ALBION PROCESS™

2020 Compendium of Technical Papers

GLENCORE TECHNOLOGY

A GLENCORE COMPANY

> “Albion Process™ exceeded expectations. Even with large swings in sulphur grade and feed rate, we maintain gold recovery in the cyanidation plant above design targets. Without Albion Process™ we would only achieve around 20% gold recovery and huge cyanide consumption, but with Albion Process™ we can achieve over 95% gold recovery and minimal cyanide consumption, giving the process plant outstanding return on investment.”
>
> – GPM Gold

Albion Process™ at a Glance

- → Real-world success in delivering over 99% leach recovery
- → Replaces large complex high-cost capital assets with simple low-cost capital assets
- → Tolerates a broader low grade more variable content feed and works where others fail
- → Incredibly fast and simple to commission and operate
- → Modular so it can be implemented in stages but maintain economic performance
- → Strongest performance guarantee in the world

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First Commercialisation of the Albion Process™ for Copper

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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.

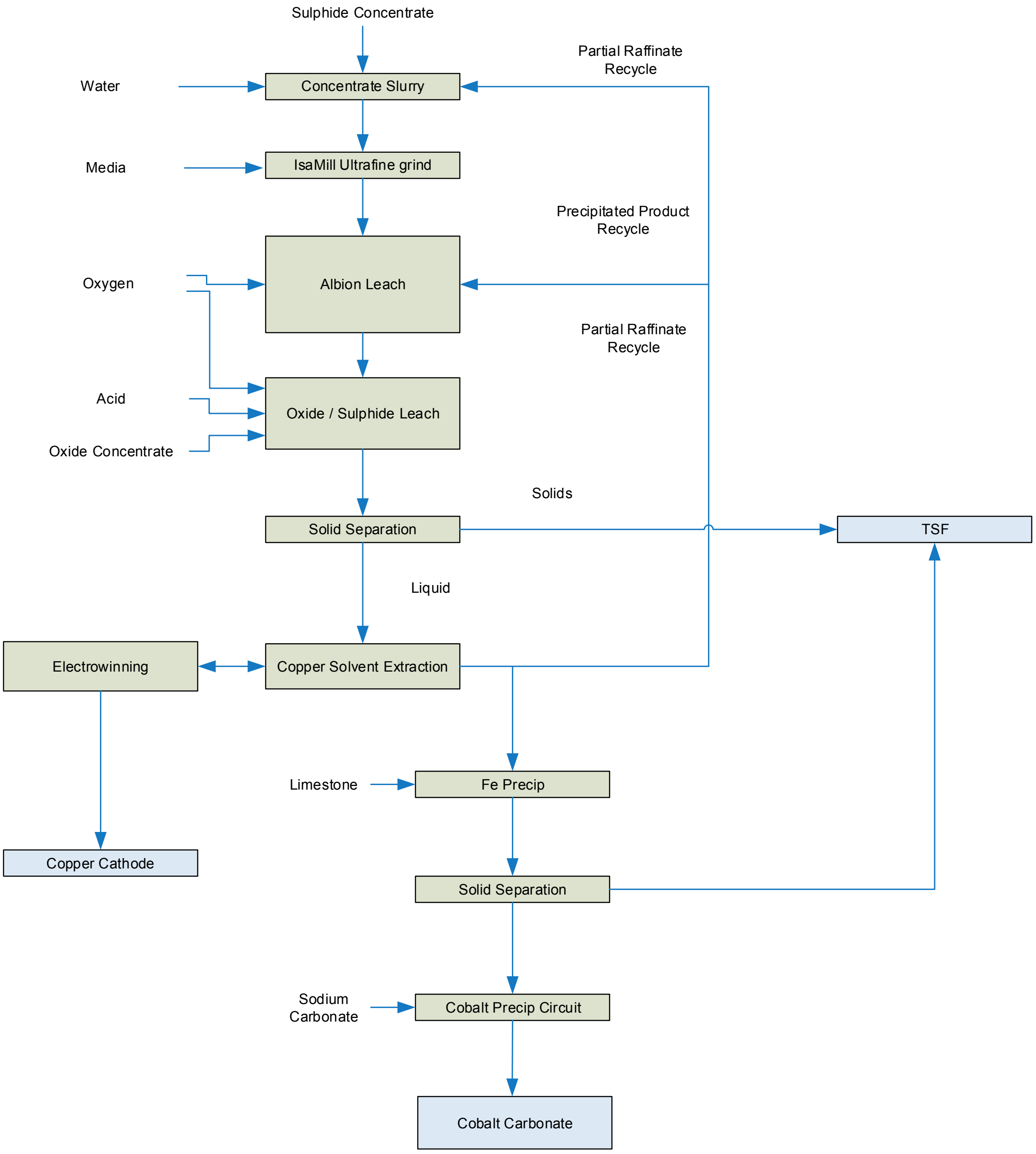

Figure 1 Sable Albion Process™ Plant Block Flow Diagram

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.

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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^{TM} manufacture			//			🟩	🟩	🟩	🟩	🟩	🟩				
IsaMill^{TM} 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.

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Ramp-up and long-term performance of the Albion Process™ plant at GeoProMining Gold Armenia

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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.

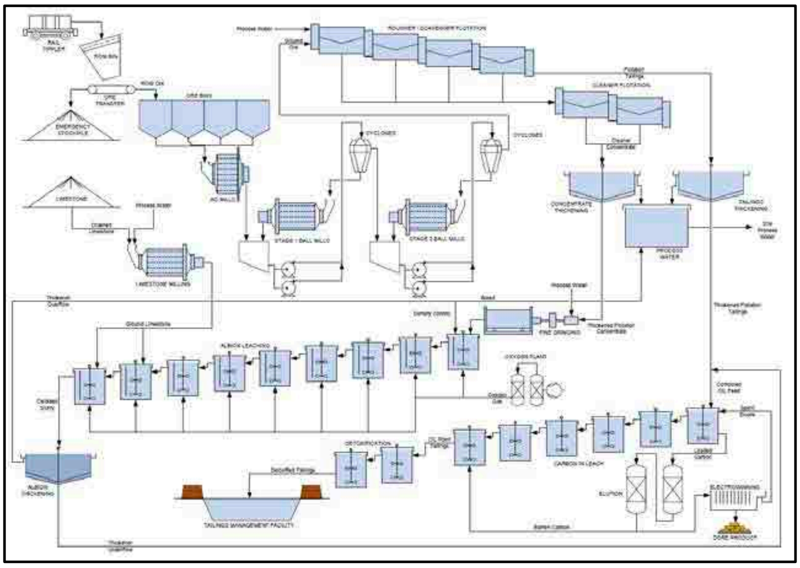

Figure 1: GPM flowsheet.

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™.

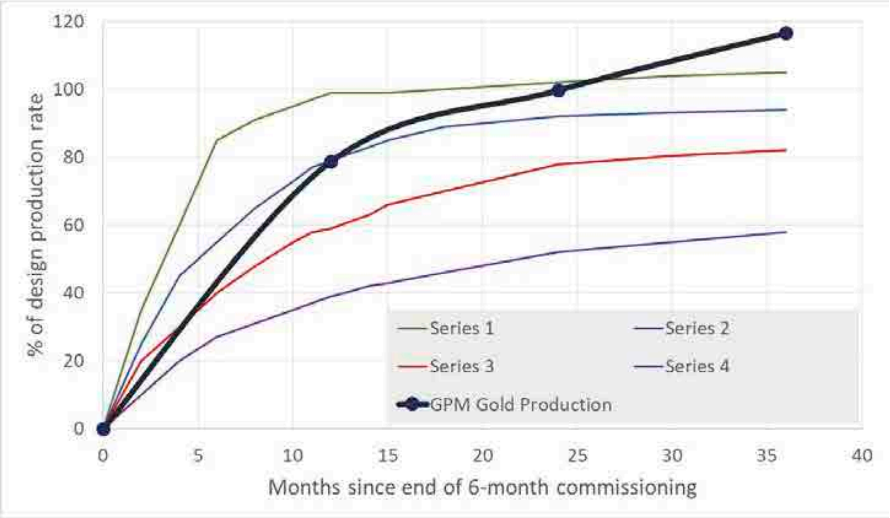

Figure 2 – McNulty curve for GPM process plant.

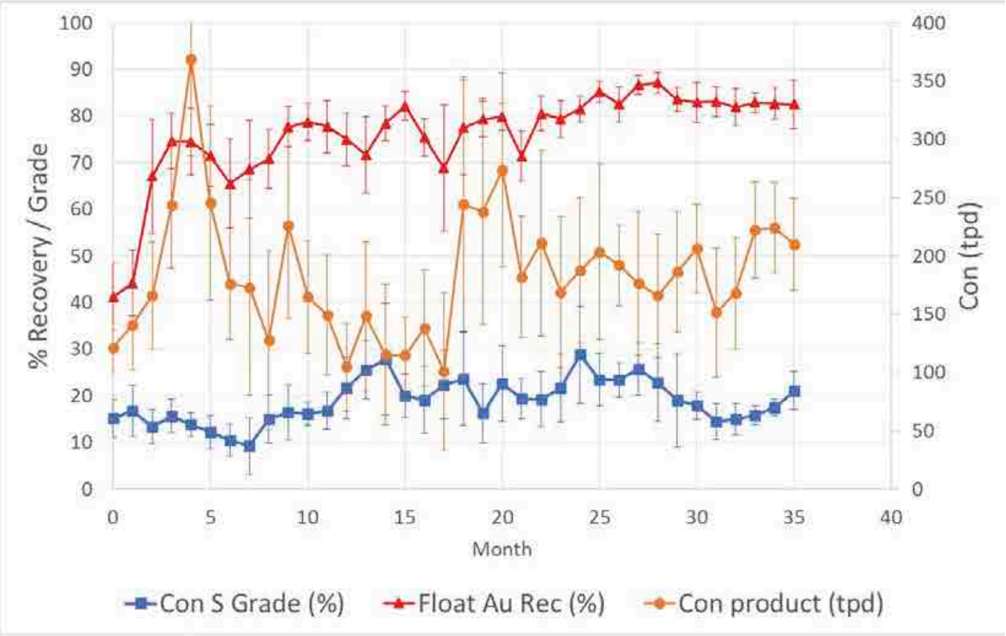

Figure 3: Concentrator performance.

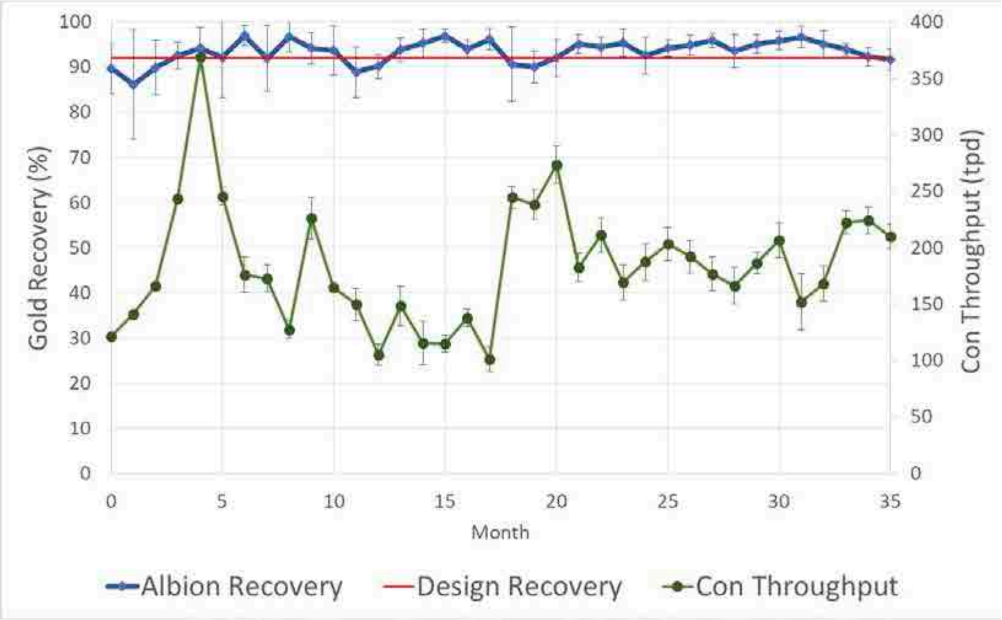

Figure 4 - Gold recovery versus throughput.

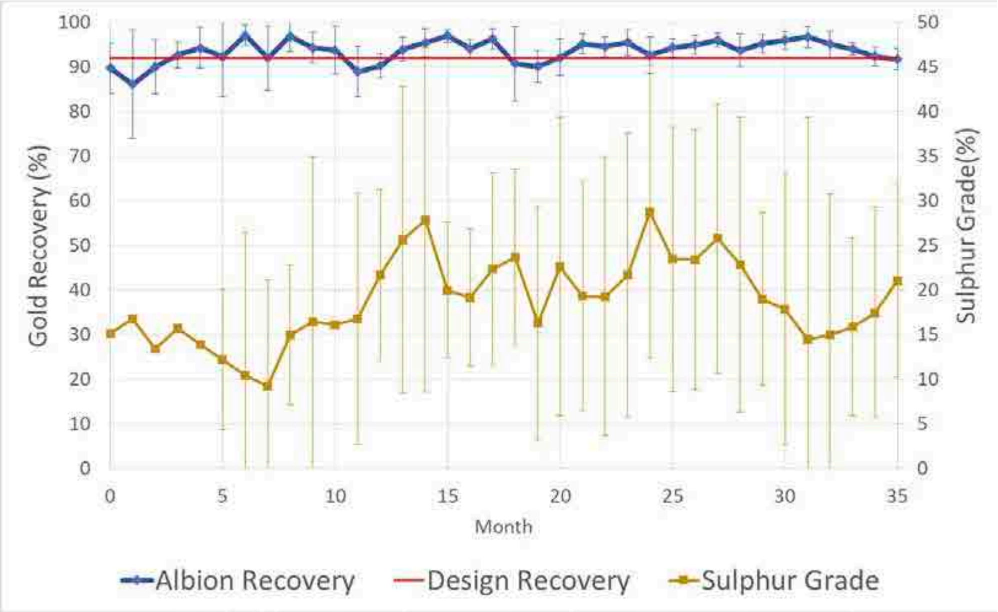

Figure 5 - Gold recovery versus concentrate sulfur grade.

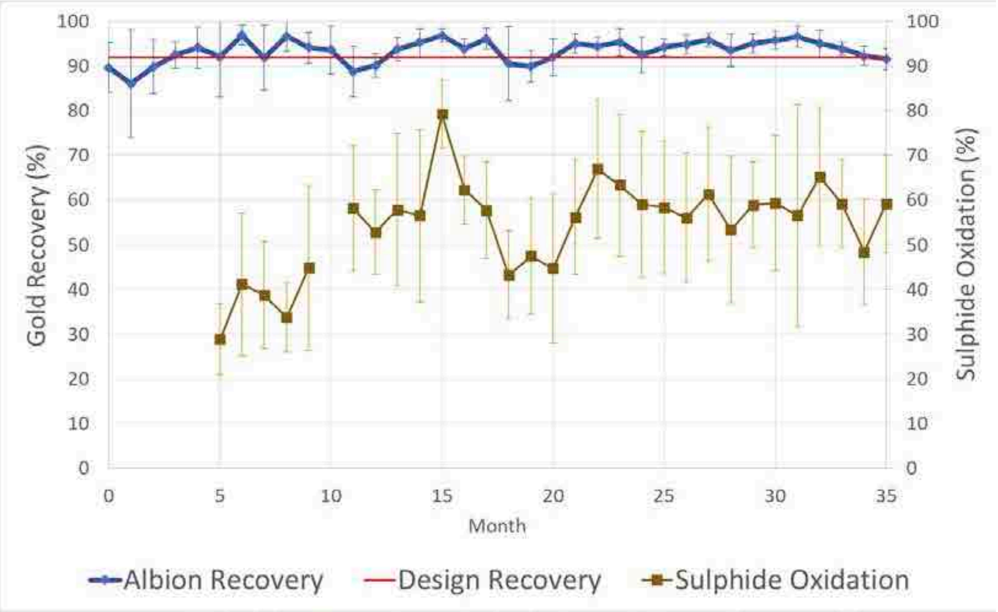

Figure 6 – Gold recovery versus degree of sulfide oxidation.

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

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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™ 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™ 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™ 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^{TM} or POx.

Albion Process™ Flowsheet

The flowsheet for the Albion Process^{TM} circuit is based on the circuit installed at GPM Gold and incorporates learning based on the performance of this circuit.

In the Albion Process^{TM} flowsheet, the flotation concentrate Thickener underflow concentrate is pumped to the IsaMill^{TM} feed pumpbox where it is combined with media before being pumped to the M3000 IsaMill^{TM}. The IsaMill^{TM} 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^{TM} maintenance activities are performed. The concentrate slurry is then transferred to one of the first three Albion Process^{TM} leach reactors. The oxidative leach circuit flowsheet consists of six, 340 m^3 Albion Process^{TM} 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^{TM} oxygen gas injection lances. The process is designed to run autothermally at or around 93ºC.

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^{TM} 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^{TM} 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^{TM} 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^{TM} 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.

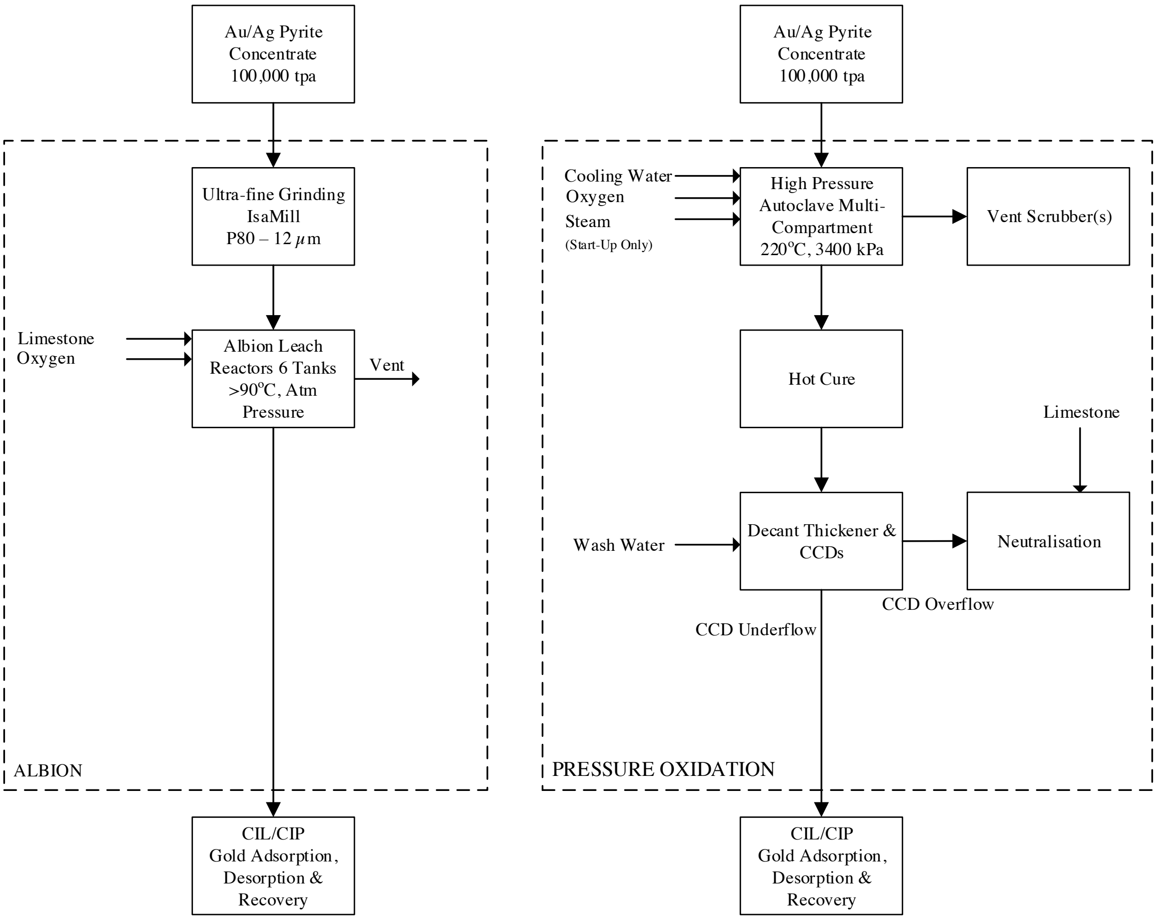

Figure 1 – Albion Process™ and POx Flowsheet Comparison

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: $2\ FeS_2 + 7\ O_2 + 2\ H_2O \leftrightarrow 2\ FeSO_4 + 2\ H_2SO_4$

$$2\,FeSO_4 + H_2SO_4 + \tfrac{1}{2}\,O_2 \leftrightarrow Fe_2(SO_4)_3 + H_2O$$

$$Fe_2(SO_4)_3 + 3\,H_2O \leftrightarrow Fe_2O_3 + 3\,H_2SO_4$$

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 93ºC 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 Flowsheet
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™ 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™ 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.

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The Albion Process™ at the GPM Gold Project – The success of a technology

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CIM Conference 2018,
Stream: A Vision For The Future - Mineral Processing In Canada

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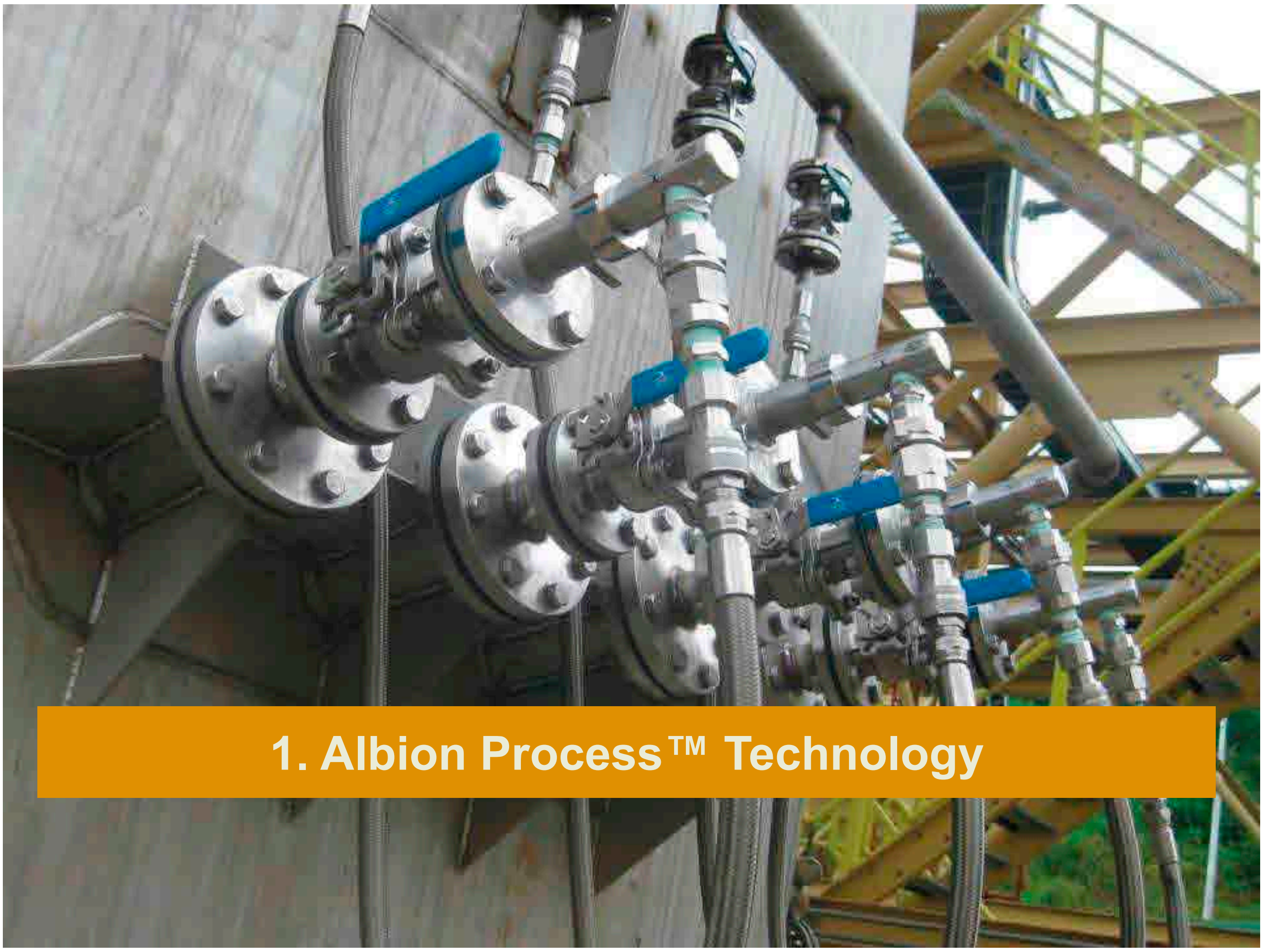
GPM Gold – The success of a technology

Outline

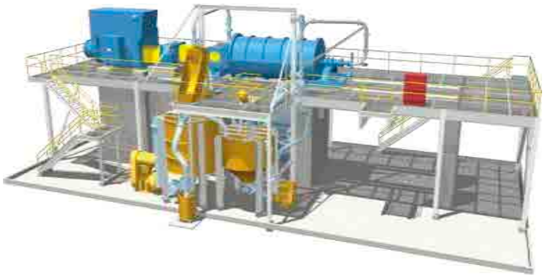
1. Albion Process™ technology
2. Performance of the Albion Process™ at GPM
3. Review of global installations
4. The Future for Refractory Gold Projects



