was a far superior choice due to the many inherent advantages, namely:

- 80,000 hrs mean time between failure (MTBF) ~ 18 years at 12 hours per day
- High accuracy and precise movements - protect the SS cathode plate
- Proven capacities 900 kg and greater –capable of performing stripping function
- Flexibility – can be programmed and or easily changed to suit a wide variety of plates, copper conditions and equipment layouts
- Installation and commissioning time – very easy program modification and installation
- Excellent safety standards – program incorporated safety features + complete operator control with reduced or eliminated manual handling.
- Low noise and clean - no hydraulic fluids

The choice of robotics itself was innovative as this was the first time robotics were used in a copper tankhouse to perform a process or task other than simple pick and place materials handling.

Tooling Design

The design of the new robotic tooling (Stripping wedge) was the next critical stage and consisted of the following features:

1: Guides – implemented to support the copper during the downwards vertical motion, ensuring the copper does not pre-strip. In turn making the new process safer as there is complete control over the copper cathode at all times.

2: Grippers – used to clamp the copper before the rotational down ending starts, again, giving the robot and the tool complete control over the copper cathode at all times. In addition, this prevents slippage and concentrates all the forces on the void or weak spot in the bottom edge of the copper envelope.

3: Rollers – designed to reduce the friction between the copper, stainless and the wedge during the downwards vertical motion. This lowers the force required to get the wedge to the bottom whilst minimising or eliminating damage to the plate, wedge and copper. Finally, with the reduction of friction there is a reduction in noise and wear on respective components.

4: Wedge – the current prototype wedge was constructed from 350 grade mild steel and 316 grade stainless steel. Optimisation of materials of construction is ongoing.

    A. Wedge Frame – consisted of three vertical members for strength and rigidity whilst also providing attachment points for the guides, grippers and other equipment.

    B. Wedge Blade – featured a bevelled tip to ensure scratching or damage is kept to a minimum. This also allowed for maximum support of the copper cathode.

5: Pneumatic components – pneumatic components were used to maintain the goal of eliminating hydraulics. In order to reduce maintenance and maintain a cleaner work environment. Further testing is currently being performed to determine if the pneumatics can produce the required strength and forces needed to carry out the stripping process across all ranges of copper deposit and production hours.

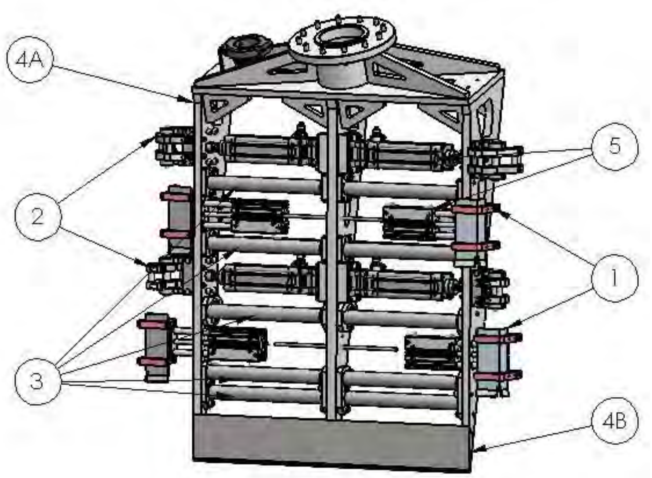


Figure 9 – ISA 2000 Stripping Wedge

Prototype Installation

On site at the Xstrata Townsville copper refinery a prototype robot cathode stripping machine was retrofitted into the original ISA 2000 waxless test stripping machine. As shown in Figure 10 and Figure 11 below, the robots greatly open up the stripping area allowing easy access for operators, maintenance, forklifts and other equipment.

Figure 10 – Before

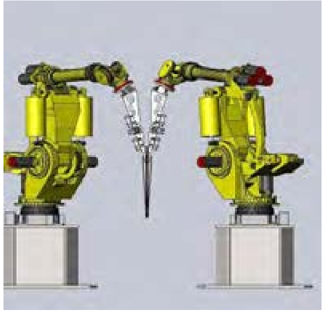
Figure 11 – After

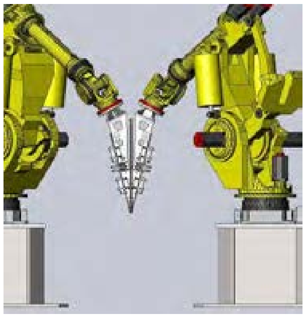
Stripping Process

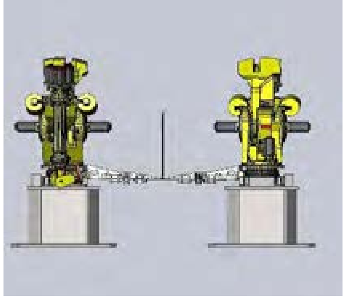
The process in which the wedge works is similar to the original ISA 2000 waxless stripping machines. However, in this case the knife, grippers and transport conveyor are all in the one apparatus. Figure 12 A to C provides a snapshot view of the movements of the new Robotic stripping wedge and its flexibility.

The permanent stainless steel cathode is presented in the stripping station already flexed and pre-opened, the robots then bring the knives in tip first until the blade is approximately 50mm past the top of the copper. At this point, the wedges start rolling to a vertical position whilst they are travelling in a downwards motion (the tips of the knives only mm away from the blade on each side). Once the wedge is down far enough to clear the hanger bar (shown by recess in wedge profile) the wedge is completely vertical and the rollers are touching the stainless steel blade keeping the knife edge away and unable to cause damage.

Figure 12B shows each of the wedges in the bottom position just before down ending. At this point the grippers are activated and the copper becomes clamped. Both wedges roll to horizontal about an imaginary axis which is perfectly aligned with the void in the copper envelope cause by the v-groove. Once horizontal, the wedges rise slightly and then pull away from each other separating the copper and transporting it away to a bundle, transfer conveyor, weighing device, reject bundle, sampling device or any other station the customer desires.







A B C

Figure 12 – Robotic Cathode Stripping sequence

Performance and Results

Testwork of separation efficiency to date has been outstanding. No flapping has been observed in the Robotic Stripping system when processing standard V- grooved stainless cathode plates, with laminated copper cathode. When processing stainless cathode plates prepared with significantly reduced V-groove condition and laminated copper deposit, a maximum 3 flapping motions were required.

If the copper did not separate on the first attempt, the robots “flap” the copper in a specific program sequence. Importantly, there is no bending in the copper and all the forces are concentrated in the same location leading to the vastly reduced flapping frequency.

With increased separation reliability, the average cycle times gets closer to the designed cycle times providing operators with satisfaction and guarantee that they will meet throughput and production requirements irrespective of copper deposit conditions.

Testwork on the Robotic Stripping prototype is ongoing, and in parallel it is used by the Xstrata Copper operating refining to process stainless cathode plates rejected from the commercial stripping machines.

Flexibility and Safety

The robots can be programmed to handle a variety of different situations minimizing manual input. They have the ability to perform a number of different processes (initiated by the push of a button) throughout the stripping cycle dependent on the copper condition and operator’s desires.

The robots also have the automatically or by operator controls to safely remove the copper with out any manual handling of the copper or entry into the stripping area. This feature greatly improves operator safety. Furthermore, if the copper sheet is to be rejected for quality reasons, whether it be one side or both, the robots have the ability to transfer the single sheets to a separate reject bundle or any other customer desired location with in the robots reach. This feature, in combination with online cathode quality scanning devices, would provide significant materials handling benefits to the operation.

Applications / Implementation

The new Xstrata Technology Cathode Robotic Stripping system is extremely flexible and is suitable for both green field and retrofit applications. It would be very well suited to implementation in:

- Plants with variable copper deposits which may be difficult to strip
- Plants where manual sorting of reject cathode copper occurs.
- Refineries with un-reliable power supply or regular power outages during plating)
- Smaller scale production operations
- Tankhouses where time efficiency is extremely important and critical to project continuation
- Conversion of older plants from wax to waxless

Ongoing Development

Xstrata Technology understands the features of the new stripping process and robotic control provide potential for advantages in other areas of the tankhouse design and operation to gain better productivity and efficiency. Scope will include optimising the permanent cathode plate design due to more precise and flexible control when using robotics to achieve reduction in operating and capital cost. In conjunction, there is potential for further improvements to automation of crane lifting and handling systems.

Xstrata Technology continues to develop these concepts as part of its’ ongoing research and development programs

Summary

The Xstrata Technology research and development program achieved the objective to create a more accommodating and universal stripping machine, for the production of spilt sheet cathode copper product. Implementation and installation of the next generation Xstrata Technology Robotic Cathode Stripping system has the following advantages:

- Reduced capital costs (reduction in conveyors and transfer equipment)
- Small foot print especially for stand alone units
- Can retrofit into existing machines with small modifications
- Easy to install and commission
- Provides operations currently producing split sheet ISA product bundles increased flexibility in the sequencing of normal harvesting operations and during smelter shutdown periods.
- Reduced noise (elimination of hydraulics, reduced impact noise from copper sheets)
- Increased time efficiency

- Flexibility with copper transport and rejection

Acknowledgments

We would like to thank all the staff involved from the Xstrata Copper, Townsville Refinery for their assistance during the testwork at CRL and recognise the extensive input by Jason Schulte, Blair Warry and Tony Ruddell from Xstrata Technology during the development process.

References

[1] Robinson, T., Davenport, W.G., Moats, M., Karcas, G. & Demetrio, S. (2007): Electrolytic copper electrowinning – 2007 world tankhouse operating data – Cu 2007, Volume V- Copper Electrorefining and Electrowinning, Eds. Houlachi, G.E., Edwards, J.D., & Robinson, T.G., Toronto, Canada, The Metallurgical Society of CIM.

[2] Moats, M., Robinson, T., Davenport, W.G., Karcas, G. & Demetrio S. (2007): Electrolytic copper refining – 2007 world tankhouse operating data – Cu 2007, Volume V- Copper Electrorefining and Electrowinning, Eds. Houlachi, G.E., Edwards, J.D., & Robinson, T.G., Toronto, Canada, The Metallurgical Society of CIM.

[3] Armstrong, W. (1999): The ISA PROCESS and its contribution to Electrolytic Copper – Rautomead Conference, Scotland.

[4] Okane, J. (1997): ISAPROCESS – Benchmarking Copper Refining Electrolysis – GDMB 33rd Metallurgical Seminar, Lunen, Germany.

[5] Donaldson, P.E. & Detulleo, J.J. (2003): Falconbridge's Kidd Copper Refinery Birthplace of the Kidd Process. An Update on the Refinery and Latest Developments – Copper 2003-Cobre 2003, Volume V- Copper Electrorefining and Electrowinning, Eds. Dutrizac, J.E. & Clement, C.G. Santiago, Chile, The Metallurgical Society of CIM.

[6] Matsuda, M., Goda, T., Takebayashi, K. & Maeda, Y. (2007): Recent Improvements at the Hitachi Refinery – Cu 2007, Volume V- Copper Electrorefining and Electrowinning, Eds. Houlachi, G.E., Edwards, J.D., & Robinson, T.G., Toronto, Canada, The Metallurgical Society of CIM.

[7] Sue Yek, G. (2008): Investigation of the Mechanical Forces of the Xstrata Technology Copper Cathode Roller Stripper – BE (MECH) Thesis, James Cook University, Townsville, Australia.

DEVELOPMENTS IN PERMANENT STAINLESS STEEL CATHODES WITHIN THE COPPER INDUSTRY

K.L. Eastwood and G.W. Whebell
Xstrata Technology
Hunter Street
Townsville, Australia 4811
keastwood@xstratatech.com.au

ABSTRACT

The ISA PROCESS™ cathode plate is characterised by its copper coated suspension bar, coupled with a blade employing austenitic stainless steel alloy 316L. The blade material has become the mainstay of the technology and has been closely copied by competing cathode designs. Improvement to the cathode plate design remains a key area for research, and ongoing developments by Xstrata Technology's ISA PROCESS™ have recently been commercialised. Two such developments are the ISA Cathode BR™ and ISA 2000 AB Cathode. The ISA BR cathode is a lower resistance cathode that has proven to enhance operating efficiencies. The AB cathode was designed to improve stripping inefficiencies in the ISA 2000 technology. These developments have now had time to mature and their long term performance will be discussed. Rising material costs and the desire to extend the operating boundaries of the standard 316L cathode plate has triggered a number of significant advances. These involve the use of different stainless steels as alternatives in some operational situations. The technical aspects and results of commercial trials on this development will also be discussed in this paper.

INTRODUCTION

The introduction of permanent stainless steel cathode technology was pioneered in the copper industry by IJ Perry and colleagues in 1978, with the introduction of the ISA PROCESS^{TM} in the Townsville Copper Refinery, Perry [1]. While a number of parallel processes have emerged since its introduction, ISA Process Technology (IPT) has continued to be the mainstay electrolytic copper process with consistent improvements and superior operational performance.

Following the introduction of the ISA 2000, recent developments that have been offered to the market are the ISA BR and AB cathode plate. The BR plate offers a low resistance electrode that has the potential to significantly reduce power costs. The AB cathode was introduced to the market to improve stripping performances particularly where tankhouses with the ISA 2000 technology were prone to frequent power outages.

A new development soon to be introduced into the market is the use of alternative steels for cathode plates. Rising costs and the desire to extend the operating boundaries of the 316L alloy triggered investigations into alternate stainless steel types and their application to various copper electrolytic processes. The steels tested have the potential to reduce the capital cost of the cathodes while still providing a technically sound cathode. Tests have been carried out in diverse operating environments from small low-cost electrowinning operations to intensively managed electro-refining plants.

ISA CATHODE BR^{TM} – "LOW RESISTANCE" PERMANENT CATHODE

BR Design

The BR cathode design extends the copper plating up to 55mm down the blade, compared to the standard ISA plate of 15-17mm. The extension of the copper plating reduces the amount of electrical resistance that exists between the copper plating and the solution line. This is achieved by reducing the distance the current has to travel through the stainless steel. The resistance of stainless steel is 74 μΩ.cm @ 50^{0}C compared to a value of 1.8 μΩ.cm @ 50^{0}C for copper.

The BR design can increase the average copper plating thickness from 2.5mm to 3.0mm. The increase in copper thickness increases the corrosion resistance of the plate. This is particularly advantageous in electrowinning applications where cathodes may be subjected to corrosive conditions.

Development of the BR cathode

Standard ISA cathodes, whilst superior in most aspects to alternate stainless steel cathode configurations have slightly higher resistance than solid copper hanger bar systems. While the lower resistance is attractive, deficiencies for some applications in the long term performance of the solid copper hanger bar were identified following a large scale trial at the Townsville Copper Refinery. Webb [2]

The initial development of the BR cathode, as reported by Webb [2], followed trials conducted to compare the performance of solid copper hanger bar cathodes and ISA PROCESS^{TM} cathodes. These trails indicated that whilst ISA PROCESS^{TM} cathodes gave higher current efficiencies of 2% to 2.4% (electro-winning and electro-refining respectively) the continuing market demand to produce copper more efficiently lead to the development of the lower resistance ISA cathode BR^{TM}.

Figure 1 – ISA cathode BR^{TM}

Additional Testing of the BR Cathode

In addition to the early work completed by Webb [2] the internal plate resistance of various cathode plates have been measured using a digital micro-ohmmeter at Townsville's Copper Refinery. To replicate cell conditions, cathode plates were plated with copper for a period of 24 hours then the resistance measured. The average plate resistances are shown in the graph below.

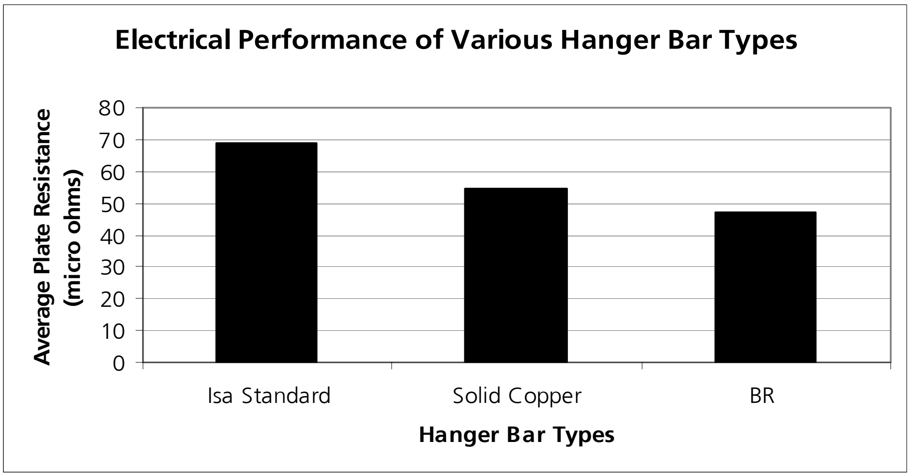


Figure 2 – Internal resistance of various hanger bar types

The measurements complemented Webb's [2] original work by confirming the ISA Cathode BR^{TM} as the electrically superior plate, by exhibiting the lowest plate resistance.

Further operational measurements have been completed at other ISA PROCESS^{TM} electrowinning facilities. All measurements confirm the previous work completed by Webb [2] at Compania Mineral Zaldivar showing a potential saving of US$100,000 per year in power costs could be achieved with the ISA Cathode BR^{TM}. This equates to a cost saving of approximately US$0.75 per tonne of copper produced.

Since the ISA BR^{TM} cathodes introduction in the early 2000's, in excess of 106,000 BR configured plates have been produced, representing approximately 450,000 tonnes of copper per year. These plates have been installed in 6 plants in Japan, USA, Philippines and South America.

THE DEVELOPMENT OF THE AB CATHODE

The ISA 2000 Waxless System

To eliminate the need for wax as a bottom and side masking agent ISA PROCESS^{TM} developed a waxless cathode design, now commonly known as the ISA 2000 technology.

The principle behind the waxless development lies within the 90^{o} v-groove machined into it. This allows separation of the enveloped cathode into two separate sheets by the ISA PROCESS^{TM} stripping machines. On a micro-scale, the copper crystals grow perpendicular to the cathode plate from opposing sides of the v-groove, causing them to intersect at right angles to each other. Where they intersect, a discontinuity in the structure is formed, resulting in weak zone, along which the copper splits.

Figure 3 details a microscope view of a copper cathode cross-section, taken from the v-groove region, magnified 10 times. The sample was polished and etched to show the copper crystal structure. The white lines indicate the orientation and direction of crystal growth.

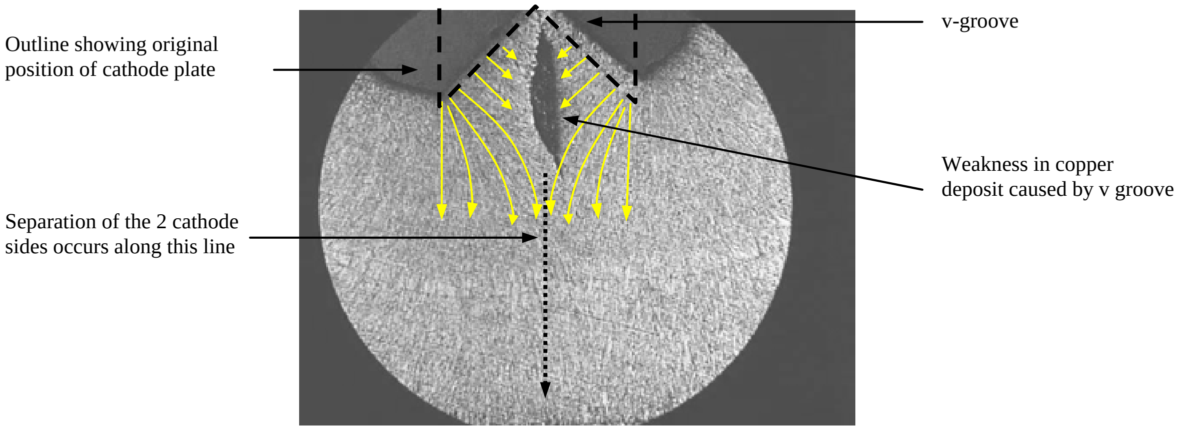


Figure 3 - Microscope view of cathode deposit cross-section in the v-groove region.

In addition to this effect it is believed in electrowinning facilities small gas bubbles (oxygen from the anodes) become trapped in the groove. This gas layer then serves as an insulating layer between the two growing faces.

The v-groove splitting mechanism combined with modifications in the stripping machines has been extremely effective in the separation of copper cathodes under most circumstances.

The AB Design

The AB cathode design has 45 degrees chamfered corners cut away from the bottom of the blade. The "v" groove runs the length of the blade and up the chamfered corners. The dimensions of the chamfered corners are 60mm x 60mm. These dimensions have been chosen to maximise the tearing action, whilst reducing the chances of bending the cathode corners or producing a ropey edge that would potentially entrap falling slimes. The tearing action produced by the corner chamfers initiates the split of the copper sheets and improves splitting if lamination has occurred.

The edgestrips designed for the AB cathode are a cross slot edgestrip with a moulded bottom plug. The internal dimensions of the plug have been specifically designed for the AB cathode and allow the edgestrip to fit securely onto the AB plate to prevent nodule growth.

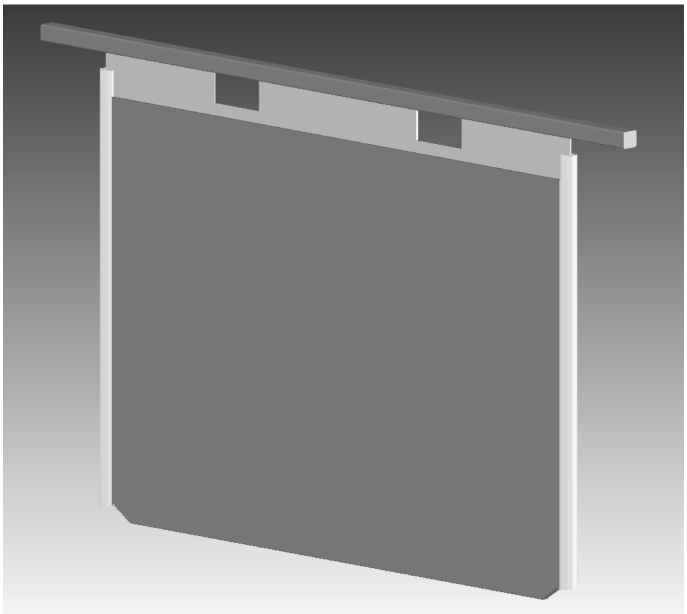


Figure 4 – AB Cathode

The AB concept was initially developed for electrorefining to:

- Reduce stripping inefficiencies if tankhouses were prone to power outages.
- Improve the life of an ISA 2000 edgestrip.
- Prevent end cap damage.

Power outages alter the initial growth pattern of the copper. When the power is resumed the new growth pattern of the crystals laminates the original copper growth. The resulting lamination can null the effect of the v groove. This can increase stripping times as cathode sheets may require additional movements in the stripping machine to split the enveloped copper. The effect power outages have on the cathodes stripping performance is dependant on the duration, the time in the cycle and the anode overlap. The AB cathode design aimed to reduce the stripping inefficiencies caused by power outages.

In the ISA 2000 cathode plate the bottom corners have square notches of steel removed. The edgestrips have moulded plastic end-plugs, which fit neatly into the voids formed by the notches in the plate. This system eliminates copper nodule growth from the bottom corners. After the early commercialisation of the ISA 2000 technology there were a large number of end cap failures as well as a reduction in the edgestrip life in some refineries. In the standard wax technology the edgestrips do not contact the traverse conveyor and therefore less stress is placed on the edgestrips. The ISA 2000 edgestrips are in direct contact with the traverse conveyor potentially increasing the stresses on the edgestrips. In the AB cathode design the edgestrips do not contact the traverse conveyor.

Experimental Testwork

IPT developed a number of prototypes of the AB cathode based on a 60mm x 60mm chamfer. The results were positive with the corner chamfer reducing stripping inefficiencies when power outages caused lamination of the copper.

At the Townsville Copper Refineries, a cell of 44 AB cathodes were installed and stripped weekly. They were monitored for stripping difficulties, defined by additional movements required by the down ending station to ultimately remove the copper cathode. These additional movements were commonly known as “flaps”. The results from these trials are summarised in Tables 1 and 2.

Table 1 - Stripping data under standard test conditions.
No power outages included.

	No. flaps / 100 strips
AB Cathode	3.3
Standard Cathode	4.2

Table 2 - Stripping data when severe laminations have occurred
during the plating cycle.

	No. flaps / 100 strips
AB Cathode	17.3
Standard Cathode	38.5

Under normal operating conditions the standard ISA 2000 cathode plates effectively separated the copper sheets and therefore the improvements were minimal.

The AB cathode improved stripping performance significantly when power outages had laminated the enveloped copper sheets. The additional "flaps" were reduced by half with an AB cathode.

The trials were then extended to include corners of varying shapes and sizes. Smaller corner cut outs as well as rounded corners were trialled. Results indicated that corner chamfers smaller than 40mm x 40 mm and rounded corners performed poorer than a standard v grooved ISA cathode under normal operating conditions. The larger chamfered corners were found to aid stripping performance however were prone to bending over time.

The AB cathode design eliminates the plastic edgestrip contacting the traverse conveyor, reducing fatigue cracking and potentially improving the life cycle of the edgestrips. AB cathodes in service in the Townsville Copper Refinery with an end cap design have been in service for 1.5 years without failure.

As the edgestrips no longer contact the base of the traverse conveyor the v groove is in direct contact with the traverse. Accelerated tests were completed to identify whether any deterioration in the v groove occurred.

An AB plate was lifted and then lowered on to the traverse conveyor a total of 728 times, equivalent to 14 years in service. The impact points on the plate were marked and the measurements were taken from these points through out the test. It was then placed back into service and the plate's performance monitored. Table 3 represents the data collected.

Table 3 - Results from v groove deterioration testing.

No of lowers	Service time	v groove measurements 3 points			Comments
0	0	1.38	1.51	1.52	
208	4 years	1.37	1.37	1.46	slight curling R/H and Mid
312	6 years	1.35	1.32	1.42	slight curling, small chips
468	9 years	1.3	1.29	1.28	curling of contact point edges
728	14 years	1.29	1.34	1.25	no observable change

After the equivalent of 14 years in service the v groove measurements decreased from 1.52 to 1.25 in this area. The tests were undertaken in the ISA PROCESS^{TM} waxless demonstration machine. This machine, unlike production machines does not have a proportional speed control valve to allow the plate to be lowered gently onto the conveyor or soft metal pads for the plate to sit on. The plate descends rapidly onto the steel conveyor therefore making the action far more severe than a production machine. The plate continued to strip successfully after the testwork. This test was completed on several occasions with similar results.

Recently commissioned with the ISA 2000 AB cathode was Sumitomo's Toyo Refinery in Japan. Approximately 22,000 plates have been manufactured and installed and are reported to be working efficiently. The edgestrip life is also being monitored.

ALTERNATIVE LOWER COST CATHODE PLATE

Current ISA blade material

The standard ISA blade is made of austenitic stainless steel, of grade 316L. It specifically contains 2-3% molybdenum for increased resistance to pitting, and has a low carbon content (0.03%) to minimise chromium carbide precipitation, or sensitisation of welded zones. This decreases the tendency for intergranular corrosion. The typical analysis of 316L is shown in Table 4, [3].

Table 4 – Blade Composition

Element	Composition %
Cr	16-18%
Ni	10-14%
Mo	2.0-3.0%
C	≤ 0.03%
Mn	≤ 2.0%
N	0.04%

The blade is 3.25 mm thick. Based on experience, this is the optimum thickness taking into consideration the plate performance, ease of manufacture and cost-effectiveness. The surface finish is a standard 2B manufacturing finish with a specific surface roughness in the range of 0.25 to 0.6 microns Ra. The high corrosion resistance of the cathode plate is provided by a very thin, tenacious, self repairing layer of chromium oxide, which forms a passive film on the blade, [3].

The development of a lower cost cathode

While the 316L alloy has been the mainstay of the technology since its development in 1978, rising costs and operational considerations in some plants have triggered the ISA PROCESS^{TM} to examine the use of alternative stainless steels in the copper electrorefining (E/R) and electrowinning (E/W) industries.

Increasing costs of metal prices, namely nickel and molybdenum continue to push 316L prices higher and therefore the need for alternative steels suitable for use in the industry was desirable.

The life cycle cost of a cathode was also a consideration in the development of an alternative cathode blade material. The ISA cathodes were developed for a 10-15 year life cycle. Some operations however have a considerably shorter life span and a more economic plate with a life cycle matching the project life would be an advantage. In addition, the extreme operating environments in some plants require plates to be replaced or repaired after 3-4 years.

The focal point of the development work has been on testing a relatively new type of duplex steel, LDX 2101 and the standard 304L stainless steel.

LDX 2101, DUPLEX STAINLESS STEEL

LDX 2101 is a low-alloyed, duplex stainless steel. Its microstructure contains approximately equal amounts of ferrite and austenite. A typical analysis of the duplex steel is Table 5, [4]

Table 5 – Blade composition

Element	Composition %
Cr	21.5%
Ni	1.5%
Mo	0.3%
C	≤ 0.03%
Mn	5%
N	0.22%

LDX 2101 has high mechanical strength due to its duplex microstructure and high nitrogen content. The corrosion resistance is in general at least as good as that of Cr– Ni grades such as 304L and in some cases as good as Cr-Ni-Mo grades such as 316, [4]. The reduction of molybdenum and nickel content reduce the overall price of the steel.

Experimental Testwork

Chemical Resistance Testing

A series of corrosion tests were completed by the Avesta Development Group. LDX 2101, in both a linished and unlinished surface finish was compared against the standard 316L steel and 304L stainless steel. The test parameters are represented in Table 6 and 7.

Table 6 – Test solutions for immersion tests

		Base	1	2	3	4	5	6	7
Cu	g/L	45	45	45	45	45	45	45	45
H_2SO_4	g/L	160	**175**	**195**	**235**	160	160	160	160
Chloride	ppm	40	40	40	40	**55**	**70**	**100**	**120**
Temp	^oC	50	50	50	50	50	50	50	50
Thiorea	ppm	2	2	2	2	2	2	2	2

Table 7 – Test solutions for immersion tests

		8	9	10	11	12	13	14
Cu	g/L	45	45	45	45	45	**15**	45
H_2SO_4	g/L	160	160	160	**175**	**195**	140	195
Chloride	ppm	40	40	40	**70**	**100**	40	100
Temp	^oC	**60**	**68**	**72**	**60**	**70**	50	70
Thiorea	ppm	2	2	2	-	-	-	-
$Fe_2(SO_4)_3$	g/L							**6**

The bold values represent the parameters that were altered from the base line test. The test steels were immersed for 30 days in all tests. Dickson [5] has shown “no uniform corrosion was detected on any of the stainless steel grades in any of the test solutions. No corrosion rate exceeded 0.01mm/year. No localised attacks, such as pitting or crevice corrosion were observed".

Strength

The LDX is superior in mechanical strength to the 316L stainless steel. The superior strength has allowed plates to be manufactured using a thinner steel sheet. The minimum thickness acceptable is still under investigation. The strength of the steel prevents the manufacturing of plates greater than 3mm thickness. The reduction in thickness, without compromising the strength of the cathode reduces the manufacturing cost of the cathodes.

Flatness Tolerance

The flatness tolerance of commercially available LDX steel is not acceptable for use as cathode plates. Development work in the manufacturing of the steel has allowed the flatness tolerances to fall within acceptable limits.

Surface Finish

The 2B finish has been a successful surface finish for the standard 316L cathode plate. Voids in the surface finish allow the copper to "key" into the steel. This facilitates the adhesion of the copper to the stainless steel in the plating and harvesting cycle, yet permits easy removal once the plate is flexed. Figure 5 is an SEM micrograph showing the 2B surface finish of 316L.

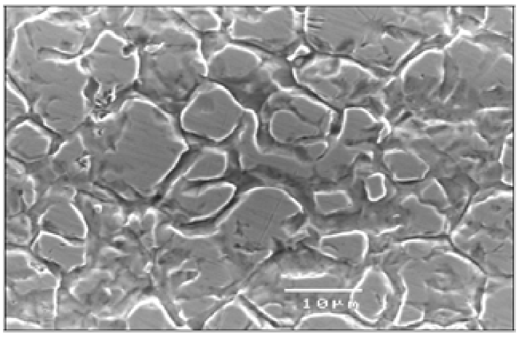


Figure 5 - SEM of 316 L surface magnified 2000x
MacDonald [6]

The nature of the duplex steel does not permit a standard 2B surface finish. A tenacious oxide layer prevents the 2B film from being achieved in the manufacturing process, MacDonald [6].

The achievable mill finishes on the duplex steel do not facilitate the successful adhesion of copper. In tests completed at the Townsville Copper Refineries pre-stripping occurred with a standard duplex mill finish. Figure 6 represents a typical mill finish of the duplex steel.

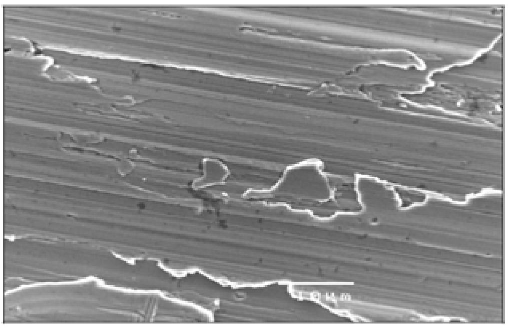


Figure 6 - SEM of LDX 2101 mill finish magnified 2000x MacDonald [6]

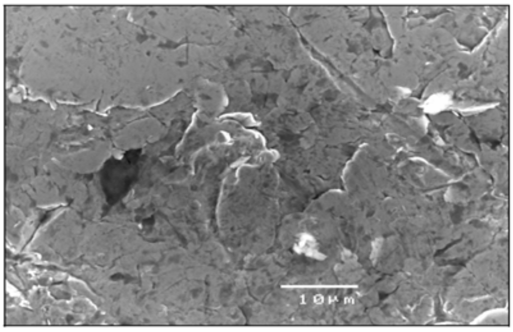


Figure 7 - SEM of LDX 2101 modified surface magnified 2000 x MacDonald [6]

The steels surface has limited voids for the copper to attach itself to. To facilitate the adhesion of copper to the LDX, IPT has developed a unique surface finish for use in the E/R and E/W of copper. An example of this modified surface is presented in Figure 7. Testwork has revealed optimum Ra (surface roughness measurement) values vary for E/R and E/W conditions due to the nature of the copper deposit. The surface finish can therefore be tailored to suit a specific operation.

Operational testing of the LDX 2101 steel

Approximately two cell loads (62 cathodes) of trial plates were manufactured with the LDX 2101 and placed into service at a small electro-winning facility in Australia.

An average analysis of the plants electrolyte conditions are shown in Table 8.

Table 8 – Average analysis of E/W tankhouse

Cu	38-45g/L
H_2SO_4	160-170g/L
Cl	20-30ppm
Mn	90ppm

The cathodes have been in service since the beginning of June 2006 and their physical properties and stripping performance have been monitored. Under standard operating conditions there have been no concerns with the plates.

304 STAINLESS STEEL

The standard 304L stainless steel with a 2B finish was tested in Townsville's Copper Refinery. The plates have been circulating for over 1.5 years with no concerns.

Accelerated corrosion tests were also completed on the steel. Table 6 and 7 represents the immersion tests completed. No uniform corrosion was detected on any of the stainless steel grades in any of the test solutions. No corrosion rate exceeded 0.01mm/year. No localised attacks, such as pitting or crevice corrosion were observed, Dickson [5].

CONCLUSION

Through ongoing development work the ISA PROCESS^{TM} has continued to improve on the original cathode plate design. Past and present testwork still suggests the ISA cathode BR^{TM} is the lowest resistance cathode on the market and the AB cathode has effectively reduced stripping inefficiencies in the ISA 2000 technology.
The development of the LDX 2101 as an alternative steel in cathode plates has the potential to reduce capital costs for end users while still providing a technically sound cathode.

ACKNOWLEDGEMENTS

The authors would like to acknowledge the assistance from Anna Iverson and associates for the corrosion work completed by them on the various steels. We would also like to acknowledge the Avesta Reserch Centre for the approval to publish the data.

In addition the authors would like to acknowledge Xstrata Technology for the time and resources to undertake this work and permission to publish this paper. This project has had input and assistance from a number of people specifically Tony Ruddell of the Xstrata Technology ISA PROCESS^{TM} development team.

REFERENCES

1. I.J Perry, J.C Jenkins and Y. Okamoto, “The Use of Stainless Steel Cathodes at Copper Refineries Pty. Ltd”, Townsville, AIME Annual Meeting, Chicago U.S.A, 1981

2. W.Webb and J.Weston, “The development of a “Lower Resistance” Permanent Cathode”, Minera Chilena March April 2003 (Spanish)
(English version available at www.xstratatechnology.com)

3. D. Peckner and I.M. Bernstein, Handbook of Stainless Steels, McGraw–Hill Book Company,USA,1977

4. Outokumpu Stainless AB, “Data Sheet LDX2101 Duplex Stainless Steel”, Sweden

5. C. Dickson, A.Iversen, A.Wallin and I. Hedstrom “Stainless Steel Cathodes in Copper Refining – Immersion Tests in Customer Defined Solutions”,Report prepared for internal distribution, Avesta Reserch Centre, 2006

6. L. MacDonald, “Stainless Steel in Copper Electrolyte Solutions”, Thesis Report, James Cook University, Townsville, Australia 2006

TANK HOUSE EXPANSION AND MODERNISATION OF COPPER REFINERIES LTD, TOWNSVILLE, AUSTRALIA

Presented at Alta Conference, Adelaide, Australia, 2000

Brendan O'Rourke
Operations Manager
Copper Refineries Pty Ltd
Townsville, Queensland 4810 Australia

CONTENTS

1.0 INTRODUCTION

Mount Isa Mines was originally founded on a series of Lead, Silver and Zinc ore bodies in 1923. In 1955 significant adjacent copper orebodies were defined. The unique polymetallic nature of the Mount Isa Mines operations commenced.

In September 1959 a subsidiary copper refinery company, Copper Refineries Pty Ltd commenced conventional "starting sheet" production and export of 40,000 tonnes per annum. A series of two major expansions lead to operating capacity of 80,000 tones per annum in 1966 and 150,000 tonne in 1975.

The first commercial introduction of the "ISA Process" permanent stripping technology was commissioned in 1978. Cathode production capacity was also increased to 170,000 tonne per annum. The vastly improved electrode alignment and accuracy inherent with the Permanent Stainless Steel Cathodes of the ISA Process then allowed the capacity of the Refinery to be increased to 210,000 tonne per annum by 1996.

The 1997 commissioning of the 51% Mount Isa Mines owned Ernest Henry Mine and the concurrent expansion of operations at Mount Isa afforded the opportunity to expand and modernise the Copper Refineries operation.

After 20 years of ISA Process tank house operation and 15 years experience in exporting the ISA Process technology Copper Refineries was in an enviable position with respect to design options for a world class facility. Experienced gained in tank house layout, machine design and operational philosophies was easily applicable.

The Refinery upgrade activities commenced in December 1996 and were completed in April 1999. An annualised capacity of 270,000 tonne per annum was achieved by July 1999.

2.0 PROCESS SELECTION

2.1 LAYOUT

Various combinations of tank house layout and design have been implemented in ISA Process tank houses. Copper Refineries was constrained in many cases due to the fact that a full retrofit had to be conducted with a minimum of disruption to the existing operation. For example the location of receipt and delivery services for anodes and cathodes were fixed. The layout selected incorporates one crane per aisle feeding to and from a centralised machinery bay.

2.2 DESIGN CRITERIA

Design objectives were related to:

- Long Term Production forecasts.
- Customer requirements.
- Copper Refineries Pty Ltd Operations Survey.
- Construction in an existing operating plant.
- ISA Process technology developments.

2.2.1 Long Term Production Forecasts

With the commissioning of the Ernest Henry mine and the expansion of the Mount Isa Copper Mine into the Enterprise orebodies an annual operating capacity of 270,000 tonne per annum was required.

Concurrent with these mine expansions the Mount Isa based copper Smelting facilities have been upgraded. Improvements include, an extra oxygen plant to provide increased primary smelting throughput, stacker reclaiming concentrate blending facility, a fourth Pierce Smith converter together with a new anode casting wheel and rotary holding vessel.

2.2.2 Customer Requirements

Accredited with ISO9002 in 1994, Copper Refineries has a proven quality system to ensure the satisfaction of our customer base. Modernisation of the cathode stripping machines allowed the introduction of improved cathode bundle appearance handling and shipping. Requirements in terms of individual cathode weight, labelling, sampling and corrugation were highlighted.

2.2.3 Operations Survey

An extensive series of surveys was conducted with all operating personnel to ensure that problem areas with the existing operation were addressed. Operational improvements were identified in material movement, machinery access, slimes handling and reagent dosing. Over 350 suggestions were logged into a database to provide for key design criteria.

2.2.4 Construction in an Existing Plant

The requirement to minimise production disruption dictated that new electrode handling equipment be commissioned prior to removal of the superseded equipment. Conversion of the electrolyte and cell systems was conducted in a series of intensive shutdowns that were coincided with planned shutdowns at the Mount Isa Copper Smelter. A key design element in the electrolyte and bus bar systems was ease of installation.

2.2.5 ISA Process Technology Improvements

The Cathode Stripping Machines have provision for improvements in the edge masking system utilised in the ISA Process. These improvements relate to the elimination of bottom wax on the stainless steel cathode blanks. This concept is now marketed as ISA Process 2000.

2.2.6 Final Process Selection

The expansion and modernisation of the Copper Refineries tank house was implemented with the following items:

- Replacement of 1024 lead lined cells with 1162 polymer concrete cells.
- Total replacement of electrolyte, slimes and DC electrical reticulation.
- Installation of section switches
- Installation of Decant Filtration and “Quick-Fill” operation.
- Centralised reagent batching and distribution.
- Replacement of Anode Scrap Machine with new machinery that incorporates single piece washing.
- Replacement of Anode Preparation Machine with new machinery incorporating lug milling and buffered storage capacity.

- Replacement of Cathode Stripping Machines with new machinery incorporating inline bundling, weighing and strapping.
- Replacement of six manually operated cranes with two automated cranes utilising pyramid based locating devises.
- Installation of a real-time cathode weighing , loading and despatch computer system

3.0 PROCESS DESCRIPTION

3.1 Modernised Refinery Operating Statistics

Capacity	270,000 tonnes per annum
Cells	1162
Electrolyte Temperature	63^0c
Cell Flow Rate	25-30 litres per minute
Anodes	45
Cathodes	44
Electrode Pitch	95mm
Maximum Anode Weight	410 kg
Maximum Cathode Weight	115 kg
Anode Cycle	21 days
Cathode Cycle	3 x 7 day or 10/11 day
Number Rectifiers	4
Current Density	305 A/M^2
Electrical Circuits	3
Electrolyte Circuits	3
Number of cranes	2

3.2 TANK HOUSE LAYOUT

The Cathode Stripping Machines were relocated from their positions at the extremities of the building to be centrally located. The anode and scrap handling functions are also located in the centre of the building.

A machinery annex was added to the building to locate all of the electrode handling machines. This design was required to meet the same side anode receipt and cathode, anode scrap despatch design requirement. Six aisle trolleys transport electrodes into and out of the machine annex. The machinery annex has two small auxiliary cranes installed to allow for heavy equipment maintenance, reject cathode plate removal and transport of stripping machine consumable items.

The incorporation of the machinery annex into the layout designs meant that the tank house only needed to be extended by one bay to accommodate the extra 138 refining cells. The extra bay was also required to facilitate the introduction of the new overhead cranes and the removal of the redundant cranes.

3.3 OVERHEAD CRANE OPERATION

Two Kunz fully automated overhead cranes handle all refinery stripping operations. These utilise a locating bale onto a fixed pyramid to achieve accurate electrode alignment. The capacity of the cranes will allow for lifts of 45 anodes at 410 kg each. The crane is capable of multiple tasks. Individual lifts of anode, anode scrap or cathode plates are possible with a dual lifting bail arrangement.

A laser-based device communicates via a central supervisory logic controller the operation status of all cranes and electrode handling machines. This communication linkage allows for safe and efficient stripping operations to be conducted and scheduled with a minimum of down time and operator intervention. Programmed sequences for all stripping operations are installed in the cranes scheduling logic controller.

3.4 ANODE PREPARATION MACHINE

Anodes are received at the Townsville Refinery from Mount Isa via rail and road. The new Anode Preparation Machine was supplied by Finland's Outokumpu Oy. The anodes as received are loaded via fork lift to the preparation machine for weighing, blade pressing, lug offset pressing and milling

Milling is both face and underside to provide improved anode verticality and current distribution. When combined with lug offset pressing it is possible to present a high weight, vertically aligned anode to the automatic Kunz crane.

The prepared anodes are then transported to a live storage facility via two automated trolleys. From the live storage area the anodes are either fed directly to the tank house Load-Out beams or placed into an assigned storage location.

The live storage feature of the installation allows the anode preparation function to be independent of the refinery stripping operations. This gives enormous flexibility to scheduling of anode receipt labour and a vast reduction in anode double handling.

3.5 ANODE SCRAP MACHINE

Anode Scrap is remove from the refinery for subsequent remelt at the Townsville site. The new Anode Scrap Machine was supplied by Mesco, Japan. Anode scrap bundles are presented to a fork lift interface for removal to the anode casting operation.

The new machine incorporates single blade washing with a three stage washing process. Three separate settling tanks are used to maximise recovery of slimes with wash water volume make-up supplied via a final rinse station. An innovative “grizzly” arrangement is located below the traverse conveyor to allow safe and efficient removal of any fallen scrap. This arrangement is serviced by one of the machinery annex auxiliary cranes.

3.6 CATHODE STRIPPING MACHINES

Two new Cathode Stripping Machine were supplied by Mesco, Japan. The new machines incorporate all features identified from customer surveys. Features that have been introduced are cathode corrugation, automatic sampling, inline sworn weighing, bundle labelling and strapping. Fully prepared cathode bundles are presented to a forklift interface.

Control of these features has been linked to an existing commercial scheduling and invoicing computer system to ensure cathode bundles are prepared to Customer Specification. Despatch and invoicing documentation is also produced by the same system.

3.7 ELECTROLYTE CIRCULATION

The operations survey identified many deficiencies in the existing circulation systems. The replacement of all the old lead-lined concrete cells afforded the opportunity to upgrade the circulation system. Polymer concrete cells were supplied by CTI Pacific Pty Ltd.

For operation at increased current densities a higher cell electrolyte flowrate is required. New circulation pumps and larger diameter piping was installed. The electrolyte piping installed is a fibre reinforced vinyl ester resin base supplied by FibreTec Australia. Over 3,500 metres of piping was installed. Individual cell feed valves provide for quick-fill operation on each refinery cell.

Decant collection and filtration has been introduced to the refinery operation. Separate decant collector boxes and Scheibler filters have been installed.

3.8 ELECTRICAL RETICULATION

The original electrical bus bar system installed at Copper Refineries Pty Ltd was not suitable for any further increase in operating current density. A 4 copper plate laminated bus bar system was installed together with full load section switches. The section switches were supplied by Hundt and Weber, Germany and are capable of full load switching at 30,000 Amps. Over 400 tonnes of copper bus bar were installed.

4.0 CONSTRUCTION

In all cases the scheduling of construction activities was dictated by cathode production requirements. This meant that cell, electrolyte circulation and bus bar system replacements had to be conducted in intensive shutdown periods, that coincided with planned production shutdowns.

The major construction milestones are listed below:

▪ April-June 1997	Crane runway beams strengthened.
▪ Sept-Oct 1997	Replacement of two electrolyte circuits – 600 cells
▪ Feb-May 1998	Construction of Machinery Bay Annex and machine runways beams.
▪ May-June 1998	Installation of Mesco supplied CSM's and ASM
▪ Aug-Sept 1998	Removal of old Cathode Stripping Machines.
▪ Aug-Sept 1998	Installation of Kunz cranes and replacement of crane rails.
▪ Sept-Oct 1998	Replacement of one electrolyte circuit - 424 cells.
▪ Sept-Oct 1998	Installation of Outokumpu supplied Anode Preparation Machine.
▪ January 1999	Removal of old APM and ASM.
▪ February 1999	Installation of additional 138 Refining Cells
▪ April 1999	Removal of six old overhead cranes

The activities were conducted with minimal production losses and over 500,000 Contractor Manhours without a lost time injury. This was only made possible due to the close cooperation and teamwork between the refinery and construction personnel.

5.0 COMMISSIONING

A dedicated commissioning team was setup to ensure a smooth transition from old to new equipment. Successful completion of several construction activities required a timely commissioning schedule for the major electrode handling machines.

The key commissioning milestones are listed below:

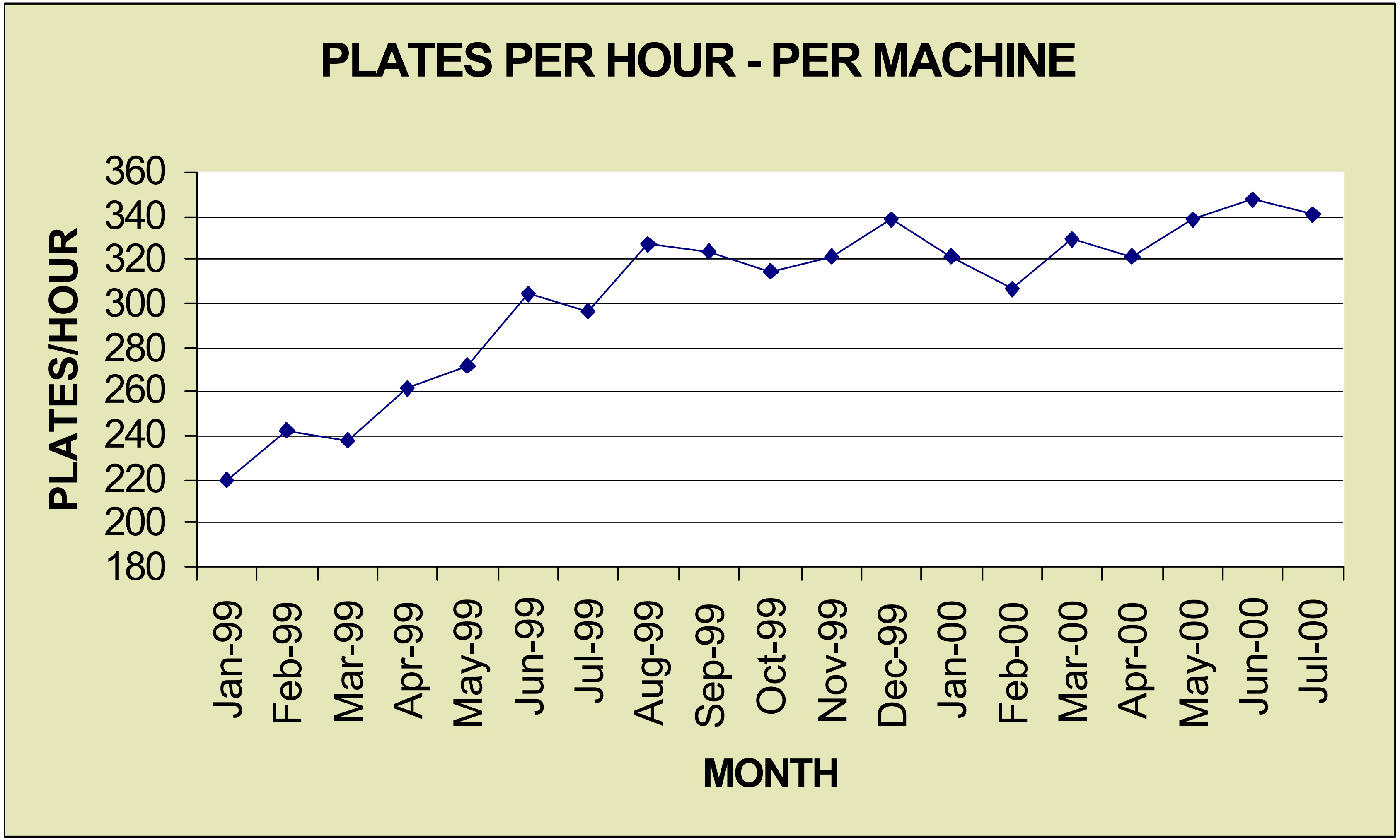
July 1998	Commissioning of Mesco supplied machines commenced.
September 1998	Acceptance tests for two Cathode Stripping Machines and Anode Scrap Machine completed.
October 1998	Partial commissioning of Kunz cranes commenced.
November 1998	Commissioning of Anode Preparation Machine commenced.
November 1998	Full-load Commissioning of Kunz cranes commenced.
December 1998	Acceptance tests for Anode Preparation machine completed.
January 1999	All Production operations converted to Kunz Cranes.
February 1999	Acceptance tests for Kunz cranes completed.

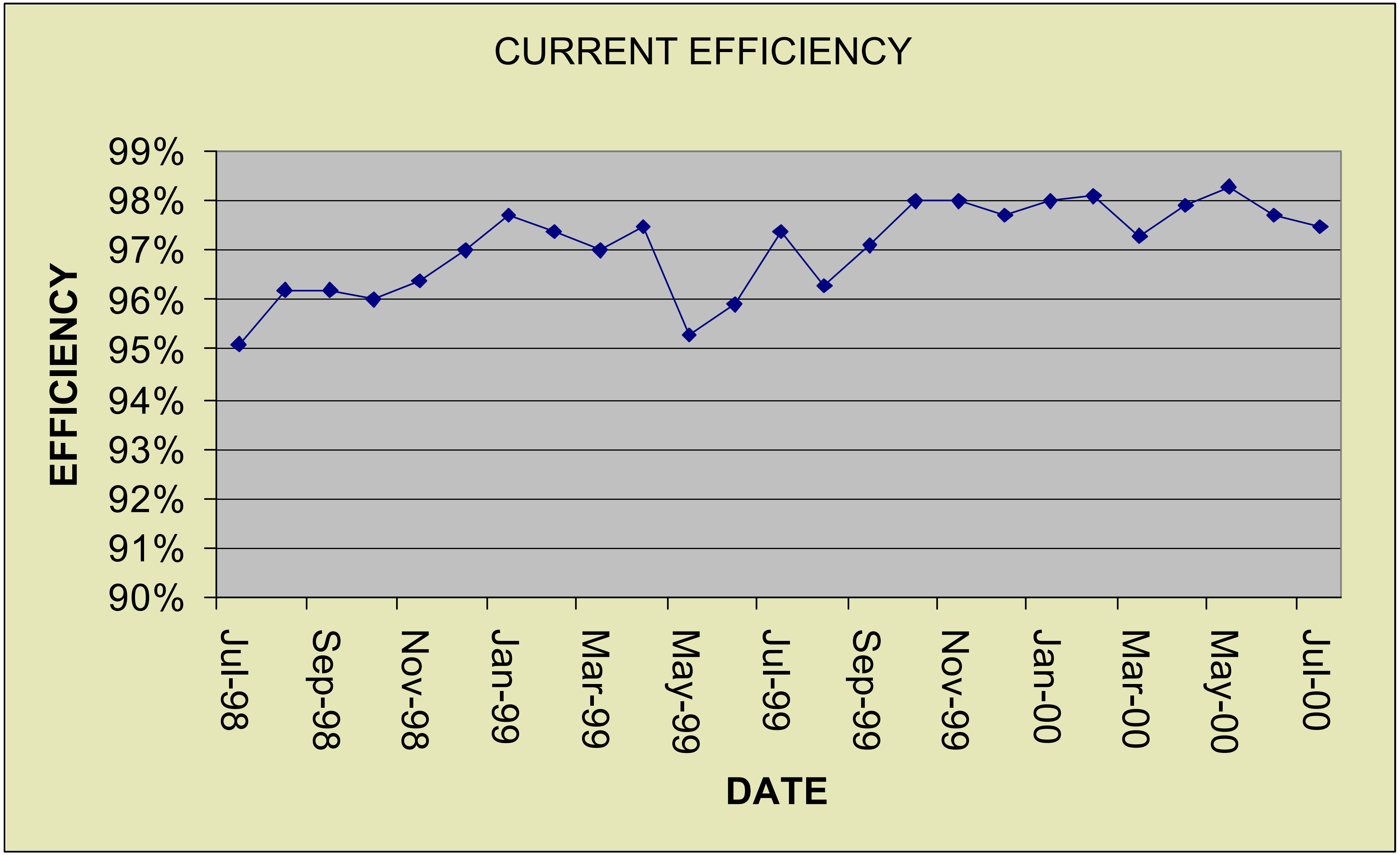
Commissioning engineers were provided by the three suppliers, Mesco, Outokumpu, and Kunz. Refinery Operating and training schedules were also developed by the Commissioning group to ensure that the overall construction schedule could be achieved.