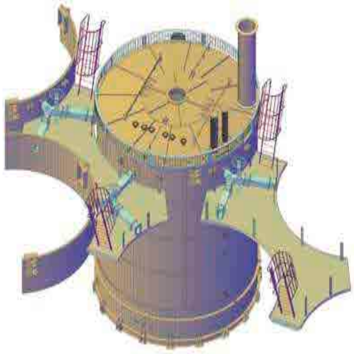
1. Albion Process™ Technology

Albion Process™

The Albion Process is a combination of mechanical and chemical liberation

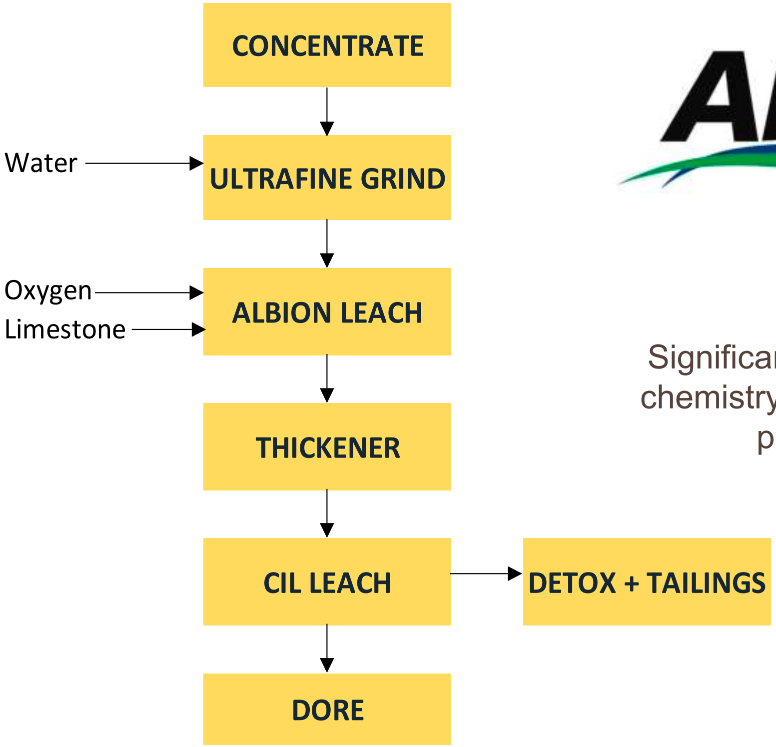


- **Ultrafine grinding:**
  - *IsaMill*^{TM} *stirred mill:*
  - *FeS_2 = 80 % passing 10 microns*
  - *$CuFeS_2$ = 80 % passing 12 – 18 microns*
  - *$Ni_9Fe_9S_{32}$= 80 % passing 10 – 14 microns*
  - *ZnS = 80 % passing 16 – 20 microns*



- **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

Limestone

ALBION LEACH

THICKENER

CIL LEACH

DETOX + TAILINGS

DORE

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

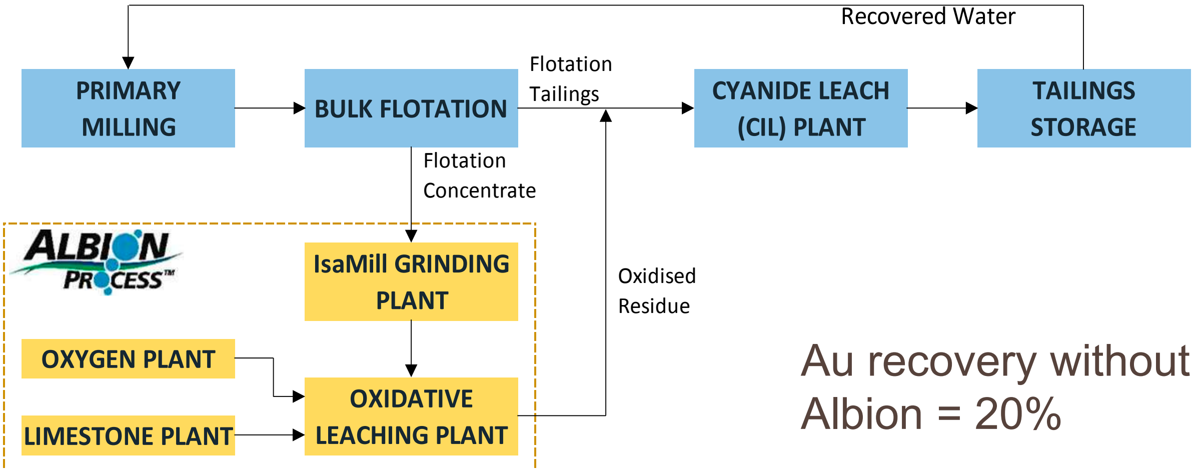
GPM Gold – The success of a technology

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

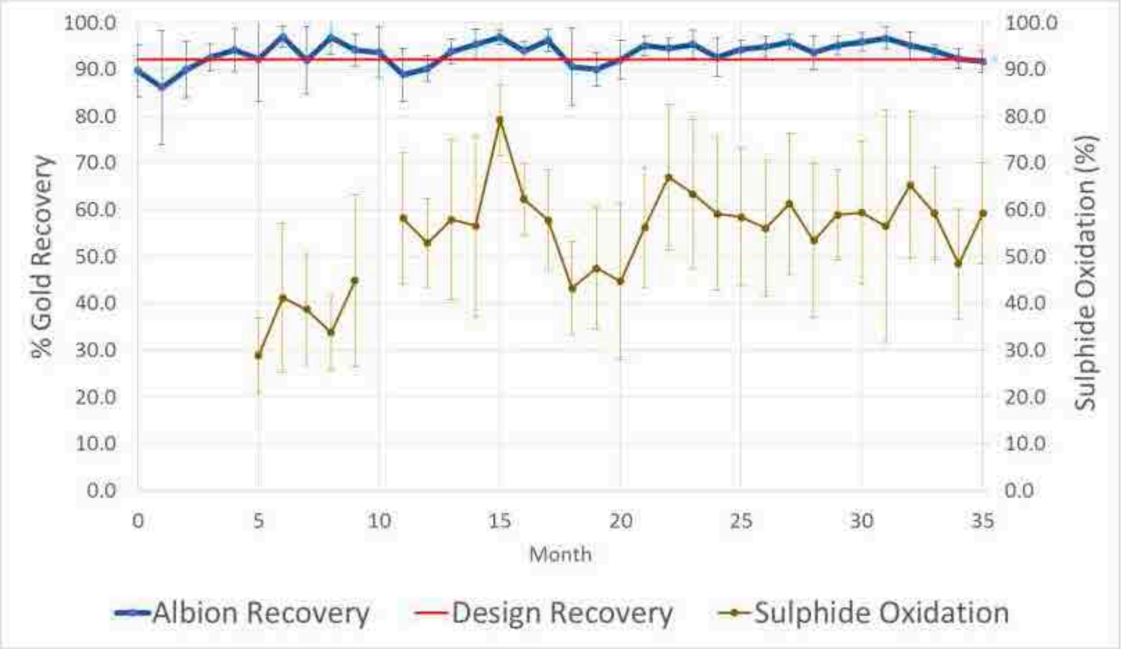


Au recovery without Albion = 20%

Au recovery with Albion = 95-98%

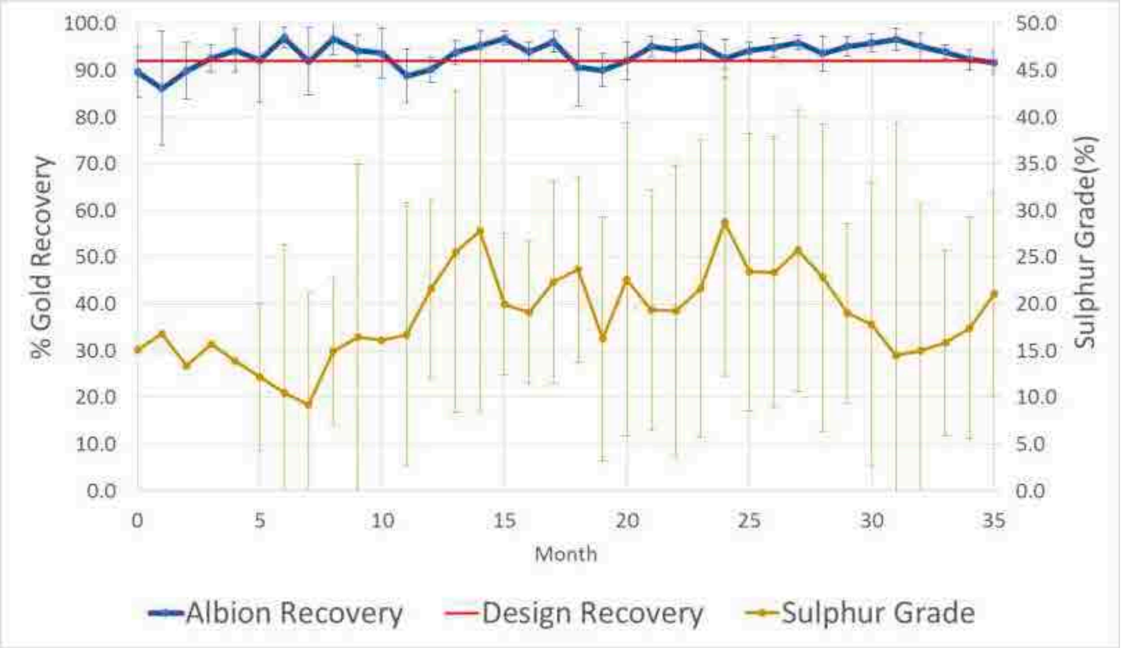
GPM Albion Process™ Plant Performance

3 years of stable gold recoveries



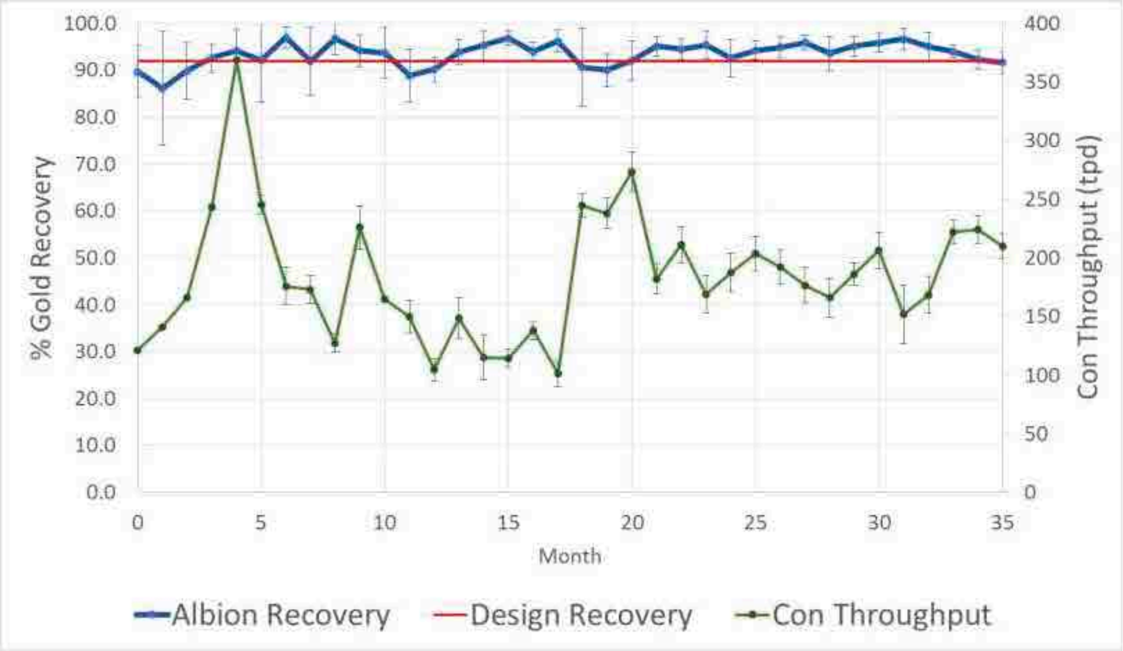
GPM Albion Process™ Plant Performance

Variable Sulphur Grades (<10% to >30%)



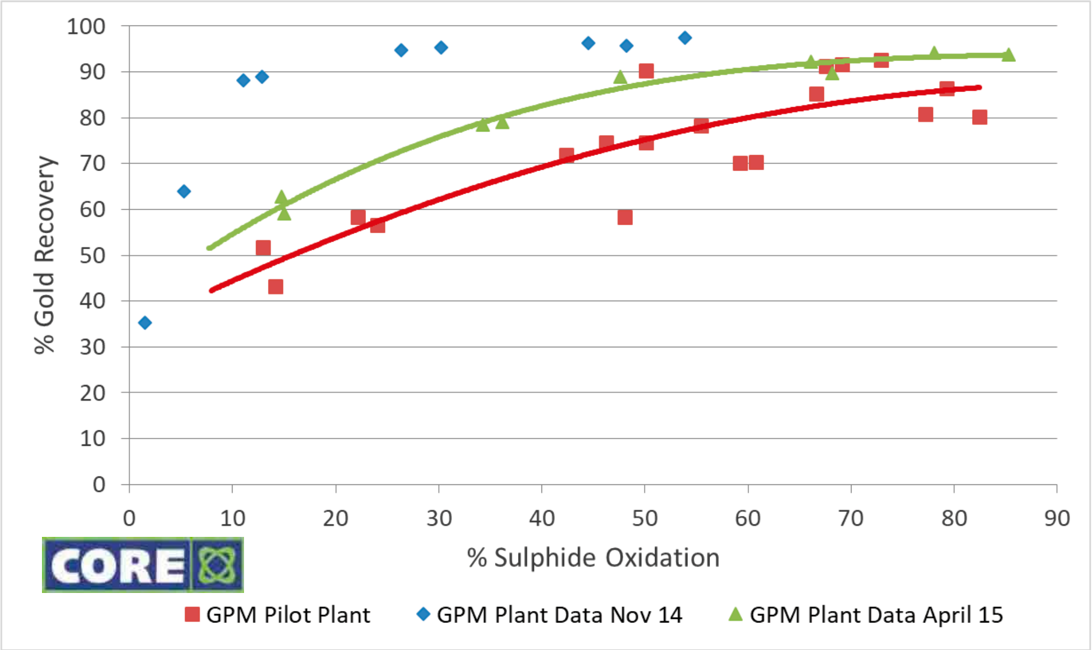
GPM Albion Process™ Plant Performance

Variable Throughput (100 – 350 tpd)



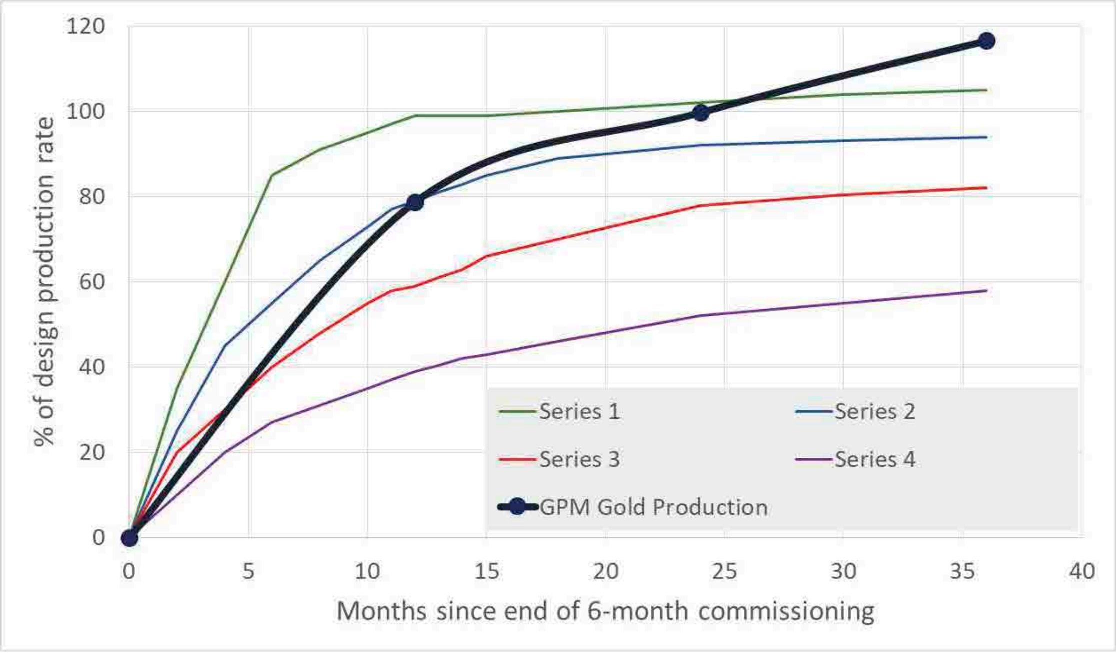
GPM Albion Process™ Plant Performance

Better than Pilot Plant



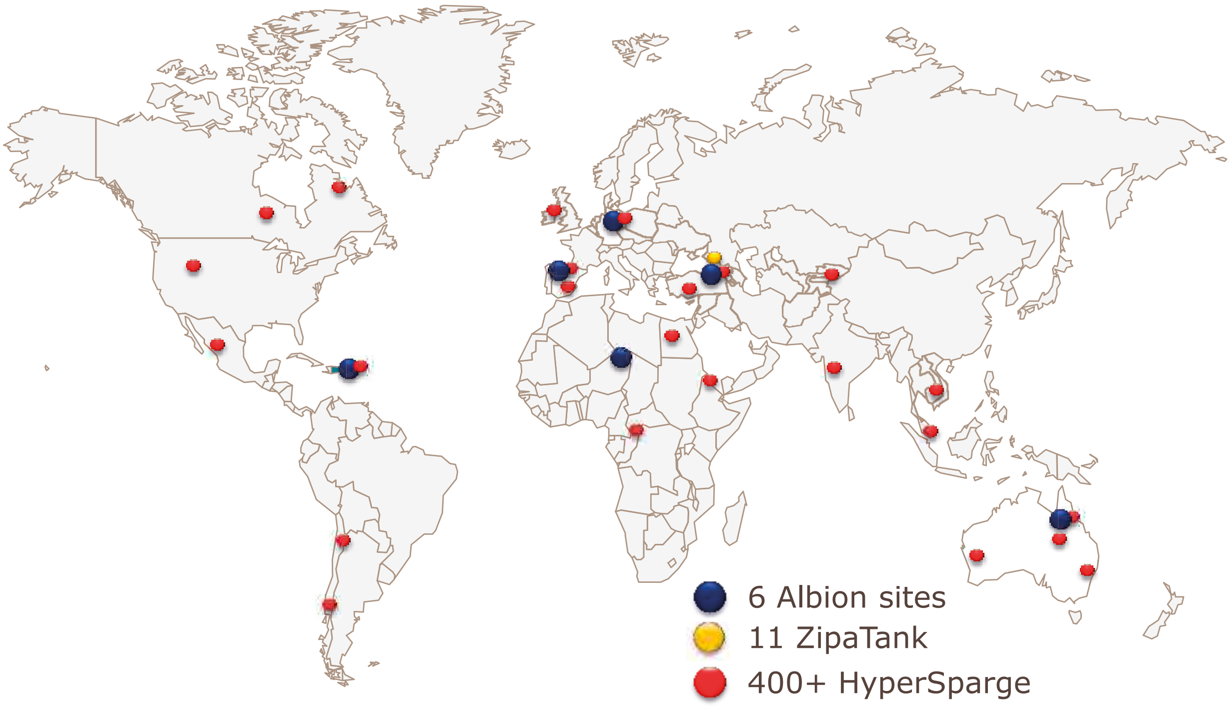
GPM Ramp Up – Relative McNulty Curve

Series 1-2 Performance



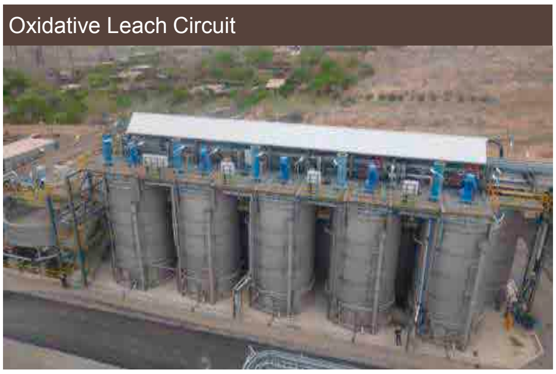
3. Albion Process™ - Global Installations

Albion Process™ Installations



GPM Gold Project – Albion Process™ Plant

Oxidative Leach Circuit



M5000 IsaMill™

Limestone Grinding Plant

Commodity – Gold

Location – Armenia

Client – GeoProMining

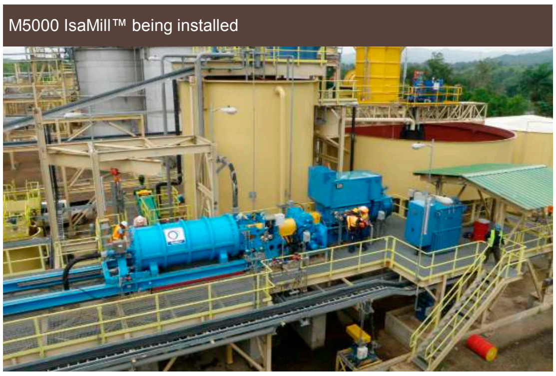
Refractory pyrite concentrate

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

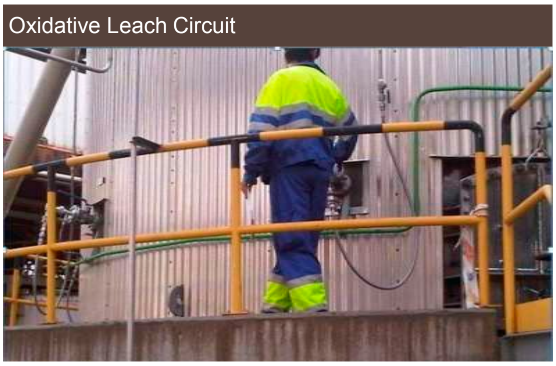
10,000 tpa copper cathode

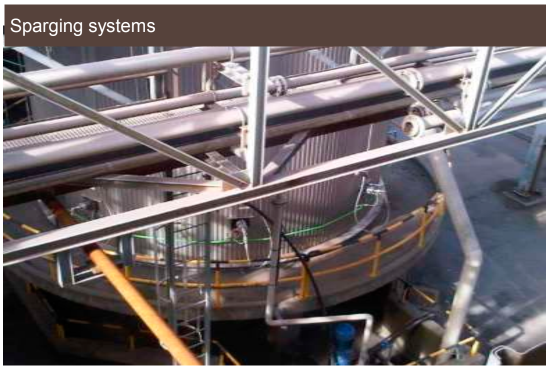
>99 % copper recovery

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

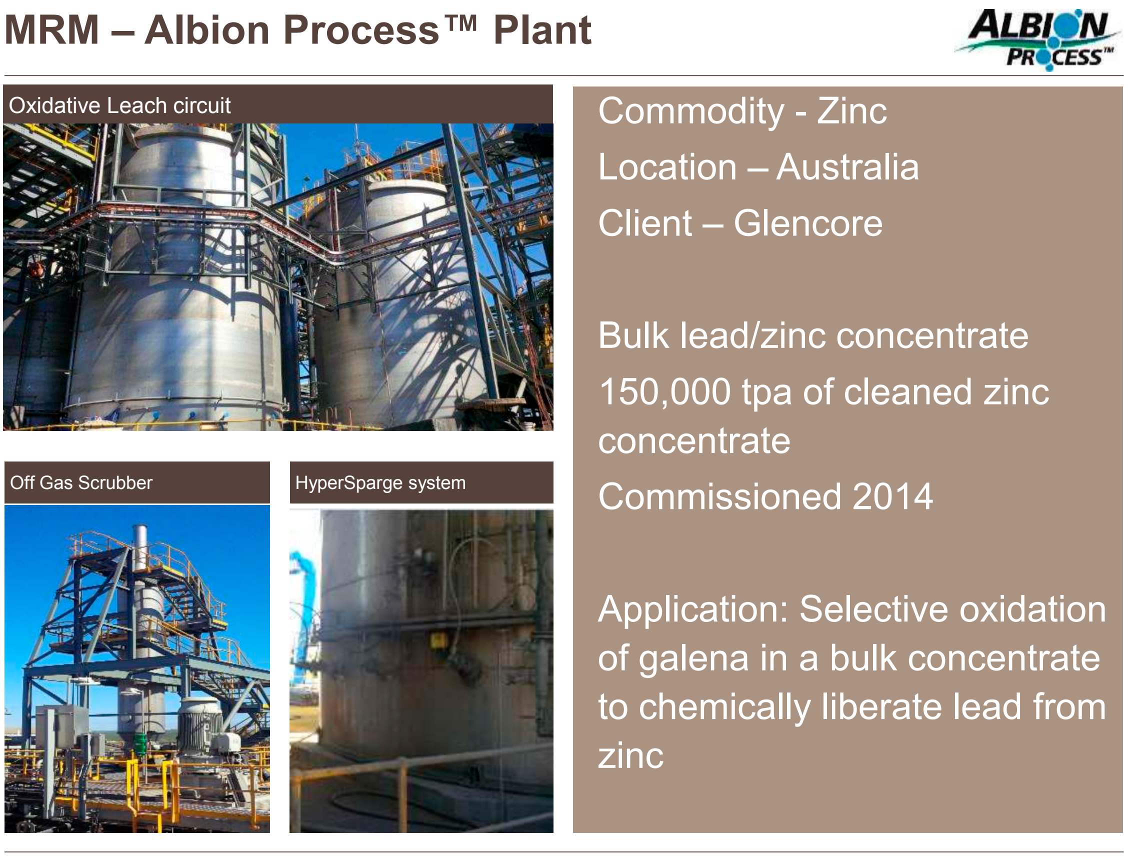

Oxidative Leach Circuit


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

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

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:

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Thank You.