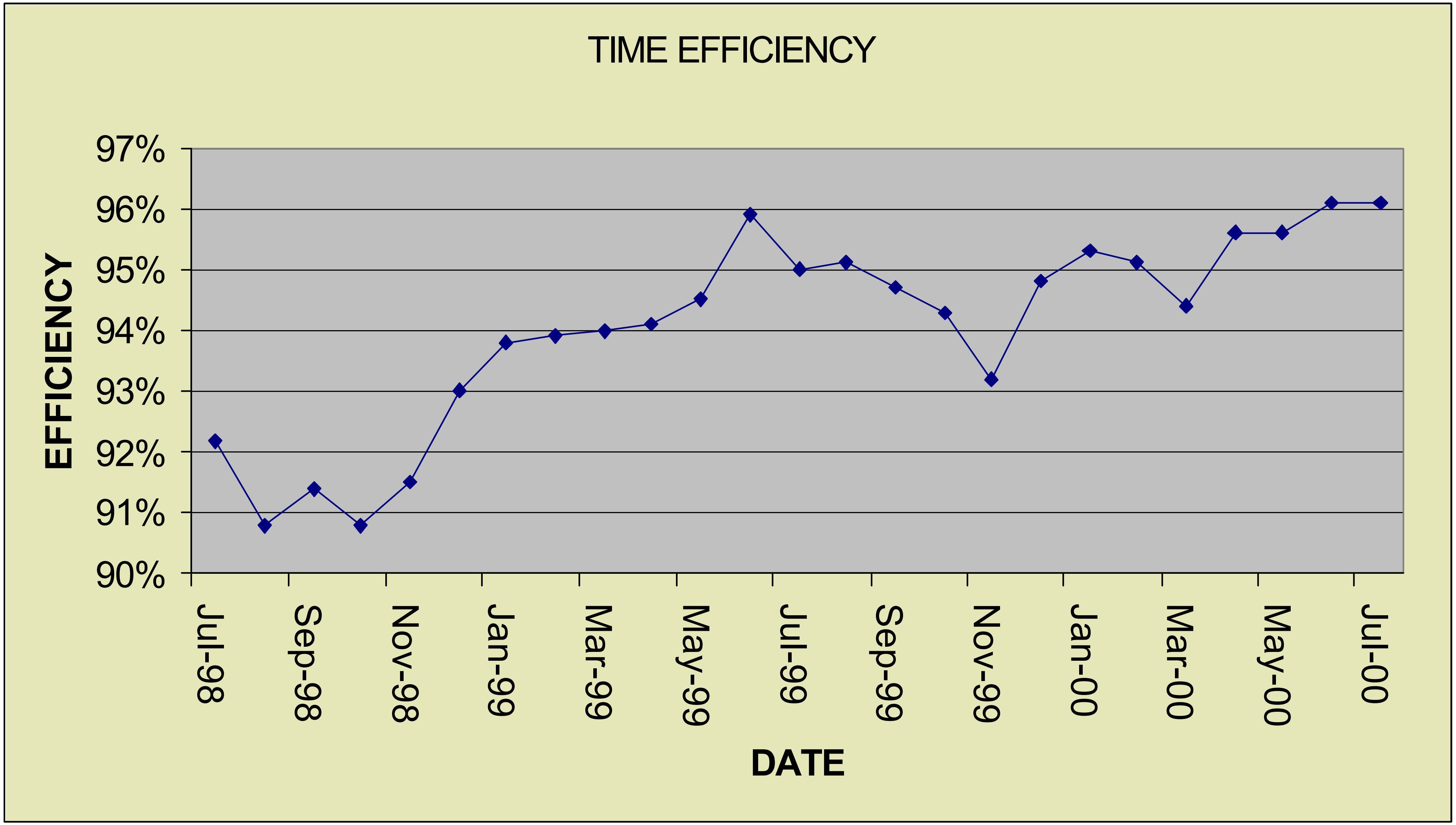
6.0 PRODUCTION AND OPERATIONAL IMPROVEMENTS

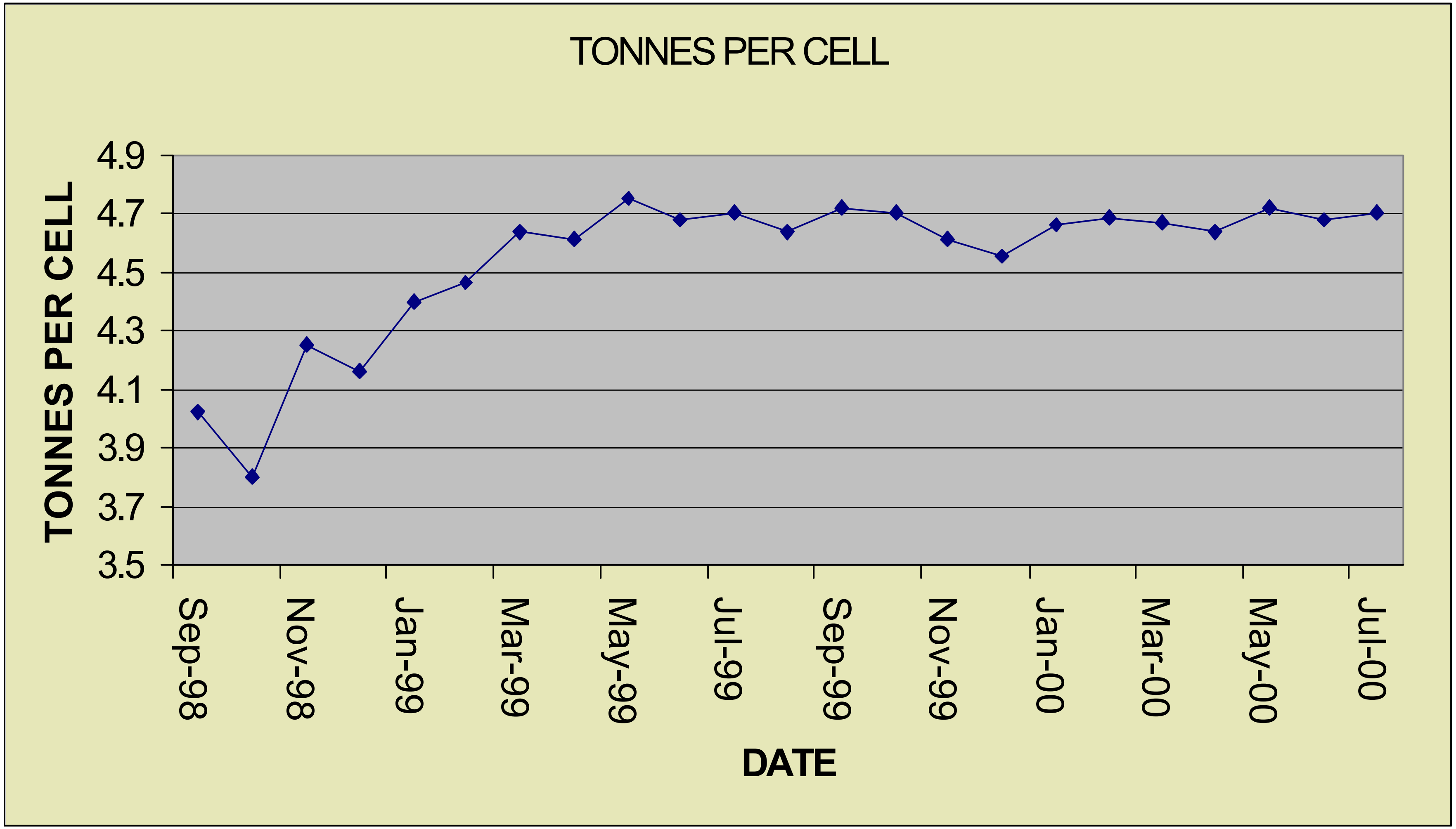
The tables below detail the production performance of the new refinery since 1998. The step change improvement in early 1999 came about by the simultaneous introduction of milled anodes and the Kunz cranes to the refinery.

Further increases in output have been due to refining intensity increases. These increases are now possible due to the removal of previous anode weight constraints. Future intensity increases are planned as the Mount Isa Copper Smelter increases output to 270,000 tonnes per annum.


PLATES PER HOUR - PER MACHINE
PLATES/HOUR
360
340
320
300
280
260
240
220
200
180
Jan-99
Feb-99
Mar-99
Apr-99
May-99
Jun-99
Jul-99
Aug-99
Sep-99
Oct-99
Nov-99
Dec-99
Jan-00
Feb-00
Mar-00
Apr-00
May-00
Jun-00
Jul-00
MONTH


CURRENT EFFICIENCY
EFFICIENCY
99%
98%
97%
96%
95%
94%
93%
92%
91%
90%
Jul-98
Sep-98
Nov-98
Jan-99
Mar-99
May-99
Jul-99
Sep-99
Nov-99
Jan-00
Mar-00
May-00
Jul-00
DATE


TIME EFFICIENCY
EFFICIENCY
97%
96%
95%
94%
93%
92%
91%
90%
Jul-98
Sep-98
Nov-98
Jan-99
Mar-99
May-99
Jul-99
Sep-99
Nov-99
Jan-00
Mar-00
May-00
Jul-00
DATE


TONNES PER CELL
TONNES PER CELL
4.9
4.7
4.5
4.3
4.1
3.9
3.7
3.5
Sep-98
Nov-98
Jan-99
Mar-99
May-99
Jul-99
Sep-99
Nov-99
Jan-00
Mar-00
May-00
Jul-00
DATE

7.0 REFERENCE

1. M Landan, H Traulsen, "The New Copper Tank House at Norddeutsche Affinerie", Copper 91/Cobre91 Volume III, Hydrometallurgy and Electrometallurgy of Copper, W C Cooper, D J Kemkpk, G E Lagos and K G Tan, Eds., Pergamon Press, Oxford, England, 1991, 517-526.

2. R Dobner, M Schwill, "Modernisation and Extension of a Conventional Tank House," Erzmetall, Vol 48, 1995.

3. J O'Kane, "ISA Process – Benchmarking Copper Refining Electrolysis," Paper presented at the GDMD 33rd Metallurgical Seminar, Lunen, Germany, November 19-21, 1997.

4. B.O'Rourke, "Copper Refineries Limited – Refinery Expansion," Paper presented at the 6th ISA Process Users Conference, Townsville, Qld Australia, June 20-22 1999.

5. W Armstrong, "ISA Process 2000", "Paper presented at the 6th ISA Process Users Conference, Townsville, Qld Australia, June 20-22 1999.

6. B.O'Rourke, "Tank House Expansion and Modernisation of Copper Refineries Limited," Paper presented at Copper-Cobre 99 Conference, Phoenix, United States, October 9-12 1999.

The development of a "lower resistance" permanent cathode

(ISA Cathode BR™)

Wayne Webb

Executive - Research & Development
ISA PROCESS Technology

Joanne Weston

Metallurgical Engineer
ISA PROCESS Technology

ABSTRACT

The use of a permanent stainless steel cathode for the electrowinning and electrorefining of cathode copper is a well-accepted technology. As the developers of this technology, the ISA PROCESS Technology (IPT) has continued to improve the technology and maintain leadership.

As the pressure to produce cathode copper more efficiently has increased in the market place, IPT has recognised the need to develop a more efficient cathode plate.

The IPT has now developed a low resistance cathode plate (ISA Cathode BR™) in response to market demand. The ISA Cathode BR™ provides lower electrical resistance by decreasing the distance the between the end of the copper plating and the solution line.

Plant trials in the Townsville ISA PROCESS refinery and in a major Chilean electrowinning plant were undertaken to prove the performance of the new cathode type BR™. Results to date indicate that significant operating cost savings are available with the use of the ISA Cathode BR™ electrodes.

INTRODUCTION

In 1978 MIM's Townsville Copper Refinery permanently changed the electro-copper industry after the introduction of the permanent stainless steel cathode.

The ISA PROCESS Technology (IPT) allowed for the reduction in manning requirements, provided superior mechanical handling systems, reduced copper inventory levels, permitted a more intense operating regime, improved cathode quality and offered a much safer electrolytic operation. It enabled the commissioning of smaller scale electrowinning operations by making them independent of starter sheet supplies.

The introduction of inherently straight permanent cathodes has necessitated the development of anode casting and preparation systems in conjunction with superior contact arrangements. These developments in combination with the ISA PROCESS have seen the introduction of operating current densities up to 350 A/M^2 and current efficiencies in excess of 98%.

The technology assures the production of better than LME “grade A” quality. The process has been so successful, in this regard, that the LME “grade A” standard, is now considered a minimum by the best producers.

The ISA Cathode BR™ is the latest ISA PROCESS cathode plate offering a low resistance electrode with the potential to reduce power costs significantly. On site trials at Compania Minera Zaldivar supported earlier test work conducted by the ISA PROCESS at Townsville, Australia.

Figure 1: ISA Cathode BR™ in service at Compania Minera Zaldivar

"Lowest resistance cathode"

1.1. The permanent cathode plate

Underlying the success of the process is the basic design of the ISA PROCESS cathode. This design, while constantly being improved over the last twenty years still remains true to the design precepts of I.J.Perry’s original concept (Perry, Jenkins & Okamoto, 1981).

The ISA cathode consists of a 304L stainless steel hanger bar which is stitch welded to a 316L stainless steel blade (3.25mm+ thick). This assembly has been proven so strong that no bond failure has occurred in a commercial operation due to either corrosion or mechanical handling.

The hanger bar assembly is then electroplated with copper to a thickness of 2.5 mm. The 2.5mm of copper provides sufficient electrical conductivity and imparts a degree of corrosion resistance. Under correct operating conditions will provide an operational life without repair in excess of 7 years in electrowinning situations and over 15 years (perhaps as much as 20) in electrorefining situations.

Deterioration of the ISA PROCESS hanger bar system occurs very slowly and is predictable. This permits operators to carefully plan for the refurbishment of their operating cathodes. No catastrophic failures of this assembly have occurred in commercial production facilities.

The diagrams below illustrate a comparison in cathode plate performance over time for the ISA cathode and solid copper hanger bar systems under electrowinning conditions. Zone A illustrates the region where there is no change in performance of the cathode. Zone B is where the performance of the cathode begins to deteriorate. ISA cathodes exposed to corrosive conditions will exhibit this slight drop off in performance. Solid copper bars will experience corrosive attack on the hanger bar to blade joint.

After hanger bar refurbishment the ISA cathode is restored to its original condition and its electrical performance is restored. The solid copper hanger bar will lose electrical performance due to corrosive attack on the joint and sudden failure is possible. The hanger bar must be reattached to the blade. Only "re-welding" of the hanger bar will restore the performance of the cathode. The advantage of the ISA cathode is that there is no sudden failure. A scheduled removal from service can occur and repair of the cathodes can be undertaken. The solid copper bar may experience a limited Zone A as the rate of progression of corrosion of the copper to stainless joint is unpredictable.

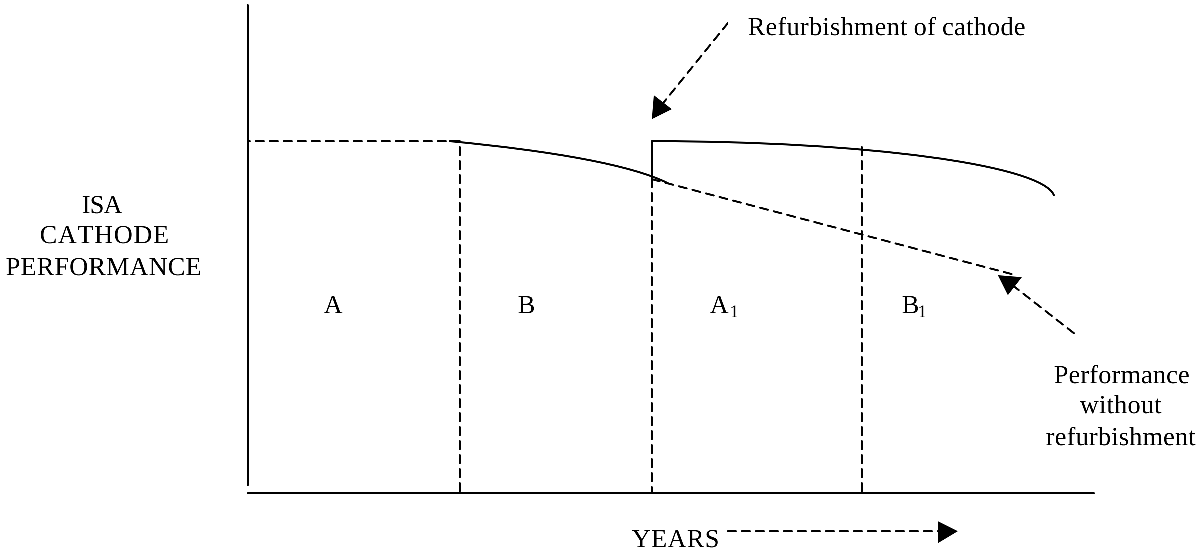


Figure 2: ISA PROCESS cathode performance

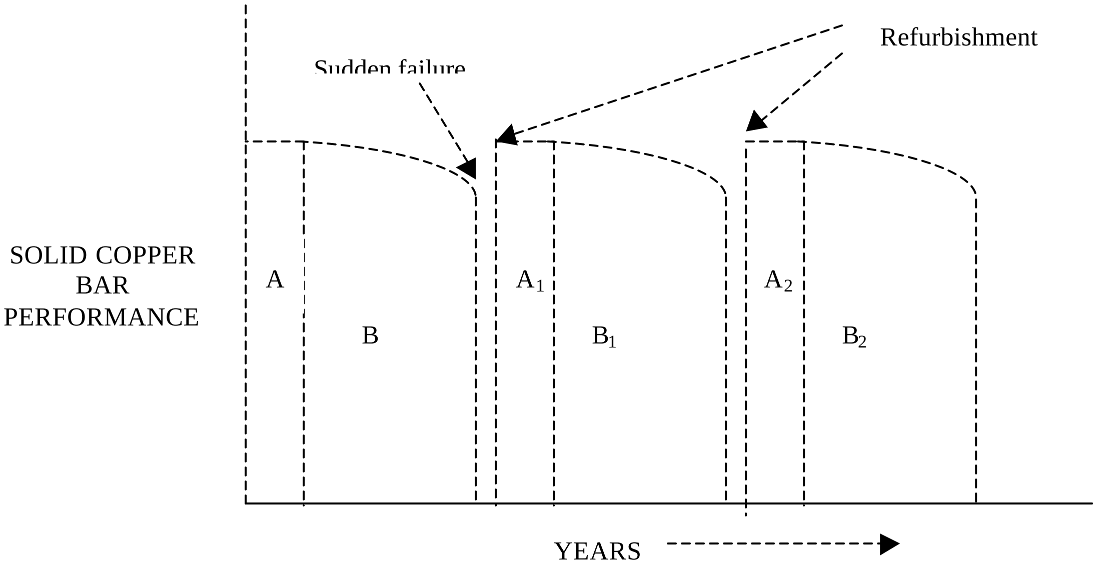


Figure 3: Solid copper hanger bar performance

The assembly of the ISA cathode provides the following key elements:-

- Great mechanical strength
- Inherent straightness, initially and in the long term
- Resistance to annealing (softening) of the hanger bar during the occurrence of electrical shorts
- Sufficient blade thickness to cope with stripping machine stresses, while maintaining the verticality necessary for high current efficiency.
- An edge masking system which is mechanically keyed into the blade.

Since commercialisation, greater than one million ISA PROCESS cathodes have been produced. They have proven to be the most mechanically robust and most corrosion resistant permanent cathode plate on the market. These characteristics are necessary to maintain high current efficiencies in the long term.

1.2.The progression of the permanent ISA cathode plate

The ISA PROCESS cathode has been continuously developed and improved since it was first manufactured in 1978. Figures 4 to 8 illustrate the progression of the ISA PROCESS system.

The first ever ISA Cathode for commercial sale was the I-beam hanger bar illustrated in Figure 4. The copper plating was an average of 1.3mm thick, this was increased to 2.5mm some years later. Prior to this Figure 5 illustrates the rounded bottom I-beam. The design improved the contact pressure and assisted with in cell verticality. The increase in average copper thickness was to improve the corrosion resistance of the hanger bar. Cathode plates with solid copper hanger bars were produced for testing in the Townsville Copper Refinery.

The next development was the move to the rectangular hollow section (RHS) stainless steel bars with a flat top as shown in Figure 6. The RHS bar gave the cathode plate greater rigidity coupled with a much simpler shape to electroplate the copper coating (Armstrong, 2001). The design was modified to a rounded RHS bar shown in figure 7. The rounded top enables a more even distribution of the electroplated copper and ensures any electrolyte will drain off the top of the hanger bar. This reduces the potential for hanger bar corrosion.

The next major development with respect to the cathode plate was the ISA2000 waxless cathode plate. This involved machining a v-groove along the bottom edge of the blade allowing for the separation of the cathode copper into two separate sheets. This eliminated the need for bottom edge wax.

Thinner RHS bars have been used in the manufacture of ISA cathodes based on customers' specific requirements. The latest cathode design, ISA Cathode BR^{TM} is shown in Figure 8. The copper plating has been extended to just above the lifting windows and has an average thickness of 3.0mm

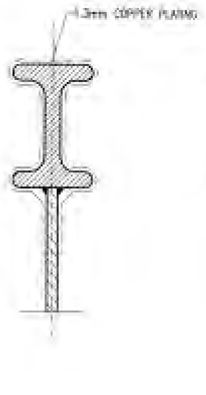


Figure 4

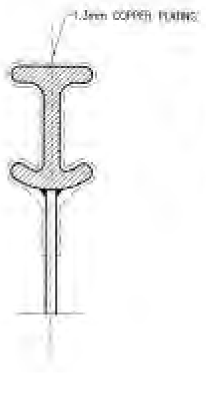


Figure 5

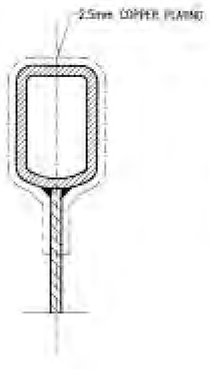


Figure 6

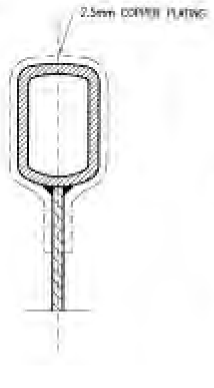


Figure 7

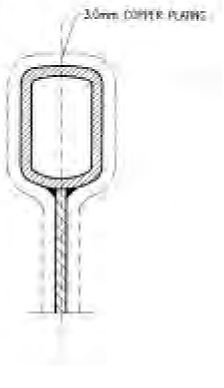


Figure 8

MARKET DEMAND FOR A "LOWER RESISTANCE" CATHODE PLATE

1.3. The earlier ISA cathodes

The earlier ISA cathodes, whilst superior in most aspects had slightly higher resistance than the solid copper hanger bar systems. In spite of the initial higher resistance, the ISA cathode will consume less power per tonne than other cathode types such as the solid copper hanger bar system. While this lower resistance is attractive, the poor long-term performance of the solid copper hanger bar led to its rejection following a large-scale trial at MIM's Townsville Copper Refinery.

Trials were conducted to compare the performance of solid copper hanger bars and ISA cathodes. Approximately 3000 cathode plates with solid copper hanger bars were manufactured and placed into service for testing. A statistical analysis on a significant proportion of tank house data was carried out. The results indicated that the solid copper hanger bars would produce a lower current efficiency than the ISA cathodes (Bailey, 1995). The difference in current efficiency was 2.4%, with crop 1 results indicating higher inefficiencies of the solid copper hanger bars (Bailey, 1995).

These differences were attributed to actual physical characteristics, such as cathode plate straightness (Bailey, 1995). The ISA cathode plate is inherently straight and maintains its integrity over extended periods of time. Even under severe shorting situations the hanger bar remains straight. This is in direct comparison with the solid copper hanger bar, which is prone to bending.

Trials were also conducted at an ISA PROCESS electrowinning plant to compare ISA cathodes with solid copper hanger bars. These trials supported the results at Townsville's Copper Refinery. They found that the ISA cathodes operated at 2% higher current efficiency than the solid copper hanger bars (4th ISA PROCESS User's Conference 1995 - Iquique).

When considering both current efficiency and conductivity, the ISA cathode plate is the most efficient cathode (Armstrong, 2001).

THE DEVELOPMENT OF THE ISA CATHODE BR™

The standard ISA PROCESS permanent cathode is electroplated with copper to approximately 15 mm down onto the blade. This ensures even flow of current into the blade and a more even initiation of the copper deposit. It was theorised that if this copper was projected further down the blade than the high resistance path through the stainless steel would be significantly reduced and a useful reduction in resistance would occur.

The ISA PROCESS Technology has now developed a low resistance cathode (ISA Cathode BR™). The concept of reducing plate resistance by depositing the copper further was theorised and subsequent calculations highlighted the impact of increasing the depth of the copper plating. An $11\mu\Omega$ per plate decrease in resistance with the copper coating down to the windows was calculated. The results showed that the further down the blade the copper coating the greater the potential energy savings.

The BR™ cathode design *differs* from the standard design in two ways:

- The copper plating extends approximately 55mm down the blade compared to 15 - 17mm.
- Increased average copper thickness of 3.0mm compared to 2.5mm.

The *benefits* of this new " high electrical performance cathode" are:

- Extension of the copper plating reduces the amount of electrical resistance that exists between the copper plating and the solution line by reducing the distance the current has to travel through the stainless steel.

- Increasing the thickness of the copper deposit provides greater corrosion resistance, particularly in electrowinning applications where cathodes may be subjected corrosive conditions. (Weston, 2002)

1.4. INITIAL TEST WORK

IPT developed some prototypes of the ISA Cathode BR^{TM}. These were tested extensively to determine the electrical characteristics of this cathode plate design. John Nielsen at James Cook University of North Queensland conducted a series of tests relating to this cathode design.

The electrical conductivities of various types of cathode plates were measured. In addition resistive models of the cathode plate were developed to show the electrical losses experienced by each of the cathode types (Nielsen, 2001)

The IPT decided that provided the copper coating does not extend below the lifting windows, no additional restriction would be placed on the operation of the cathodes and on the compatible stripping machines. The concept proved correct and further finite element analysis suggested that compared to a standard cathode, gains as high as 38% could be achieved (Nielsen, 2002). The deeper down the blade the copper coating is deposited the greater the reduction in resistance.

OPERATIONAL TESTING OF THE BR^{TM} CATHODE

Testing of the original cathodes by the ISA PROCESS in the Townsville electrorefinery was commenced and is ongoing. The cathodes have achieved the targeted reduction in resistance while easily fulfilling the operational requirements of the electrorefining process.

A decision was made to extend the testing to electrowinning operations. Following a small test of the BR^{TM} cathode design by a large Chilean electrowinning operation it was agreed to extend the test. Testing of a full cell load of ISA Cathode BR^{TM} electrodes commenced in October 2002.

1.5.Test Data - Compania Minera Zaldivar

A full cell load of trial plates were manufactured to the ISA Cathode BR^{TM} design and placed into service at Compania Minera Zaldivar (CMZ).

Figure 9. Cell load of ISA Cathode BR^{TM} plates on trial at Compania Minera Zaldivar

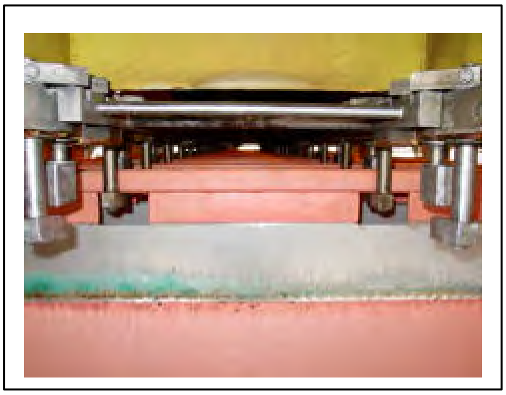


Figure 10. ISA Cathode BR^{TM} plates on trial at Compania Minera Zaldivar

Since these cathode plates have been in operation their electrical performance has been monitored and the results to date are confirmation of the earlier test work. Current distribution and internal plate resistance measurements have been recorded on a regular basis.

The potential difference between the hanger bar and solution line was measured which enabled the internal plate resistance to be calculated. The results indicate that ISA Cathode BR^{TM} electrodes are a lower resistance cathode plate in comparison with ISA cathodes approximately 7 years old. Figure 11 highlights the difference in performance of the new ISA Cathode BR™ with the older style ISA cathode.

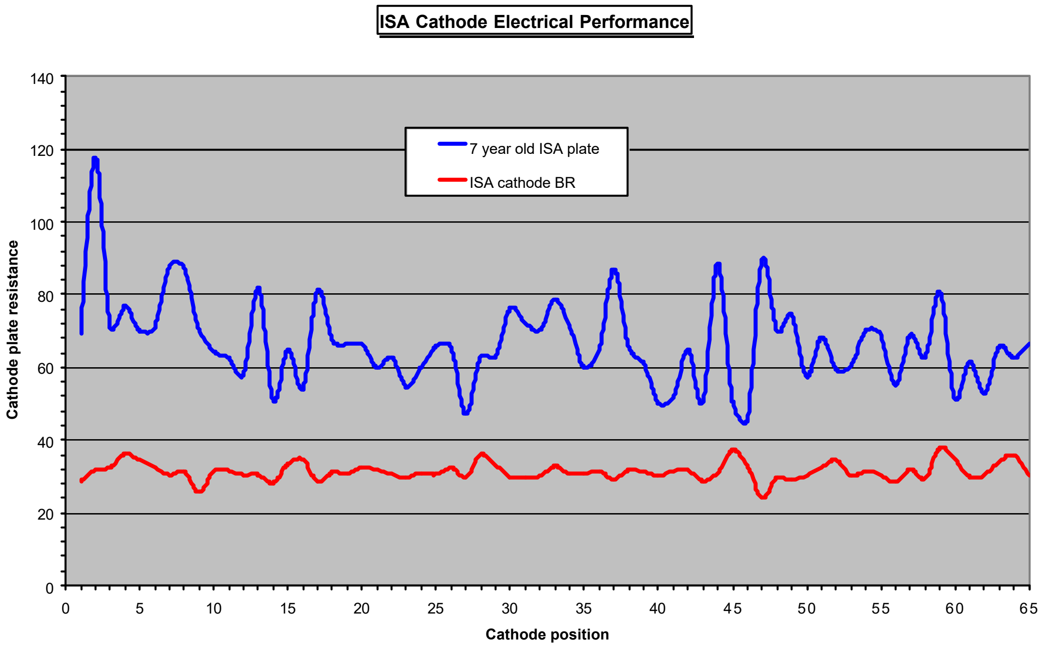


Figure 11: Graph illustrating the difference in cathode plate performance of different cathode types

Further measurements were taken at Compania Minera Zaldivar to compare the performance of a number of cathode plate designs. The results are illustrated in Figure 12. The cathode types that were used in the comparison were as follows:

- ISA Cathode BR™
- Solid copper hanger bars
- Thicker solid copper bars
- Wrapped solid copper hanger bars

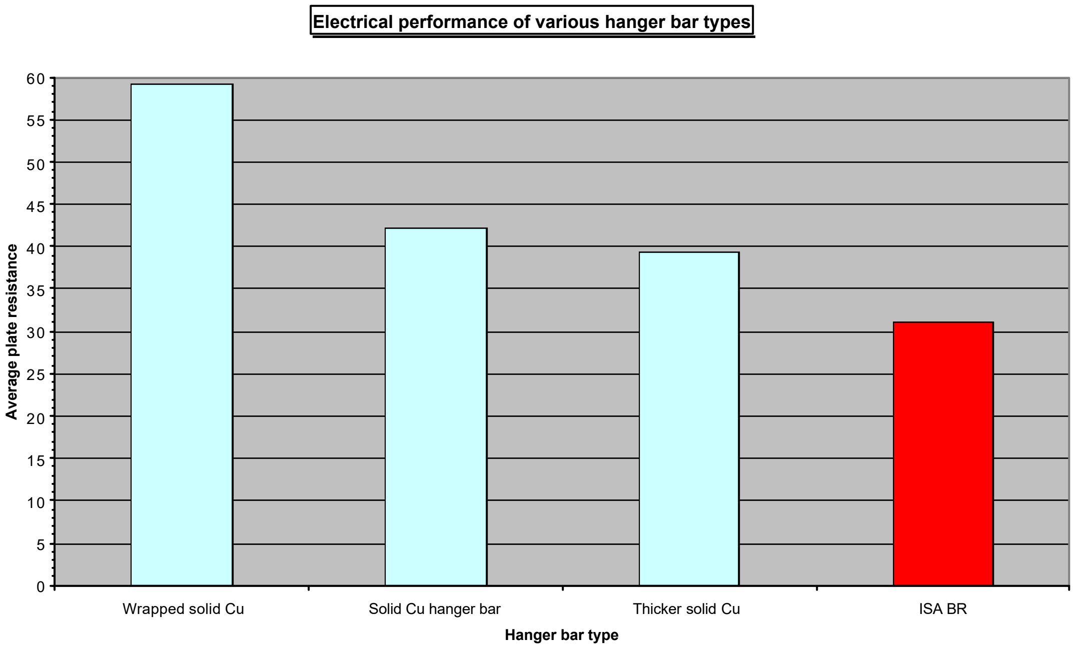


Figure 12: Internal resistance of various types of cathode plates

The results from the last set of measurements taken at CMZ on the 8th January 2003 clearly show that the ISA Cathode BR™ electrodes are electrically superior, exhibiting the lowest plate resistance.

POTENTIAL COST SAVINGS

Measurements recorded at Compania Minera Zaldivar illustrate that the ISA Cathode BR™ has the potential to reduce their power costs by approximately US$100,000 per year in a plant of this size. This is the magnitude of saving that the new ISA Cathode BR™ can provide due to their significantly lower plate resistance resulting from the increased depth of copper plating. (Data supplied by C. Pasten, 2002). This is in comparison 7-year-old ISA cathodes and solid copper hanger bars. These measurements were taken on the 15-16th November 2002. A full cell load containing the new ISA Cathode BR™ was measured along with three other standard cells which contained seven-year-old ISA cathodes and solid copper hanger bars.

Calculations showing potential power savings

Potential from contact to solution line	Average difference: 23mV
Average current per cathode plate	660 Amps
Power per cell	989 Watts per cell
Total number of cells in plant	368
Power cost	US$ 0.032 / kWh
Total power savings across the plant	364kW
Total power COST savings per year	US$102,036

Earlier tests and calculations based on data from the Townsville copper refinery also showed that significant cost savings could be made through the application of deeper copper plated hanger bars.

CONCLUSION

The permanent ISA cathode plate has undergone significant changes since its development. The latest development is the ISA Cathode BR™. Past and present test work suggests it is the most efficient cathode plate available, offering significant savings in power costs. The copper plating now extends to just above the bottom edge of the window and the average copper coating thickness has been increased. The ISA Cathode BR™ is a low resistance, mechanically and chemically stable cathode plate.

REFERENCES

Armstrong, W. (27th May - 1st June 2001). Seventh ISA PROCESS Users' Conference, Phoenix Arizona, USA. Session 1. *ISA PROCESS Developments.* MIM.

Bailey, D. (20th July 1995). Internal Memorandum. *Comparison of current efficiencies of CRL solid and RHS hanger bar mother plates.* Copper Refineries Pty. Ltd.

Nielsen.J. (2001- 2002). *Electrical conductivity tests of stainless steel cathode plates.* Report prepared for internal distribution. Series of reports covering tests conducted during the period.

Perry.I.J, Jenkins.J.C, & Okamoto,Y. (23rd Feb 1981). *The use of permanent stainless steel cathode sat Copper Refineries Pty Ltd. Townsville.*

Fourth ISA PROCESS Users' Conference, Iquique - Chile, (3rd -7th April 1995). Session 3, Topic 2 - *Cathode Mother Plate Performance*. MIM Technology Marketing Limited and Lanzco.

Pasten, C. (2002). E-mail correspondence regarding measurements taken at Compania Minera Zaldivar.

Weston, J.L. (14th October 2002). Internal Memorandum. *Summary of plating trials for Zaldivar.* MIBU - Townsville Operation.

ACKNOWLEDGMENTS

The authors would like to acknowledge the assistance and support provided by Guillermo Merino from Compania Minera Zaldivar, Christian Pasten from Lanzco. The authors would also like to acknowledge the permission to publish actual plant data from Compania Minera Zaldivar and results from John Nielsen's electrical conductivity tests.

OPERATIONS CONTROL IN XSTRATA TECHNOLOGY TANK HOUSES

C. Phan
Xstrata Technology
Box 5484, Mail Centre
Townsville, Queensland 4810
Australia
cphan@xstratatech.com.au

M. Oellermann
VRT Systems
Level 1/1 Gardner Close, Milton
Brisbane, Queensland 4064
Australia
marko@vrt.com.au

ABSTRACT

This paper outlines opportunities provided for enhanced management and process control by Xstrata Technology (XT). XT has the core advantage of offering the potential for vastly improved electrode handling systems. This fundamental characteristic has enabled XT to develop management and process control systems that are modular, minimalist in their complexity and tailored to a tankhouses operating environment. A review will be presented on the current status of management and process control in XT tankhouses covering the central issue of electrode handling as well as electrolyte circulation systems, voltage monitoring, cathode plate tracking, and product tracking. Particular reference will be made of the recent development by XT and their technology partners, VRT Systems, of a Radio Frequency Identification (RFID) based solution for tracking permanent cathode plates. The system encompasses fixed readers identifying plates passing through the stripping machines, and the use of sophisticated tracking and database software enabling the development of cathode asset management and operational strategies. The performance of individual plates and batches of plates can be tracked through a full service history.

INTRODUCTION

Xstrata Technology (XT) has the core advantage of offering the potential for vastly improved electrode handling systems. This fundamental characteristic has enabled XT to develop management and process control systems that are modular, minimalist in their complexity and tailored to a tankhouses operating environment. This modularity enables control systems to be tailored to the tankhouses operating environment and customer requirements.

TANKHOUSE MANAGEMENT

XT tankhouses deliver high intensity refining, maximum time efficiency and maximum current efficiency via managing and integrating all aspects of the process. Xstrata Technology Tankhouse Management System (XT-TMS) covers:

- Assets
- Energy (i.e. efficiency, etc…)
- Safety & occupational hygiene
- Environment (i.e. green house gases, etc…)
- Costs

PROCESS CONTROL

Within the XT-TMS the central issues of process control require monitoring and control specifically include:

- Electrode handling
  - Cranes
  - Anode preparation
  - Anode scrap washing
  - Cathode stripping
  - Cathode preparation
  - Anode scrap remelting
- Product tracking
  - Bundling
  - Copper cathode
  - Batch labelling
  - Packaging
  - Slimes
- Electrolyte circulation systems
  - Laboratory analysis
  - Online analysers
  - Circulation systems
  - Slimes preparation
  - Reagent management
  - Cell voltage and condition monitoring
  - Auxiliary – boilers, power, steam etc…
- Asset management
  - Cathode plate tracking
  - Maintenance

The XT-TMS centralises control from a single point irrespective of the level of automation. It can provide basic control over general operations or enable sophisticated interlocking control over the entire system encompassing machines and processes. External information from laboratory testing and product sampling is also available to complete the control loop.

Another sophisticated feature is the maintenance of historical records enabling reporting and performance tracking over time, a key feature of the system.

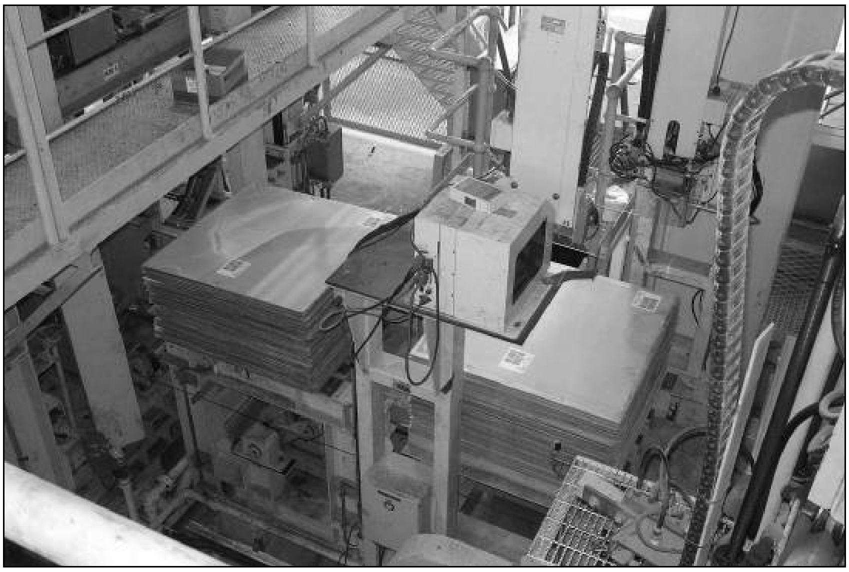


Figure 1 – Copper cathode labelling station

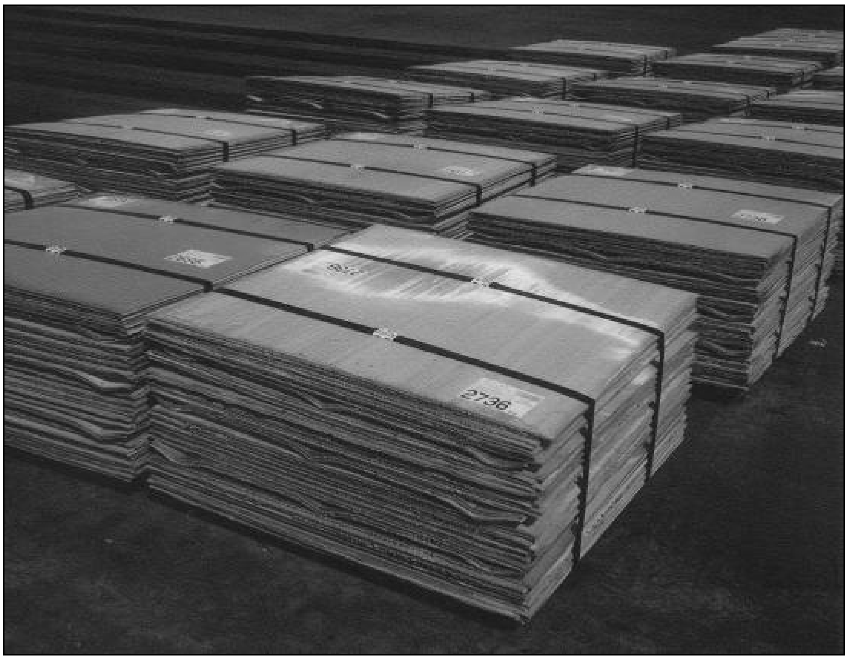


Figure 2 – Copper cathode bundle strapped and labelled

Distributed Control Systems

The Distributed Control System (DCS) monitors and maintains the operational parameters of the tankhouse. It is an automated system that will continue to operate the tankhouse effectively, without human input, while all aspects of the operations are functioning. The DCS is a computerised system that enables operational parameters to be maintained. Changes to the operation parameters, shutdown, start up and monitoring of the refining process can be done from Visual Display Units (VDUs) in the Tankhouse Control Room. Figure 3 shows an example of a graphic screen from a VDU. The DCS is made of three components:

(1) Process Control Unit (PCU),
(2) Distribution Control Unit (DCU), and
(3) Operator Stations (VDUs).

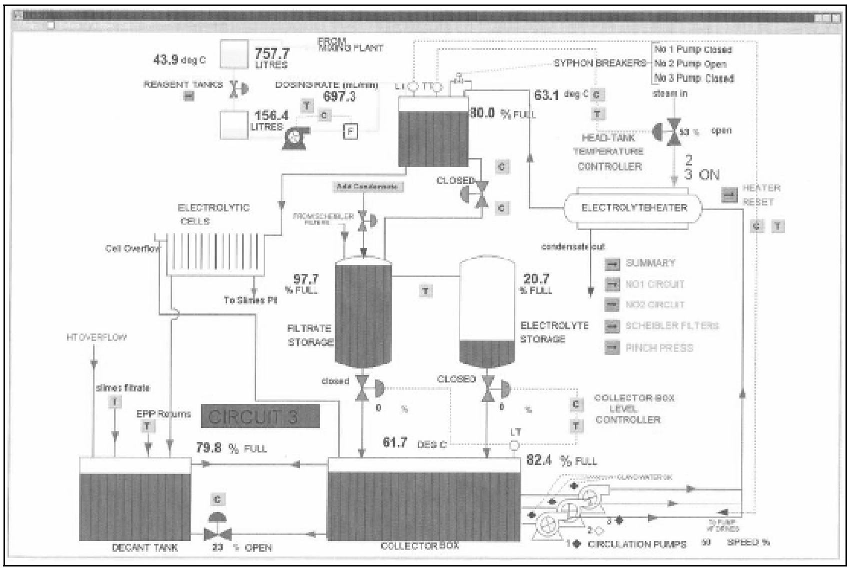


Figure 3 – Graphic screen from a Visual Display Unit

Tankhouse Machines

The major tankhouse machines comprise the Cathode Stripping Machine (CSM), see Figures 4 and 5, Anode Preparation Machine (APM), see Figure 6 and the Anode Scrap Machine (ASM), see Figure 7. Although these machines communicate and operate in line with other tankhouse units such as the overhead crane, their modular setup enables each machine to operate as a stand alone unit without affecting other operating units. Throughput capacities and automation levels for these machines can be designed to suit individual customers.

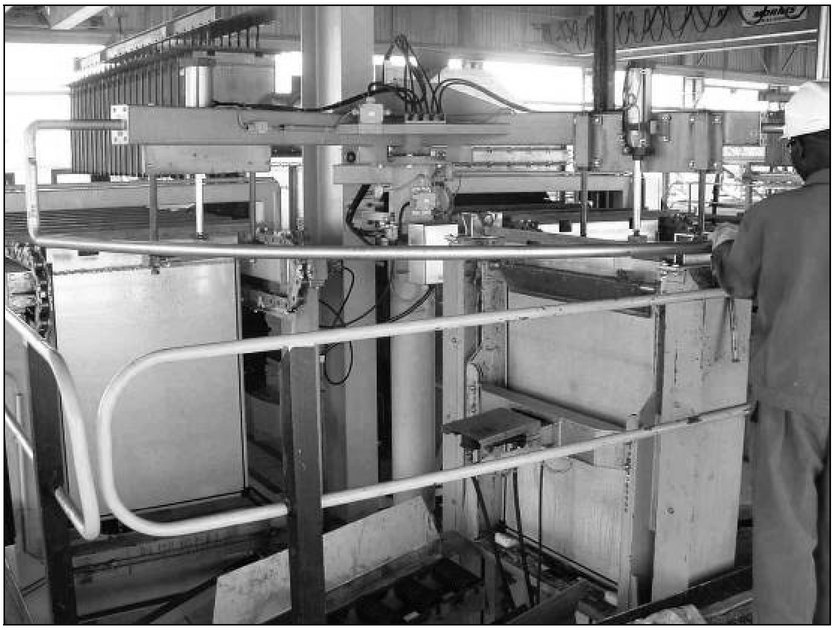


Figure 4 – Low automation Cathode Stripping Machine

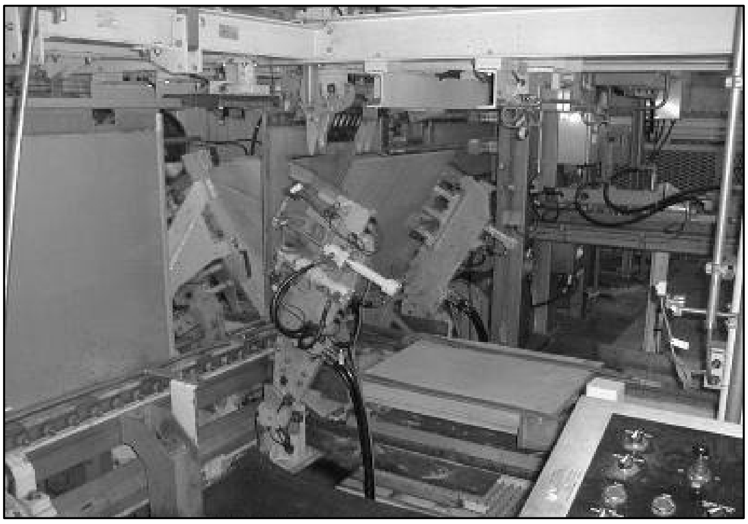


Figure 5 – High automation Cathode Stripping Machine

Figure 6 – Anode Preparation Machine

Figure 7 – Anode Scrap Machine

Tankhouse Overhead Crane

The cranes can be operated in either manual mode or automatic mode. In automatic mode, programs are stored in the crane's DCS and the crane will move automatically to either perform strip operations or strip and scrap operations for the programmed sections. The key function of the overhead crane provides reproducible results with respect to superior electrode handling and alignment. In manual mode, the operator overrides the automatic controls to operate the crane to fix problems encountered during an automatic operation. Manual mode is also used to perform other routine lifting operations in the Tankhouse.

Figure 8 – Tankhouse Overhead Crane

Voltage Monitoring

In order to achieve an efficient refining operation in a tankhouse, it is important to detect all abnormal cell voltages and take corrective action as soon as possible. There are several methods to detect short circuits; these include Gauss and Hall-Effect Meters, multimeters, thermal imaging and computer cell voltage monitoring. These methods can be used individually or in combination with each other. The type of short circuit detection method required is generally dependant on labour cost versus plant capacity, and losses due to current efficiency.

XT offers all the above mentioned methods for shorts detection. The Cell Voltage Monitoring System (CVMS) XT offers is used to measure, record and monitor the voltage across each cell. Each cell voltage is measured and compared against preset limits. If the voltage is out of limits an indication is given. Each CVMS is individually designed to match the specific requirements of the particular tankhouse application and is available as a complete package. Figures 9 and 10 are graphic displays from the XT-TMS extracted from a CVMS showing section and cell voltages.

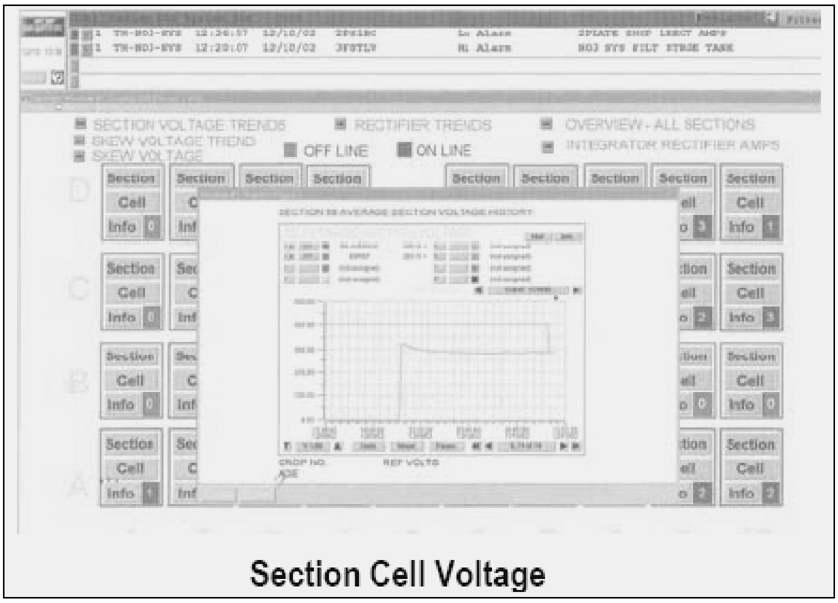


Figure 9 – Graphical display of section cell voltage

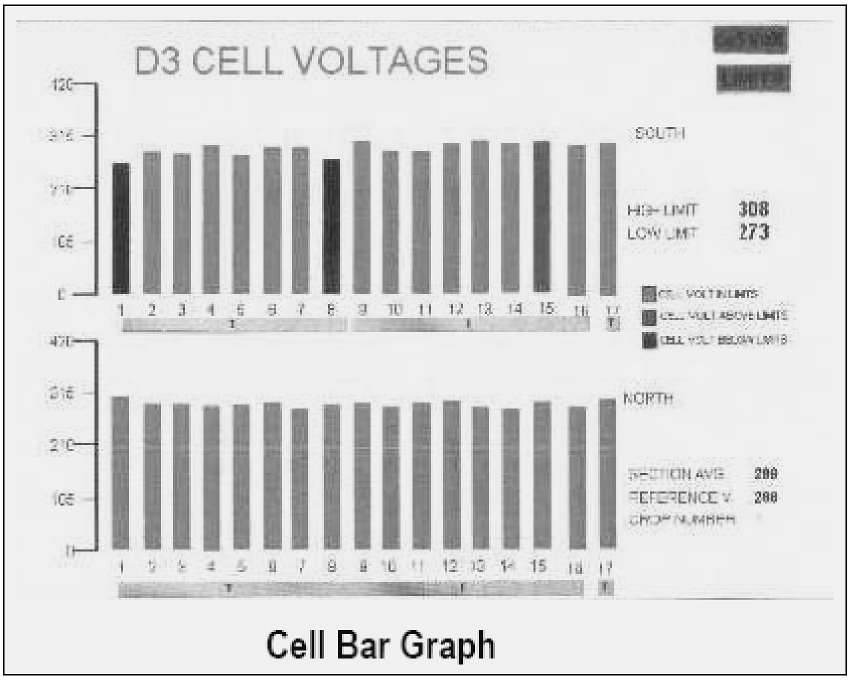


Figure 10 – Bar Graph of section cell voltage

RADIO FREQUENCY IDENTIFICATION (RFID) SYSTEM

Introduction

XT and their technology partners, VRT Systems, have developed an RFID based solution for tracking permanent cathode plates. The system encompasses fixed readers identifying plates passing through the stripping machines and the use of sophisticated tracking and database software enabling the development of cathode asset management and improved operational control strategies, see Figure 11. The performance of individual plates and batches of plates can be tracked through a full service history.

Some of the opportunities that have been identified are:

- Tracking current efficiencies for individual and groups of cathode plates.
- Tracking weights of individual cathodes with significant implications for improvements in tankhouse operations.
- Tracking edge strip performance and failure rates.
- Tracking cathode quality by means of cathode to anode positioning (in EW operations where anodes are fixed).
- Distinguishing between old and new plates.
- Tracking cathode plate performance in segregated circuits.
- Tracking cathodes plates requiring the use of plate repair facilities.

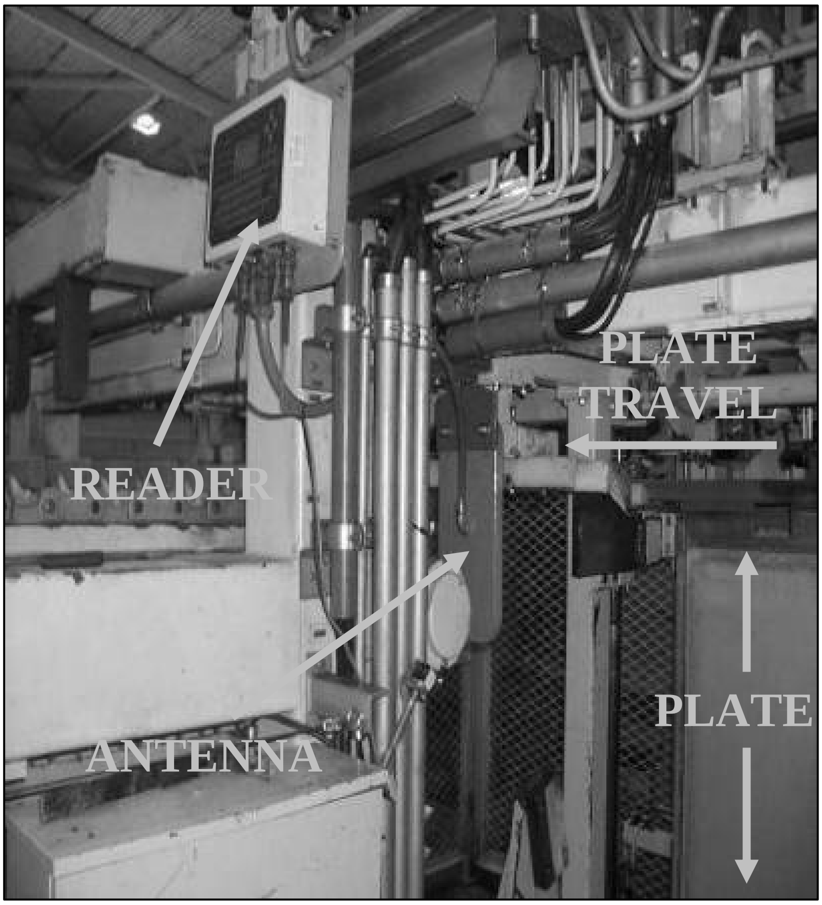


Figure 11 – RFID reader mounted to a Cathode Stripping Machine

Development Phases / Methodology

The extreme operating environment resulted in a focus to date on resolving the issues around reliability of hardware associated with readers, tags and tag housings.