MOUNT ISA MINES NECESSITY DRIVING INNOVATION

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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.

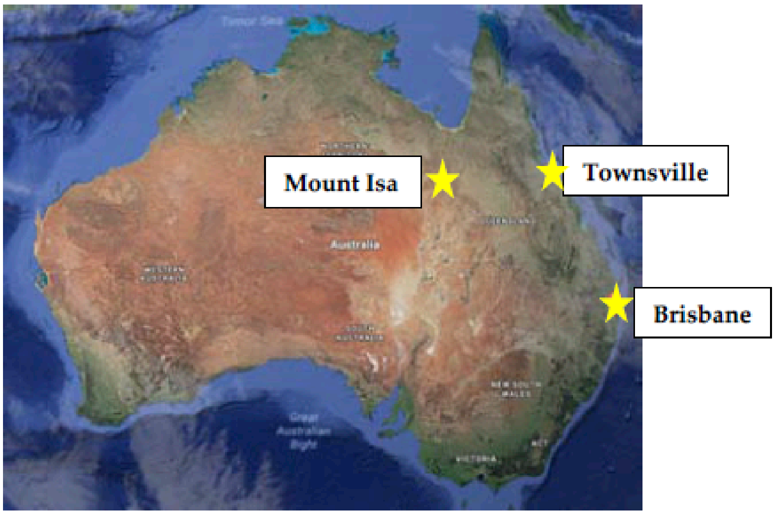

Figure 1 – Location of Mount Isa and Townsville relative to Brisbane the nearest Capital City

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.

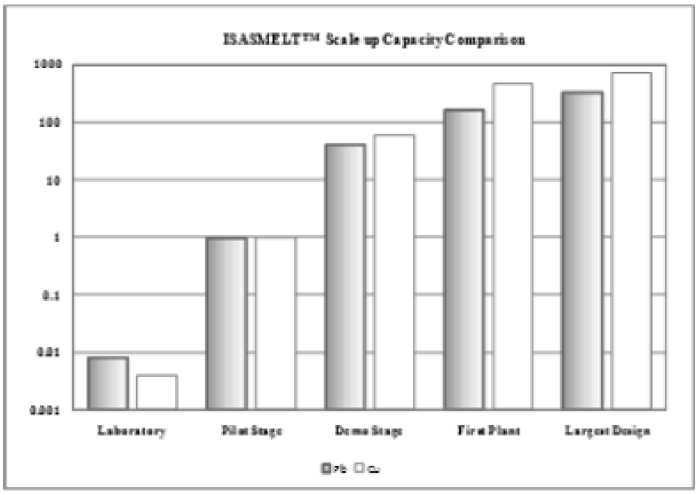

Figure 3 – Lead and copper ISASMELT™ Scale up comparison

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

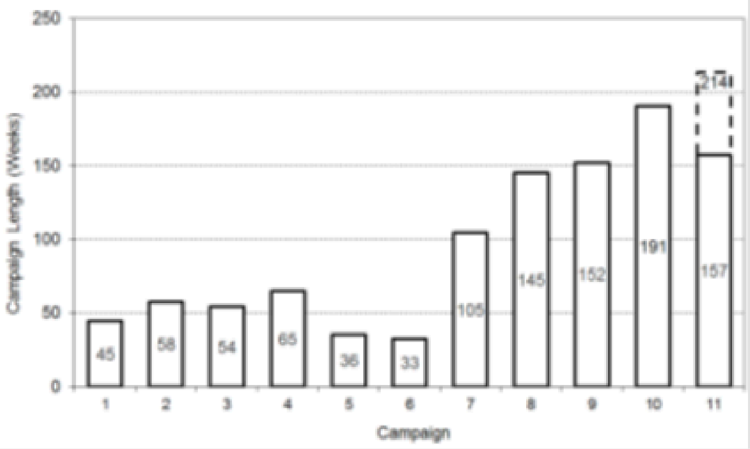

Figure 5 – Mount Isa copper ISASMELT™ plant campaigns (as of 2013)

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.

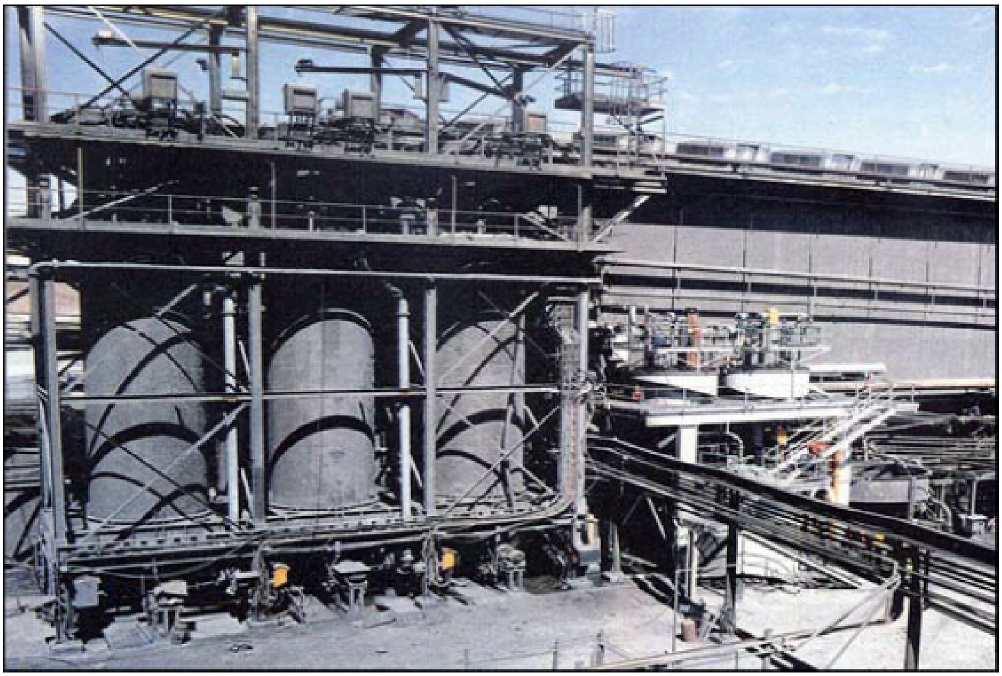

Figure 6 – Jameson Cells compared to columns of the same capacity at Mount Isa

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.

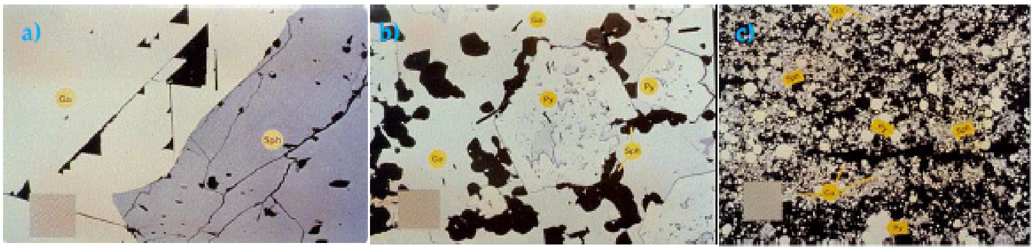

Figure 7 – Photomicrograph of a) Broken Hill ore b) Mount Isa ore c) McArthur River ore

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^{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.

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Oxygen Mass Transfer in the Albion Process™: from the Laboratory to the Plant

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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.

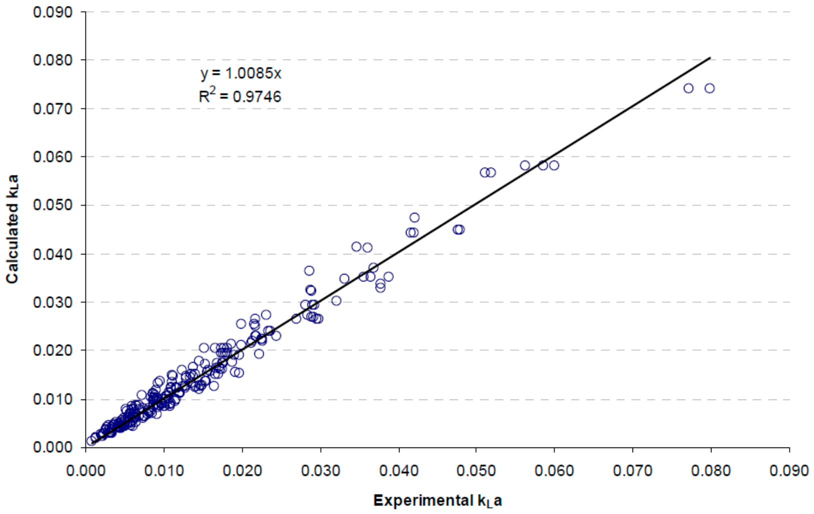

Figure 1 Experimental and calculated k_La for a system using GT experimental methodology

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.

$$W = (n/(n-1))mRT[(P_2/P_1)^{(n-1)/n} - 1 \qquad [4]$$

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.

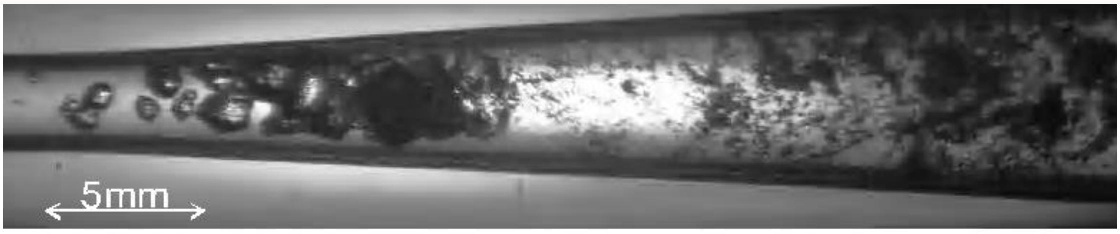

Figure 2 Collapsing bubbles crossing the shock wave of a supersonic nozzle (Kawamura et al., 2004)

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.

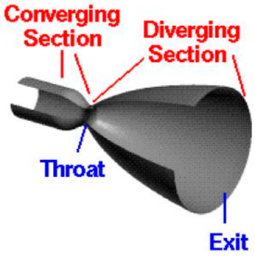

Figure 3 Cross section of a convergent-divergent nozzle (NASA, 2015)

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.

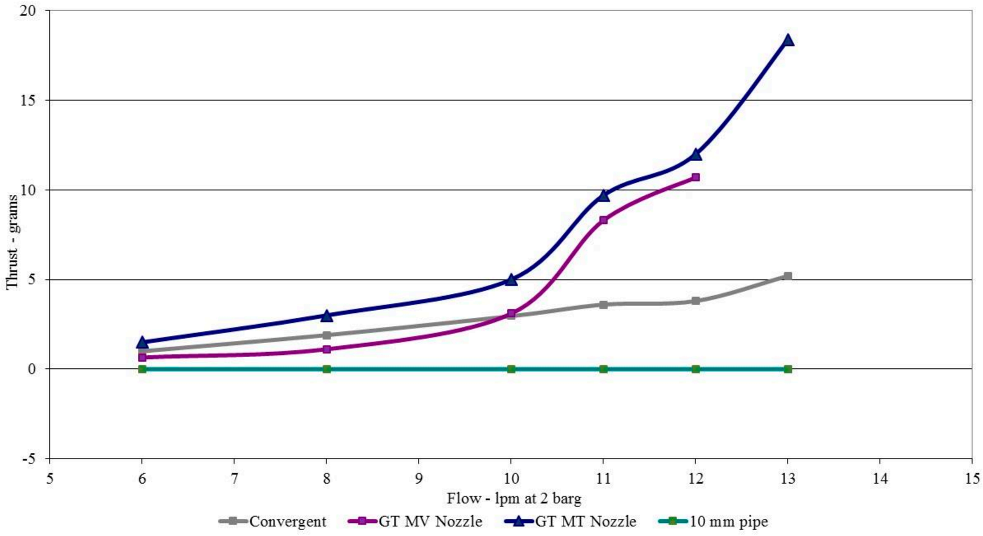

Figure 4 Developed thrust measured at 200kPa

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 (kLa)

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 kLa 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 kLa was more pronounced, and was up to 300 % of the value observed for the conventional open pipe and 50% compared to a converged nozzle.

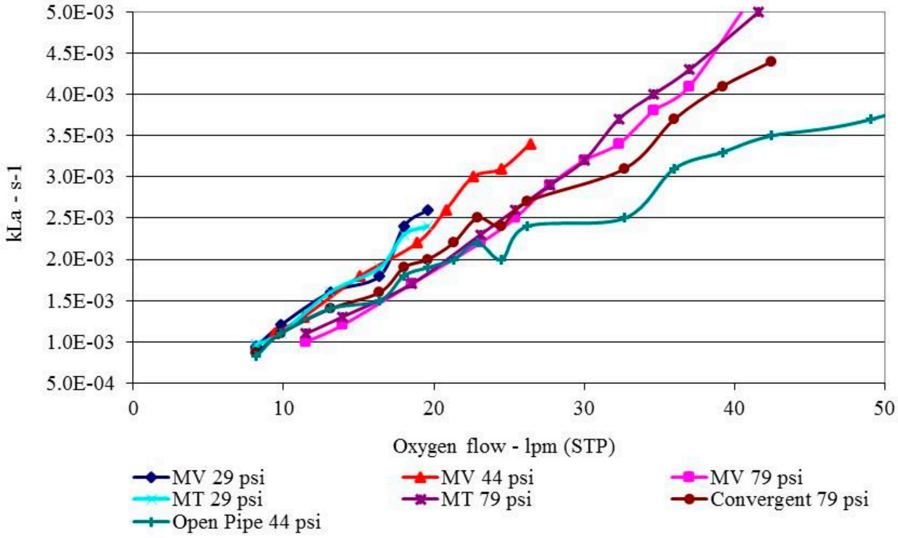

Figure 5 kLa at different supply pressure

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 kLa, 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.

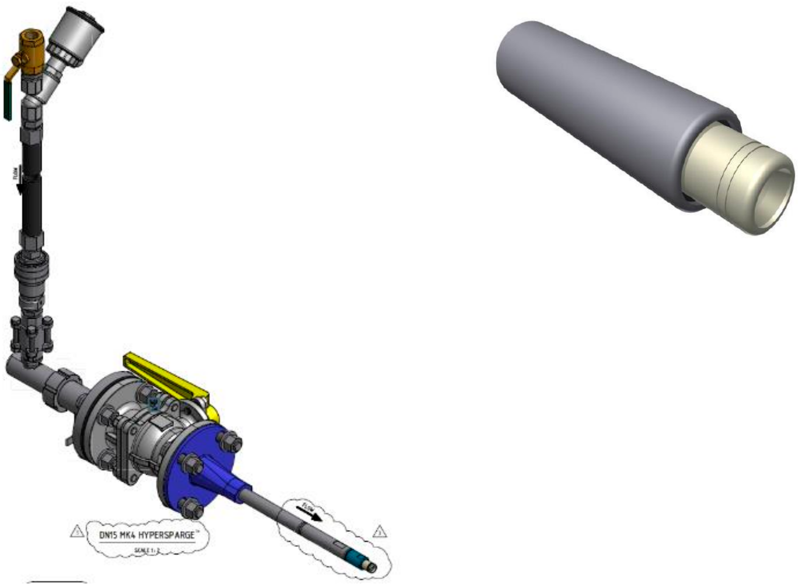

Figure 6 HyperSparge™ supersonic gas injector and nozzle

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.

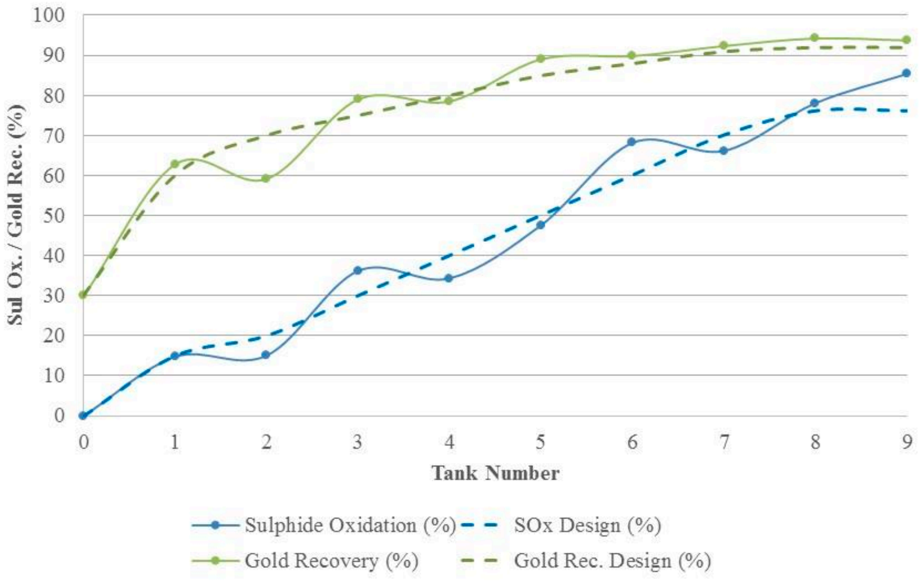

Figure 7 Sulphide Oxidation survey of GPM Gold Albion Process™ Plant.

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.

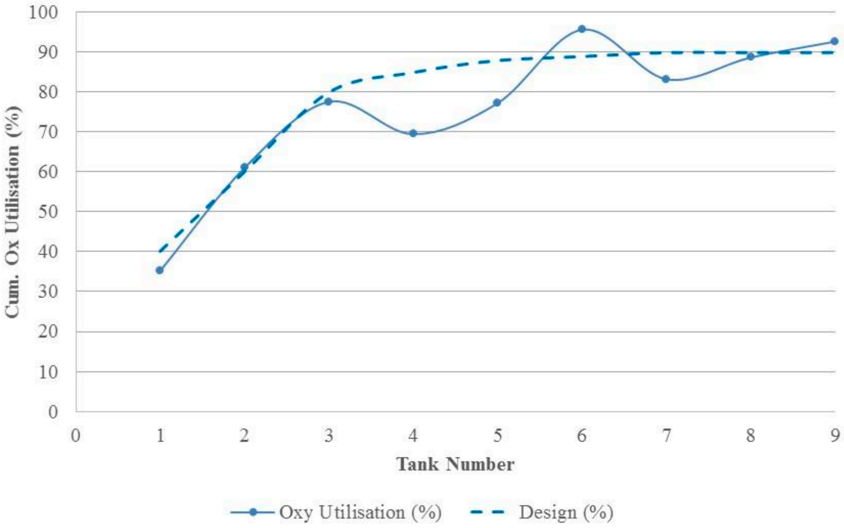

Figure 8 Oxygen utilisation at the GPM Gold Albion Process™ Plant

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.

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ECONOMIC RECOVERY AND UPGRADE OF METALS FROM MIDDLING AND TAILING STREAMS

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ABSTRACT

As mine head grades decline and orebodies become more complex, traditional mineral processing techniques and flowsheets to achieve saleable concentrate become more difficult to design and construct. Mines with lower quality concentrates or concentrates with penalty elements are particular under threat. The economics of these operations are far more susceptible to metal price, concentrate treatment terms and the availability of other, cleaner concentrates. Additional value may be realised for these orebodies through improved recovery by producing a low grade middling concentrate for further processing, in conjunction with a saleable concentrate.

The most cost effective way to reduce impurity levels is to do so as early as possible in the mining value chain. Technologies such as fine grinding and fine particle flotation are well established as effective methods for impurity rejection in mineral processing. What is normally overlooked is how a hydrometallurgical process could also be integrated in the overall flowsheet to achieve higher overall recovery at the mill. In the base metals environment, this is mainly because hydrometallurgical processes are associated with production of metal or use of expensive and toxic precipitating agents once the minerals of interest are solubilised. These processes can be very expensive, particularly with rising power costs and poor economies of scale in capital costs associated with low production rates from middling streams.

Glencore Technology (GT) has recent experience in the treatment of middling and low grade concentrate streams as well as tailings streams to compliment a concentrator flowsheet in a refractory gold and base metals setting. The value proposition is the isolation of a low grade middlings concentrate from the primary circuit or the tailings stream for upgrading to an intermediate product with an equal or higher grade than the primary concentrate to allow blending for sale. This allows plants to operate on a more favourable part grade-recovery curve while avoiding the expense of metal production. For existing operations this is particularly attractive since it can be added on with no process interruptions.

Two case studies are examined showing flowsheets and costings to arrive at the value proposition of the GT low grade treatment flowsheet.

KEYWORDS

Albion, IsaMill, Jameson cell, Glencore

INTRODUCTION

One significant challenge facing the gold and base metals mining industries is the globally observed trend of reducing mined head grades. This challenge is faced with existing operating assets and presents a significant hurdle in the justification for new projects. Figure 1 shows global trends in mined head grade from the mid 1800's to 2010 (CSIRO 2015).

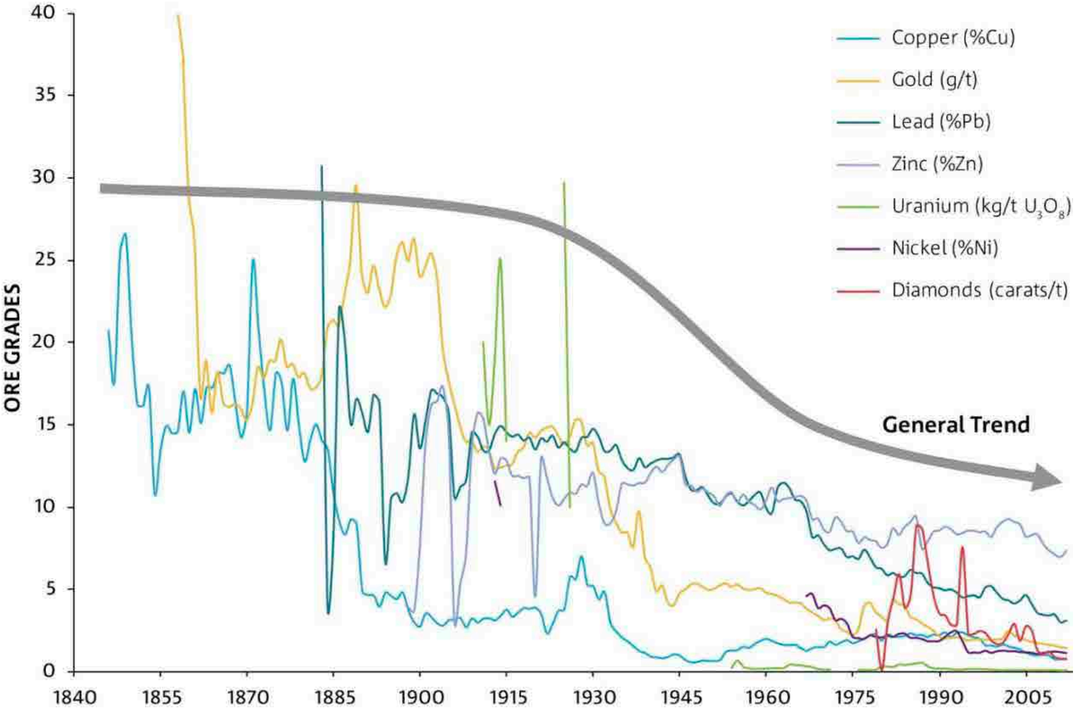

Figure 1 – Mined head grades for base metals and gold.

New projects or orebodies within existing operations, where reasonably good head grades can be maintained, tend to defy this globally observed trend through a corresponding increase in geological and metallurgical complexity. This is important for miners because metal input can be maintained without increasing milling rates however for a given flowsheet the quantity and quality of metal outputs will be compromised. While throughput of high-grade, complex orebodies can maintain input metal units they may not achieve target grade or recovery, and may even introduce penalty elements to the concentrate (Munro 2015), effecting economic viability.

By way of example, challenges faced with complex ore treatment in a flotation concentrator where the concentrate would be directed to a smelter include:

- Inability to find an economic operating point on the grade recovery curve such that both grade or recovery is not achieved for economic sale to a smelter
- Sacrificing the final product concentrate grade by inclusion of a middling concentrate that serves to increase overall recovery but negatively impacts grade and introducing penalty elements
- Inability to separate the economic minerals in an orebody rendering the production of a bulk concentrate with poor terms of sale to a smelter

Figure 2 illustrates the general trend of how grade versus recovery curves have become less defined as recovery increases.