Early development work has centred on configuration of the readers to ensure that all necessary data is captured. Positioning of the readers on the Cathode Stripping Machine (CSM) will be dependant on individual layout/setup of the machine. The aim is to achieve a closed loop system such that every individual cathode plate can be traced to a particular location.

Ideally there would be one reader on the inlet of the machine to read every plate that passes through the CSM, one reader to capture the plate once it has been stripped and transferred by the transfer conveyor and ready to go back into the tankhouse, one reader on the reject conveyor to capture all plates that are rejected for any particular reason, such as damaged edge strips or sticky copper.

New plates or repaired plates that are returning to the tankhouse need to be read before going into service. This can either be done by feeding the plates through the CSM (which is required if it is a bottom wax operation), or the plates can be read using another fixed or handheld reader and the information transferred onto the database.

Various tag designs with respect to tag housing material and tag profile have been trialled to determine the most suitable design. The tag housing most suitable will be one that fits the following criteria: ease of fitting to the cathode plate, material suitability to handle the corrosive and harsh tankhouse environment, and longevity of tag life.

Tag Position on Plate

Environments with high levels of metal pose specific challenges for RFID implementation and generally require special metal-capable RFID tags. Metal-capable RFID transponders usually rely on a surface field-effect to operate and have specific mounting position and orientation requirements to achieve sufficient read reliability. In contrast, the RFID solution developed by XT and VRT allows tags to be mounted in a hole punched in the plate, enabling a single tag to be scanned from either side of the plate. Additionally this reduces the overall tag profile and provides maximum flexibility in tag placement on the plate surface.

XT has identified several locations on the cathode plate to insert the tag. Location will be dependant on individual layout of the CSM. Tags can either be fitted to the centre of the plate just below the copper plating extended from the hanger bar (this is the ideal position, however, this may not suit all CSM layouts), see Figure 12. The position of the tag can be modified to suit both the CSM layout and cathode plate configuration of individual customers.

Since the plates operate in a closed system and no data needs to be stored on the actual tags, a read-only tag inlay has been selected. These have a guaranteed globally-unique 16 digit number which is assigned at time of manufacture. The tag inlays are mounted in a protective housing specifically designed for the conditions found in acidic electrolytic processes.

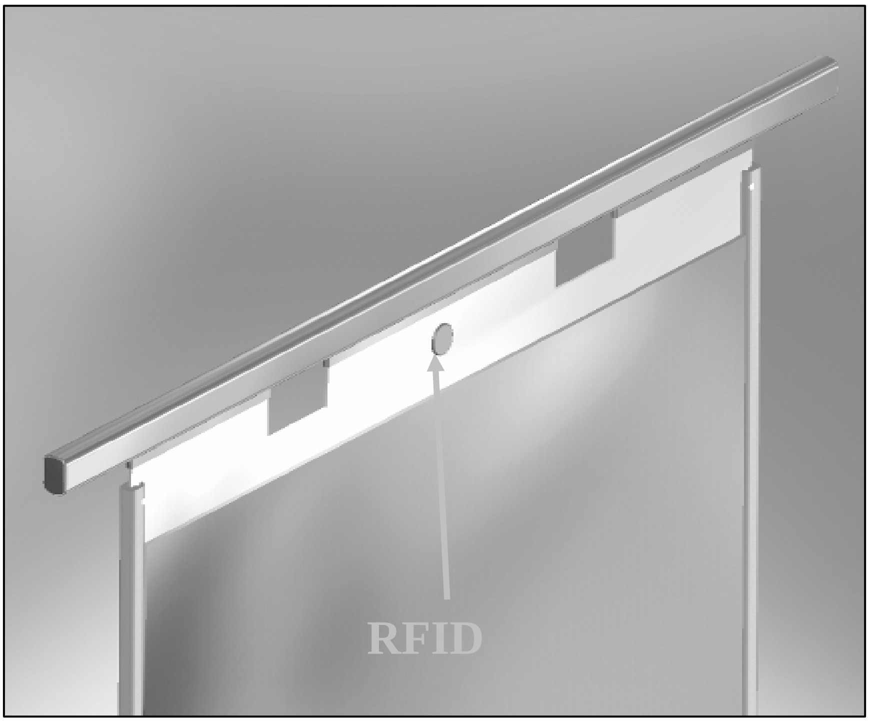


Figure 12 – Schematic of RFID tag positioned on stainless steel cathode plate

The System

The RFID readers and tags form part of a broader plate tracking and asset management system under development by XT. The system comprises RFID tags and readers, coupled with a controller running Plate Asset Management Software to provide a complete tracking and asset management solution. The system is intended to offer asset owners the means to analyse plate performance through life, identify and locate problematic or under-performing plates, determine the optimal end-of-life time for plate batches, and trace plates through offsite or contracted maintenance.

Tracking Functionality

The plate tracking system incorporates a number of key capabilities:

- Identity Preservation – Fundamental to whole-of-life tracking is the preservation of plate identity. The system separates the plate ID from its RFID tag so that tags can be replaced and the tracking system retains a consistent plate identity (serial number) for analysis purposes.
- State Tracking – The system can be configured to record plates in any number of states, (e.g. “in service”, “stripping”, “rejected – awaiting repair”, “quarantined”, “in repair”, “awaiting return to service” and so on). This enables analysis of duty cycles and overall asset utilisation.
- Location Tracking – The system tracks individually identified plates through any number of discrete logical locations. The locations are configurable to support various tank house capacities and layouts, and other on-site and off-site maintenance locations, including mobile locations such as vehicles. Locations are defined in a hierarchy and can be configured to the resolution required – for example: site/tank-house/section/cell, or site/area/workshop.

- Plate Grouping – As plants use plates of various ages (and sometimes of slightly different types as improvements to plate design are made), it is beneficial to compare plates with similar age or type. The system tracks plates by plate batch/group, so that queries and analysis can be performed by each batch of plates within a plant.

Additional Data Capture

RFID readers and tags provide essential data for tracking of plates through the system, but can only reveal limited statistics surrounding cycle times and failure rates. In order to provide further value, the plate tracking system is designed to allow additional qualitative data to be captured along with associated key process elements:

- Plate characteristics: manufacture date, batch number, material thickness, edge strip type etc.
- Cell conditions: profile of electricity consumption, electrolyte details etc.
- Plate transfer details (e.g. removal of plate from cell): weight of copper (when removed at stripping machine).
- Inspection and maintenance details: Reason for rejection, maintenance carried out etc.

The capture of this data is enabled by providing interfaces to external data capture systems:

- Facilities for interfacing to a range of PLC, DCS and instrumentation systems via optional embedded supervisory control and data acquisition (SCADA) software. This provides opportunities for interfacing to automated and semi-automated crane handling systems to provide cell-level plate tracking in the tank house.
- Screen based manual input (console and web browser based).
- Interfaces to wired and wireless field devices (PDA and mobile RFID scanners) for plate inspection and maintenance activities.

Equipment

The RFID readers are purpose-built industrially hardened, sealed units with onboard data logger, onboard keyboard and screen, and a serial interface to the tracking system. The units support a read range of 200mm, read speed of 70 milliseconds, and operating temperature range of -30 °C to 70 °C.

Additional handheld RFID scanners are being used for manual plate scanning, for survey and pilot system verification. These are battery-operated units designed for the tank-house environment and all-weather outdoor use (-30 °C to 70 °C). They can support RF wireless, cable or blue-tooth connectivity.

Xstrata Copper – Townsville Operations Configuration

The current development system is operating on two stripping machines; each fitted with fixed RFID readers on the CSM inlet and reject ports. The tracking software infers plates returned to the tank-house (those scanned on the inlet port, but not rejected).

The plate tracking software is currently running on an industrial PC server, connected to the fixed RFID scanners via serial line drivers.

Performance

The RFID readers are modified from existing industrial process application, so are already field-proven in harsh environments. The RFID tag inlays have proven reliable and read rates in the current configuration are high for an RFID application of this type.

The current RFID tag housing under trial is a single-piece "snap-in" polymer design that does not require adhesive or fasteners to affix to the plate. There have been some issues with material deformation causing a small number of tags to come away from the plate. This is due to in-process degradation of the polymer used to manufacture the prototype units, however, other materials with known resilience in a tank-house environment are now being tested. Further developments underway may see the configuration of the tag housing change.

The plate tracking and asset management software is still being refined, but has proven to be stable and reliable.

The read reliability of the system is exceptional with a readability of 99.95% to date. It is believed this a significant improvement on other systems currently available on the market with quoted accuracies of >95%, [1].

Development

The most price-sensitive component in the plate asset management system is the RFID tag, and is consequently an area of continuing development and refinement. Improved tag housing designs and materials are being tested, with the goal of minimising installed cost, while at the same time maximising tag longevity.

The plate asset management software is being refined and tailored to suit the requirements for plate asset management and process improvement in XT tankhouses. This will likely be an area of ongoing development as further opportunities for improvement are identified.

Further development work is being undertaken to improve the system packaging. The system may be available as a server based solution, but efforts are being considered to package the tracking engine software in a rugged appliance-based form factor:

- Suitable for rack or DIN-rail mounting in electrical control cabinets.
- Ethernet connectivity, removable non-volatile memory card for configuration storage, network backup and/or remote database.
- Zero- or minimal- configuration to set up – allow tracking engine to be swapped out by field staff for easy maintenance and repair.
- Web browser based user interface for configuration and operation.
- Facilities for Open Database Connectivity (ODBC) for ad-hoc analysis and reporting capabilities. Acknowledgments

CONCLUSIONS

Xstrata Technology has developed management and process control systems that are modular, minimalist in their complexity and tailored to a tankhouses operating environment. XT offers a complete tankhouse management system with respect to electrode handling as well as electrolyte circulation systems, voltage monitoring, cathode plate tracking, and product tracking.

Developments by Xstrata Technology and their technology partners, VRT Systems, of an RFID based solution for tracking permanent cathode plates have proven to be highly successful. The read reliability of the system is exceptional with a readability of 99.95% offering a significant improvement on other systems currently available on the market.

ACKNOWLEDGMENTS

The authors would like to thank Xstrata Technology for allowing them the time and resources to undertake this work and publish this paper. Special thanks to the Xstrata Copper Refinery in Townville for allowing the trials to be perform in their tankhouse. This project has had input and assistance from a number of people at Xstrata Copper not least of which is Noel Kimlin, Les Brand, Bruce Hall and Peter Hall as well as Kellie Eastwood and Tony Ruddell from the Xstrata Technology development team.

REFERENCES

1. CSC, http://www.csc.com/solutions/supplychainmanagement/casestudies/2213.shtml sited 20 December 2006.

2. F. Westlake, L. Brand and B. Jenkins, “Shorts Detection and Correct”, Xstrata Copper Training Package TRA-641101, Townsville Operations, Queensland, Australia, 2003.

3. L. Law and D. Fowles, “Distributed Control Systems”, Xstrata Copper Training Package TRA-641003, Townsville Operations, Queensland, Australia, 2003.

4. F. Westlake, G. Lynch and P. Clifford, “Cathode Stripping Machine”, Xstrata Copper Training Package TRA-638301, Townsville Operations, Queensland, Australia, 2003.

5. F. Westlake, D. Robertson, P. Cutts and P. Kopittke, “Kunz Crane Operations”, Xstrata Copper Training Package TRA-638101, Townsville Operations, Queensland, Australia, 2006.

Hydrometallurgy 106 (2011) 113–118

Contents lists available at ScienceDirect

Hydrometallurgy

journal homepage: www.elsevier.com/locate/hydromet

Statistical analysis of the effect of operating parameters on acid mist generation in copper electrowinning

Reza Al Shakarji [a], Yinghe He [a,*], Simon Gregory [b]

[a] *School of Engineering and Physical Science, James Cook University, QLD 4811, Townsville, Australia*
[b] *Xstrata Technology, Copper Refinery Limited, Hunter St, QLD 4811, Townsville, Australia*

ARTICLE INFO

Article history:
Received 27 July 2010
Received in revised form 13 December 2010
Accepted 15 December 2010
Available online 23 December 2010

Keywords:
Copper electrowinning
Acid mist
Oxygen evolution
Surface tension
Viscosity
Mist suppressants

ABSTRACT

Acid mist is generated during the final stage of hydrometallurgical metal refining processes including the electrowinning of copper. In this study, the effect of five process parameters and their interactions on the amount of acid mist generated is analysed quantitatively. The amount of acid mist generated was measured under 32 different operating conditions. It was found that solution's temperature and mist suppressant chemical FC-1100 had significant effect on the amount of acid mist generated. More than 90% of the variations in the acid mist generation can be explained by changes in these two parameters and their interaction. To a lesser extent, electrical current density and solution acidity also affected the total amount of acid mist generated. The anode's age and most of the 3, 4, and 5-way parameter interactions were found to have negligible influence on the amount of acid mist. Overall, acid mist was found to increase with temperature and current density. In contrast, increasing the viscosity of the solutions tends to decrease the amount of acid mist.

1. Introduction

Electrowinning is an electrochemical process that is used to extract metal from its solution, and is extensively used in the production of copper. More than 20% of the world's primary copper is produced through electrowinning (Davenport et al., 2002).

In the electrowinning of copper, a direct electrical current is passed between an anode and a cathode that are submersed in a copper-rich solution (Robinson et al., 1994). At the inert anode, water molecules are electrolysed and oxygen bubbles are formed on the surface of the anode. These oxygen bubbles grow and eventually detach from the surface and rise in the bulk of the solution. These bubbles burst at the free surface of the solution and produce highly acidic droplets; of which the fine ones become airborne and form an acid mist throughout the tankhouse of the electrowinning plant.

Acid mist is highly corrosive and results in the corrosion of cathode plates, anode's hanger bar, tankhouse equipment and building structures. Acid mist also poses a serious health hazard and causes extreme discomfort to the skin, eyes and respiratory systems of the tankhouse workers (HSIS, 2009). The Occupational Safety and Health Administration (OSHA) recommends a time-weighted average (TWA) exposure limit of 1 mg of sulphuric acid per m^3 of air, and a short term exposure limit (STEL) of 3 mg m^{-3} (OSHA, 2003).

There have been many attempts to eliminate or minimize acid mist in copper electrowinning operations (Mella et al., 2006). Polyethylene balls, suction hoods, mats, brushes and wipers, chemical reagents and forced ventilation are examples of such attempts (3M, 2007; Davis and Eng, 2002; Hooper, 2008; Mella et al., 2006; Sunwest, 2004). Qualitatively, the use of chemical reagents such as FC-1100, Mistop, Dowfroth, and alkylated ethoxylates has been rated as the most effective method of suppressing acid mist (Bender, 2010). However, there have been no systematic studies to quantitatively compare the effect of different operating parameters, including the use of a chemical reagent, on the amount of acid mist generated. Most of the published works to date, have only examined the effect of one or two parameters individually without considering any possible interaction effects on the amount of acid mist (Alfantazi and Dreisinger, 2003; Cheng et al., 2004; Hosny, 1993; San Martin et al., 2005a,b; Sigley et al., 2003).

This paper examines, quantitatively, the relationship between the amount of acid mist generated and five operating parameters. These parameters are the age of the anode, electrical current density, solution temperature, sulphuric acid concentration of the solution, and the presence of a typical chemical mist suppressant (i.e. FC-1100). The results are useful for the design of more efficient methods or systems for acid mist minimization at electrowinning tankhouses.

---

* Corresponding author. Tel.: +61 7 47814270; fax: +61 7 47816788.
E-mail address: yinghe.he@jcu.edu.au (Y. He).

0304-386X/$ – see front matter. Crown Copyright © 2010 Published by Elsevier B.V. All rights reserved.
doi:10.1016/j.hydromet.2010.12.013

2. Methodology

2.1. Equipment set up

The copper electrowinning process was replicated in a bench-scale cell. This cell (C2 in Fig. 1) was constructed of 10 mm thick clear acrylic and had a capacity of 6 L. During each test, electrochemical reactions resulted in continuous copper depletion and acid generation in the solution. Therefore, a peristaltic pump was utilized for gradual and continuous addition of fresh solution, from C1 container to C2, to maintain consistency in the composition of the solution. A horizontal slit in the side of C2 was utilized to keep the level of the solution in the container constant.

Four submersed heaters were placed in the corners of C2 to keep the solution at a constant temperature for the duration of each test. To replicate industrial operations, Pb–Ca–Sn alloy and 316 L stainless steel were used as anode and cathode, respectively (Houlachi et al., 2007). For each set of experiments a fresh batch of solution was synthesized that contained 45 g L^{-1} Cu, 15 mg L^{-1} Guar gum, 20 ppm Cl and 100 ppm Co. The synthesized solution was similar to the electrolyte solutions found in most copper electrowinning tankhouses worldwide (Robinson et al., 1994). The sulphuric acid concentration in the solution, however, was one of the five selected variables and its concentration differed from that of a typical copper electrolyte solution.

Nitro cellulose filters were used to capture acid mist. These filters were 47 mm in diameter and had a pore size of 0.45 μm which ensured the capture of very fine acidic droplets. For each experiment, a fresh filter paper was installed at 45 mm above the free surface of the electrolyte solution inside the electrowinning cell (C2). The filter was held by an inverted funnel (F in Fig. 1) and connected to a vacuum pump (VP) via a pneumatic tube. For precise air flow measurements, the drawn air was dehumidified by passing it through an enclosed flask that contained silica beans (D). A flow meter (FM) with a built in valve was installed on the pneumatic tube to ensure a constant air flow of 5 L min^{-1} through the filter for the duration of each experiment.

At the end of each test, the used nitro cellulose filter was removed from the cell and placed in 25 mL of deionized water and stirred for about 60 min. The pH of the solution was then measured and the amount of the captured acid was calculated as grams of sulphuric acid per cubic meter of air drawn through the filter.

2.2. Experimental design

The main goal of the present work was to compare, quantitatively, the effect of different operating parameters on the amount of acid mist generated during a typical copper electrowinning process. To fully explore the main effect of each individual parameter as well as any possible interaction effects, the 2 K factorial method was utilized to determine the required experimental conditions (Montgomery, 2005). In this method, the examined parameters are tested at two levels (low and high). The bigger the difference between the low and high levels of a parameter, the more reliable its effect measurements would be (Montgomery, 2005). Table 1 illustrates the examined parameters and their low and high level values.

The values shown in Table 1 were selected so that the midpoint between low and high levels of each parameter represented the typical value used in most copper electrowinning tankhouses (Houlachi et al., 2007). For example, the temperature of the electrolyte solution is usually kept at about 45 °C. Thus, the low and high limits of temperature for the experiments were set at 30 °C and 60 °C, respectively.

Based on the 2 K factorial method, 32 parameter combinations were required to fully examine the effect of five parameters at two levels. For reliable data analysis, a minimum of two replicates were needed for each test condition. Hence, 64 tests were conducted to determine the influence of each parameter on the generation of acid mist and also to identify any significant interactions amongst the selected parameters.

3. Results and discussion

3.1. Descriptive statistical analysis

In all the 64 experiments at 32 different operating conditions, the least amounts of detected acid mist were 0.03 and 0.02 mg acid m^{-3} air. These two tests were repeats, conducted with an old anode at low current density in a solution with low acidity, low temperature and high concentration of FC-1100.

The highest amounts of detected acid mist were 114.4 and 117.1 mg acid m^{-3} air. These two repeats were conducted with a new anode at high current density in a solution with low acidity, high temperature and no FC-1100.

The test results proved to be highly repeatable. The average relative error between the repeats of medium–high acid mist concentration tests was 18%. The smallest relative error was 0.001%. For the very low acid mist concentration tests, however, the average relative error between the repeats was 41%. This high relative error was essentially due to the resolution limit of the measurement method.

To distinguish the effect of individual parameters, the average amount of collected acid mist for each level of a parameter was calculated. The bar graph in Fig. 2 illustrates the results where each bar represents the average amount of acid mist collected from 32 independent experiments.

In Fig. 2, it can be seen that highest amounts of acid mist were generated when the solution was kept at high temperature or when no FC-1100 was added to the electrolyte solution. To a lesser extent,

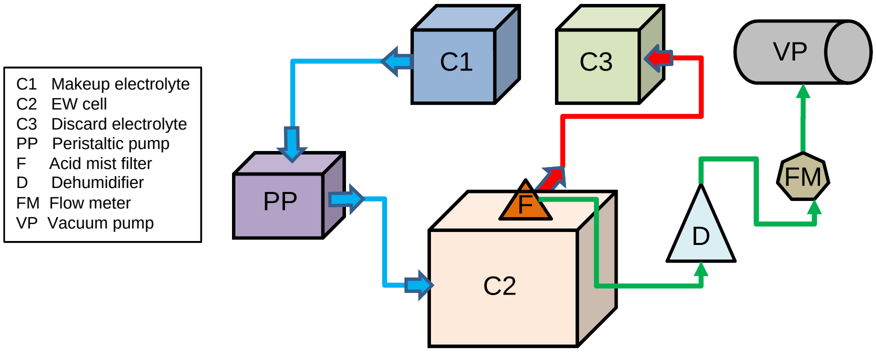


Fig. 1. A schematic view of equipment setup.

Table 1
The selected test variables and their values.

No	Examined parameter	Low	High
1	Anode age (months)	0	6
2	Current density (A m^{-2})	200	400
3	Temperature (°C)	30	60
4	Acidity (g L^{-1})	100	250
5	FC-1100 (ppm)	0	30

high current density, low sulphuric acid concentration in the solution and new anode also favoured higher acid mist generations.

High temperature solutions produced the most amounts of acid mist. Temperature is known to affect the rheological properties of fluids. Surface tension and viscosity of the electrolyte solution were known to strongly influence the final size and the burst process of the oxygen bubbles (Xie et al., 2009; Xu et al., 2009). Thus, to find an explanation for the strong influence of temperature on acid mist, the effects of temperature on both the surface tension and the viscosity of the solution were evaluated and the results are discussed in detail in Section 3.3.

The second highest amount of acid mist belonged to the tests where no FC-1100 was added to the solution. The addition of 30 ppm of FC-1100 reduced the acid mist amount from 47.2 to 7.3 mg acid m^{-3} air. As will be seen later, this reduction in the amount of acid mist in the presence of FC-1100 is believed to be caused by a change in the burst mechanism of bubbles at the free surface of the solution through changes in the surface viscosity and surface elasticity.

The third highest acid mist concentration belonged to high current density experiments. Based on Faraday's law, electrical current is directly related to the rate of chemical reactions that occur during an electrochemical process (Harris, 2007). Decomposition of water molecules at the surface of the anode is the main anodic reaction that takes place during the copper electrowinning process. Therefore, based on Faraday's law, doubling the electrical current density will double the total volume of the generated oxygen bubbles.

Our previous experimental work suggested that changing the electrical current density had negligible effect on the final sizes of the bubbles that detached from the anode (Al Shakarji et al., 2010). Consequently, doubling the current density will approximately double the number of oxygen bubbles generated. This increase in the number of bubbles (per unit time) with current density results in a net increase in the number of bubbles that burst at the free surface of the solution which in turn produces a higher number of airborne acidic droplets (i.e. acid mist) per unit time per unit surface area of the solution.

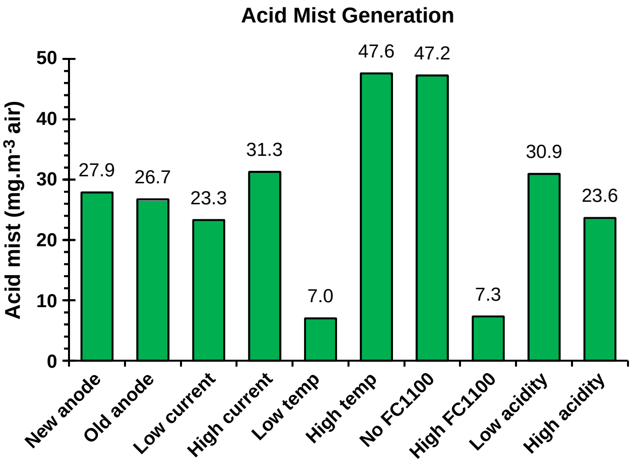


Fig. 2. Mean acid mist generation at different operating conditions.

3.2. Quantitative statistical analysis of test parameters on acid mist generation

A 5-way ANOVA analysis was conducted on the raw experimental data to determine the full effect of test parameters on the amount of acid mist generated. This analysis returned p values less than 0.05 (i.e. significant at 95% level) for 18 of the 32 possible parameter combinations. Cohen classifies the magnitude of a parameter's effect into three categories of small, medium and large (Cohen, 1988). These categories correspond to R^2 values of 0.01, 0.09 and 0.25, hence accounting for 1%, 9% and 25% of the total variance, respectively (Cohen, 1988). To quantify the influence of the aforementioned 18 conditions on acid mist generation, the R^2 value was calculated for each case based on its Pearson's r value (Field, 2005). The top ten cases (with respect to R^2) are shown in descending order in Fig. 3.

The quantitative analysis (Fig. 3) shows that temperature, FC-1100 and the interaction of these two parameters are the most influential parameters in determining the amount of acid mist generated. To a lesser degree, current density and solution acidity also affected the generation of acid mist. Anode's age with an R^2 value of 8×10^{-5} was determined to be a parameter with negligible effect on the amount of acid mist generated.

The ANOVA analysis returned an overall adjusted R^2 value of 0.986, which meant 98.6% of the variation in acid mist recorded in the series of experiments could be explained by changes in the five parameters. The uncontrolled variables and instrumentation errors accounted for only 1.4% of the measured acid mist variations which implied the experiments were conducted at a highly controlled environment.

The influence of individual test parameters as well as that of important interactions (shown in Fig. 3) on acid mist is discussed in more details in the following sections.

3.2.1. Effect of temperature and FC-1100 on acid mist

Based on Cohen's classification, both temperature (T) and FC-1100 (S) parameters with R^2 values of 0.378 and 0.366, respectively, proved to have large effects on the amount of acid mist generated. The interaction of these two influential parameters (T*S) with an R^2 value of 0.159 had a medium effect on acid mist. These effects can be seen graphically in Fig. 4.

Fig. 4 also shows that the presence of FC-1100 in the electrolyte solution strongly influenced the effect of temperature on the amount of acid mist generated. In the absence of FC-1100, a 30 °C increase in the solution temperature increased the amount of acid mist by almost 67 mg m^{-3} of air whereas in the presence of FC-1100 the same increase in the solution temperature resulted in only 14.3 mg more acid mist per m^3 of air.

The vast difference in the slopes of the two lines in Fig. 4 indicated the strong interaction between temperature and FC-1100. The T*S interaction alone accounted for nearly 16% of the variations seen in the amount of acid mist generated. Overall, temperature (T), FC-1100 (S), and T*S have a combined R^2 value of 0.903 which means that

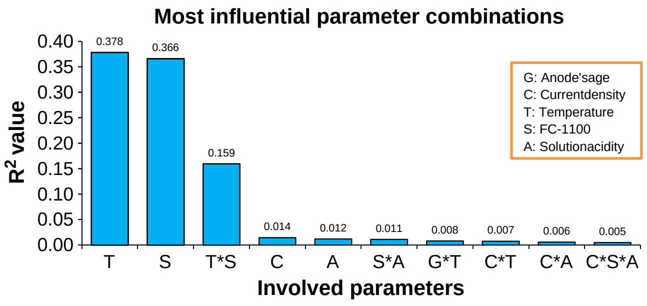


Fig. 3. Quantified influence of different parameters and parameter combinations on the amount of acid mist generated.