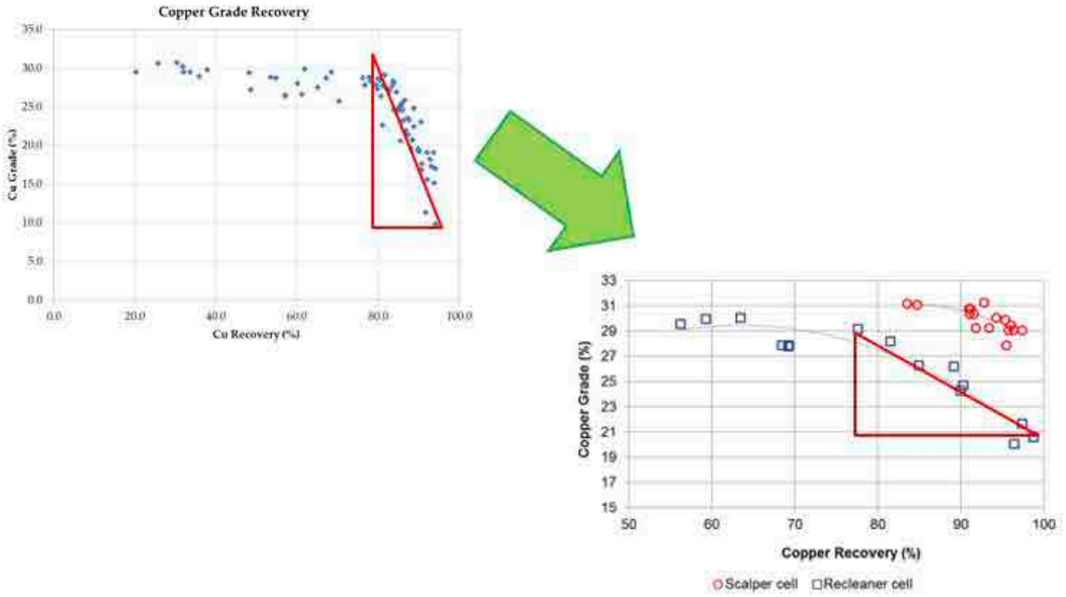

Figure 2 – Less defined grade recovery curves lead to lower concentrate grade to achieve target recovery

In general, the gradient of the grade recovery curve has decreased resulting in a non-optimal operating position on the curve. This is a reflection of the general increasing complexity of ore to maintain mined head grade. The reason the ore complexity is reflected in Figure 2 is that minerals become more difficult to separate from one another and are recovered together, (Young 1997).

Certain ore-types cannot be upgraded with mineral processing techniques to produce a concentrate for downstream treatment in smelters. One example of these ores are the highly weathered or oxidised ores that are treated via heap leaching or whole ore leaching for copper and cobalt recovery. This processing method suits certain ore types and where sufficient infrastructure can be established at low cost. For example, heap leachable ore, requires ore with minimum levels of competency and permeability when it is stacked. The mineralisation must be such, that it is readily acid leached, with sufficiently low net acid consumption to be economic. A separate plant is then required to recover the copper once leached into solution, and then solvent extraction and electro-winning plants are required to produce saleable copper cathode. This flowsheet is conditional on the availability of reliable and relatively cheap power.

Ore types that don't fit neatly into conventional flotation or heap leaching flowsheets have traditionally been relegated to waste. These ores are generally referred to as Complex Ores. Increasingly, these ore types can no longer be viewed as waste due to the contained metal content and high costs of pre-stripping this material where it overlies more economic deposits. Often these pre-stripping costs can make a project uneconomic unless metal can be recovered from this waste material.

Treatment of Complex Ores Through the Concentrator

During the mining boom of 2004 to 2009 the challenges of increasing metal prices coupled with the increasingly long lead times to bring a project to production led to mining companies adopting a strategy of a standard concentrator design (Combes 2011). The standard concentrator would allow significant improvement in the design and procurement phases of a project, and allow projects to be implemented faster to take advantage of rising metal prices. A common circuit used in a standard concentrator design is reflected in a simplified flowsheet in Figure 3.

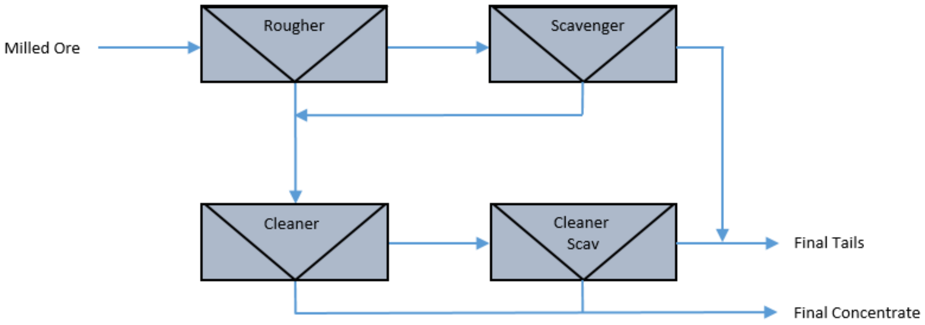

Figure 3 – Conceptual flotation plant

The standard concentrator is a valid concept where a number of concentrators are envisaged to be built across projects with identical geological and metallurgical characteristics. The standard concentrator is still an excellent concept where there will be some variation between projects where some slight modification to the standard concentrator can be tolerated. When processing complex ores, however, where a single deposit may have multiple complex metallurgical domains, it becomes very difficult to design a single flowsheet that can treat all ore types while maximising economic performance.

Figure 2 shows that treatment of ore with increasing complexity in a set flowsheet will result in a compromise between recovery and grade or contamination of the final concentrate with deleterious metals and gangue. This can significantly impact the economics of a mining operation, (Munro 2015).

Modification of the Concentrator for Complex Ores

Modifications made from the conventional flowsheet for the treatment of complex ores generally comprises two approaches. The first is increasing the extent of grinding. This is based on the premise that when grade or recovery of an economic mineral is not obtained it is not liberated from gangue and does not have exposed surfaces to float. While there is some focus on the primary grind, usually when ore complexity increases, concentrate re-grinding, using a fine grinding mill such as the IsaMill™ is included in the circuit (Burford 2007). This style of flowsheet is reflected in Figure 4.

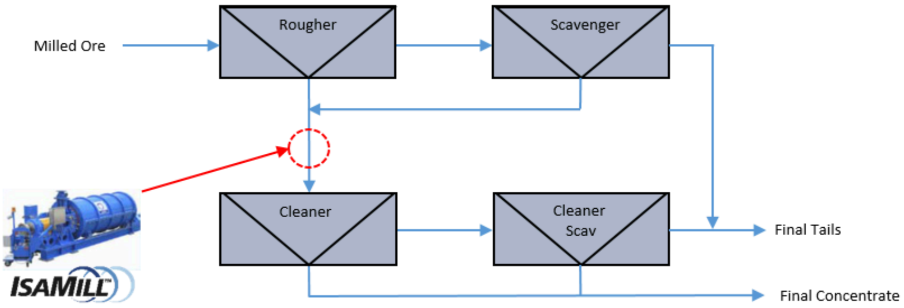

Figure 4 – Modification of the conventional flowsheet for fine grinding

The addition of concentrate regrinding to a flotation circuit as shown in Figure 4 is based entirely on increasing liberation. Once the minerals are liberated they must then be floated which can create further issues as finer particles have slower flotation kinetics compared to courser particles, and require more residence time to achieve the same recovery.

The second modification to the standard flowsheet commonly used is an increase of residence time through the installation of more flotation capacity or modification of the circuit configuration as reflected in Figure 5.

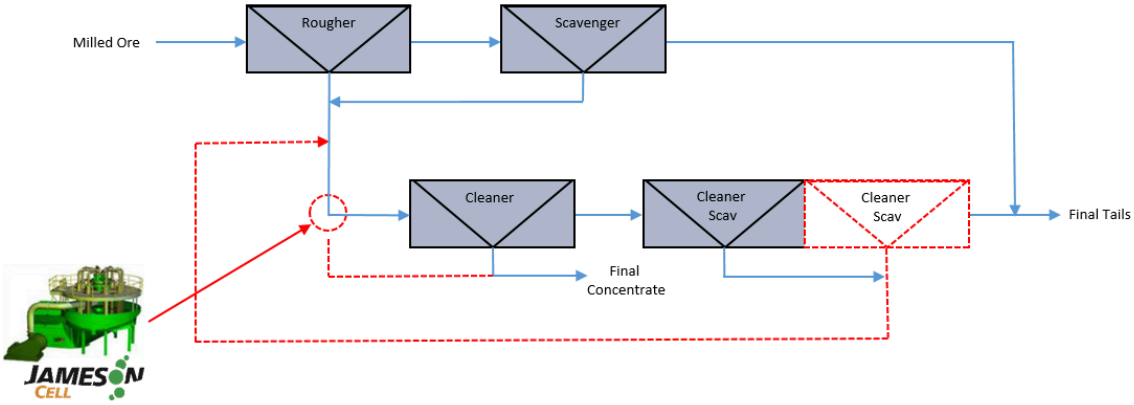

Figure 5 – Modification of the conventional flowsheet for circuit configuration or longer residence time

For fine particle flotation, a properly designed circuit may have the benefits of reducing recirculating loads, reducing reagent demand, as well as improved metallurgy. Examples such as McArthur River Mines routinely produce average zinc concentrate less than 10um in size, (Pease 2004). Alternatively, different flotation equipment can be added to the flowsheet such as the Jameson Cell which is well documented in the literature and proven in the field for improved fine particle flotation compared to a mechanical cell (Young 2006).

The basic modifications to the conventional flotation flowsheet are all validated when they result in improvements in the grade and recovery of the valuable metals to economic levels. Some extremely complex ore types, however, will still not respond fully to such modifications and there is a need for a more encompassing approach.

A further complication occurs when designing a new project or modifying an existing operation for the treatment of a more complex ore, where the circuit design is based on the treatment of the most complex metallurgical domain in the orebody. This leads to circuit complexity that is not needed for a large percentage of the ore treated. Due to the mining sequence and certain ore types not stockpiling well, the more complex ores will not be treated in discrete campaigns. The result can be the installation of excess flotation and re-grinding capacity that is not utilised all of the time, resulting in an inefficient use of capital.

GT Process for Treatment of Complex Ores

Over the past 20 years there have been significant advances in technology and equipment in the fields of mineral processing and hydrometallurgy in the mining industry. GT has been at the forefront of these advances with the following technologies:

- IsaMill^{TM} – A high efficiency fine grinding technology in a horizontally stirred mill utilising inert ceramic media
- Jameson Cell^{TM} – A high intensity pneumatic flotation machine with no moving parts generating fine bubbles
- Albion Process^{TM} – Fine grinding followed by atmospheric leaching technology for refractory and base metals concentrates, including low cost recovery of base metals from solution to high grade concentrates with low grade reagents

The Albion Process^{TM} is a patented technology developed by Glencore in 1994. The Albion Process^{TM} consists of two key steps. The first step is ultrafine grinding of a sulphide concentrate, using Glencore Technology's IsaMill^{TM}, to particle sizes in the range 80 % passing 10 – 15 microns. The second step is an oxidative leach of the finely ground sulphides at atmospheric pressure to breakdown the sulphide matrix and liberate base and precious metals prior to metal recovery. There are currently five Albion Process^{TM} plants operating globally in base and precious metals duties.

As a response to the increasing ore complexity, GT proposes a flowsheet that is a combination of these recent advances in mineral processing and hydrometallurgy processes. A conventional flotation flowsheet is still adopted when designing for new projects or existing operations encountering increasing ore complexity but with the addition of a hydrometallurgical processing step to deal with the low grade concentrates bled from the flotation circuit to smooth our variations in plant operation. One version of the concept is illustrated in Figure 6.

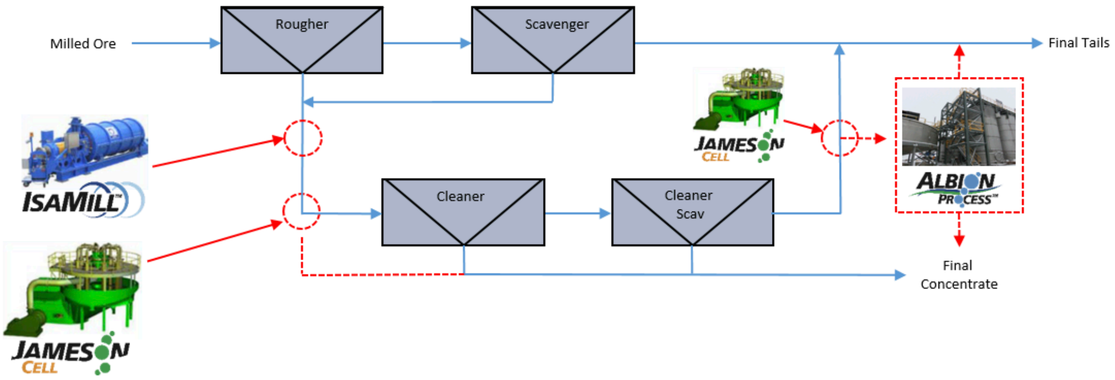

Figure 6 – GT concept for complex ore treatment

Figure 6 shows how the conventional flowsheet may be modified in an example where the concentrate reporting to the cleaner circuit is treated through an IsaMill™ and then a Jameson Cell prior to the Cleaner bank, obtaining grade, however the recovery is not at target levels meaning that further cleaning must be employed to achieve recovery. The Cleaner and Cleaner Scavenger banks provide further recovery, however the combined concentrate is below target, resulting in an overall dilution of the concentrate grade. In the modified flowsheet, a Jameson cell treats the Cleaner Scav tailings to recover a low grade concentrate. This low grade concentrate is bled from the circuit and processed in a dedicated hydrometallurgical plant. Complex middling particles recovered in the Jameson cell are removed from the circuit and are not recirculated through the flotation plant.

The fine grinding stage prior to the cleaning circuit allows for high pull rates from the Rougher and Scavenger, improving primary circuit recovery. The use of the Jameson Cell allows for good quality concentrate to be produced after fine grinding, with the wash water on the cell reducing the recovery of non-gangue particles.

The operation of the Cleaner and the Cleaner Scav enables the operation to balance the grade and recovery to be achieved from the circuit. Too high a recovery from this circuit recovers not only wanted liberated valuable mineral, but also the locked and complex particles towards the end of the circuit, unnecessarily diluting the concentrate. There is also the possibility of penalty elements that could be recovered in the concentrate with too high recovery rates. Therefore recovery needs to be controlled to prevent these particles from being recovered and left in the Cleaner Scav tails.

The Cleaner Scav tails are treated with a Jameson Cell, further increasing circuit recovery, but targeting complex particles that cannot be collected to concentrate in their current state due to the low concentration of valuable minerals.

The concentrate collected from the Jameson Cell doesn't need to be high grade to be economically treated through the Albion Process™. Grades down to 5% copper in concentrate have found to be economic. One issue with including the Albion Process™ in the flowsheet is how to recover the metals that are leached into solution at low cost. This is achieved through a process developed by GT for zinc, copper, nickel and cobalt where either lime or limestone is used to continuously precipitate the base metals at controlled pH. A common problem with this type of precipitation process is the co-precipitation of gypsum. GT has developed procedures to overcome this issue by carefully controlling the process conditions such that the gypsum grows to particle sizes significantly coarser than the base metal oxides, and can be separated by hydrocyclone, as illustrated in Figure 7.

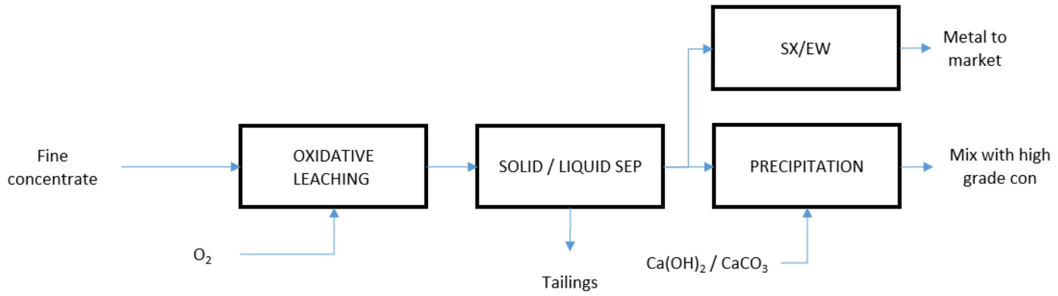

Figure 7 – Integration of the Albion Process™ into the mineral processing flowsheet

When applied to a copper circuit, the feed material to the Albion Process™ would be a low grade 5% copper middling concentrate which is then leached to solution and precipitated as a 45% copper oxide concentrate. This high grade intermediate can then be sold for use in a range of industries, or alternatively blended with the final concentrate product for sale.

CASE STUDIES FOR THE GT PROCESS

Two case studies are presented to illustrate the concept and high level information has been provided on the incremental improvement that can be achieved by incorporating the Albion Process™ into the concentrator flowsheet for processing a low grade middling stream.

Case 1 – Copper NW Queensland

Case 1 relates to an opportunity for brownfields expansion of mining and concentrator operations at a mine in North Queensland. The ore complexity at the operation will increase for a short duration due to the need to mine through mainly transitional/weathered ore zones with a variable base of oxidation, resulting in some primary zones intermixed with the transitional ores. The variation in the contact zone between the transitional and primary ores is such that the ores cannot be separated and must be treated together. They are in a quantity such that the contained metal units from both transitional and primary ore must be recovered to justify the overall project.

The main copper bearing minerals comprise native copper, chalcopyrite and chalcocite with minor chrysocolla and malachite. The sulphide gangue comprises mainly pyrite with minor pyrrhotite, galena and sphalerite.

The ore presented to the process typically grades 1 to 2% copper with varying mineralogy. Within the feed some of the ore types can be recovered to a concentrate grading greater than 25% copper, however for the pure transitional ores, the maximum copper concentrate grade is only 5% copper. The transitional concentrates, however, contain predominantly leachable minerals, and lend themselves to be separated to a middlings stream for separate hydrometallurgical processing. A simplified flowsheet is illustrated in Figure 8.

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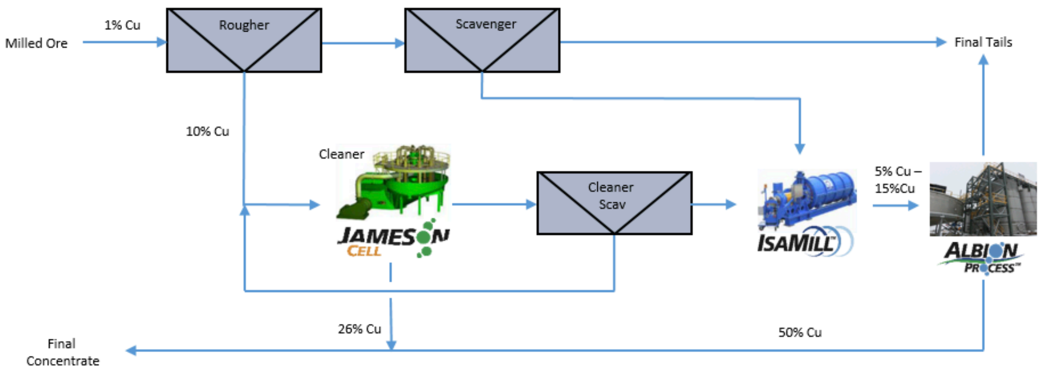

Figure 8 – Copper treatment flowsheet

The simplified flowsheet in Figure 8 indicates a pathway for liberated copper sulphide mineralogy to report to a final cleaner concentrate to achieve on specification concentrate at +26% copper grade for sale. The material that is not amenable to upgrading, diluted by both sulphide and non-sulphide gangue is recovered in both the scavenger concentrate and the cleaner tailings as a 5% Cu middling concentrate that is treated through the Albion Process™. Even at low feed grades the hydrometallurgy treatment option is economical since the final intermediate copper oxide produced grades approximately 50% copper and overall circuit grade and recovery are maximised.

The mineral processing and hydrometallurgy flowsheet adds between 4% to 30% copper recovery at target grade depending on what material is treated. The project allows access to a further 6.0 Mt of high grade primary ore. The treatment of the transitional cap alone has an IRR of approximately 25%.

Case 2 – Zinc NW Queensland

This application of the GT flowsheet to zinc processing is treatment of historical zinc tailings. It was acknowledged that significant zinc was contained in the tailings but when the tailings were re-floated a zinc grade of around 10% had to be accepted for any economic recovery levels due to both sulphide and non-sulphide gangue. A number of hydrometallurgy flowsheet options were considered to treat the low grade concentrate, however these options were marginal economically due to the high cost of installing expensive processing equipment to recover the final zinc metal product. The GT approach overcame these economic and technical limitations. A simplified flowsheet is illustrated in Figure 9.

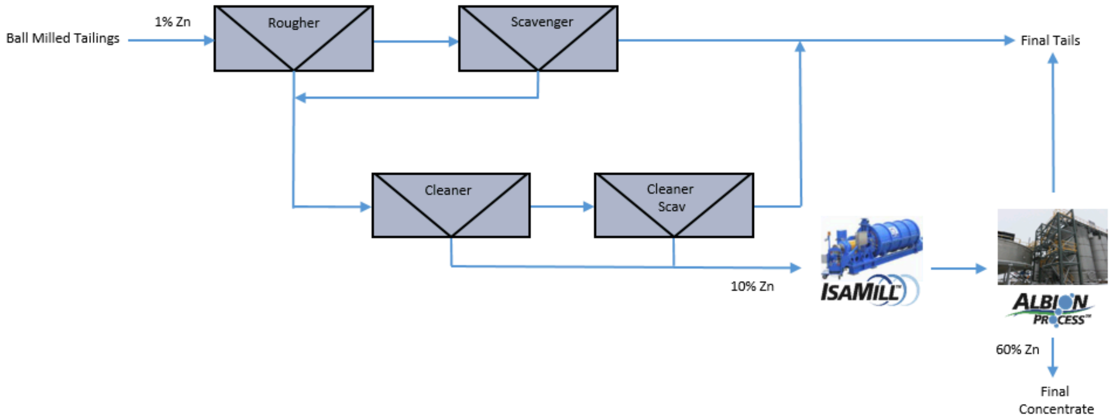

Figure 9 – Zinc treatment flowsheet

The flotation tailings can be re-floated in a conventional flowsheet to produce a 10% zinc concentrate at 90% recovery. The resulting concentrate is fed to an IsaMill^{TM} for grinding to 80% passing 20 micron or below and then to the Albion Process^{TM} oxidative leach for the extraction of zinc. The oxidative leach achieves zinc recoveries of up to 99.5%. Glencore has installed two Albion Process^{TM} plants to recover zinc from a bulk concentrate, in Spain and Germany, and has experience in the design of these plants (Hourn 2012).

Once in solution, rather than producing metal, the GT precipitation process was used with either lime or limestone to produce a zinc oxide at a grade between 50 to 60% zinc. Strategically, since most zinc roast-leach-electrowinning plants are limited at the roaster, such a zinc product can be treated by conventional zinc refineries to maximise cellhouse capacity or operate during roaster downtime.

The plant was sized to produce 100ktpa contained zinc metal but could be easily scaled down if required. The project IRR was 30%.

CONCLUSION

GT has developed novel flowsheet configurations for the treatment of complex ores through minor modification to conventional flowsheets with minimal process disruption and integration with hydrometallurgy unit operations. The flowsheets presented are just examples and many other variations are possible.

In the mining value chain, value can be most easily added when complexity in ores can be overcome at the earliest part of the chain as practicable. This starts in the mine with grade control, understanding metallurgical domain definition and optimising blast patterns. In the mineral processing sphere this can start with screening and dense medium separation through to grind size and reagent use. In extractive metallurgy this can mean blending with different feeds, additional plant and equipment through to by-product waste disposal. The final product is sold in the market relative to the product quality of other producers. The authors recognise that most value can be added through addressing complex ore treatment as early as possible in the value chain and this paper focuses on blending hydrometallurgical techniques with minerals processing to address ore complexity in the concentrator.

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TECHNOLOGY FOR PROCESSING OF REFRACTORY GOLD-CONTAINING CONCENTRATES BASED ON ULTRAFINE GRINDING AND ATMOSPHERIC OXIDATION

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ABSTRACT

Autoclave oxidation, bacterial and chemical liberation, roasting, acid oxygen processes, atmospheric oxidation prior to cyanidation are applied to process refractory sulphide concentrates. It is obvious that the processes with the maximum effect regarding gold recovery are those in which complete destruction of sulphides takes place and gold becomes accessible for further cyanidation through liberation from sulphide "capsula". But the significant recovery increase is often accompanied by a process complexity, high expenses for equipment purchase and maintenance. It makes material processing cost-inefficient. Process optimization has led to progress in ultrafine grinding, bacterial and autoclave oxidation processes. In our opinion sulphide atmospheric oxidation processes are the most interesting as oxidation there runs without high temperature (up to 100 ^{o}C) and excess pressure (open units), and with acid environment close to neutral. The process of sulphide concentrates treatment named "Albion" (Albion Process™) represents a combination of ultrafine grinding and oxygen oxidation without pressure and external heating. The process has a number of advantages: sodium cyanide consumption decrease during gold leaching and high ecological compatibility. TOMS institute in collaboration with Glencore Technology Pty ltd (Australia) tested Albion technology aiming at liberation of finely disseminated gold from flotation concentrate produced from refractory ore of one Russian deposit. The testing has shown that Albion technology makes it possible to process refractory products with quite small capital investments. It enables involving medium and small refractory gold deposits into processing.

KEYWORDS

Gold, atmospheric oxidation, refractory concentrate, leaching

INTRODUCTION

The considerable part of the world Au-bearing reserves is present by refractory ores with finely disseminated gold. Gold in such ores is often associated with sulphide minerals and locked in so called "sulphide capsula" since it is closed inside dense sulphide grains and inaccessible to recovery by cyanidation. To treat refractory sulphide concentrates containing gold a number of processes were developed and they can be divided into three groups (Figure 1):

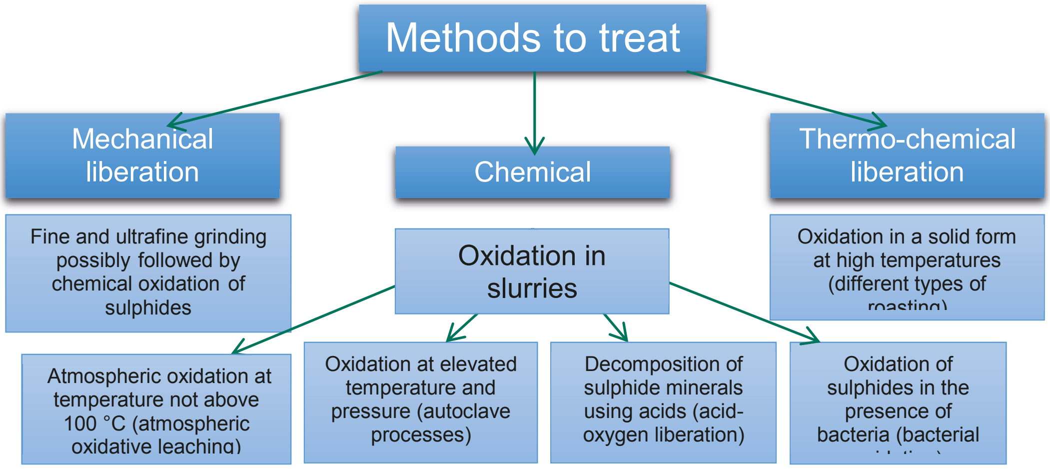


Figure 1 – Methods to liberate finely disseminated gold

Chemical liberation methods to recover finely disseminated gold mean either complete destruction of gold-bearing sulphides (autoclave leaching, bacterial-chemical leaching, acidic-oxygen processes etc.) or partial alteration of the specified minerals (mild atmospheric oxidation prior to cyanidation).

The technologies aimed at complete destruction of sulphide minerals provide for the best Au recovery since they allow one to almost completely liberate finely disseminated gold, which makes it accessible to the subsequent cyanidation. However, these processes are usually characterized by arrangement complexity, high equipment investments and operating costs. It is quite often that the resulted oxidation products contain compounds which will react with NaCN in the course of cyanidation causing its high consumption. This along with the other complicating factors makes the material processing less attractive in terms of cost efficiency.

Continuous search for technologies devoid of the disadvantages takes place resulting in improvements of ultrafine grinding and advance of bacterial and autoclave oxidation methods. The authors consider atmospheric oxidation of sulphides as the technology of the most interest. This process assumes that oxidation proceeds at temperatures not exceeding 100°C, without excessive pressure and under acid conditions close to neutral.

ALBION PROCESS™ DESCRIPTION

One of such atmospheric oxidation technologies is Albion Process™, developed by Xstrata Pty Ltd (Glencore Technology Pty Limited), Australia. This technology is a combination of ultrafine grinding and oxidative leaching without pressure and external heating. The process sequence is schematically shown in Figure 2.

Grinding of a concentrate is conducted in stirred mills to about 80% -10 um size. Due to ultrafine grinding of the concentrate minerals surface area increases significantly. Mechanical and chemical activation of sulphides results in the enhanced rate of the subsequent oxidation.

Oxidation of the finely ground concentrate is conducted in a series of agitated vessels at atmospheric pressure. Oxygen of technical grade is introduced to the leach slurry to oxidize the sulphide minerals. Intensive chemical reaction leads to self-heating of the slurry higher than 80°C but lower than the boiling point. The Albion Process™ is an autothermal technology.

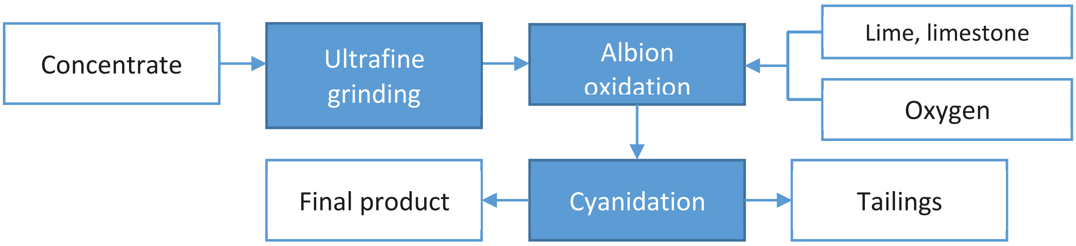


Figure 2 – Schematic flowsheet to treat refractory Au-bearing sulphide concentrates using Albion Process™

The creators of the process indicate that sulphide minerals will leach according to the reactions below.

$$2FeS_2 + 7.5O_2 + H_2O = Fe_2(SO_4)_3 + H_2SO_4 \tag{1}$$

$$2FeAsS + 5.5O_2 + H_2O = 2HAsO_2 + 2FeSO_4 \tag{2}$$

$$2FeSO_4 + 0.5O_2 + H_2SO_4 = Fe_2(SO_4)_3 + H_2O \tag{3}$$

$$HAsO_2 + Fe_2(SO_4)_3 + 2H_2O = H_3AsO_4 + 2FeSO_4 + H_2SO_4 \tag{4}$$

As was earlier mentioned the Albion Process™ is carried out under relatively neutral conditions: pH of the slurry is higher than 4.0. Limestone, the cheapest reagent, is used to neutralize sulphuric acid generating according to equations (1) and (4). In this way the sulphides leaching in the course of the Albion Process™ can be essentially described by the following reactions.

$$2FeS_2 + 7.5O_2 + 9H_2O + 4CaCO_3 = 2FeOOH + 4CaSO_4{\cdot}2H_2O + 4CO_2 \tag{5}$$

$$FeAsS + 3.5O_2 + 2H_2O + CaCO_3 = FeAsO_4 + CaSO_4{\cdot}2H_2O + CO_2 \tag{6}$$

The Albion Process™ has several advantages over conventional oxidative processing of sulphide minerals. One of the benefits is a relative inertness of the oxidation products towards cyanide and their environmental compatibility. Since the Albion Process™ runs on under relatively neutral conditions almost all sulphur passing into solution precipitates as gypsum. Therefore no suplphur accumulation occurs in the solution (S concentration in the solution is maintained less than 2 g/L). So the oxidative leaching of sulphides results in generation of sulphates instead of elemental sulphur as it happens in the processes with lower pH level (bacterial oxidation, acidic-oxygen oxidation etc.). Elemental sulphur (if any in the oxidation product) will actively react with CN^- in the course of the following cyanidation that leads to thiocyanates formation and consequent high consumption of NacN. In this context the Albion Process™ application allows one not only to liberate finely disseminated gold but as well to ensure relatively low reagent consumption during the following cyanidation of the oxidized material.

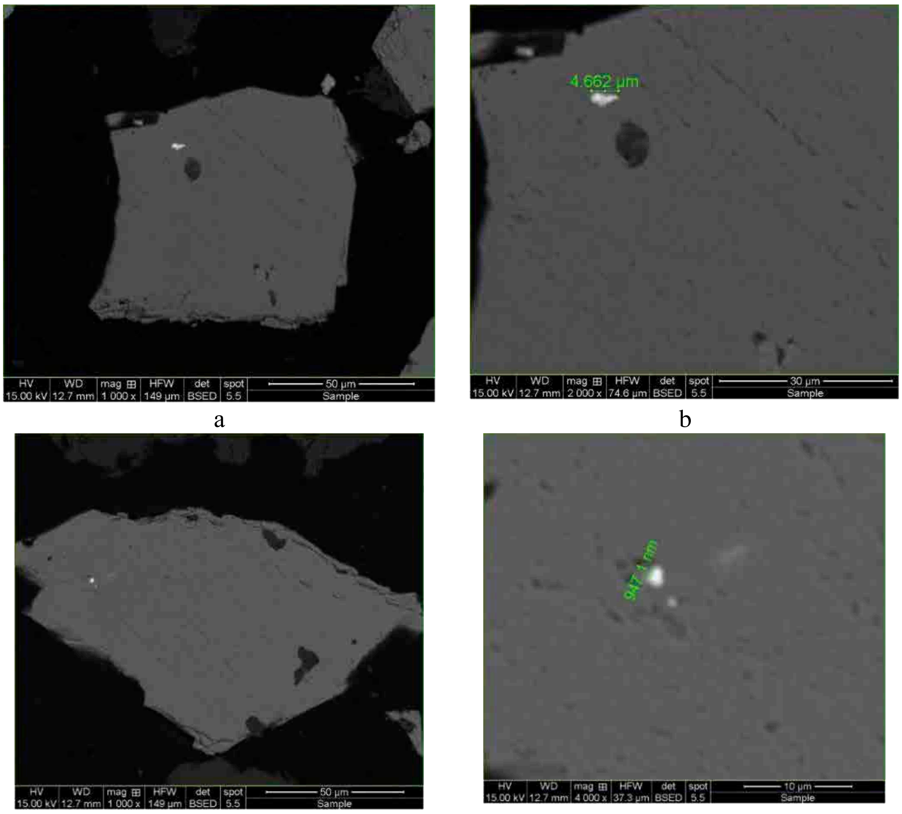
The elevated temperature (higher than 80°C) of the Albion Process™ provides precipitation of iron and arsenic from the solutions with formation of goethite and scorodite which have crystalline structure. These minerals are very inert towards NaCN. Moreover crystalline scorodite is the most stable and environmentally safest form to store As-containing wastes. The processes running at lower temperatures than the Albion Process™ may result in precipitation of iron and arsenic in the form of iron hydroxides and amorphous scorodite as well as jarosite and complex sulphates. The presence of these compounds in the oxidized material leads to the increased consumption of NaCN during gold leaching. Amorphous scorodite contributes to partial arsenic dissolution during storage of leaching tailings, which has adverse environmental effects.

RESULTS AND DISCUSSION

TOMS institute conducted testing of the Albion Process™ aimed at liberation of finely disseminated gold from the flotation concentrate resulted from processing of a refractory ore of one of Russian gold deposits.

Concentrate description

Gold grade of the head flotation concentrate was 29.6 g/t. The concentrate size was 80% -80 um. Mineral composition of the concentrate comprises 17% of rock-forming minerals, 59% of pyrite and 24% of arsenopyrite. Gold size (Figure 3) in the concentrate of interest was less than 10-15 um, some gold particles were detected with size of less than 1.0 um.



a b

a b

Figure 3 – Au particles in sulphide minerals:
a – general view of a grain, b – zoomed

Cyanidation of the head concentrate

Based on the material composition study for the concentrate and on sorption cyanidation tests (Table 1) it was determined that amount of cyanide-leachable gold in the concentrate was about 20-25%, basic part of gold (more than 70%) in the concentrate is finely disseminated into sulphide minerals and inaccessible to leaching solutions.

Table 1 – Data from sorption cyanidation tests for flotation concentrate prior to oxidation (cyanidation conditions: NaCN rate – 2 g/L; cyanidation time – 24 h; slurry pH – 10.5; solids in slurry – 40%; sorbent load (active carbon Norit RO 3515) – 8% of slurry volume)

Concentrate size, um	Au grade, g/t		Au recovery, %	Reagent rate, kg/t of concentrate		
				NaCN		
	Head material	cake		Total added	Total consumed	CaO
80% -80 (head)	29.6	23.7	19.9	3.0	1.8	3.7
		23.5	20.6	3.0	1.8	3.7
80% -20		23.5	20.6	4.0	2.8	4.7
		23.2	21.6	4.0	2.8	4.7
80% -10		22.1	25.3	6.5	5.2	6.5
		22.0	25.7	6.5	5.3	6.5

Concentrate grinding

A grindability test in Netzsch-IsaMill M4 was carried out to estimate power consumption to grind the concentrate. The testing was performed according to the method of Glencore Technology – the stirred mill manufacturer. The testing data indicated that 87 kW-h/t specific power was required to grind the concentrate to 80% -10 um. Media specific consumption in the course of the ultrafine grinding equaled 0.9 kg/t (based on stirred mills operational practice).

Concentrate oxidative leaching

The ground concentrate was leached in a special unit manufactured under the guidance of the Albion's creators. The oxidation time was 48 hours. Figure 4 shows the main results from the testing.

Cyanidation of the oxidized concentrate

The oxidized concentrate was cyanided under the same conditions as the concentrate prior to oxidation. Consumption rate of NaCN for cyanidation of the concentrate previously leached in Albion unit was 19.8 kg/t.

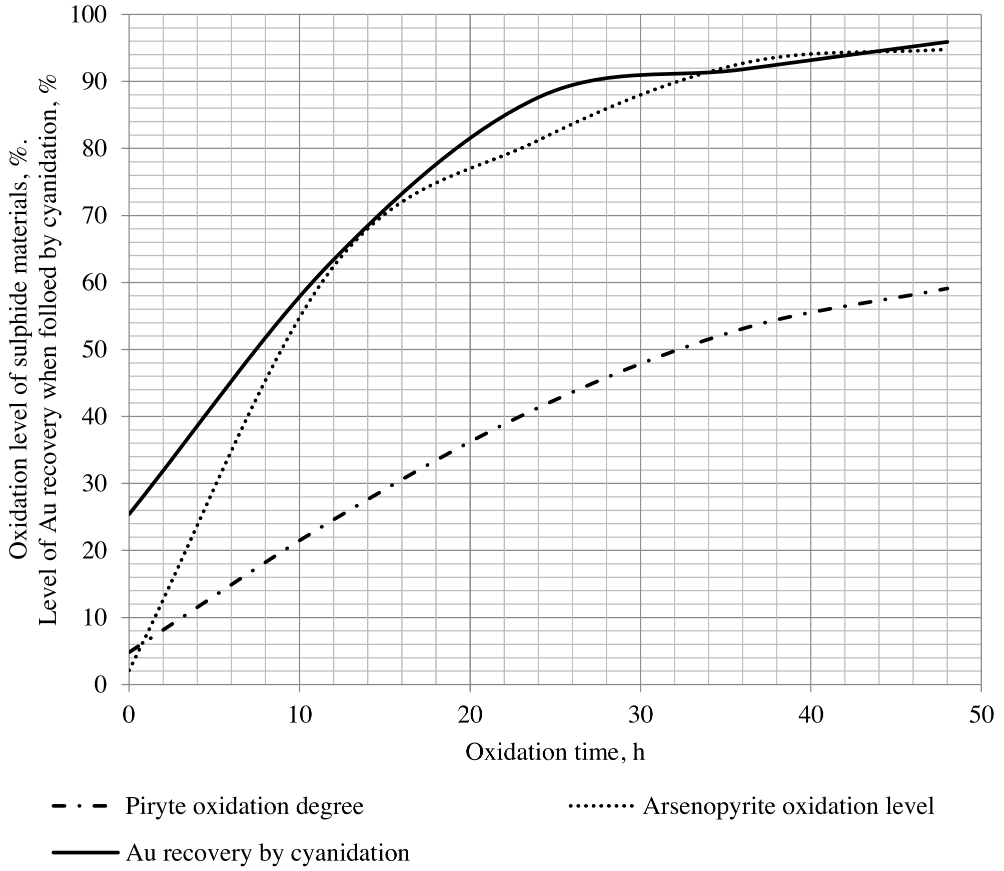


Figure 4 – Basic data from the concentrate processing according to Albion technology

Table 2 represents comparison of data from the concentrate processing by the Albion Process™ and by autoclave oxidation.

Table 2 – Comparison of the data from Albion Process™ and from autoclave oxidation

Process	Oxidation cake mass pull, %	Sulphide oxidation level (based on S), %	Specific consumption of reagent, kg/t of head concentrate			Au recovery at the following cyanidation, %
			O_2	Limestone	Lime	
Autoclave oxidation*	81	более 98	520	670	94	97.2
Albion	190	64	360	660	11	95.9

* autoclave oxidation conditions: 210°C, O_2 addition at 10 bar, oxidation for 1 h

CONCLUSIONS

The test data confirm that the Albion Process™ allows efficient processing of concentrates containing finely disseminated gold at lower operational costs (consumption of O_2, limestone, and lime) and at level of Au recovery comparable to autoclave oxidation technology.

The Albion Process™ makes it possible to arrange processing of refractory products at significantly lower capital costs. In contrast to the competitive processes (autoclave and bacterial oxidation) Albion technology does not require constant presence of high-quality specialists as well as there is no need to build processing plants with high concentrate throughput or a separate plant to oxidize and cyanide concentrates from several sites. The Albion Process™ enables arrangement of a small-scale on-site plant, which allows involving of small and medium refractory gold deposits in processing.

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

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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™ plant to oxidize the refractory concentrate. Glencore Technology (GT) provided the Albion Process™ plant as a technology package.

In July 2014, the progress of the GPM Gold Project Albion Process™ 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™ 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™ 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.

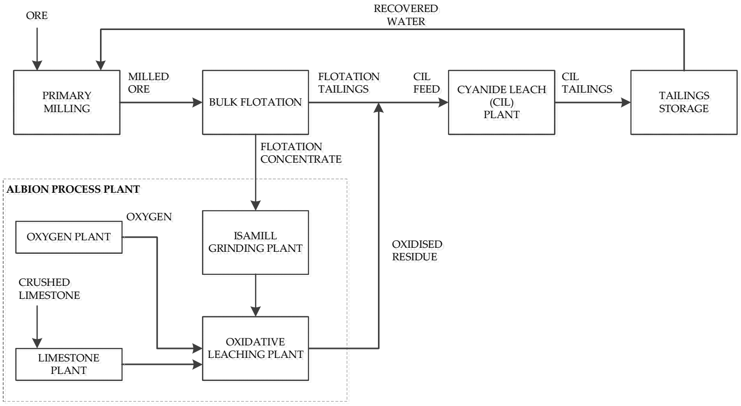

Figure 1 – GPM Gold Project overall flow sheet.

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.

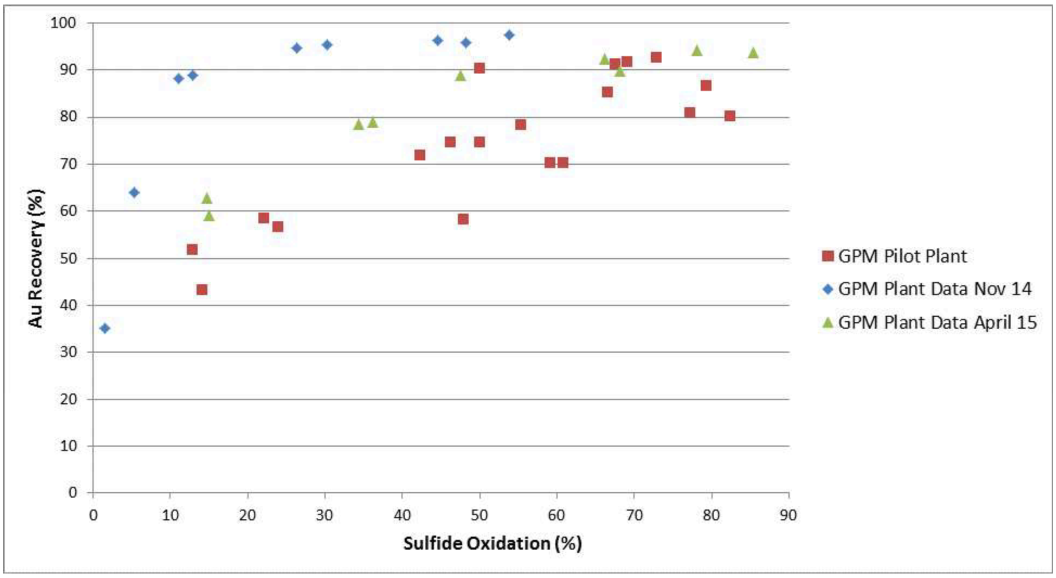

Figure 7 – Sulfide oxidation against gold recovery comparing pilot plant data and actual operating data.

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.

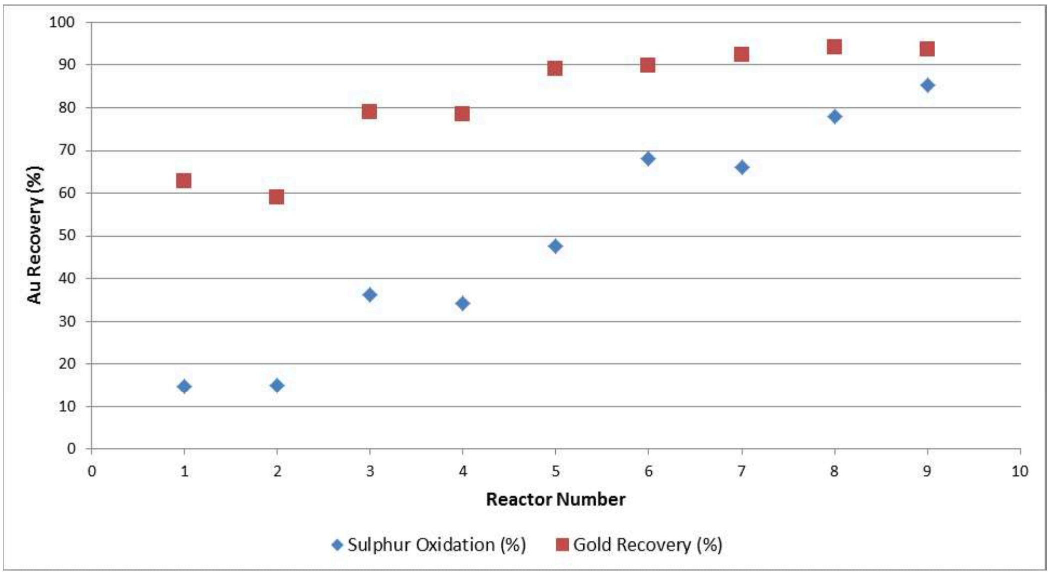

Figure 8 – Profile of sulfide oxidation and gold recovery down the leaching train. Plant is on reduced rates.

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.

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EVALUATING PROCESS OPTIONS FOR TREATING SOME REFRACTORY ORES

By

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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.

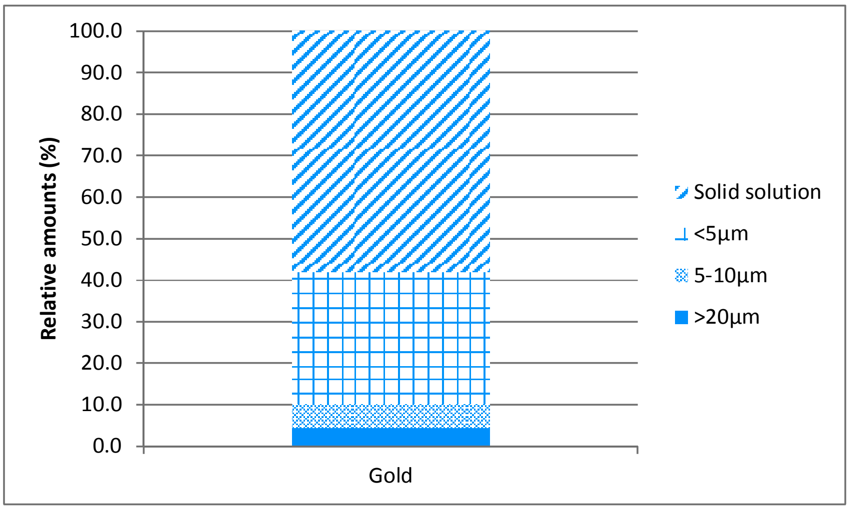

Figure 1: Gold deportment in Ore A

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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.

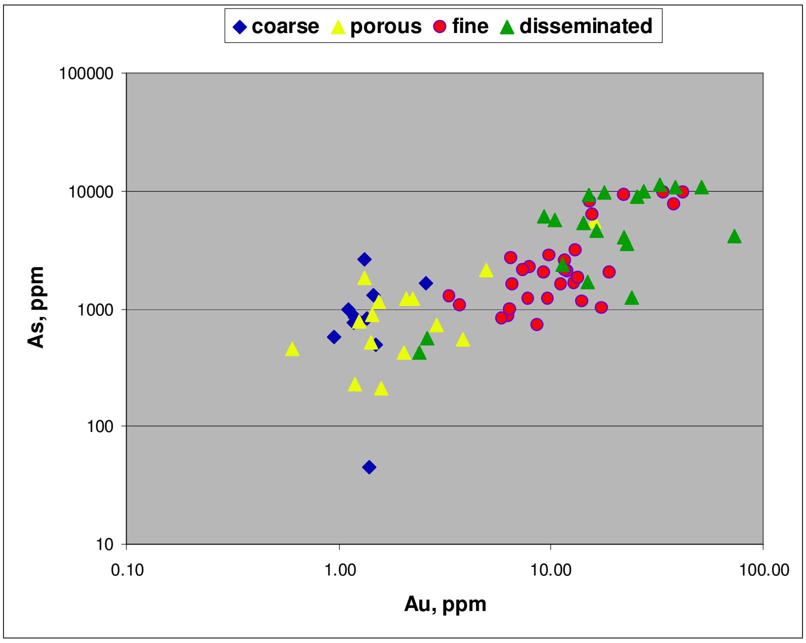


Figure 2: Correlation between measured concentrations of sub-microscopic gold and arsenic in different morphological types of pyrite (from 74 point analyses)

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.