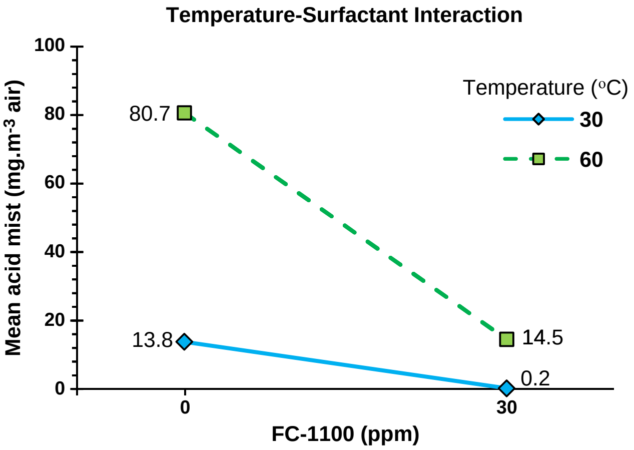


Fig. 4. The temperature–FC-1100 interaction based on mean acid mist values.

changes made in these three parameters are sufficient to explain more than 90% of the variations seen in the amount of acid mist generated.

3.2.2. Effect of current density on acid mist

Current density with an R^2 value of 0.014 was the fourth influential parameter in the amount of acid mist generated. In absolute terms, the effect of current density on acid mist is much less than that of temperature and FC-1100 and based on Cohen's classification the effect of current density is considered small. Nevertheless, in relative terms, current density is an important factor as the amount of acid mist generated at 400 A m^{-2} was 34% more than that generated at 200 A m^{-2} (Fig. 2).

To compare the magnitude that current density and FC-1100 affect the amount of acid mist generated, the averaged acid mist measurements were plotted at different current densities with or without the presence of FC-1100. The results are shown in Fig. 5.

Fig. 5 confirms again a substantial reduction in acid mist at both low and high current densities when FC-1100 is added to the electrolyte solution. Since there is no significant difference in the slope of the two lines in Fig. 5, it means that current density has no significant influence on the effect of FC-1100 on acid mist generation, i.e. little interaction effect between the current density and FC-1100.

Fig. 5 also shows that, regardless of the presence of FC-1100, an increase in current density increases the amount of acid mist generated. However, the acid amount did not double when the current density was doubled. This is because acid mist is produced from the burst of bubbles that ejects droplets into the air of which some become airborne. The amount of mist generated from the simultaneous bursts of two neighbouring bubbles is known to be less than the sum of mist amount from the bursts of two individual bubbles separately, due to interferences between the bursts of the bubbles. At a higher current density, the likelihood for a number of neighbouring bubbles to burst simultaneously is higher. Consequently, even though the number of oxygen bubbles must be doubled when the current density is doubled, based on Faraday's Law and that the bubble size does not change with current density, the acid mist amount is expected to be less than double.

3.2.3. Effect of solution acidity on acid mist

The ANOVA analysis returned an R^2 value of 0.012 for the solution acidity which meant this parameter was the fifth most influential parameter in acid mist generation, which is comparable to that of the current density (0.012 vs 0.014). This means that the acidity of the solution is approximately as influential as the current density in acid mist generation.

The C*A interaction is listed as one of the top ten influential parameters (Fig. 3). This interaction is confirmed from the considerable difference in the slope of the two lines shown in Fig. 6. For high acidity solutions, doubling the current density from 200 to 400 A m^{-2} only resulted in 13% increase (equivalent to 0.087 standard deviations) in the amount of acid mist. In contrast, doubling the current density in low acidity solutions resulted in a staggering 54% increase (equivalent to 0.395 standard deviations) in the amount of acid mist.

Fig. 7 shows the effects of solution temperature and acidity on the acid mist amount. It again confirms the profound effect that temperature has. Further, the two lines in Fig. 7 have similar slopes (1.46 for low acidity solutions and 1.24 for high acidity solutions), which suggests that the interaction between temperature and acidity is negligible.

The plots in both Figs. 6 and 7 reveal an interesting phenomenon that low acidity solutions consistently resulted in higher amount of acid mist than high acidity solutions did, regardless of the applied current density and temperature. This apparent counter-intuitive result will be explained in detail in Section 3.3 below.

3.3. The role of surface tension and viscosity in acid mist generation

The previous two sections presented statistical observations on the influence of operational variables and their interactions on the amount of acid mist generated. This section offers mechanistic explanations to the results obtained in this investigation.

3.3.1. Effect of surface tension

Qualitatively, temperature and FC-1100 are both known to affect the surface tension of a solution. To quantify their effect on the

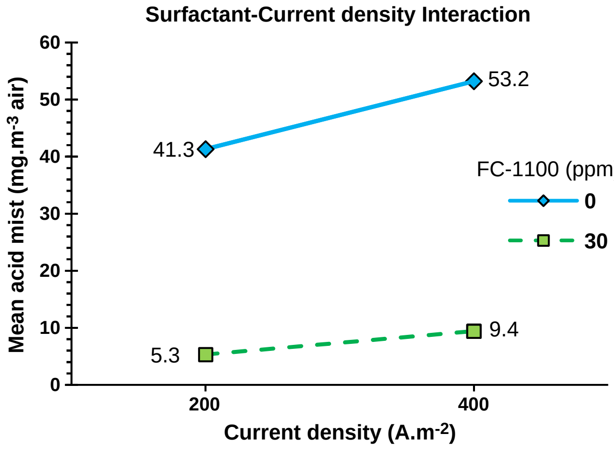


Fig. 5. The effect FC-1100 on acid mist at different current densities.

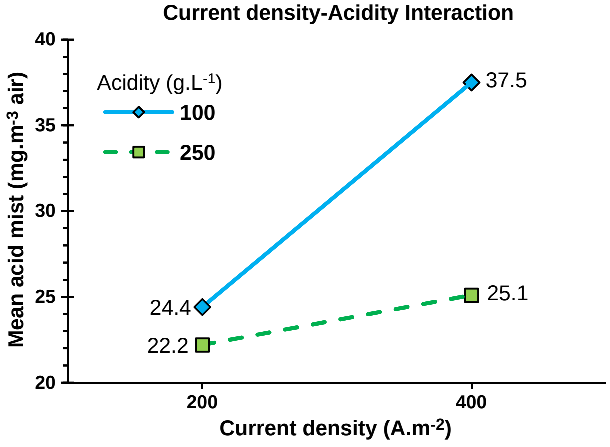


Fig. 6. The effect of solution acidity on acid mist at different current densities.

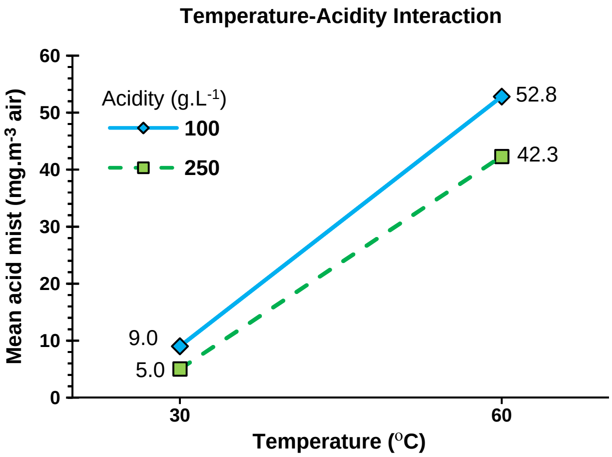


Fig. 7. The effect of solution acidity on acid mist at different temperatures.

electrolyte, the surface tension of 7 samples was measured by the Wilhelmy's plate method at two different temperatures. The averaged results are shown in Fig. 8.

It can be seen that temperature strongly influences the surface tension of the electrolyte in the absence of FC-1100. On average, the surface tension decreases by 32% when the temperature is raised from 30 to 60 °C in the absence of FC-1100. In contrast, in the presence of FC-1100 at 30 ppm, the temperature change has a negligible effect (less than 3%), as shown in Fig. 8. These observations are similar to those seen in Fig. 4 in that, in the absence of FC-1100, the difference caused in the amount of acid mist by temperature change is much larger. Both figures demonstrate a strong interaction between temperature and FC-1100.

It is important to note that, while both increasing solution temperature and addition of FC-1100 cause a significant decrease in the surface tension of the electrolyte, their effect on acid mist generation is the opposite. For example, at 30 °C the addition of 30 ppm FC-1100 reduces surface tension to 44 mN m^{-1} and results in 0.41 standard deviations reduction in acid mist generation. Increasing the temperature from 30 to 60 °C, in the absence of FC-1100, also reduces the surface tension to almost the same value (43 mN m^{-1}). However, this has resulted in an increase of 2.01 standard deviations in acid mist amount.

In the absence of a surface active agent, lower surface tension or lower surface energy, means lower amount of energy required for the generation of new surfaces. In other words, at a lower surface tension, it is more likely to generate higher number and smaller size of liquid droplets from the burst of bubbles. In addition, previous studies also show that the amount of liquid droplets produced from the burst of gas bubbles increases exponentially with decreasing diameter of the bubbles (Liow et al., 2007; Liow and Gray, 1996). That is, the burst of smaller gas bubbles produces more liquid droplets, or mist. In this study, when the temperature is increased, the surface tension is reduced. This has resulted in not only an increase in the likelihood to produce a higher number and smaller size of liquid droplets, but also a reduction in the size of the oxygen bubbles detaching from the anode, causing an increase in the amount of acid mist produced.

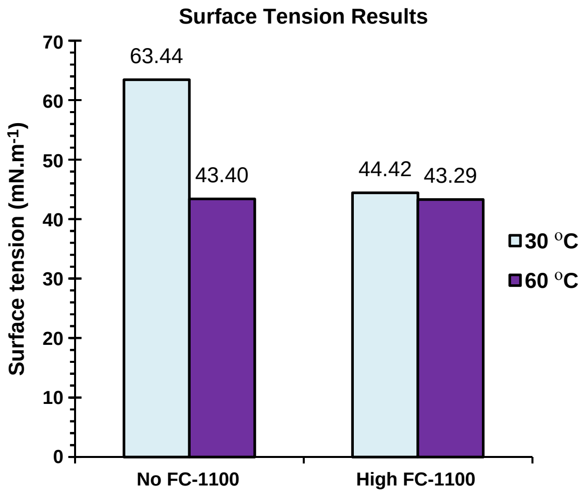


Fig. 8. The effect of temperature and FC-1100 on surface tension.

In contrast, the reduction in surface tension from the presence of surface active agents not only reduces the final bubble sizes but also changes the bubble burst mechanism at the free surface of the solution. When a bubble rises through the bulk solution and reaches the free surface of the solution, a thin film is produced in the top of the bubble. The stability and lifetime of this thin film is influenced by a number of factors such as surfactant concentration, surface diffusion, surface tension gradient, and drainage rate (Pugh, 1996). Generally, drainage rate decreases with the increase in bulk viscosity, surface viscosity and surface elasticity of the solution (Pugh, 1996). The latter two factors can be increased significantly by the presence of surfactant molecules at the liquid–gas interface (Pugh, 1996). Therefore, it is proposed that the presence of FC-1100 molecules in the solution reduces acid mist mainly via its effects on the thin film drainage rate through a change in the surface viscosity and surface elasticity rather than its effect on final bubble sizes. The slower film drainage and higher surface elasticity in the solutions containing FC-1100 result in the generation of a smaller number of airborne acid droplets (i.e. lesser amounts of acid mist).

3.3.2. Effect of viscosity

To investigate the counter-intuitive inverse relationship between the solution acidity and the amount of acid mist generated shown in Figs. 6 and 7, the viscosity of 12 electrolyte samples (7 low acidity and 5 high acidity solutions) was measured at three different temperatures. The averaged viscosity measurements for the two types of the electrolyte solution are shown in Fig. 9.

Fig. 9 shows as expected that the viscosity of a solution increases with the increase of its acid concentration but decreases with an increase in temperature. The results shown in Figs. 6 and 7, that is, more acid mist is generated by low acid concentration solutions regardless of the applied current density and temperature, can be

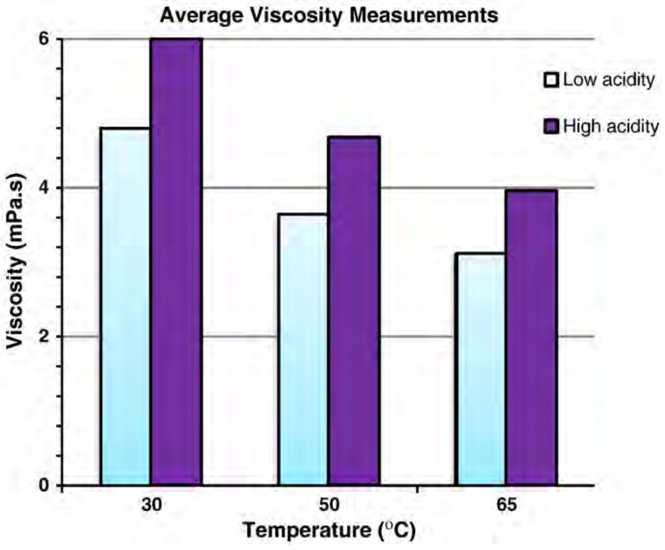


Fig. 9. The effect of solution acidity on its viscosity at different temperatures.

explained by that the amount of acid mist generated is inversely related to the viscosity of the solution.

While no quantitative relationships between the liquid viscosity and the droplet size and size distribution from the burst of bubbles can be offered at present, it is certainly true qualitatively that, the burst of more viscous liquid films or bubbles will produce less number, but in larger sizes, of droplets. This is simply because more viscous liquid films are more difficult to break-up. Further, based on Stokes' law, the terminal velocity of a rising bubble in a liquid is inversely related to the viscosity of the liquid due to the higher drag force exerted on the bubble by the surrounding liquid. As such, the average ascending oxygen bubble velocity in a more viscous liquid is lower than that in a less viscous liquid. Bubbles reaching the free surface at a lower speed would have less dynamic energy to “throw” the droplets from the bubble burst and make them airborne. The results shown in Figs. 6 and 7 imply that, at the high acidity level (high viscosity), the number of electrolyte droplets that become airborne is much less than that from the low acidity level.

It must be noted, however, that the substantial increase in acid mist with increasing temperature is not solely due to the decrease in viscosity. Based on Figs. 7 and 9 it can be seen that a 20% decrease in viscosity due to lower acid concentration in the solution results in 0.22 standard deviations increase in acid mist whereas the 34% decrease in viscosity due to temperature rise results in 1.22 standard deviations increase in acid mist. This significant increase must be attributed to the compound effect of a simultaneous decrease in both the viscosity and surface tension of the solution caused by the temperature increase.

4. Conclusions

The effect of five different operating parameters on acid mist generation was analysed using a full matrix experimental design. The temperature of the solution and the presence of FC-1100 in the solution proved to be the most influential parameters in the amount of acid mist generated. More than 90% of the variations in the acid mist generation can be explained by changes in the two parameters and their interaction.

To a lesser extent, electrical current density and solution acidity also affected the total amount of acid mist generated. Anode's age and most of the 3, 4, and 5-way parameter interactions were found to have negligible influence on the amount of acid mist.

Overall, acid mist was found to increase with temperature and current density. In contrast, addition of FC-1100 to the solutions decreased the amount of acid mist. However, it is critical to note that it is the ability of FC-1100 to increase the surface elasticity and surface viscosity, not its ability to reduce surface tension, that is responsible for the reduction of acid mist generation. The bubble burst mechanism at the free surface of the solution, which is mainly influenced by surface elasticity and surface and bulk viscosity of the solution, proved to be a critical factor in the amount of acid mist generated.

References

3M, 2007. FC 1100 acid mist suppressant. Xstrata Technology Bulletin. 3M Chile, Townsville Australia, pp. 1–19.

Al Shakarji, R., He, Y., Gregory, S., 2010. Measurement of bubble size distribution in copper electrowinning process by image analysis. In: Harre, J. (Ed.), 7th International Copper Conference. GDMB, Hamburg Germany, pp. 1237–1251.

Alfantazi, A.M., Dreisinger, D.B., 2003. Foaming behavior of surfactants for acid mist control in zinc electrolysis processes. Hydrometallurgy 69 (1–3), 57–72.

Bender, J.T., 2010. Evaluation of mist suppression agents for use in copper electrowinning. In: Harre, J. (Ed.), 7th International Copper Conference. Electrowinning and -refining. GDMB, Hamburg Germany, pp. 1271–1280.

Cheng, C.Y., et al., 2004. Evaluation of saponins as acid mist suppressants in zinc electrowinning. Hydrometallurgy 73 (1–2), 133–145.

Cohen, J., 1988. Statistical power analysis for the behavioural science. Lawrence Erlbaum Associates Inc, New Jersey.

Davenport, W.G., King, M., Schlesinger, M., Biswas, A.K., 2002. Extractive Metallurgy of Copper. Elsevier Science Ltd, London.

Davis, J.A., Eng, P., 2002. Budget quotation for a cross flow ventilation system of Sepon Copper Ltd. DESOM Systems Inc, Ontario USA.

Field, A.P., 2005. Discovering statistics using SPSS: (and sex, drugs and rock ‘n’ roll). ISM Introducing Statistical Methods, SAGE, London.

Harris, D.C., 2007. Quantitative Chemical Analysis. Freeman and Co, New York.

Hooper, G., 2008. In: Gregory, S. (Ed.), Quotation on 3M Matting. Blackwoods Townsville, Australia.

Hosny, A.Y., 1993. Electrowinning of zinc from electrolytes containing anti-acid mist surfactant. Hydrometallurgy 32 (2), 261–269.

Houlachi, G.E., Edeards, J.D., Robinson, T.G., 2007. Copper Electrorefining and Electrowinning, V. Canadian Institute of Mining, Metallurgy and Petroleum, Montreal, Quebec Canada.

HSIS, 2009. Exposure Standard Documentation. Hazardous Substances Information System, Australia. Available: http://hsis.ascc.gov.au/DocumentationES.aspx?ID=578. [Accessed 5 March 2009].

Liow, J.L., Frazer, A., He, Y., Eastwood, K., Phan, C., 2007. Acid mist formation in the electrowinning of copper, Chemeca. Chemeca, Melbourne Australia.

Liow, J.L., Gray, N.B., 1996. Experimental study of splash generation in a flash smelting furnace. Metallurgical and Materials Transactions B: Process Metallurgy and Materials Processing Science 27 (4), 633–646.

Mella, S., Rodrigo, V., Lillo, A., 2006. Copper electrowinning in the absence of acid mist. SAME Ltd, Santiago Chile.

Montgomery, D.C., 2005. Design and Analysis of Experiments. John Wiley & Sons Inc, New York.

OSHA, 2003. Sulfuric acid exposure limits. Occupational Safety & Health Administration, Washington DC. Available: http://www.osha.gov/dts/chemicalsampling/data/CH_268700.html. [Accessed 3 March 2009].

Pugh, R.J., 1996. Foaming, foam films, anti-foaming and de-foaming. Advances in Colloid and Interface Science 64, 67–142.

Robinson, T., Lang, J., Isbell, L., 1994. Cerro Colorado Copper Company SX-EW Plant. ISA PROCESS Tankhouse, Colorado USA.

San Martin, R.M., Otero, A.F., Cruz, A., 2005a. Use of quillaja saponins (*Quillaja saponaria* Molina) to control acid mist in copper electrowinning processes: Part 2: pilot plant and industrial scale evaluation. Hydrometallurgy 77 (3–4), 171–181.

San Martin, R.M., Otero, A.F., Figueroa, M., Escobar, V., Cruz, A., 2005b. Use of quillaja saponins (*Quillaja saponaria* Molina) to control acid mist in copper electrowinning processes: Part 1. Laboratory scale evaluation. Hydrometallurgy 77 (3–4), 163–170.

Sigley, J.L., Johnson, P.C., Beaudoin, S.P., 2003. Use of nonionic surfactant to reduce sulfuric acid mist in the copper electrowinning process. Hydrometallurgy 70 (1–3), 1–8.

Sunwest, 2004. The most comprehensive solution to tankhouse emissions problems. Sunwest Supply, Arizona USA. Available: http://www.sunwesttec.com/brushes.html. [Accessed 13 April 2008].

Xie, G.X., et al., 2009. Effect of liquid properties on the growth and motion characteristics of micro-bubbles induced by electric fields in confined liquid films. Journal of Physics. D. Applied Physics 42 (11), 12.

Xu, Q., et al., 2009. Effects of surfactant and electrolyte concentrations on bubble formation and stabilization. Journal of Colloid and Interface Science 332 (1), 208–214.

The link between operational practice and maximising the life of stainless steel electrodes in electrowinning and electrorefining applications

1. J. Weston
Mount Isa Mines Limited – ISA PROCESS^{TM} Technology
Hunter Street, Stuart, Queensland, Australia 4814
Joanne.Weston@mim.com.au

2. W. Webb
Mount Isa Mines Limited - ISA PROCESS^{TM} Technology
Hunter Street, Stuart, Queensland, Australia 4814

ABSTRACT

The use of a permanent stainless steel cathode plate in both electrowinning and electrorefining applications has long been accepted as a proven technology for cathode copper production. Mount Isa Mines Limited is the original inventor of this technology and experience has shown these cathodes can have a life of more than fifteen years in electrorefining and ten years in electrowinning.

Correct management of the operating system and attention to detail will ensure the life of the electrode in both applications. Chloride levels in electrolyte, current density, the use of shorting frames, current distribution, harvesting patterns, and cathode-plating cycles all affect cathode plate life.

The aim of this paper is to discuss common problems encountered with electrode management, with a particular focus on electrowinning plants. Proven practical control measures, possible management solutions and operating parameters that extend the cathode operating life will be discussed. These recommendations are based on test work conducted by the ISA PROCESS^{TM}, operating experience and knowledge gained through cooperation with licensees.

INTRODUCTION

The introduction of the permanent stainless steel cathode plate in 1978 was a revolutionary development in the copper industry. This occurred in 1978 when Mount Isa Mines (MIM) undertook a complete modernisation from starter sheet to permanent cathode technology in their Townsville Copper Refinery.

The ISA PROCESS^{TM} Technology is based upon a superior cathode plate design and cathode-stripping machine. The ISA cathode plate consists of a stainless steel hanger bar that is stitch welded to a stainless steel blade. The hanger bar and part of the blade are then electroplated with copper to provide maximum electrical conductivity and corrosion resistance.

Currently ISA PROCESS^{TM} has fifty-six licensees, of which thirty-five are electrowinning plants and twenty-one are electrorefining plants. Approximately 33% of the world's copper production is produced using the ISA PROCESS^{TM}. The ISA cathode plate has applications in both electrowinning (EW) and electrorefining (ER) applications. Cathode plate longevity is directly related to operating conditions.

Operators of both electrowinning and electrorefining plants are facing similar challenges, including:

- Achieving and maintaining target copper production levels.
- Ensuring chemical and physical cathode copper quality.
- Ensuring a safe working environment for all employees.

The operational methodology to achieve these goals can at times be detrimental to the long-term service life of the cathode plate.

THE STAINLESS STEEL PERMANENT CATHODE PLATE

The latest ISA PROCESS^{TM} cathode plate design is the ISA Cathode BR^{TM} and is the lowest resistance cathode plate available in the market. The ISA Cathode BR^{TM} is based on the proven ISA design. The copper coating is thicker to provide greater corrosion resistance and the plating has been extended to the base of the lifting windows to significantly reduce electrical resistance. These improvements will provide a greater service life, but only good management practice will ensure the extended operating life of the permanent cathode plates.

LIFE EXPECTANCY OF THE ISA PROCESS^{TM} CATHODE PLATE

The ISA PROCESS^{TM} is based on the premise that stainless steel cathode plates are a capital item not a consumable. Cell house management in both electrowinning and electrorefining will influence the effective life of this capital investment. It has been proven that ISA PROCESS^{TM} cathodes will perform without significant repair, for greater than ten years in electrowinning and in excess of fifteen years under electrorefining conditions.

Cathode plate life may be defined as the operating period in which the cathode plate will produce LME grade A copper at high operational efficiency. The integrity of the cathode plate assembly must be maintained. This requires attention to plate straightness, verticality, integrity of the welded joints, condition of the copper coating, stripability and overall electrical performance.

The life expectancy of the standard ISA PROCESS^{TM} plate in electrorefining is currently beyond fifteen years in well-managed refineries. These plants do not commit to major cathode repairs except through unusual misadventure.

This is not the case in electrowinning plants because the ability to maintain a stable, ongoing production regime within the cell house is often compromised. The solvent extraction plant, heap leaching operation and in some cases the mining operations have a significant impact of the ability to operate the electrowinning plant. Subsequently the condition of the cathode plate can be compromised. The life expectancy of an ISA PROCESS^{TM} cathode plate is up to ten years without repair compared to less than three with other cathode types.

EXTERNAL FACTORS INFLUENCING ELECTROWINNING OPERATIONS

Mining, leaching and solvent extraction processes can have overflow effects on the electrowinning operations that are not within the direct control of the EW operators. These effects can be transient, whilst others may be longer term and site specific. Failure to react promptly to such occurrences can impact on cell house operations, which may affect the integrity of the cathode plates.

Mining operations

The impact of variable mineralogy and ore grades is well recognised. Experience at Girilambone Copper Company in NSW showed that a change in crushing, stacking and heap leach irrigation methods was necessary to optimise recovery from lower grade ores with the changing ore types (Dudley, Bos & Readett, 3).

Variability of gangue minerals in heap leaching operations can be just as significant. Gangue materials may contain higher silicates, a larger percentage of fines or produce more fines in crushing (due to lower mechanical strength), higher chlorides and iron minerals (for example pyrite). These characteristics may result in:

- Reduced recovery rates and overall recovery, resulting from lower heap permeability.
- Lower heap permeability can also lead to leach solution ponding, wash outs resulting in higher suspended solids in the PLS.
- High chloride and iron levels in the PLS.

These changes can affect crud build up and impurity transfer (Cl, Mn & ferrous) to the cell house. Elevated chlorides in the cell house will increase the potential for corrosion of the cathode plate and high iron transfer will reduce the current efficiency.

Heap leaching operations

Management of the heap leaching operations is crucial in maintaining the quantity and quality of the PLS solution being delivered to the solvent extraction plant. The composition of soluble salts (other than copper) in the PLS can vary and is generally site specific. The heap leach system can be used to "filter" out many of the unwanted suspended solids in the leach solution prior to the solvent extraction plant.

Close monitoring of the heap leach system to reduce the amount of solids entering the SX plant and achieve the desired recovery rates is required. Below are general areas which require attention to achieve these targets:

- Irrigation methods - wobbler versus dripper irrigation for temperature control and recovery.
- Application rates should match heap permeability in order to reduce the occurrence of ponding and the likelihood of wash outs.
- Heap maintenance to detect quickly any problems with application rates, line failures, washouts.

Suspended solids and low copper grades in the PLS increase the difficulty of the solvent extraction operation. Suspended solids generally result in crud formation, increasing the phase break times and ultimately lead to higher entrainment levels passing through to the cell house.

Solvent extraction plant

The copper grade of the PLS can be utilised to maintain production, reduce iron transfer, organic losses, crud build up and the loads on the electrolyte filters. High copper grades in the PLS combined with low extraction rates (provided production targets are achieved) will:

- Allow lower PLS flow rates - resulting in reduced aqueous and organic entrainment due to more effective phase disengagement.
- Minimise ferric iron transfer, therefore maximising current efficiency.

- Result in longer residence times in PLS, ILS ponds which reduces suspended solids and therefore crud formation.

Crud formation can lead to increased aqueous entrainment. This can result in increased chloride levels in the electrolyte feeding the tankhouse which at elevated levels can lead to pitting corrosion of the stainless steel blade.

Increased sulphuric acid concentrations and high levels of organic entrainment in the electrolyte can be detrimental to the cathode plate, resulting in increased levels of corrosion. Higher primary flow rates can lead to larger organic losses to both the raffinate stream and to the strong electrolyte phase also increasing the likelihood of corrosion. In plants that are operated at high extraction efficiencies with low copper PLS grades, ferric iron transfer can become a significant operational problem for the electrowinning plant

SOME PROBLEMS EXPERIENCED IN ELECTROWINNING PLANTS

Repair of mechanical damage to the cathode plate is a common problem for both electrowinning and refining operator's. Problems in EW plants tend to be more severe than those of an ER plant and as a result their cathode plates generally last longer.

Problems experienced in some operations include corrosion of the stainless steel blade, corrosion of the copper coating on the hanger bar, corrosion of brazed copper to stainless joints, solution line corrosion and maintaining plate geometry.

Corrosion of copper plating and brazed copper to steel joints

Experience and plant data indicates that two of the most common problems experienced in electrowinning plants are:

- Corrosion of the brazed joints of solid copper hanger bar system
- Corrosion of the copper of the hanger bars

It is common to experience failure of the braze between the solid copper hanger bar and the plate after a relatively short time (as illustrated in Figure 1). When the hanger bar is bent and cracking of the braze occurs, crevice corrosion is accelerated through the crack. The braze material becomes anodic and corrodes, where rapid failure is possible.

The rate of corrosion of this type will be dependent upon weld quality and material choice. Corrosion of this joint results in higher resistances and poor cathode efficiency. Re-brazing of the solid copper hanger bar to the blade will repair this poor performance and prevent ultimate failure of this joint. ISA PROCESS^{TM} cathodes are not subjected to this type of failure.

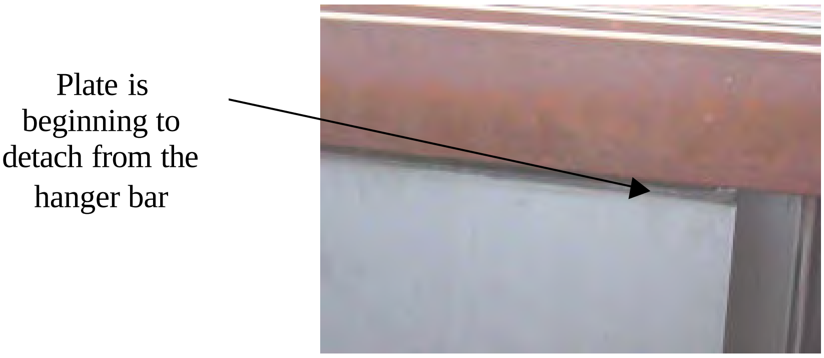


Plate is beginning to detach from the hanger bar

Figure 1: Corrosion of brazed joint between solid Cu bar and stainless steel plate

In some environments ISA cathodes may experience corrosion of the copper coating. This can be accelerated if the coating has been mechanically damaged in some way. The corrosion mechanism is galvanic. Often incorrect use of shorting frames can initiate this type of corrosion. This is illustrated in Figure 2 where the shorting frame has damaged the hanger bar and corrosion of the copper coating has resulted.

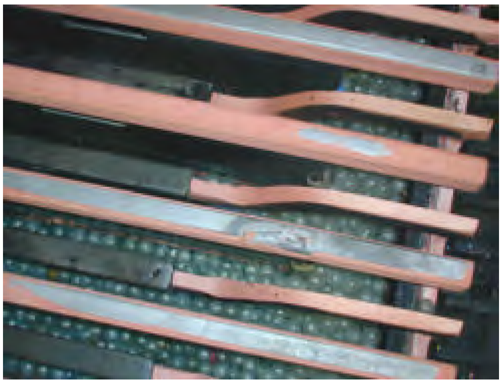


Figure 2: Corrosion of copper coating after 7 years operation

Galvanic corrosion

A review of the mechanism of galvanic corrosion, illustrates that it will occur when two dissimilar metals are connected in an electric circuit and immersed in electrolyte (Hayes, 4). The point at which the two dissimilar metals are joined is generally where the corrosion will occur first. Copper is more reactive and will act as an anode compared to the passive 316 stainless steel will take the part of the cathode (Craig and Pohlman, 2).

Corrosion of the stainless steel blades

The superior corrosion resistance of stainless steel is a result of the formation of a thin passive layer on the surface of the cathode plate. This passive layer mainly comprises of chromium oxide. If this layer is damaged and unable to rebuild itself, corrosion will occur in the unprotected areas (Hayes, 4). Great effort is then required to remove copper from corroded blades often resulting in damage to the cathode plate. Following copper removal, cathode plates are re-straightened and buffed to restore a useable surface. Figure 3 illustrates surface corrosion of the stainless steel blade.

316 stainless steel is corrosion resistant within a range of electrolyte conditions. Temperature, acid concentration and chloride levels are the key factors when considering the corrosion resistance of 316 stainless steel.

Figure 3: Surface corrosion of the cathode mother plate

Pitting and crevice corrosion

Pitting and crevice corrosion are a type of localised corrosion, where there has been a localised breakdown of the passive film. Pits are formed that are very small in comparison with the overall surface, and subsequently corrode very quickly (Aspahahani & Silence, 1). Once the localised breakdown of the protective film occurs an electrolytic cell is effectively created. The anode is the pit (small area) and the surrounding larger protected surface acts as the cathode. As the corrosion continues down the pit it is less likely that the surface in the pit is able to re-passivate.

Pitting corrosion is common in neutral and acid solutions that contain chlorides (and bromides). The chloride ions facilitate the pits by breaking down the passive film in localised areas, this is particularly so in the presence of surface imperfections. These imperfections could include such things as suspended solids, sulphides and micro crevice imperfections (Hayes, 2001).

Crevice corrosion occurs under the same conditions as pitting corrosion but only requires a small crevice for corrosion to be initiated. This is because capillary forces apply, and solution is drawn into the crevice creating an environment with stagnant solution. The supply of oxygen to the surface of the plate is restricted and the passive oxide layer is not replenished and corrosion continues to occur.

Low pH's, high chlorides, high temperatures and stagnant solutions are conducive to pitting and crevice corrosion.

Solution line corrosion

Solution line corrosion is often characterised by pits above and around the solution line (illustrated in figure 4). Pitting may be due to chlorides and organic transferred from the SX plant. It may also be due to incorrect addition of chlorides for reagent purposes, further exaggerated by the presence of organic residue in the electrolyte.

Unnecessary buffing of the entire cathode plate to overcome this 'tightness' is frequently undertaken. This practice can accelerate the surface corrosion and may result in 'prestripping' of the newly buffed plates.

Figure 4: Solution line corrosion

Pre-stripping of the copper cathode

The ISA PROCESS^{TM} cathodes have a surface finish that will allow for the necessary mechanical handling of the cathodes without pre-stripping. It will also enable the easy separation of the copper in the cathode-stripping machine that has been specifically designed for that purpose.

When cathode plates are buffed excessively, pre-stripping may occur. ISA PROCESS^{TM} recommends a standard surface finish for refurbished blades, which is different to that of a brand new plate. The tendency to polish the plate to a 'mirror like' surface must be avoided. It should be noted that cathodes up to 110kg per side are routinely produced in ISA PROCESS^{TM} plants without any pre-stripping problems.

Difficult to strip copper - "sticky copper"

At times it is difficult to strip the copper successfully from the cathode mother plate. This can be due to pitting corrosion of the blade or thin copper deposits. ISA PROCESS^{TM} recommends an average copper cathode weight in excess of 40kg to ensure easy stripping. Without sufficient weight and thickness it may be extremely difficult to flex the copper from the plate. The copper tends to simply bend and flex with the stainless steel plate.

Sticky cathodes are often removed by manual or more aggressive mechanical means. These techniques can result in dents, bends and severe scratching of the blades. All of which will affect the plate verticality and future corrosion resistance.

INTERNAL CELLHOUSE OPERATING PARAMETERS

Within the cell house there are opportunities for operational choices that can extend the service life of the cathode plate. Where possible these practices should be employed. Those that result from limitations of the original design need to be compensated for where practical.

Harvesting sequences in electrowinning operations - current density

The operating current density at the time of copper initiation and nucleation has a significant effect on the copper quality and stripability. The chosen harvesting cycle will determine the initiation current density. There are two types of live harvesting sequences commonly employed in electrowinning:

- Harvesting the cells by lifting every third cathode
- Harvesting the cells in blocks of cathodes in thirds

<u>Harvesting by lifting every third cathode</u>

This harvesting sequence ensures that current is maintained through each anode and the subsequent current density of the cathode plates is not excessively high. Cathode currents in adjacent cells are virtually unaffected, anode currents are maintained, easy access for above cell inspection and wash down of cathodes and contacts. The results of which is a much higher cathode quality and the life of the lead anode is prolonged by a more steady and constant method of operation.

Harvesting in one third blocks

The difference lies within the current distribution of the anodes. In effect every time that one third of the cathodes are removed from the cell for stripping, that block of anodes is effectively turned off. Consequently the anodes are frequently cycled which can lead to premature spalling of the anode. This will result in reduced anode life and a greater potential for lead contamination in the cathode copper. Once lead contamination of the cathode is detected the logical step is to remove the lead from the bottom of the cells or clean individual anodes. To remove lead from the cells a shorting frame is required which leads to further anode cycling.

This method utilises a smaller crane bale and enables a shorter inlet and outlet conveyor. Loading and unloading of electrodes into the cell is simplified, resulting in a more rapid turnaround of electrodes through the cell house.

Stripping cycles

The initiation current density on bare stainless steel is extremely important when choosing stripping schedules. At high current densities in excess of the normal operating range it is possible to form nascent hydrogen. This occurs when the limiting current density for cupric ions has been exceeded. This highly reactive form of hydrogen can react with the chromium oxide film that protects the stainless steel. Once this occurs corrosion of the plate is possible, resulting in "sticky" cathodes.

If one entire cell load of cathodes is stripped one after the other in blocks, bare stainless steel is exposed to much higher current densities. It is preferable to strip one group (every third cathode) moving around the cell house, then stripping the next group the following day and so on. With this approach no bare stainless steel is exposed to the higher current densities while one group is out of the cell being stripped. In this situation if the limiting current density for cupric is exceeded and nascent hydrogen is formed, the stainless steel blade cannot be corroded because it will already have a skin of copper over it. It may be possible that the copper will be slightly corroded and perhaps result in a rougher copper deposit.

The harvesting method and stripping sequence determine the initiation current density. At lower current densities corrosion of the stainless steel blade is unlikely and the possibility of rough copper deposits is reduced.

Use of shorting frames

Shorting frames are commonly used in electrowinning operations, to take cells off line for lead scale removal and anode cleaning. The correct choice, design and proper use of the shorting frame is important in avoiding damage to the cathode plate's hanger bar.

Uniform current distribution is critical during the operation of a shorting frame. Significant damage can result from uneven current distribution resulting in elevated temperatures and melting of the bar at the point of contact. Subsequently, penetration of the copper coating will accelerate the corrosion and this is illustrated in figure 5. Extreme temperatures may also soften and weaken the braze of the solid copper hanger bar systems, making them susceptible to corrosion attack at the joint.

Certain types of shorting frames are able to compensate for variation in cathode type where the bars do not sit at the same height and should be used.

Figure 5 Damage caused to hanger bar from incorrect use of shorting frame

Frequent use of shorting frames increases the rate of anode cycling, the likelihood of lead contamination of the copper cathode and damage to the hanger bar.

Choice of acid mist control systems

There are various types of acid mist suppression systems employed by EW operators throughout the world. These include, cell covers, ventilation systems, mist suppressant reagents, bb's and balls and hoods. Associated with each of these are elements that can cause some damage to the cathode mother plate in the form of corrosion. The extent of these problems will be dependent on the specific management practices within the cell house.

Cell covers tend to limit the operation of the cell house by dictating harvesting sequencing due to the necessary removal and placement of the cell cover/hoods. The longer cathodes are out of the cell means the remaining cathodes are subjected to higher current densities for extended periods. This can result in corrosion of the stainless steel and rough copper deposits.

The use of Bb's or balls in scavenger cells requires constant cleaning to remove organic carryover. Failure to do so may result in corrosion of the stainless steel at the solution line. Organic carryover can result in an increase in acid concentration, copper depleted solution in this region that is conducive to corrosion. A thick layer of bb's/balls in the presence of high chlorides can also result in solution line corrosion.

Cathode plate maintenance

Plate straightness and verticality is of upmost importance in maintaining uniform current distribution and efficiency. A cathode plate that is not hanging vertical will result in uneven current distribution, which can lead to further problems with short-circuiting and cathode quality. Significant damage to the plates can occur if these are not straightened correctly.

Surface finish is the key factor affecting stripping performance. The initial 2B finish provides sufficient adhesion so that the cathode can be transported safely throughout the tankhouse but easily stripped by ISA PROCESS™ cathode stripping machines. Only when the copper is proving to be too difficult to strip by the normal means, should buffing of the plates to restore the surface condition be considered. As a general rule it is said that plates which do not machine strip, but are able to be hand stripped do not require surface conditioning.

The reason some plates do not machine strip is due to thin copper deposits rather than poor surface condition. In the case of thin deposits, rather than the copper flexing (shearing) off the cathode plate the copper moves with the stainless steel blade. Aging anodes, poorly maintained electrode geometry, poor current distribution and high current densities all contribute to poor physical quality cathode and may necessitate early cathode stripping.

The correct refurbishment methods are crucial. Buffing should be conducted with the appropriate buffing media, fitted to the correct machines and operated at the recommended speeds for which the media was designed. Only those areas that have a damaged surface should be buffed.

Over polishing can often result in pre-stripping of electrodes, creating a serious safety concern in the tankhouse and repercussions in current distribution. If buffing is conducted at machine speeds that are too high there is a tendency to scratch the plate. This can lead to crevice corrosion of the stainless steel blade.

Copper deposit quality

As current density and anode age increases the ability to maintain even current densities is compromised. Poor current distribution can lead rough copper deposits, increasing the probability of lead contamination and the formation of direct shorts.

Direct shorts will result in anode distortion and increase spalling of the lead anode. Subsequently, lead contamination in the cathode will increase. To overcome this problem, operators may choose to produce lighter cathodes.

Light cathodes are not as easily stripped and subsequently stripping performance may deteriorate. This deterioration may result in the perception that the plates require buffing. The amount of cathode buffing increases, which can then result in pre-stripping.

MACHINE AND DESIGN CONSIDERATIONS FOR EW PLANTS

Machine reliability, crane cycles and stripping cycle times are crucial to the successful operation of a cell house and maintaining the integrity of the stainless steel cathodes. It is important that all of the equipment operates in sequence with one another, so that one piece of element does not delay the entire process. Cell turnaround requirements and work scheduling ultimately dictates the machine capacity in the design for a new EW plant.

The crane must provide a continuous supply of cathodes to the stripping machine so that it can operate at design capacity. If the crane is unable to keep up with the stripping machines, cathodes will spend longer out of the cells than necessary. Cycle times need to compliment one another to ensure this does not occur and that cathodes have adequate time in the wash chamber. On the other hand the crane should not wait over the inlet conveyor with a rack of cathodes if the machine is not coping with the throughput. It should wait over the cell and remove the cathodes when the conveyor is able to fit another rack of cathodes. This will also ensure that cathodes are out of the cells for the shortest time possible.

Regular machine and equipment maintenance is absolutely critical for the smooth operation of the cell house and service life of the cathode plate. The design of all ISA PROCESS^{TM} cathode stripping machines ensures that there is no damage to the ISA cathode plate and associated edge strips whilst maintaining design throughput.

CONCLUSION

Cathode plates are a capital item and should not be considered as a consumable. Careful management of the electrowinning (and electrorefining) operation will ensure the longevity of the stainless steel cathode plate and enable the production of commercially saleable cathode copper. Management practices from mining through to the electrowinning plant all play an important role in the maintenance and preservation of the permanent cathode plate.

The ISA PROCESS^{TM} is an integrated system that combines a robust cathode plate with a safe and high performance stripping machine design. When combined with good process management the long-term effectiveness of licensed facilities is ensured.

REFERENCES

Book References

1. A.I. Asphahni and W.L. Silence "Pitting Corrosion" in Metals Handbook, Ninth Edition, Volume 13 Corrosion, ASM International, United States of America, 113-114.

2. Craig and S.L. Pohlman "Introduction" in Metals Handbook, Ninth Edition, Volume 13 Corrosion, ASM International, United States of America, 83-85.

Edited Conference Proceedings

3. K.A. Dudley, D.R. Readett and J.L. Bos, "Hyrometallurgical Processing of Copper Ores, Heap leach, solvent extraction and Electrowinning at Girilambone Copper Company", in Proceedings of the AusIMM MinPREX 2000 Conference, Melbourne, Australia.

4. Hayes, "Corrosion of Stainless Steel", Paper presented at the Seventh ISA Process User's Conference, Phoenix, Arizona, USA, 27th May - 1st June 2001.

Hydrometallurgy 109 (2011) 168–174

Contents lists available at ScienceDirect

Hydrometallurgy

journal homepage: www.elsevier.com/locate/hydromet

The sizing of oxygen bubbles in copper electrowinning

Reza Al Shakarji [a,1], Yinghe He [a,*], Simon Gregory [b,2]

[a] School of Engineering and Physical Sciences, James Cook University, QLD 4811, Townsville, Australia
[b] Xstrata Technology, Copper Refinery Limited, Hunter St, QLD 4811, Townsville, Australia

ARTICLE INFO

Article history:
Received 12 May 2011
Received in revised form 29 June 2011
Accepted 29 June 2011
Available online 12 July 2011

Keywords:
Copper electrowinning
Oxygen evolution
Imaging
Bubble size
Contact angle
Surface tension

ABSTRACT

Oxygen bubbles are formed on the anode in the copper electrowinning process. Burst of the bubbles produces acid mist in the tankhouse. While it is well acknowledged that the amount of acid mist is related to the size of the bubble, no systematic measurements have been made to quantify bubble size and its relationship with materials and process variables. This paper presents results of bubble size measurement under different operating conditions. For each of the operating conditions tested, bubbles were detected in a wide size distribution ranging from 20 μm to more than 400 μm in diameter. Statistical analyses on the measurement results showed that addition of FC-1100, a surfactant widely used in copper electrowinning to suppress acid mist, and solution temperature were the two most influential test parameters on the bubble size followed by the age of the anode. In contrast, current density and solution acidity had negligible effect.

1. Introduction

Electrowinning is an electrochemical process that is used to extract and refine metals such as Cu, Zn, Ni, Mn and Cd from their solutions. In most cases, the solution is acidic and oxidation of water molecules at the anode is the common counter reaction which results in the liberation of oxygen bubbles:

$$2H_2O_{(l)} \rightarrow O_{2(g)} + 4H^+_{(aq)} + 4e^- \quad (1)$$

The generation of oxygen bubbles generally requires a high overpotential and accounts for a considerable percentage of the energy consumption in metal electrowinning tankhouses. Moreover, the generated oxygen bubbles burst at the air/solution interface, producing highly acidic droplets of which the fine ones become airborne and form an acid mist throughout the tankhouse of the electrowinning plant.

Acid mist is highly corrosive and results in the corrosion of cathode plates, anode hanger bars, tankhouse equipment and building structures. It also poses a serious health hazard and causes extreme discomfort to the skin, eyes and respiratory systems of tankhouse workers. The effects of different process parameters on the amount of acid mist have been discussed in detail in our earlier publication (Al Shakarji et al., 2011).

Previous research results suggest that the amount of acid mist being generated is related to the size of the bursting bubbles (Liow et al., 2007). For electrowinning processes, although it is known that some operating parameters such as solution temperature can influence the size of the bubbles, there have been no systematic studies to quantify the effect of operating parameters on the size and size distribution of bubbles.

This paper reports results of bubble size measurement under different operating conditions. It also examines, quantitatively, the relationship between the median bubble size and five operating parameters in copper electrowinning. These parameters are anode age, electrical current density, solution temperature, sulphuric acid concentration of the solution, and the presence of a typical chemical mist suppressant such as FC-1100.

2. Experimental

2.1. Equipment setup

The copper electrowinning process was replicated in a bench-scale tank (EW cell). The equipment used in this investigation and its arrangements are shown in Fig. 1.

A peristaltic pump and four submersible heaters were used to maintain consistency in the composition and temperature of the solution for the duration of each test. To replicate industrial operations Pb–Ca–Sn alloy and 316 L stainless steel were used as anode and cathode, respectively (Houlachi et al., 2007). Areas of 100 mm wide and 150 mm long of the sides of electrodes facing each other were exposed to the solution and the rest of the immersed electrodes were isolated from the solution. This arrangement allowed

---

[*] Corresponding author. Tel.: +61 7 47814270; fax: +61 7 47816788.
E-mail addresses: reza.alshakarji@my.jcu.edu.au (R. Al Shakarji), yinghe.he@jcu.edu.au (Y. He), sgregory@xstratatech.com (S. Gregory).
[1] Tel.: +61 422347562; fax: +61 7 47816788.
[2] Tel.: +61 7 47589530; fax: +61 7 47589501.

0304-386X/$ – see front matter © 2011 Elsevier B.V. All rights reserved.
doi:10.1016/j.hydromet.2011.06.010

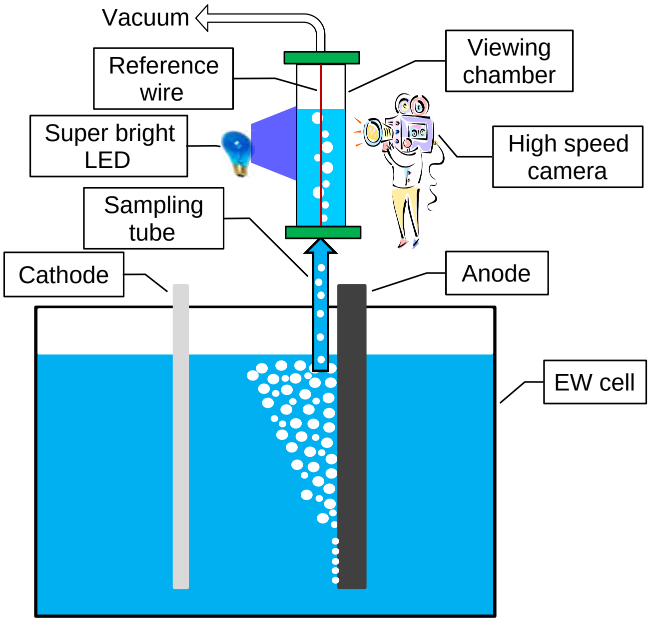


Fig. 1. A schematic view of equipment setup (not to scale).

a precise and controlled electrical current flow through the solution via the specified electrodes surfaces.

For each set of experiments a fresh batch of solution was synthesized containing 45 g L^{-1} Cu, 15 mg L^{-1} Guar gum, 20 ppm Cl, 100 ppm Co, and various amounts of sulphuric acid. The concentration of Cu, Guar gum, Cl and Co in the synthesized solution was similar to the electrolyte solutions found in most copper electrowinning tankhouses worldwide (Robinson et al., 1994). The sulphuric acid concentration in the solution, however, was one of the five selected variables and its concentration differed from that of a typical copper electrolyte solution.

To measure bubble sizes via image analysis, the generated oxygen bubbles were transferred from the EW cell to a viewing chamber. A high-speed camera (RedLake MotionXtra HG-100K) with a special macro lens and super bright blue LEDs were utilized to capture images of the rising bubbles in the viewing chamber.

2.2. Experimental design

The main goal of the present work was to compare, quantitatively, the effect of different operating parameters on bubbles sizes during a typical copper electrowinning process. To fully explore the main effect of each individual parameter as well as any possible interaction effects, the 2 K factorial method was used to determine the number of required experimental conditions (Montgomery, 2005). Based on this method, the test parameters were examined at two levels (low and high). Table 1 lists the five selected parameters and their low and high level values.

The values shown in Table 1 were selected so that the midpoint between low and high levels of each parameter represented the typical value used in most copper electrowinning tankhouses (Houlachi et al., 2007). For example, a sulphuric acid concentration in the electrolyte solution of about 170 g L^{-1} is typically used in most copper electrowinning operations. Thus, the low and high limits of solution acidity for the experiments were set at 100 and 250 g L^{-1}, respectively.

Table 1
The selected test variables and their values.

No	Examined parameter	Low	High
1	Anode age (months)	0	6
2	Current density (A m^{-2})	200	400
3	Temperature (°C)	30	60
4	Acidity (g L^{-1})	100	250
5	FC-1100 (ppm)	0	30

Based on the 2 K factorial method, 32 parameter combinations were required to fully examine the effect of five parameters at two levels. For reliable data analysis, a minimum of two replicates were required for each test condition. Hence, 64 tests were conducted to determine the significance of each parameter and also to identify any significant interactions among the selected parameters.

For each of the 64 experiments, at least two sets of high resolution photos were taken by the high-speed camera. The images were taken at 10 min intervals at a rate of 250 frames per second while the experiment was being conducted. To ensure that the same bubbles were not accounted for more than once, only one photo was selected per 100 consecutive frames (i.e. 3 to 5 photos were selected from each set of images of each experiment) for image analysis.

2.3. Bubble sizing process

A flexible silicon tube (sampling tube in Fig. 1) was used to transfer the bubble-containing solution from the EW cell to the viewing chamber. One end of the sampling tube was placed about 10 mm underneath the free surface of solution near the surface of the anode while the other end was connected to the bottom of the viewing chamber. By applying a gentle vacuum to the viewing chamber, the bubble containing solution was drawn from the EW cell to the viewing chamber at low flow rates. The large internal diameter of the sampling tube (5 mm) and the low solution flow rate, from the EW cell to the viewing chamber, ensured no bubble breakage or coalescence occurred during the bubble transfer process. Thus, bubbles observed in the viewing chamber were true representatives of those generated in the EW cell.

The high speed camera was then used to capture 800 × 600 pixel images. Each image taken from the rising bubbles in the viewing chamber contained about 100 bubbles on average (more than 300 bubbles in some cases). A MATLAB based program was developed to automate the analyses of the captured images. The developed program was able to process the images in grey scale format directly from the camera. The software was designed to take into account the uneven background illumination and to distinguish the boundaries of each bubble from its surroundings.

A thin copper wire, 130 μm in diameter, was fixed inside the viewing chamber adjacent to the solution's entrance. The diameter of this wire was used as a reference to determine the size of the bubbles captured in each image. The developed software was able to process each image in a few seconds and export the raw data of bubble diameters (in μm) to an Excel file. The software also indicated the detected bubbles with blue lines along the boundaries of each bubble and displayed the bubble diameter (in μm) next to it (Fig. 2). These processed images were used to visually inspect the results of the bubble sizing process.

To ensure the accuracy of the developed bubble sizing program, ImageJ (an open source image analysis software) was used to manually determine the bubble sizes of three randomly selected images. The results of the automated and manual bubble sizing were statistically identical.

3. Results and discussion

3.1. Overall statistical analysis

In total, more than 54,000 bubbles were sampled and their sizes measured. The diameter of the bubbles ranged predominantly from 26 μm up to 182 μm, on average, for each of the 64 experiments.

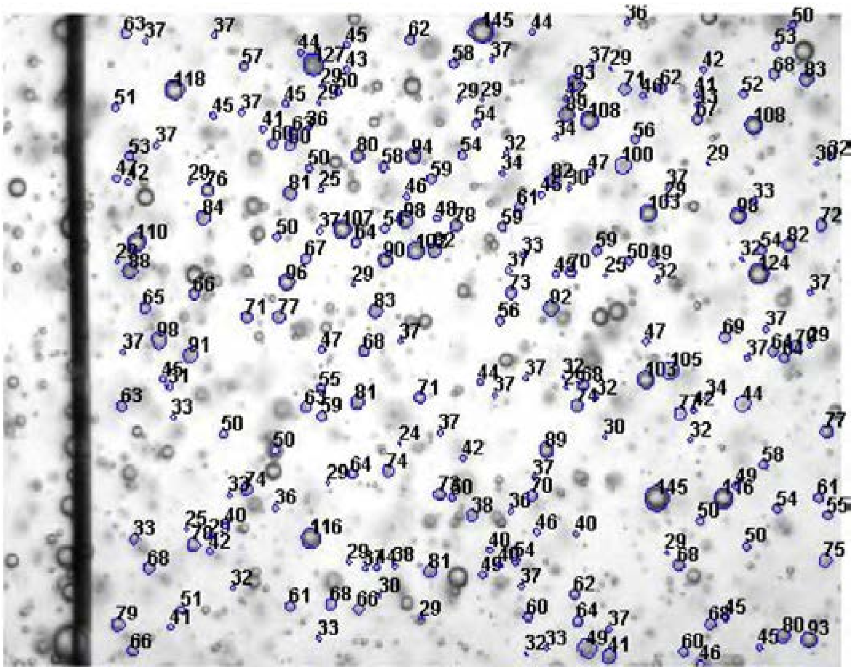


Fig. 2. A processed image with 250 detected bubbles and the reference wire (the vertical black line) appearing on the far left.