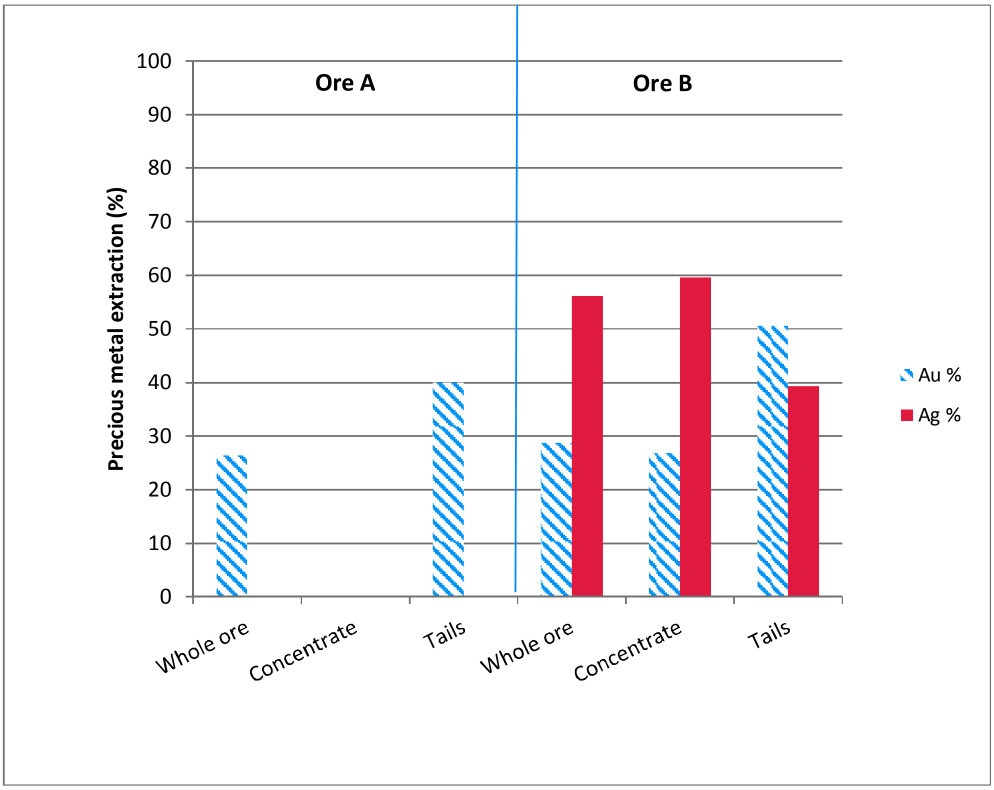


Figure 3: Precious metal extraction following cyanidation on both Ore A and Ore B samples

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.

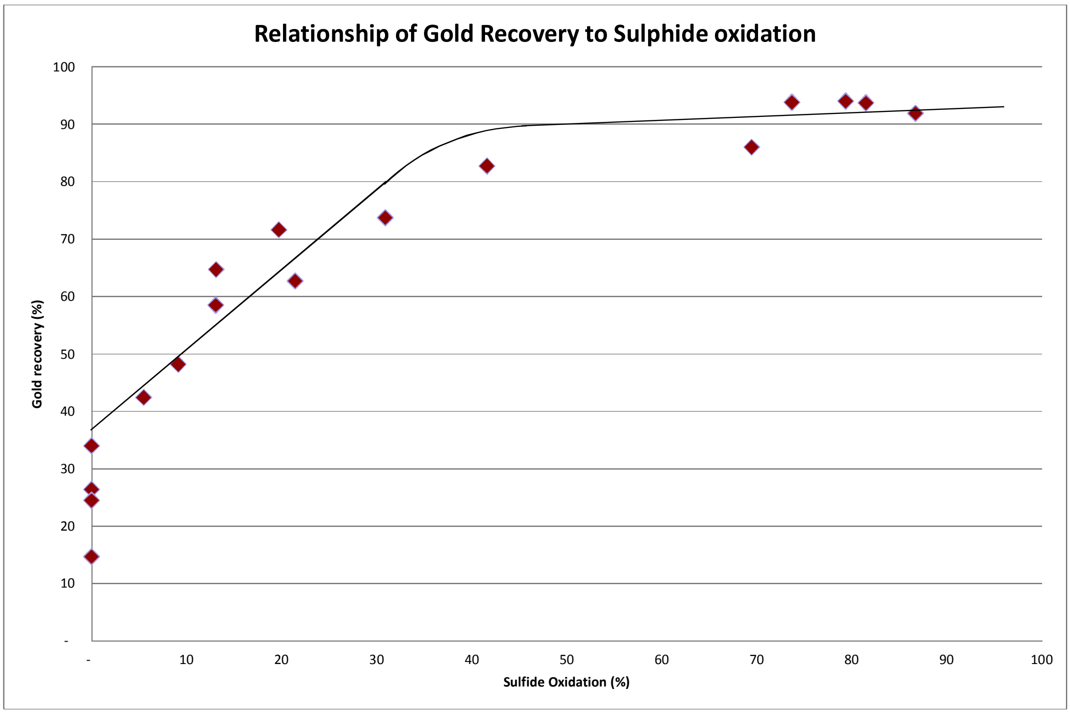


Figure 4: Relationship of gold recovery to sulfide oxidation required for Ore A and concentrate to be amendable to cyanidation

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.

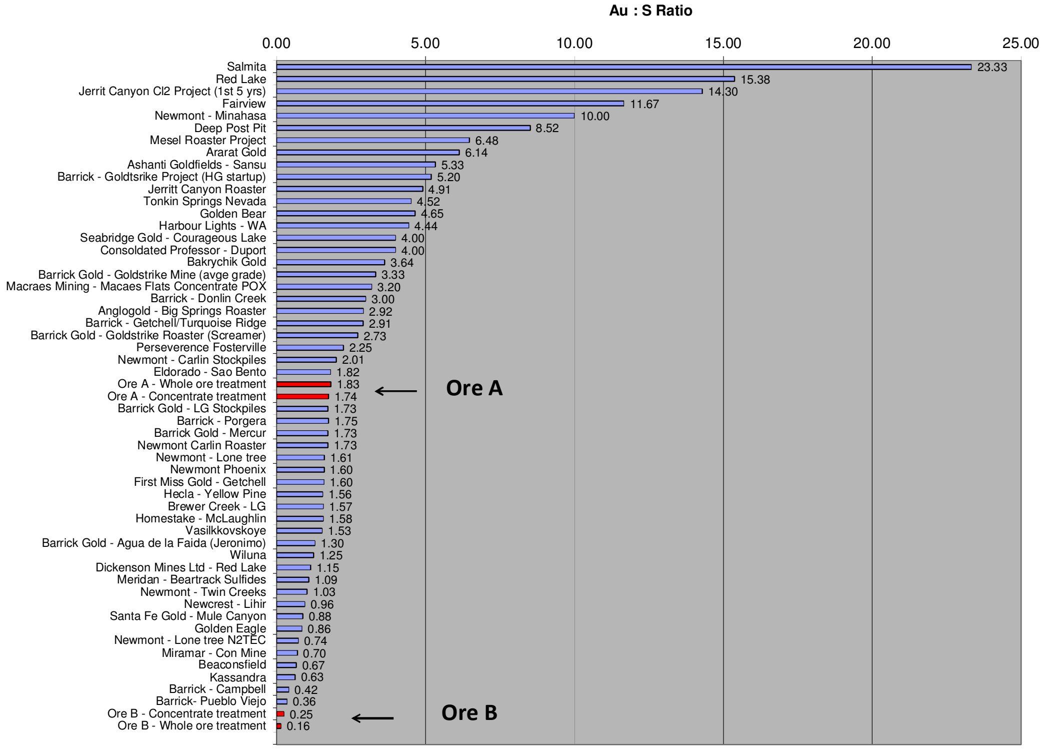


Figure 5: Comparison of gold to sulfur ratio for the two ore types and various existing and previous process operations around the world

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 ($Fe_4(AsO_4)_3\text{-}(OH)_x(SO_4)_y$) 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. Ultrafine 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℃. 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.

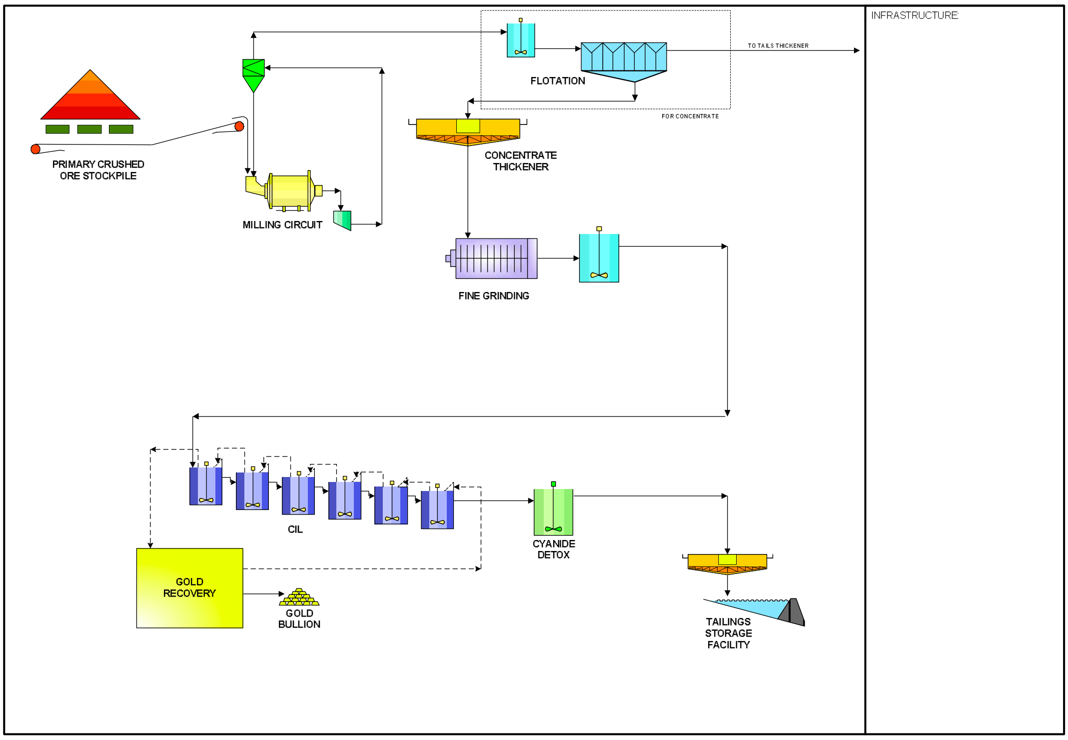


Figure 6: Ultra fine grinding circuit for whole ore and concentration (flotation circuit inclusive)

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.

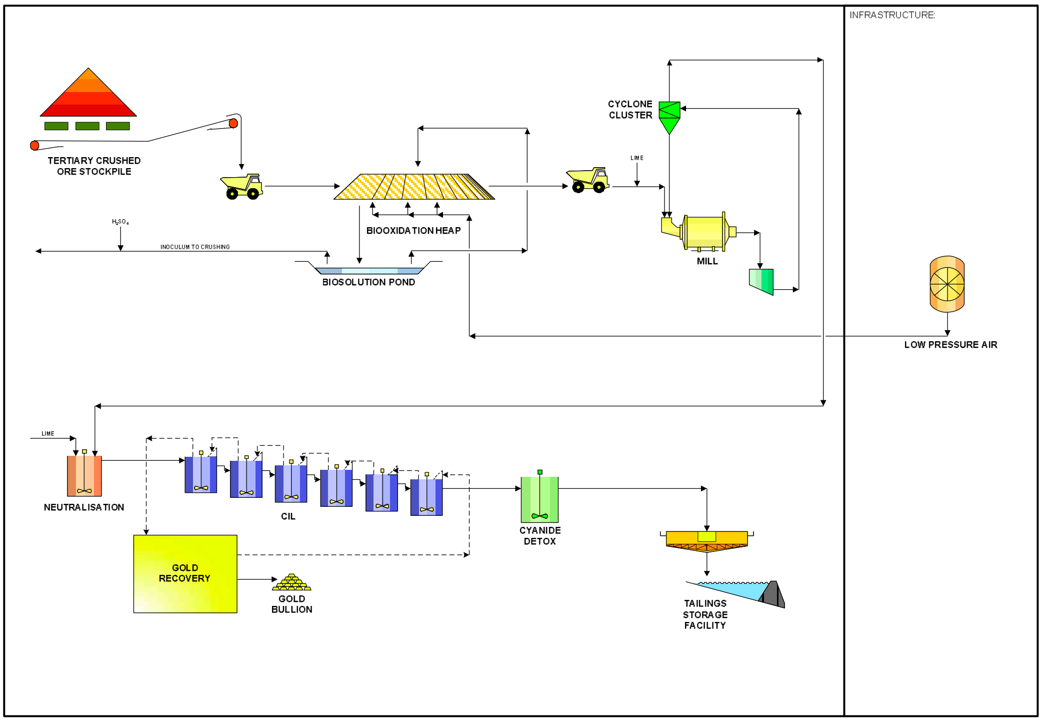


Figure 7: Biological Oxidation Heap Leaching – Whole Ore

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.

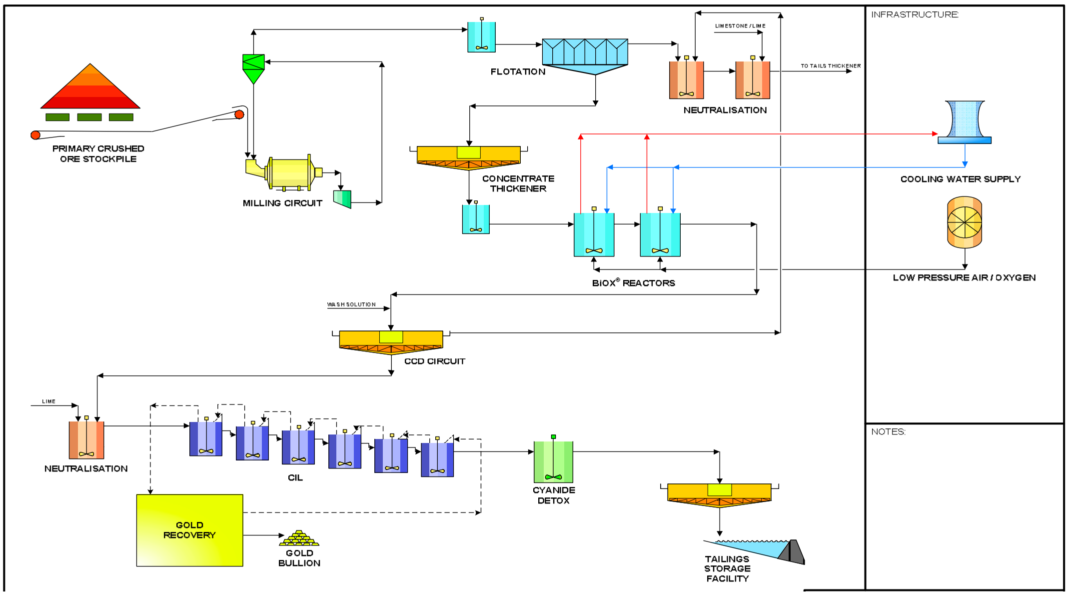


Figure 8: Agitated Tank Bioleaching – Concentrate

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°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.

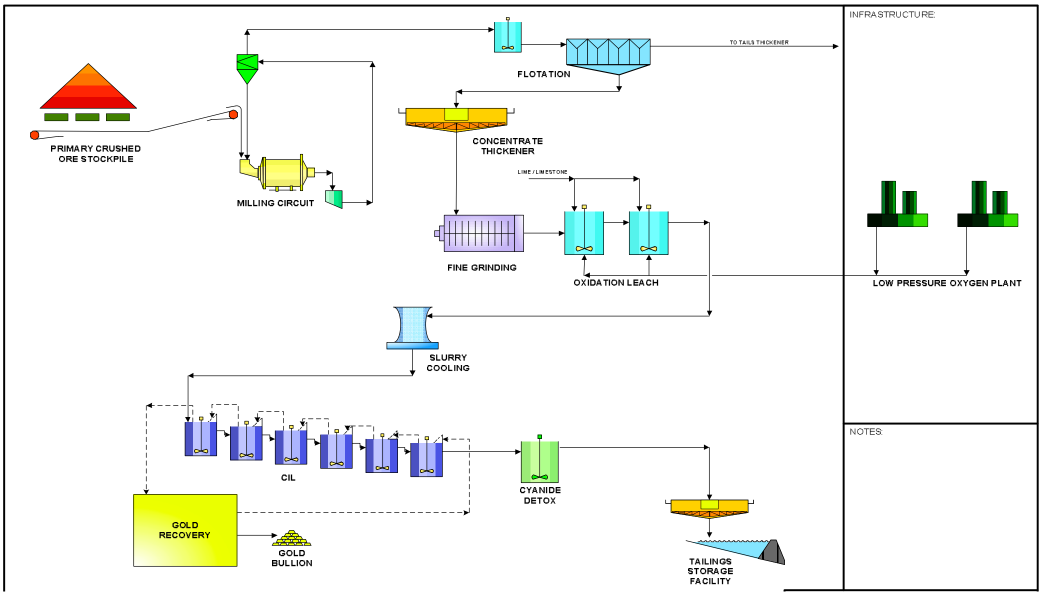


Figure 9: Albion Process – Concentrate

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.

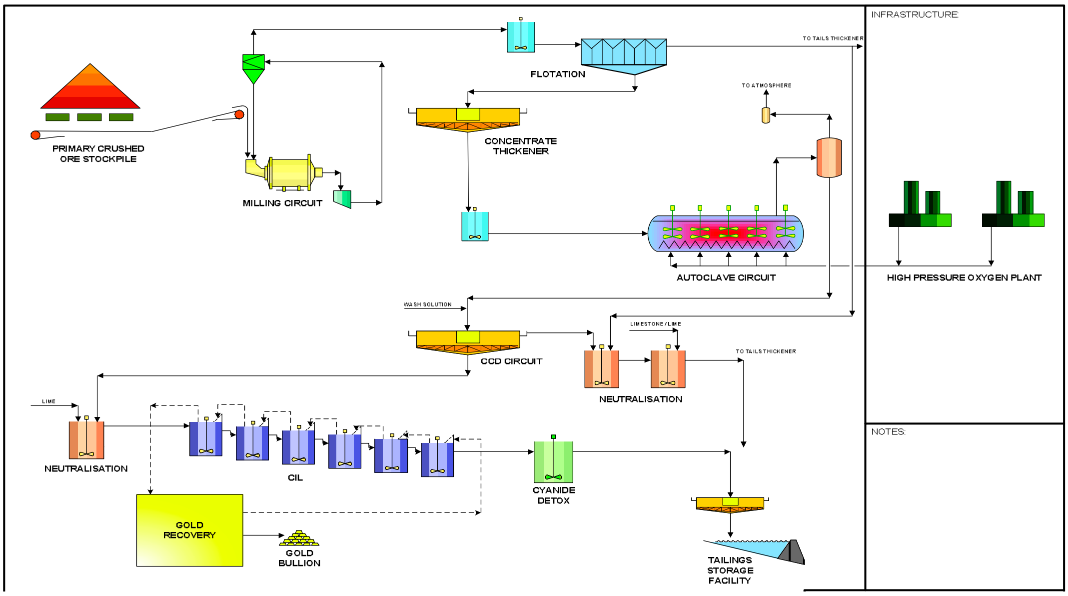


Figure 10: High Temperature Pressure Oxidation

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.

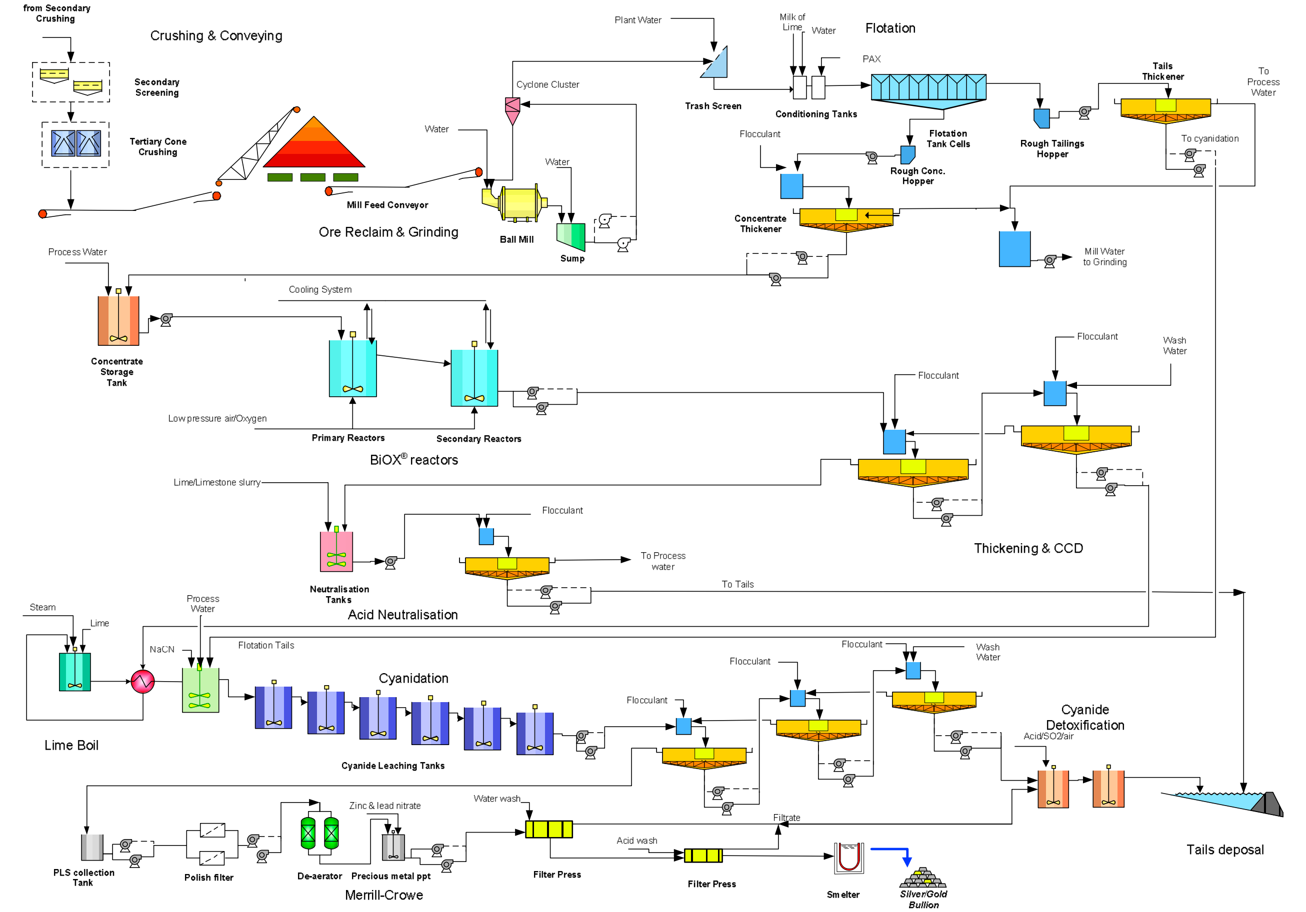


Figure 11: Agitated Tank Bioleaching – Concentrate

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.

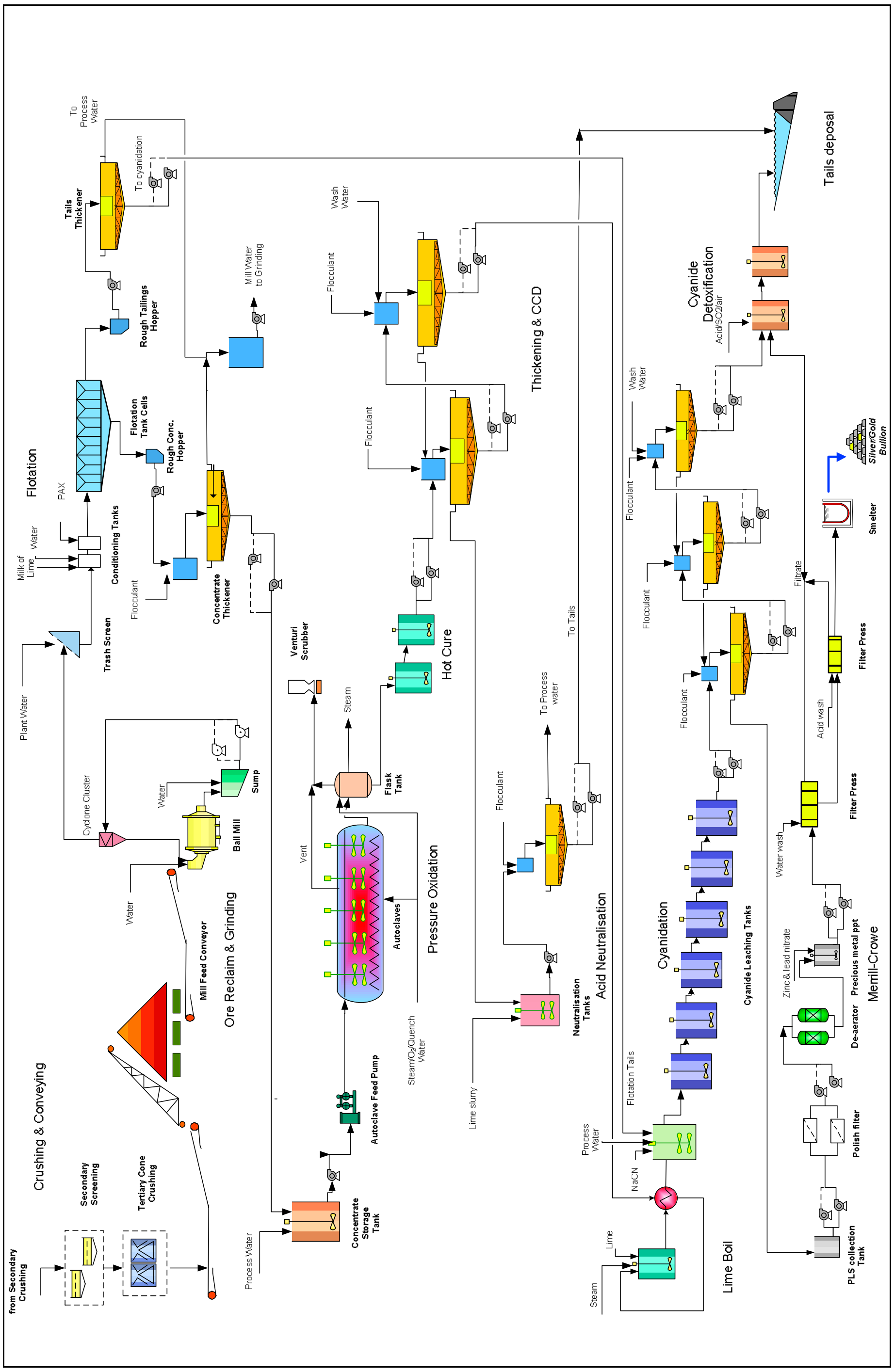


Figure 12: High Temperature Pressure Oxidation – Concentrate

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.

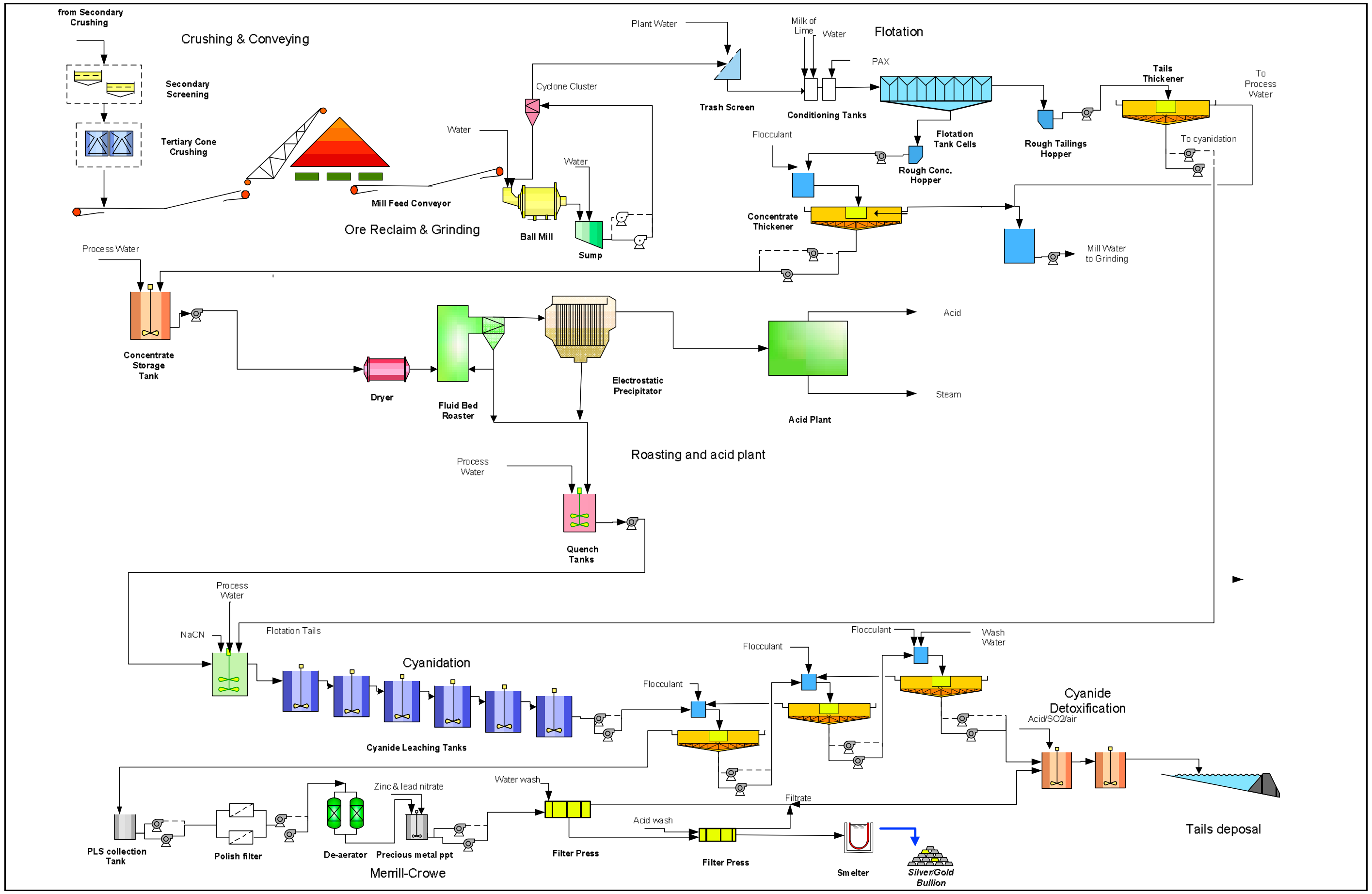


Figure 13: Roaster – Concentrate

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

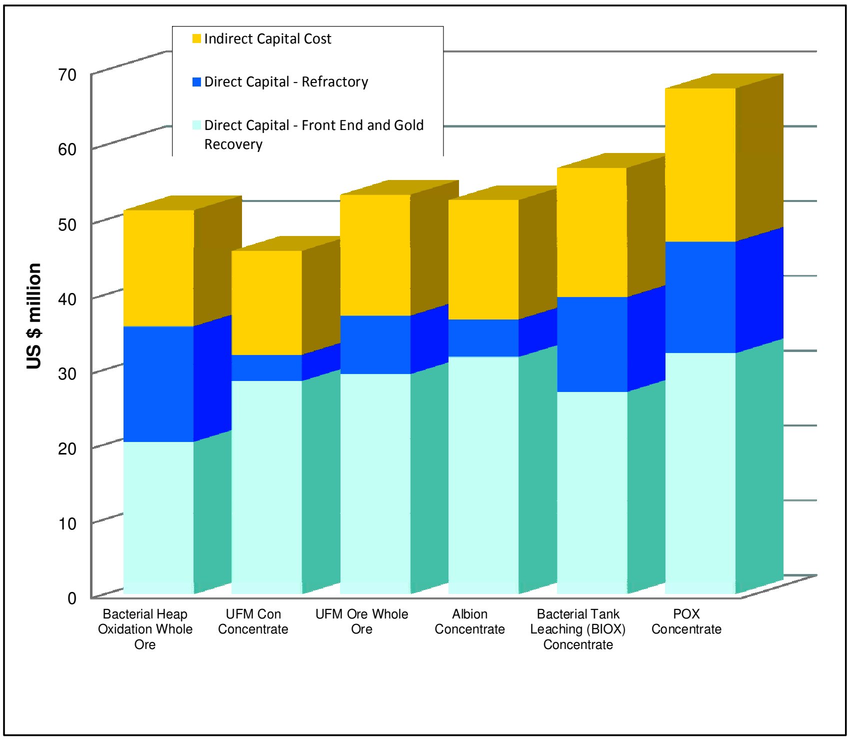
OPTION		1	2	3	4	5	6
PROCESS		**Bacterial Heap Oxidation Whole Ore**	**UFM Con Concentrate**	**UFM Ore Whole Ore**	**Albion Concentrate**	**Bacterial Tank Leaching (BIOX) Concentrate**	**POX Concentrate**
Plant Design Data							
Ore Processed	t / yr	600,000	600,000	600,000	600,000	600,000	600,000
Concentrate 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.



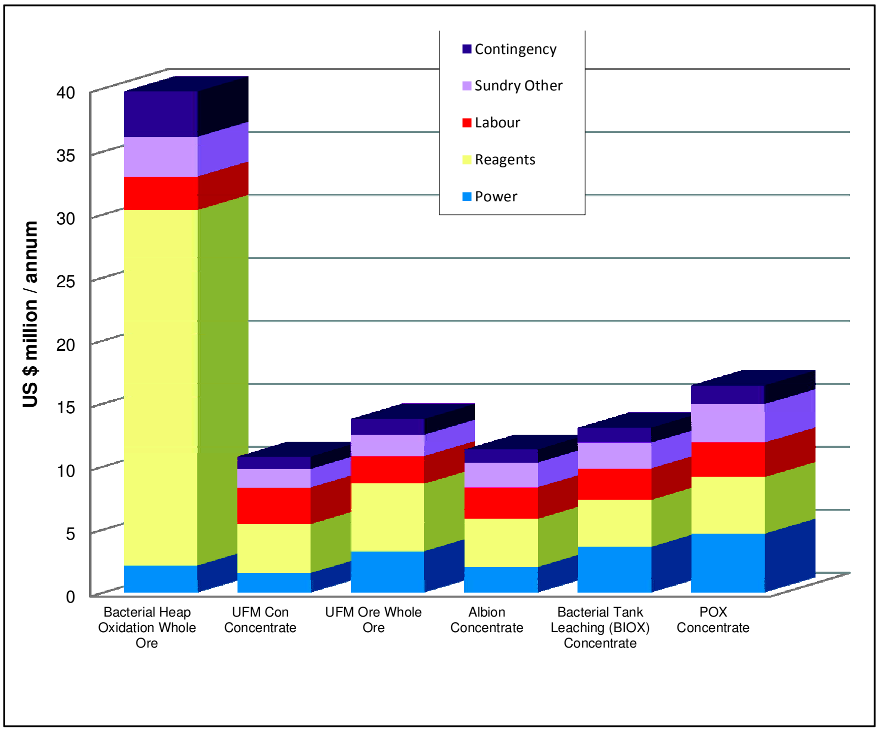


Figure 14: Capital (a) and Operating (b) costs shown graphically for Ore A

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.

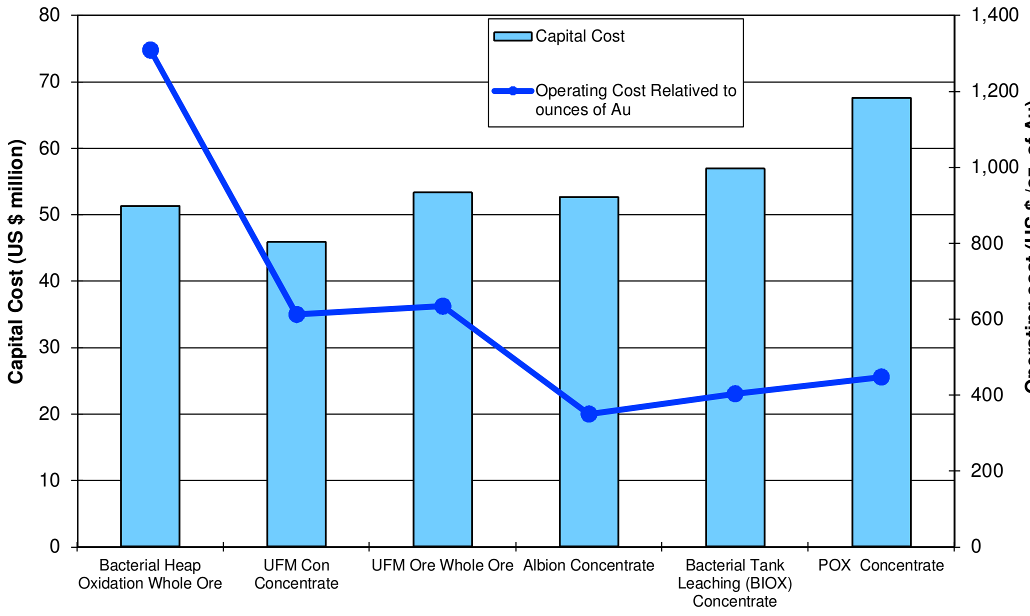
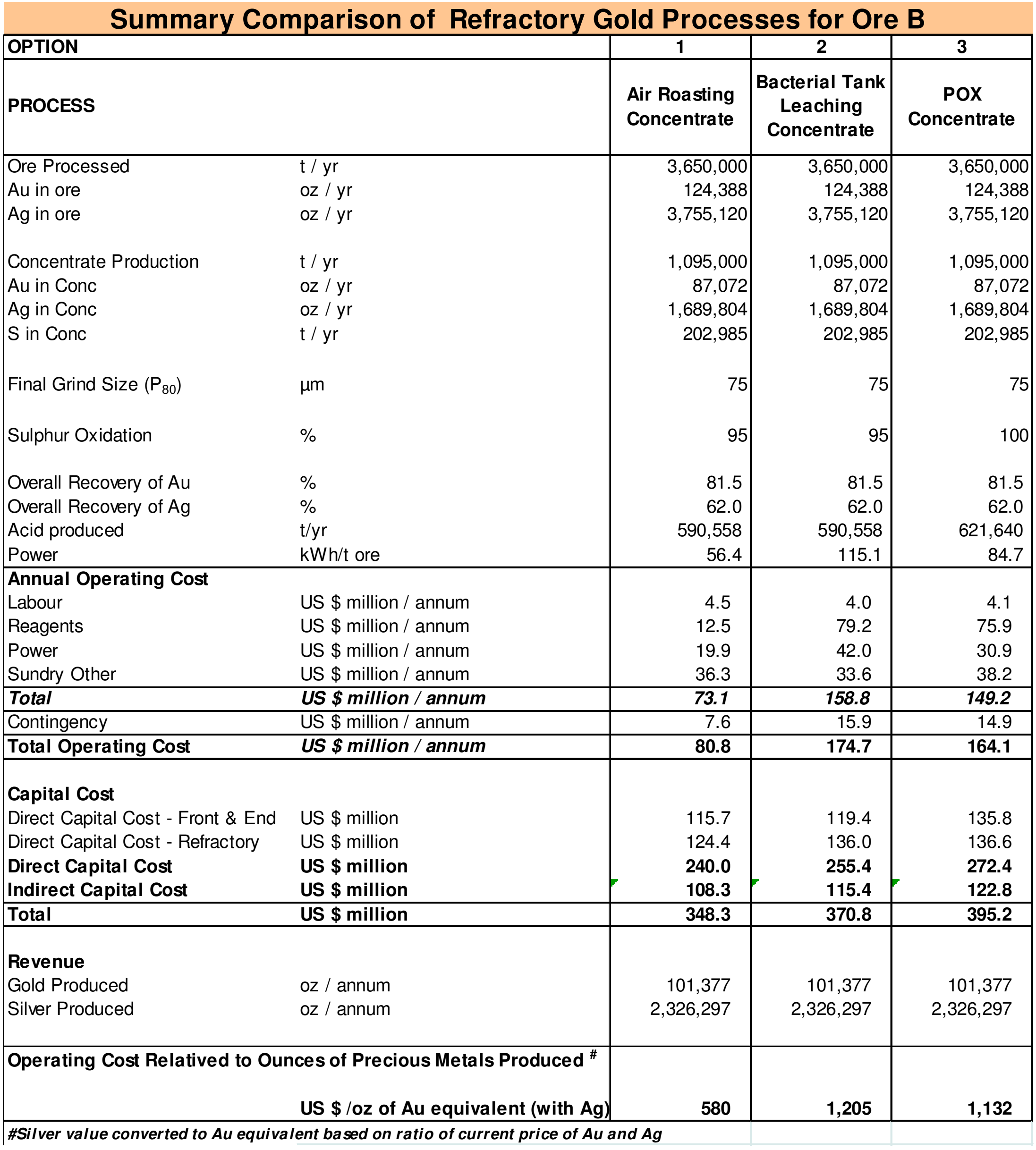


Figure 15: Comparison of conceptual Operating cost to ounces of gold produced versus capital costs for the six process options

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%.

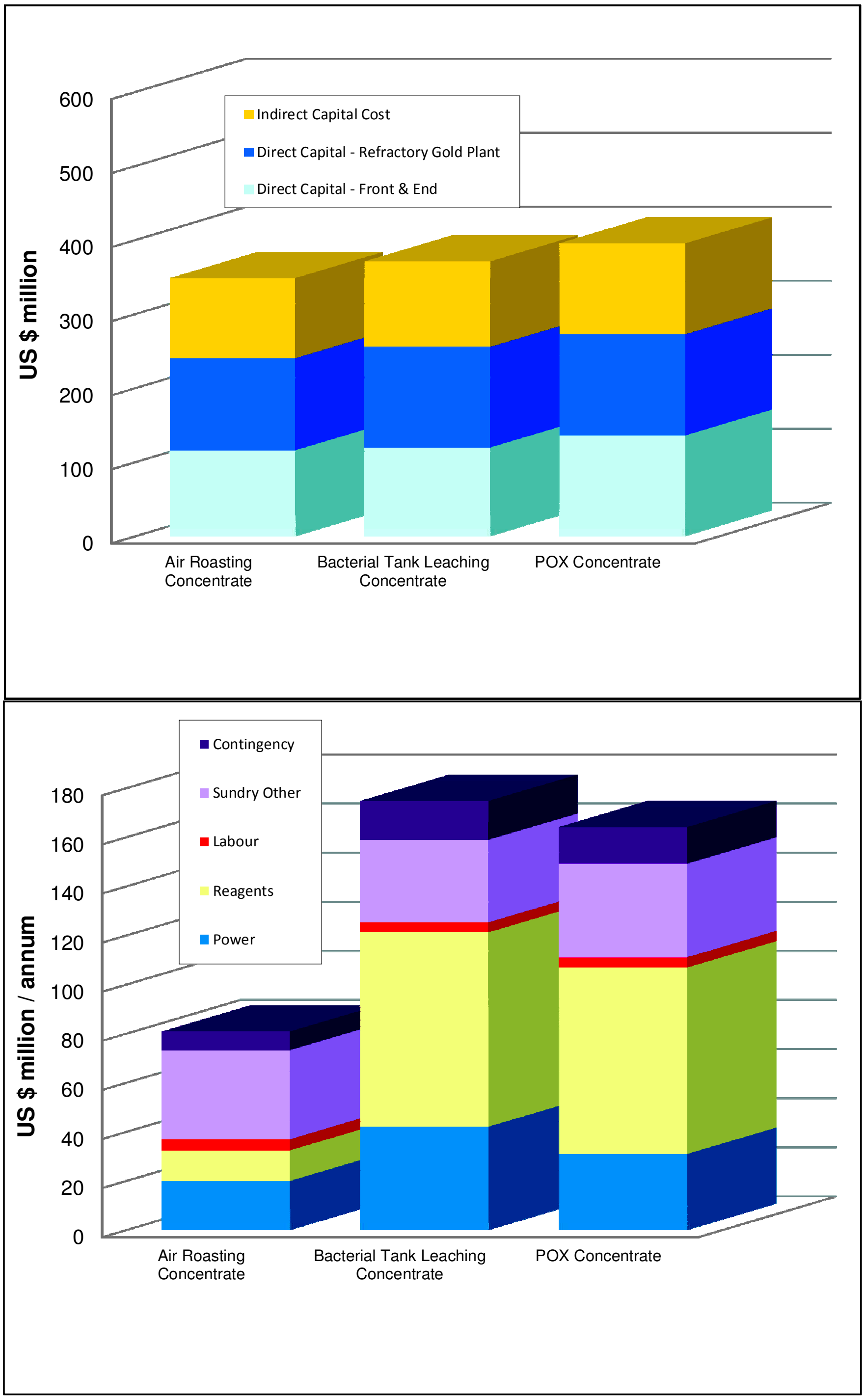
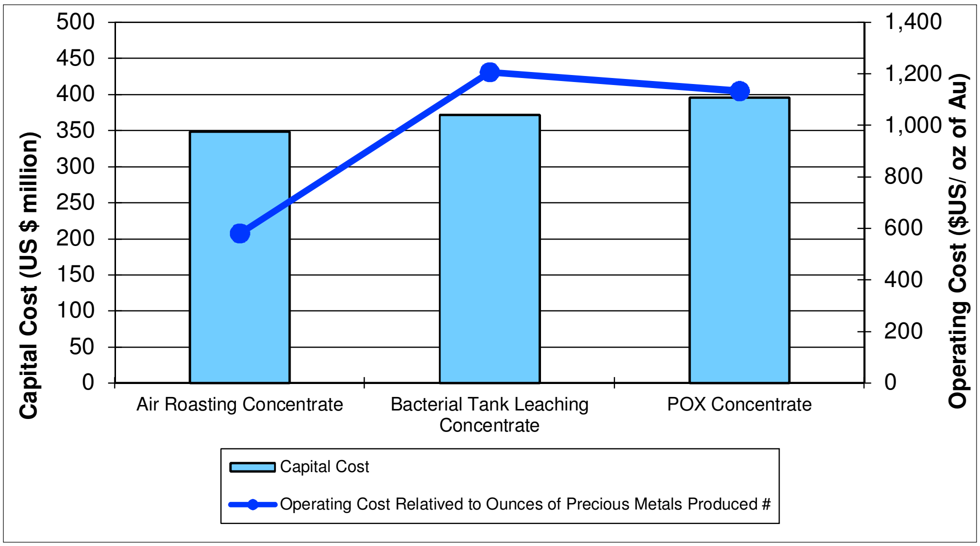


Figure 16: Capital (a) and Operating (b) Cost Estimate for Ore B

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.

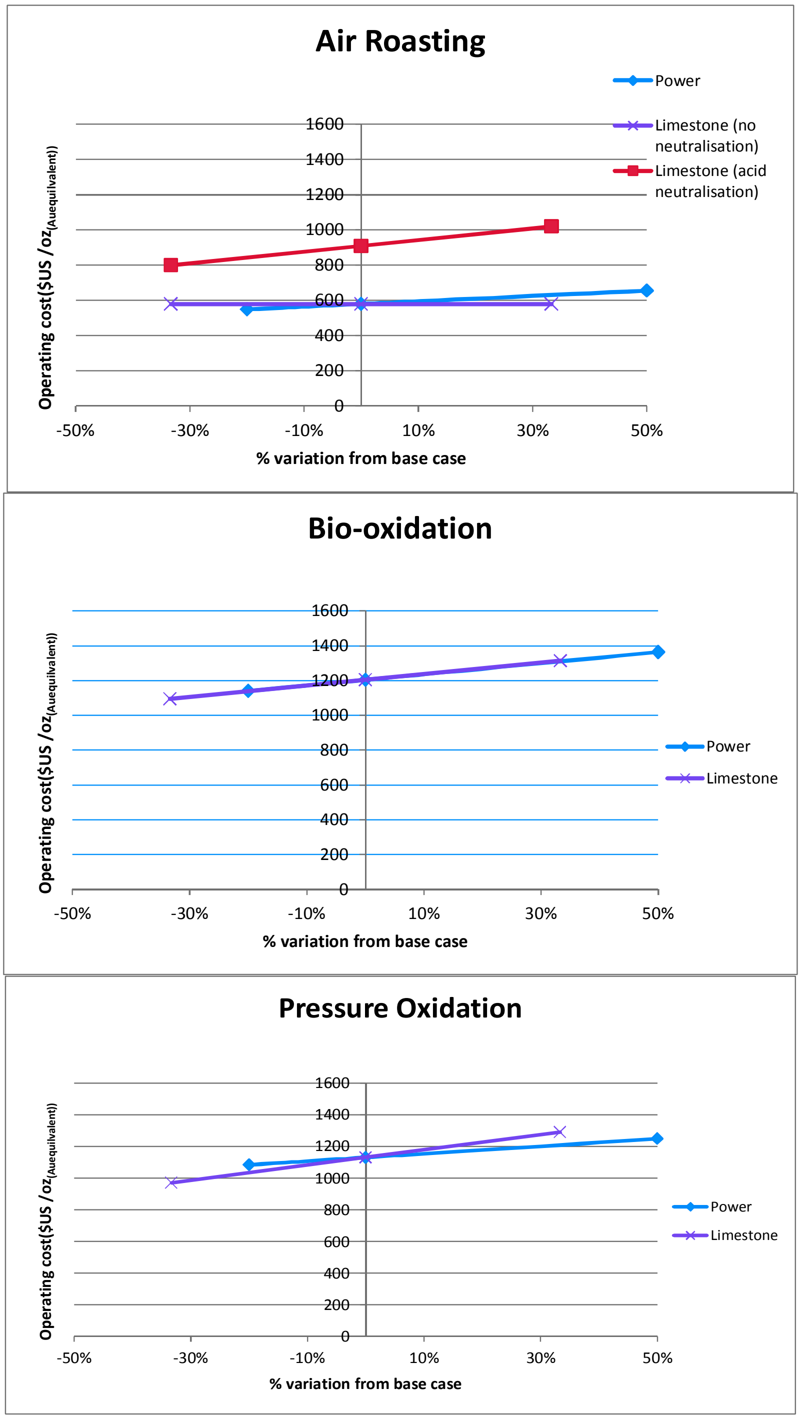


Figure 18: Sensitivity analyses of limestone and power costs for three process options (base case, Power $0.10 kWh, Limestone $75/t)