However, there were cases where the smallest detected bubble was 20 µm and the largest one was 404 µm. On average, 50% of the detected bubbles had a diameter of 53 µm or less (i.e. the grand median value). A typical bubble size distribution and cumulative percentage graph is shown in Fig. 3.

There was a relatively strong positive skewness (1.7 on average) in all the collected data samples, similar to that shown in Fig. 3. Therefore, median values were selected instead of mean values to represent the average bubble size for each experiment. The small difference between the medians of replicates of a test condition, which ranged from 0.00 µm up to 5.24 µm with the average being 2.17 µm, was a good indication of the high repeatability of the conducted experiments.

Fig. 4 illustrates the median bubble diameter (the bold black line inside each box), the lower and upper quartiles (the lower and upper edges of each box), the lower and upper extremes (the lower and upper whiskers), and also the outliers (the stars) for each of the 32 experimental conditions. The values of median bubble diameters ranged from 43 µm to 66 µm. Regardless of the applied operating conditions, 95% of the generated bubbles were less than 106 µm in diameter. It must be noted that the outliers were neither the result of error in the bubble sampling process nor the image analysis procedure. Live video footage taken from the surface of an immersed operating anode proved the existence of such unusually large bubbles. More details about these outliers can be found in our previous publication (Al Shakarji et al., 2010).

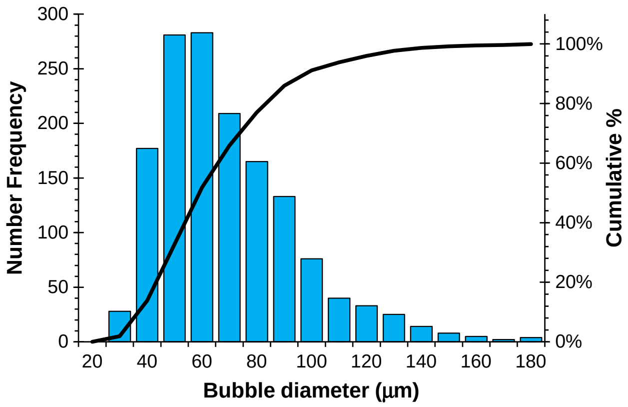


Fig. 3. Bubble size distribution and cumulative percentage graph for experiment No 9 (parameter values: new anode, 200 A m^{-2} current density, 60 °C solution, no FC-1100 and 100 g L^{-1} sulphuric acid concentration).

A 5-way ANOVA analysis was performed on the raw experimental data to determine the full effect of test parameters on the average bubble size. To quantify the magnitude of the effect of each of the 32 parameter combinations, Pearson's r value was calculated for each case based on its ANOVA's mean square value (Field, 2005). Cohen classifies the magnitude of a parameter's effect into three categories of small, medium and large (Cohen, 1988). These categories correspond to Pearson's r values of 0.1, 0.3 and 0.5 respectively (Cohen, 1988). Parameter combinations with Pearson's r value greater than 0.10 are shown in Fig. 5.

FC-1100 and solution temperature, having the highest r values in Fig. 5, were the most influential parameters in determining the final bubble sizes. To a lesser extent, anode age also influenced the final bubble sizes. Current density and solution acidity, with Pearson's r values of 0.04 and 0.05 respectively (not shown in Fig. 5), had negligible effect on the detachment size of bubbles.

The full factorial ANOVA analysis not only examined the effect of individual parameters but also all the possible parameter interactions. This analysis returned an R^2 value of 0.951, which meant the variations made in the test parameters of the conducted experiments could explain more than 95% of the bubble size variations measured in the test series.

3.2. Effect of individual parameters on bubble size

To assess the effect of individual test parameters on the bubble size, the average median bubble diameter for each level of a parameter was calculated. Each bar in Fig. 6 represents the averaged median bubble diameter from 32 independent experiments.

3.2.1. FC-1100

FC-1100 was the only parameter that had a large effect, based on Cohen's classification, on the average bubble sizes. The presence of FC-

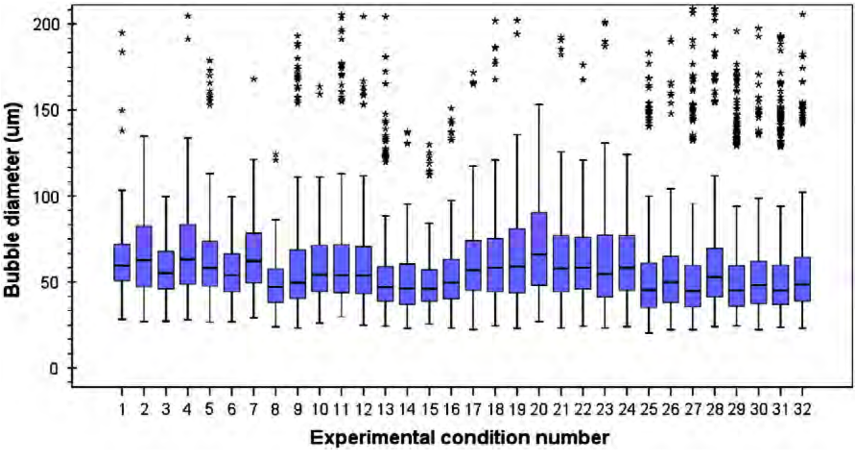


Fig. 4. Bubble size distribution under 32 different operating conditions.

1100 in the solution reduced the average bubble diameter by 9.3 μm. This result was largely attributed to the ability of FC-1100 to reduce the surface tension of the electrolyte from 63 mN m^{-1} to 44 mN m^{-1}. The near parallel solid lines in Fig. 7 implied that, within the tested range, the effectiveness of FC-1100 in reducing the final bubble sizes was independent of the solution temperature.

The bubble size distribution, in addition to the average bubble size, was also considerably reduced in the presence of FC-1100 (Fig. 8). It was proposed that the presence of FC-1100 molecules reduced the surface tension of the solution to a degree that resulted in less variation in the three phase contact angle on the anode, leading to the generation of more uniformly sized bubbles (i.e. smaller standard deviation).

3.2.2. Temperature

Solution temperature with a Pearson's r value of 0.30 was the second most influential parameter in determining the final size of a bubble (Fig. 5). Changing the solution temperature caused a relatively large difference in the average bubble diameter. Raising the temperature of the solution from 30 °C to 60 °C resulted in 3.7 μm reduction in the average bubble sizes, equivalent to 0.150 standard deviations (Fig. 6).

Similar to the presence of FC-1100, when temperature of the solution was raised from 30 °C to 60 °C, surface tension of the electrolyte solution decreased from 63 mN m^{-1} to 44 mN m^{-1}. Therefore, it was postulated that the decrease in the average bubble size at higher temperature was mainly due to the decrease in the surface tension of the solution. The large reduction in surface tension due to temperature rise was also expected to result in less variation in the three phase contact angle, leading to the generation of more uniformly sized bubbles. This expectation was confirmed by data presented in Fig. 9 where bubble size distribution was noticeably less (smaller box and whiskers) in high temperature electrolyte solution.

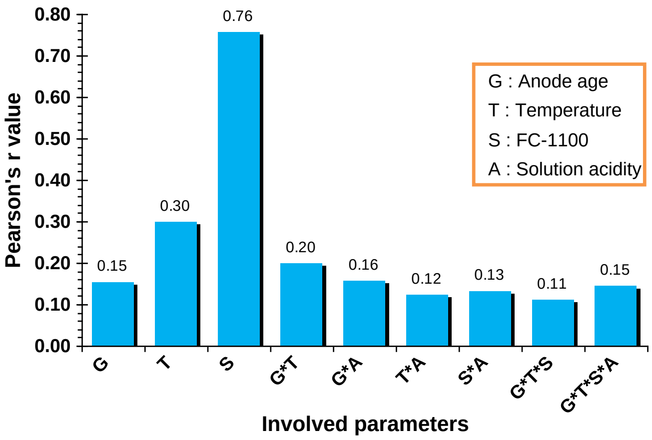


Fig. 5. Quantitative comparison of the most influential parameters and parameter interactions on the average bubble size.

3.2.3. Anode age

The experimental results shown in Fig. 6 suggested that bubbles generated on an old anode were, on average, 2.0 μm (0.089 standard deviations) larger than those generated on a new anode. The ANOVA analysis resulted in a Pearson's r value of 0.15 for the anode age which meant this parameter had a small/medium effect on the final size of a bubble. It was postulated that anode age affected the detachment size of a bubble by influencing the liquid–solid–gas contact angle which in turn influenced the balance of the forces acted on the bubble. SEM images revealed that the surface of an old anode was highly uneven and covered with voids and micro cracks whereas the surface of a new anode was relatively smooth and even (Fig. 10).

It was likely that the small cavities and cracks on the surface of the old anode provided much higher number of nucleation sites for the bubbles, leading to “forced” coalescence of the bubbles during their growth on the anode immediately prior to their detachment due to insufficient distances between neighbouring bubbles on the anode. Similar observations were reported by Huet where mean detachment radius of electrolytic bubbles increased with the surface roughness of nickel electrodes (Huet et al., 2004).

It is important to note that, regardless of the age of the anodes or operating conditions, bubble size distribution was relatively large, ranging from 20 μm to more than 400 μm (Fig. 4). The heterogeneities in both the chemical composition and the surface roughness of the anodes were believed to be the main reasons for such wide bubble size distributions. The anodes were Pb–Ca–Sn cold rolled alloys, either the surface roughness or the distribution of the three elements was very unlikely to be even on the surface of the anodes after the rolling process.

3.2.4. Solution acidity and current density

It is worth noting that, within the ranges tested in this study, both the solution acidity and the current density had negligible influence on the bubble size. Solution acidity could affect the bubble size only through the change it could cause in the solution viscosity. Kazakis et al. reported a noticeable difference in the average bubble size only when the viscosities of the test solutions were considerably different, e.g. the difference between their viscosities was more than 600% (Kazakis et al., 2008). Rheological tests of the copper electrolyte solutions, however, showed that high acidity solutions were only 21% more viscous than low acidity solutions. Therefore, it was suggested

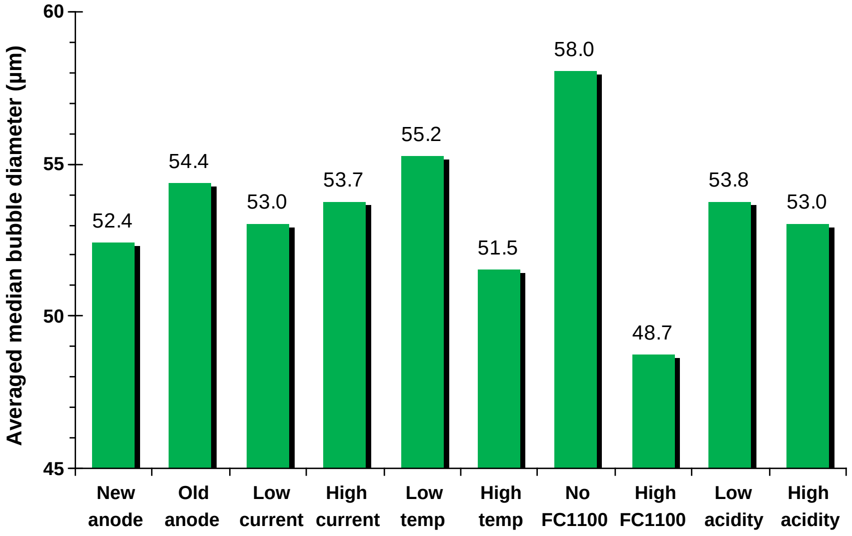


Fig. 6. Average bubble diameter at different operating conditions.

that a much larger difference in the solution acidity, and hence the viscosity, was required to observe a noticeable difference in the average size of the bubbles generated during the copper electrowinning process.

Based on Faraday's law, increasing the electrical current density would increase the amount of oxygen produced at the anode (Harris, 2007). Consistent with our results presented previously, within the range of 200 and 400 A m^{-2}, a change in the current density had caused only a change in the bubble nucleation rate and hence the number of bubbles generated per unit time. The final size of the bubbles changed little with current density (Al Shakarji et al., 2010).

3.3. Effect of parameters' interaction on bubble size

Few publications in this field studied the interactions that exist among the test parameters. The full factorial ANOVA analysis showed that more than 22% of variations in bubble sizes recorded in the test series were due to the interactions between the test parameters.

With a Pearson's r value of 0.20, the anode age–temperature interaction (G*T) was identified as the most significant parameter interaction. The results are shown in Fig. 11.

It can be seen that, on an old anode, increasing the temperature of the solution from 30 °C to 60 °C resulted in a relatively large decrease (more than 0.247 standard deviations) in the average bubble diameter. In contrast, the same increase in the solution temperature resulted in a much smaller decrease (about 0.049 standard deviations) in the average bubble diameter on a new anode.

The G*T interaction not only affected the average size but also the size distribution of bubbles. Size distribution of bubbles formed on the old anode was much wider at low solution temperature in comparison to that of high solution temperature (i.e. larger box and whiskers at low solution temperature). In contrast, the temperature of the solution had a relatively small effect on the size distribution of bubbles formed on a new anode (i.e. similar box and whiskers sizes at low and high temperatures).

The decrease in the average bubble size and its distribution with temperature rise was due to surface tension reduction as explained earlier. Consistent with observations made by other researchers (Drelich and Miller, 1994; Meiron et al., 2004), the strong interaction between anode age and solution temperature was believed to be mainly due to the effect of anode's surface morphology on the three phase contact angle.

The effect of the anode's surface morphology (i.e. anode age) on bubble size and its distribution was much more evident at low solution temperature (Fig. 11) since the overall surface tension was higher at lower temperature.

One of the other important interactions was the T*A interaction. The decrease in the average bubble size and its distribution due to temperature rise was not the same for the low and high acidity solutions (Fig. 12). The average size and size distribution of bubbles were more strongly affected by temperature rise in low acidity solutions than in high acidity solutions.

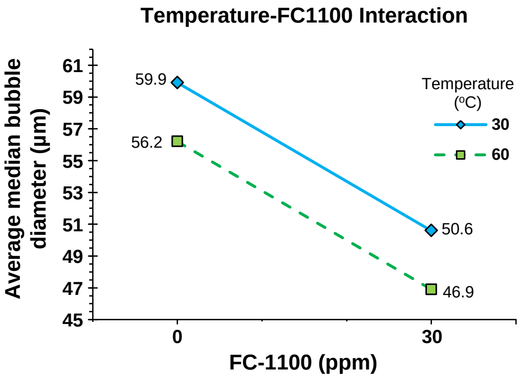


Fig. 7. The effect of FC-1100 on bubble size at different temperatures.

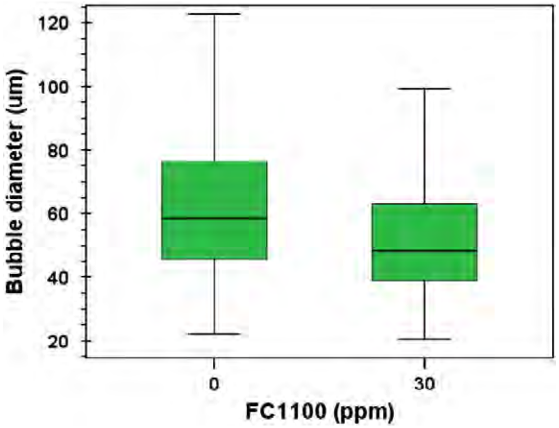


Fig. 8. Effect of FC-1100 on bubble size distribution (not including outliers).

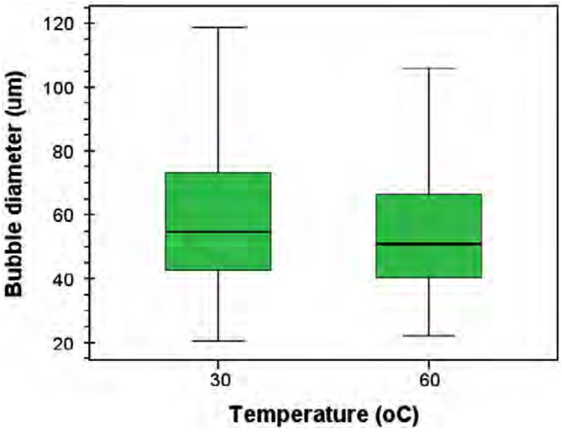


Fig. 9. Effect of solution temperature on bubble size distribution (not including outliers).

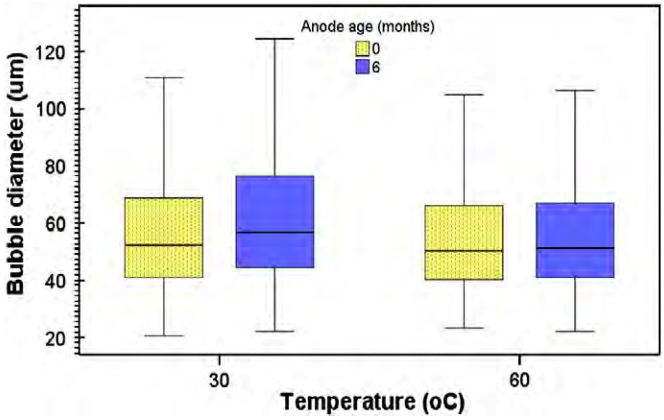


Fig. 11. The interaction between anode age and temperature.

Surface tension and viscosity were both known to influence the final size of a bubble (Xie et al., 2009; Xu et al., 2009). Temperature affected both surface tension and viscosity whereas acidity only affected viscosity. Therefore, the T*A interaction was expected to be the result of surface tension and viscosity interaction. The surface tensions of low and high acidity solutions were measured at two different temperatures (Fig. 13).

It is general knowledge that surface tension of a liquid is a function of not only its composition but also its temperature. Measurements showed that surface tension of a low acidity solution was higher than that of a high acidity solution at 30 °C (Fig. 13). However, surface tensions of the two solutions were almost the same at 60 °C. In other words, temperature had a more profound effect on the surface tension of low acidity solutions than that of high acidity solutions. These observations were similar to those seen in Fig. 12 in that, in low acidity solutions, the differences in the average bubble size and bubble size distribution caused by temperature rise were much larger. Both figures demonstrated a strong interaction between temperature and solution acidity.

The surfactant–solution acidity (S*A) interaction had a Pearson's r value of 0.13. The interaction between FC-1100 and solution acidity is illustrated in Fig. 14.

In the absence of FC-1100, although bubble size distribution was slightly larger in high acidity solutions, the average bubble sizes generated in the low and high acidity solutions were very similar. In the presence of FC-1100, bubbles generated in high acidity solutions were on average 0.101 standard deviations smaller in diameter than those generated in low acidity solutions. Xu et al. also reported similar observations when studying the effect of NaCl concentrations on bubbles generated from a sparger in the presence of fixed amounts of sodium dodecyl sulphate surfactant (Xu et al., 2009). The higher ionic concentration at high acidity solutions was speculated to result in closer packing of the surfactant molecules by reducing the repulsive forces between them (Xu et al., 2009). In our case, bulk concentration of FC-1100 in high acidity solutions might be lower, resulting in more FC-1100 molecules packing at the interface. This can be inferred from the fact that the average surface tension of low acidity solutions is higher than that of high acidity solutions. The closer packing of the surfactant molecules at the oxygen/solution interface lowers the surface tension and results in smaller bubbles.

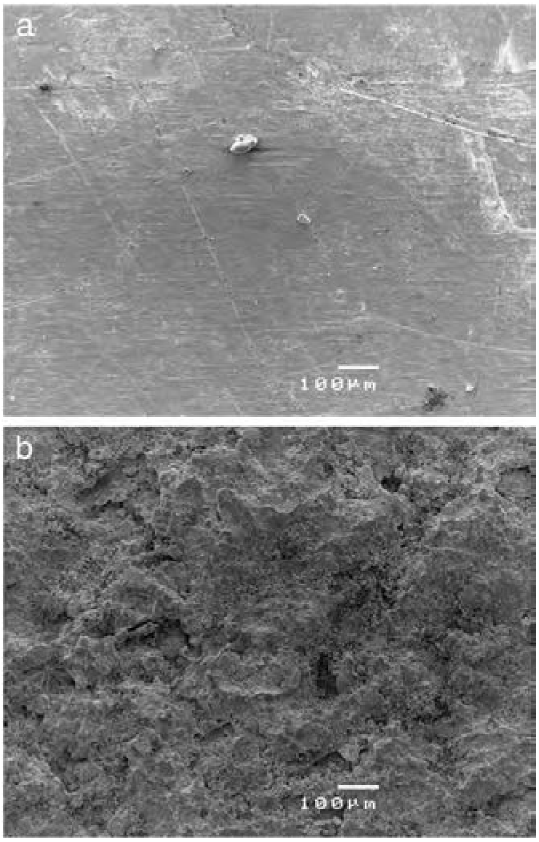


Fig. 10. The surface morphology of (a) a new anode (b) an old anode.

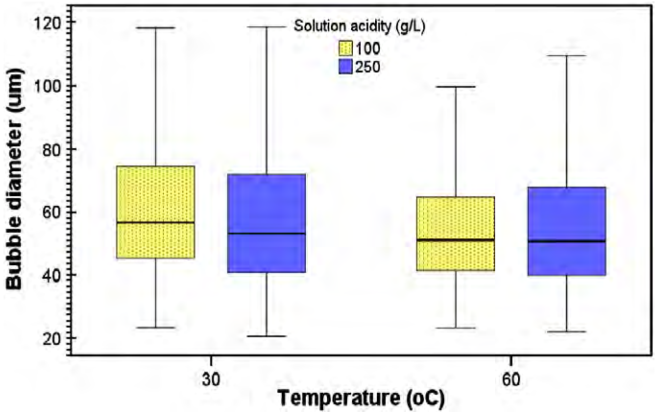


Fig. 12. The interaction between temperature and solution acidity.

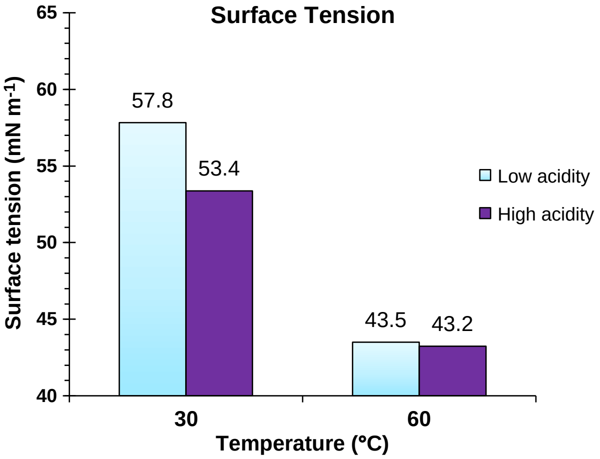


Fig. 13. Effect of temperature and acidity on surface tension.

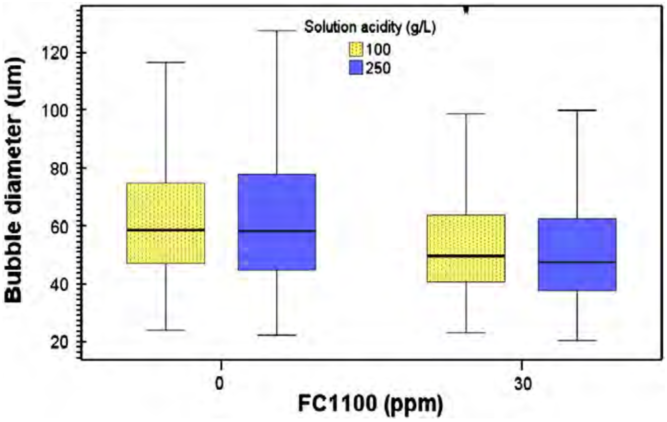


Fig. 14. The interaction between FC-1100 and solution acidity.

4. Conclusions

A high-speed camera and image analysis software were used to measure the size of the bubbles generated during the copper electrowinning process. Measurements, made under a variety of operating conditions, showed that of the five test parameters, FC-1100 and solution temperature had the most significant effect on the average size of the oxygen bubbles. Bubble size and bubble size distribution were noticeably reduced in the presence of FC-1100 or when the solution temperature was increased. FC-1100 and solution temperature both influenced bubble sizes by reducing the overall surface tension of the solution. To a lesser extent, anode age also influenced the detachment size of bubbles. Older anode generated larger bubbles and wider bubble size distributions mainly due to its surface roughness variations.

The data produced by the present work can be used to investigate the correlation between bubble size and the amount of acid mist, facilitating the design of strategies for minimizing acid mist generation in the tankhouse. The data can also be used to calculate the rate of bubble generation and release under a variety of operating conditions. Such information would be extremely useful in simulating the flow pattern of electrolyte solution between electrodes, which can be utilized to enhance the design of the electrodes and cells leading to improved electrolyte flow and copper quality at higher current densities.

Acknowledgment

The authors gratefully acknowledge the partial financial and technical support of Xstrata Technology. We also would like to thank Mr Tony Ruddell and Mr Curtis Arrowsmith who assisted us with the design and fabrication of the apparatus used in the test work.

References

Al Shakarji, R., He, Y., Gregory, S., 2010. Measurement of bubble size distribution in copper electrowinning process by image analysis. In: Harre, J. (Ed.), 7th International Copper Conference. GDMB, Hamburg Germany, pp. 1237–1251.

Al Shakarji, R., He, Y., Gregory, S., 2011. Statistical analysis of the effect of operating parameters on acid mist generation in copper electrowinning. Hydrometallurgy 106 (1–2), 113–118.

Cohen, J., 1988. Statistical Power Analysis for the Behavioural Science. Lawrence Erlbaum Associates Inc, New Jersey.

Drelich, J., Miller, J.D., 1994. The effect of solid surface heterogeneity and roughness on the contact angle/drop (bubble) size relationship. Journal of Colloid and Interface Science 164 (1), 252–259.

Field, A., 2005. Discovering Statistics Using SPSS. SAGE Publications Ltd., London.

Harris, D.C., 2007. Quantitative Chemical Analysis. Freeman and Co., New York.

Houlachi, G.E., Edeards, J.D., Robinson, T.G., 2007. Copper Electrorefining and Electrowinning, V. Canadian Institute of Mining, Metallurgy and Petroleum, Quebec Canada.

Huet, F., Musiani, M., Nogueira, R.P., 2004. Oxygen evolution on electrodes of different roughness: an electrochemical noise study. Journal of Solid State Electrochemistry 8 (10), 786–793.

Kazakis, N.A., Mouza, A.A., Paras, S.V., 2008. Experimental study of bubble formation at metal porous spargers: effect of liquid properties and sparger characteristics on the initial bubble size distribution. Chemical Engineering Journal 137 (2), 265–281.

Liow, J.L., Frazer, A., He, Y., Eastwood, K., Phan, C., 2007. Acid Mist Formation in the Electrowinning of Copper. Chemeca, Chemeca, Melbourne Australia.

Meiron, T.S., Marmur, A., Saguy, I.S., 2004. Contact angle measurement on rough surfaces. Journal of Colloid and Interface Science 274 (2), 637–644.

Montgomery, D.C., 2005. Design and Analysis of Experiments. John Wiley & Sons Inc, New York.

Robinson, T., Lang, J., Isbell, L., 1994. Cerro Colorado Copper Company SX-EW Plant. ISA PROCESS Tankhouse, Colorado USA.

Xie, G.X., Luo, J.B., Liu, S.H., Guo, D., Li, G., Zhang, C.H., 2009. Effect of liquid properties on the growth and motion characteristics of micro-bubbles induced by electric fields in confined liquid films. Journal of Physics D — Applied Physics 42 (11), 12.

Xu, Q., Nakajima, M., Ichikawa, S., Nakamura, N., Roy, P., Okadome, H., Shiina, T., 2009. Effects of surfactant and electrolyte concentrations on bubble formation and stabilization. Journal of Colloid and Interface Science 332 (1), 208–214.