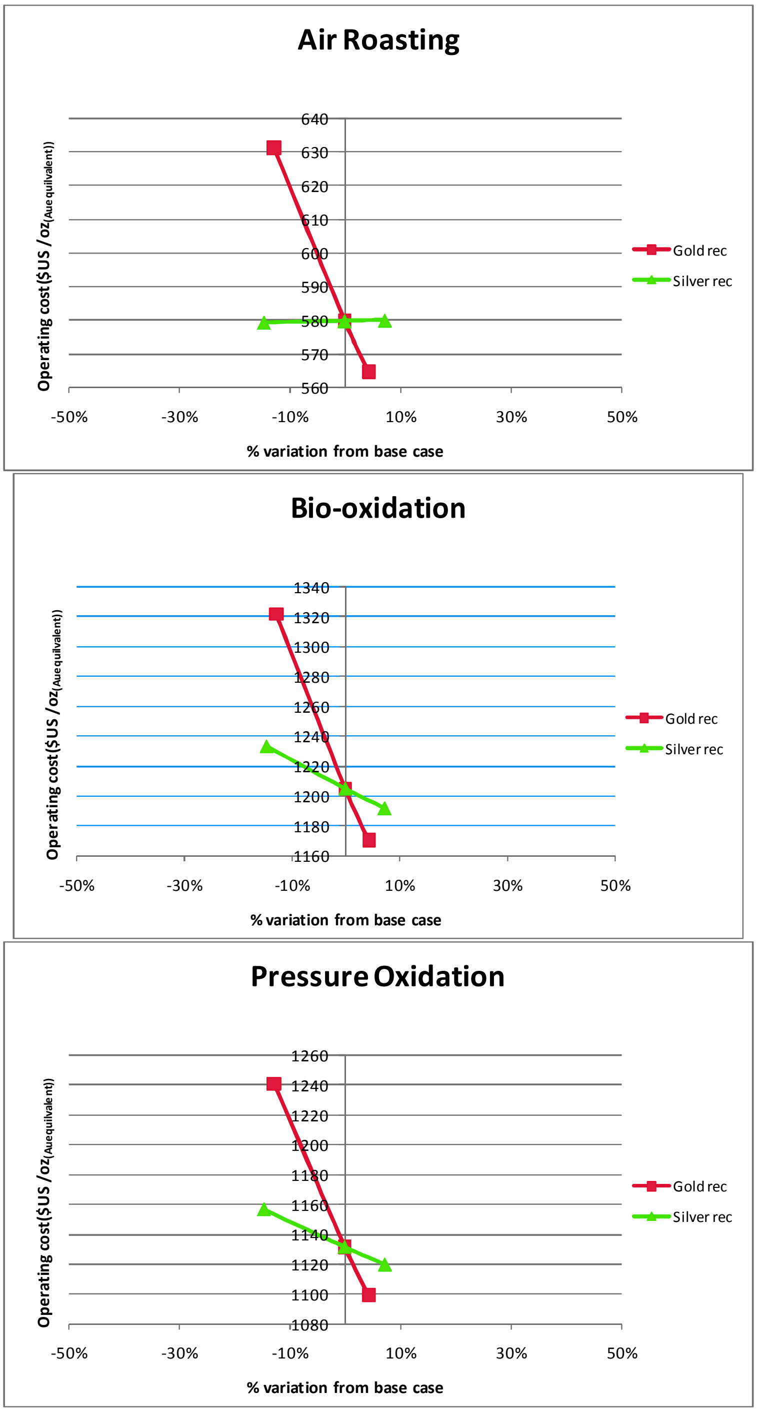


Figure 19: Sensitivity analyses of gold and silver recovery for all three process options (assuming gold and silver extraction variability from oxidised concentrate only, base case Au 95% and Ag 90% in extraction)

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

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COMMERCIALISATION OF THE ALBION PROCESS

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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

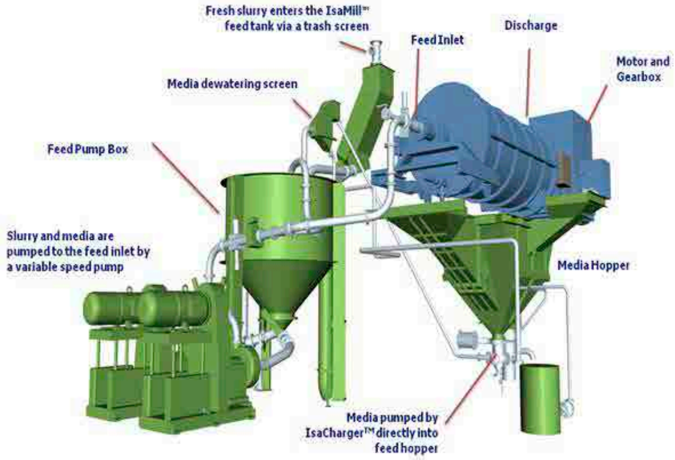

Figure 1
IsaMill$^{\text{TM}}$ Key Components

The IsaMill$^{\text{TM}}$ 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$^{\text{TM}}$ 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$^{\text{TM}}$ 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.

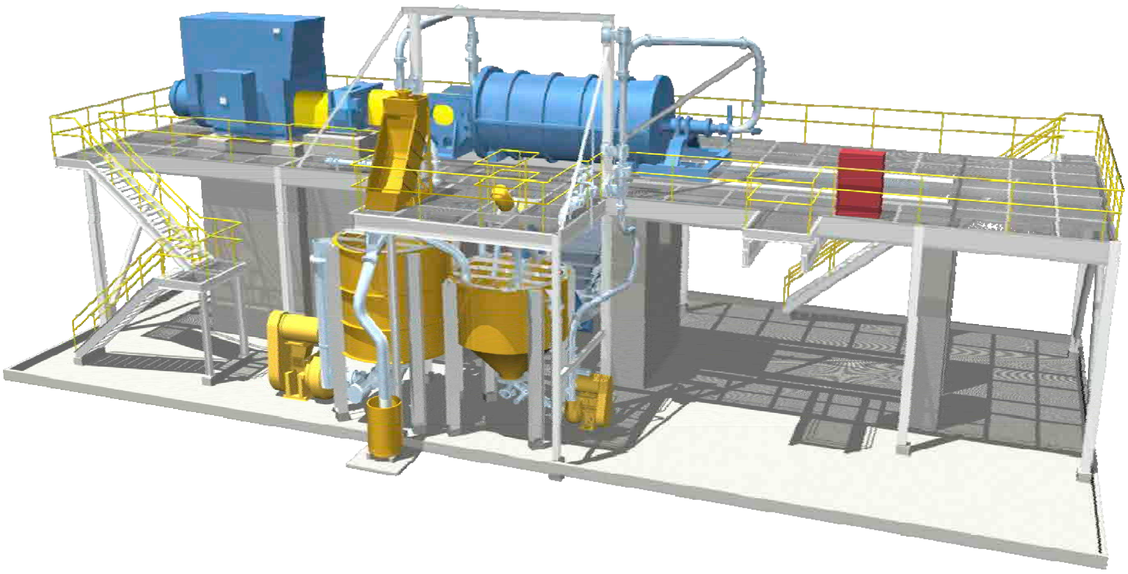

Figure 2
IsaMill$^{\text{TM}}$ Grinding Plant Layout

The IsaMill$^{\text{TM}}$ 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$^{\text{TM}}$ 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$^{\text{TM}}$ and the associated grinding plant. Maintenance of the IsaMill$^{\text{TM}}$ 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$^{\text{TM}}$.

Scale-up of the IsaMill$^{\text{TM}}$ is straight forward. Laboratory test results are directly scaled to commercial size with 100% accuracy. The IsaMill$^{\text{TM}}$ 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.

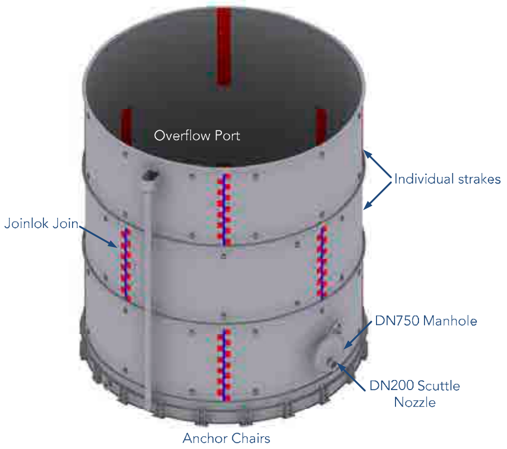

Figure 3
Modular Albion Leach Reactor

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.

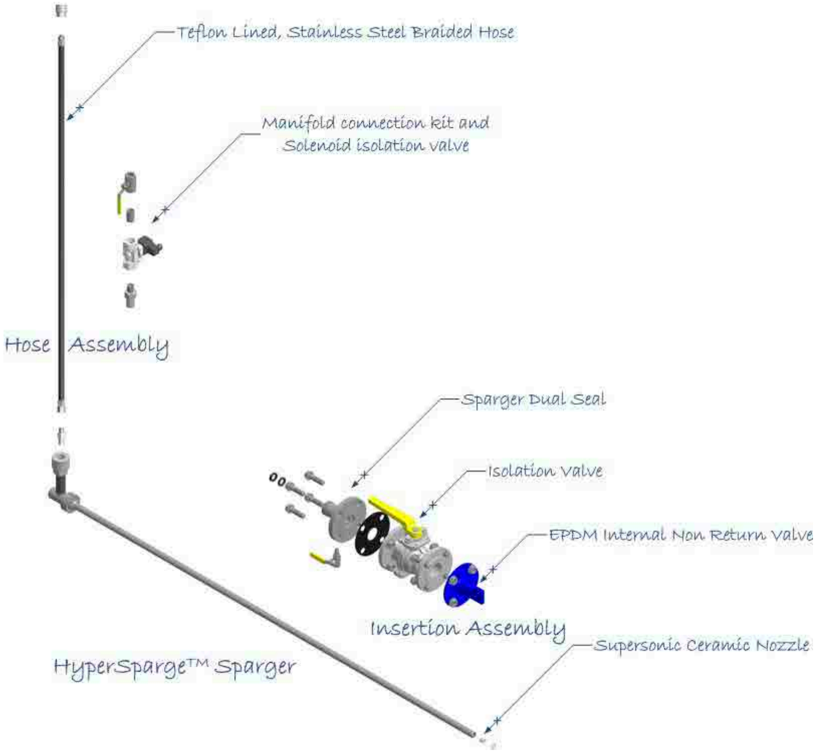

Figure 4
HyperSparge System

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:

$$FeS_2 + 15/4O_2(g) + 9/2H_2O + 2CaCO_3 = FeO.OH + 2CaSO_4.2H_2O + 2CO_2 \qquad (C)$$

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:

$$FeAsS + 7/2O_2(g) + 4\ H_2O + CaCO_3 = FeAsO_4.2H_2O + CaSO_4.2H_2O + CO_2 \qquad (D)$$

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:

$$AgAuTe + 2O_2(g) = TeO_2 + AuO + AgO \qquad (E)$$

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:

$$CaCO_3 + H^+ = HCO_3^- + Ca^{2+} \qquad (F)$$

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™ 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™ 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™ oxidative leach is shown in Figure 6.

The lead residue from the Nordenham Albion Process™ 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.

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.

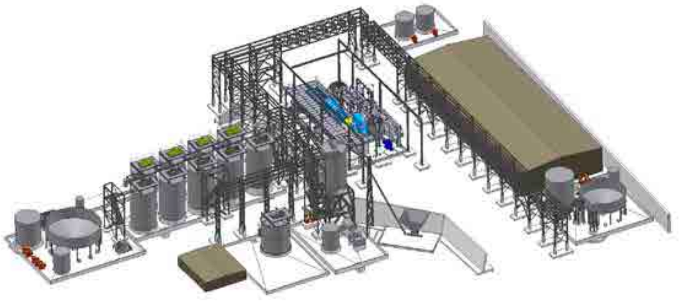

Figure 9
GPM Gold Albion Process™ Plant

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.

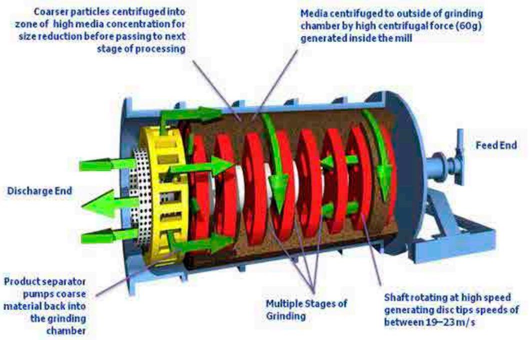

Figure 10
IsaMill^{TM} Operating Mechanism

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 $k_L a$. It is the $k_L a$ 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™ 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™ development program. The $k_L a$ 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™ 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:

$k_L a$ = 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.

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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

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Acknowledgments:

1. Abstract

With increasing levels of gold production coming from refractory sulphide ores and a chronic skills shortages in the mining industry there is a need for a simple, robust and lower cost treatment route for refractory gold ores.

Xstrata and Highlands Pacific have recently agreed to offer their Albion Process technology under license as a new lower cost alternative to existing refractory gold treatment technologies. Core Resources, a company founded by former Xstrata executives, has entered into arrangements to undertake further development and commercial licensing of the Albion Process technology.

As part of this commercialisation process, Aker Kvaerner Australia (AKA) worked in partnership with Core Resources to carry out a comparative capital and operating cost study using three different refractory gold treatment technologies for the gold ores of an advanced exploration project in North Queensland, Australia. This paper provides background to the Albion Process along with a summary of the benefits of using this novel technology at the North Queensland project, compared with bacterial oxidation or pressure oxidation. It also draws on AKA's extensive knowledge of bacterial and pressure oxidation plants and provides qualitative and quantitative assessments of the different technologies.

2. Introduction

Xstrata's Hydrometallurgical Research Laboratories was commissioned by the owner of the North Queensland project to conduct pre-feasibility metallurgical testwork on ore samples from the project. The upgraded flotation concentrate samples were tested using pressure oxidation, bacterial oxidation and the Albion Process.

Parameters from this testwork provided the basis of the design criteria for the study into a refractory gold treatment plant with a scale of approximately 100,000 oz/yr of gold, using each of these technologies.

The testwork was intended to determine the recovery performance of the North Queensland ore from each of these technologies and then to provide comparative data on operating and capital costs.

3. Albion Process – A New Lower Cost Technology

Metals such as copper, nickel, zinc and gold may be recovered from ores or concentrates by hydrometallurgical techniques, which involve the leaching of the valuable mineral in a chemical solution. In some instances, the mineral containing the valuable metal can not be leached efficiently using conventional techniques, and the mineral is called *refractory*.

Modern methods for liberating valuable metals from refractory minerals have involved complex processing of the minerals at high temperatures and/or pressures to facilitate leaching. These methods, such as roasting and pressure leaching, are capital intensive and, in the case of roasting, produce environmentally damaging effluent streams, requiring further processing. Other processes such as bio-oxidation are also capital intensive and difficult to manage.

The Albion Process is a highly attractive process to treat refractory sulphide minerals to recover valuable metals housed in the sulphide lattice. The process consists of ultrafine grinding of the mineral, followed by oxidative leaching at atmospheric pressure in open tanks. The flowsheet utilises existing, commercially proven technology in its unit operations. The Albion Process is therefore an important technical breakthrough.

The Albion Process was developed by MIM Holdings (now Xstrata) and a former MIM subsidiary Highlands Gold, to treat refractory base metal and gold ores. The technology has been tested extensively at pilot plant scale. The Albion Process is currently being evaluated for use in several world class mining projects by Xstrata, as well as by companies external to the Xstrata group.

The process was initially designed to operate under acid conditions for the extraction of base metals, possibly with precious metals by-products (the “Acid Process”). However during the development process, it was recognised that oxidation of pyrite (and tellurides) could be achieved more effectively under alkaline conditions, and that this was well suited to treatment of precious metals ores containing mainly pyrite. This led to a parallel development of the technology for mainly refractory pyritic gold ores, with separate patents lodged for this branch of the technology (The “Alkaline Process”).

3.1 Description of the Albion Process technology

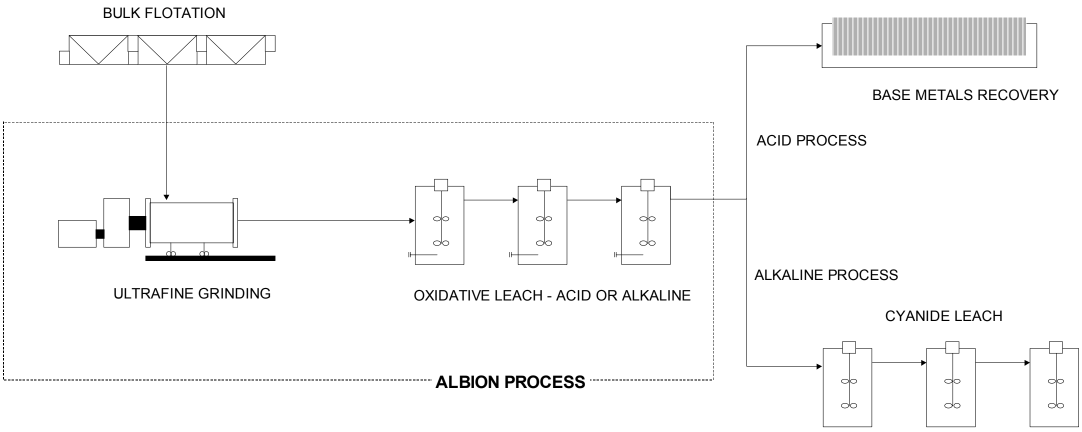


Figure 1 - Albion Process schematic flowsheet

3.1.1 Acid Process

The acid process is used to recover valuable metals such as copper, zinc, nickel and cobalt that are in the form of sulphide mineral concentrates. Precious metal by products such as gold can also be recovered from the leach residue.

The flowsheet involves oxidative leaching of a finely ground concentrate at atmospheric pressure. The process of ultrafine grinding, normally to 80 % passing sizes of below 20 microns, results in a high degree of strain being introduced into the mineral lattice which "activates" the mineral, facilitating leaching. The rate of leaching is also enhanced, due to the dramatic increase in the mineral surface area. Ultrafine grinding of sulphide minerals to a particle size of 80 % passing 8 – 12 microns will also eliminate passivation of the mineral by sulphur precipitates, as the leached mineral will disintegrate prior to the precipitate layer becoming thick enough to passivate the mineral.

Sulphide metals such as chalcopyrite, pyrite, or arsenopyrite are oxidised in the acid process, with ferric ions being the main oxidising agent.
The major leach reaction is:

$$MeS + Fe_2(SO_4)_3 = MeSO_4 + 2FeSO_4 + S^o$$

As the leach is carried out at atmospheric pressure, sulphides are preferentially oxidised to elemental sulphur.

The resulting ferrous iron is re-oxidised by oxygen to regenerate the ferric ions:

$$2FeSO_4 + 1/2O_2 + H_2SO_4 = Fe_2(SO_4)_3 + H_2O$$

Depending upon the feed mineral composition, some or all of the acid and the ferrous/ferric ions are obtained from dissolution of iron minerals (usually pyrite) in the feed. Metals such as copper may then be recovered by solvent extraction and electrowinning. Unwanted metals such as iron and arsenic are selectively precipitated and removed in a neutralisation stage following the leach. Contained gold (having been liberated through the oxidation of the sulphide minerals in which it was previously encapsulated), may be cyanide leached.

The process can be applied to the recovery of base metals such as copper, zinc, nickel and cobalt from sulphide concentrates.

3.1.2 Alkaline Process

The alkaline process is used to recover precious metals such as gold, silver and platinum group minerals that are present in refractory or carbonaceous ores that cannot be processed by standard cyanide leaching. The alkaline process is utilised where there is no requirement to recover base metals from the concentrate. The alkaline process is suited to minerals such as pyrite, arsenopyrite, selenide, or telluride which have encapsulated precious metals.

Refractory gold bearing sulphides, such as pyrite, liberate both iron and acid when oxidised. In the alkaline process, this iron and acid is continually neutralised and precipitated from the leach by the addition of an alkali such as limestone. The continual removal of the leach reaction products from solution means that the leach progresses rapidly and very high levels of sulphide oxidation can be achieved. Ultrafine grinding of the minerals to minus 20 microns or less prevents passivation by these precipitated leach products, as the leaching minerals are consumed before a sufficient layer of precipitates can form. The general leach reaction would be:

$$FeS_2 + 15/4O_2 + 9/2H_2O + 2CaO = FeO.OH + 2CaSO_4.2H_2O$$

At the same time, carbonaceous minerals that might interfere with the gold recovery processes are rendered inert by the precipitation of gypsum and iron oxides on the carbonaceous surfaces. As the subsequent cyanidation stage occurs in alkaline conditions, there is no need to neutralise the leach slurry prior to the cyanide leach, and so CCD washing or filtration of the leach slurry is not required.

4. Comparison of Albion Process against Bacterial oxidation or Pressure oxidation on a North Queensland ore.

4.1 Project Parameters

For the purposes of the comparative study the project location chosen was green-fields Australia, with access to regional infrastructure. This would be representative of the North Queensland project, and would also illustrate the project economics in other similar locations.

The project scale chosen was 740,000 tonnes per annum (tpa) of ore with a feed grading 4.6 g/t Au and 11.6 g/t Ag. Bulk flotation of the ore produced an upgraded concentrate as feed to the oxidation stage at a rate of 18 tph, grading 22 g/t Au, 28 g/t Ag, 22% S, and 5% As. Gold production was up to 100,000 oz/yr Au, depending on the process recovery.

The study examined the production and treatment of this refractory sulphide gold concentrate by three different oxidation technologies to produce a cyanide leachable slurry suitable for gold recovery in a conventional CIL/CIP circuit.

4.2 Ore Mineralogy

The dominant host mineral in the ore sample tested in the work program was chlorite-clay altered, high-potassium quartz gabbro. Abundant coarse pyroxene was present, altered to chlorite and kaolinite. Minor host minerals included rutile and ilmenite.

Sulphides were widely disseminated in the host rock, in silicified wall rocks, and also occurred in veinlets. Subhedral-anhedral <0.5 mm pyrite (5 vol%) was the commonest sulphide, and occurred in the least silicified host rocks, whereas <150 micron subhedral arsenopyrite (1 vol %), the other major sulphide in the ore, was concentrated in the quartz veins, the siliceous wall rock, and in siliceous alteration patches. No free gold was found, suggesting that a significant amount of the gold was present as submicroscopic gold in arsenopyrite.

4.3 Test Results

Bench scale test work conducted at Hydrometallurgy Research Laboratories demonstrated the following recoveries and consumable requirements from the North Queensland ores:

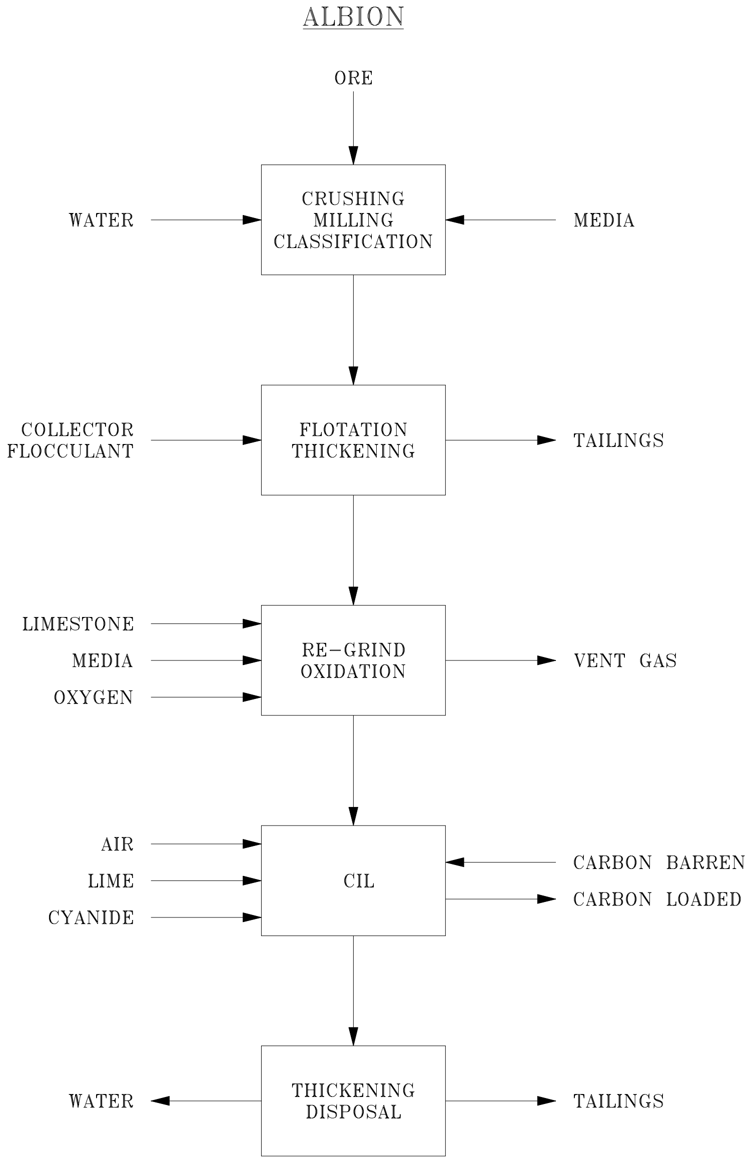
Table 1 - Process Test Results and Consumable Requirements

	Albion Process	**Pressure Oxidation**	**Bacterial Oxidation**
Leach recovery - Au %	92	94	78
Leach recovery - Ag %	75	4	25
Leach oxidation - sulphide %	70	96	65
Consumables			
Oxygen tpa	62,357	82,000	0
Limestone tpa	83,818	111,757	111,427
Lime tpa	14,209	6,824	2,069
$(NH_4)_2SO_4$ tpa			1,480
H_3PO_4 tpa			296
KOH tpa			814
Antifoam tpa			74
Flocculant tpa		3.8	
Sodium cyanide tpa	1,500	394	1,931
Activated Carbon tpa	18	7	5

4.4 Process description

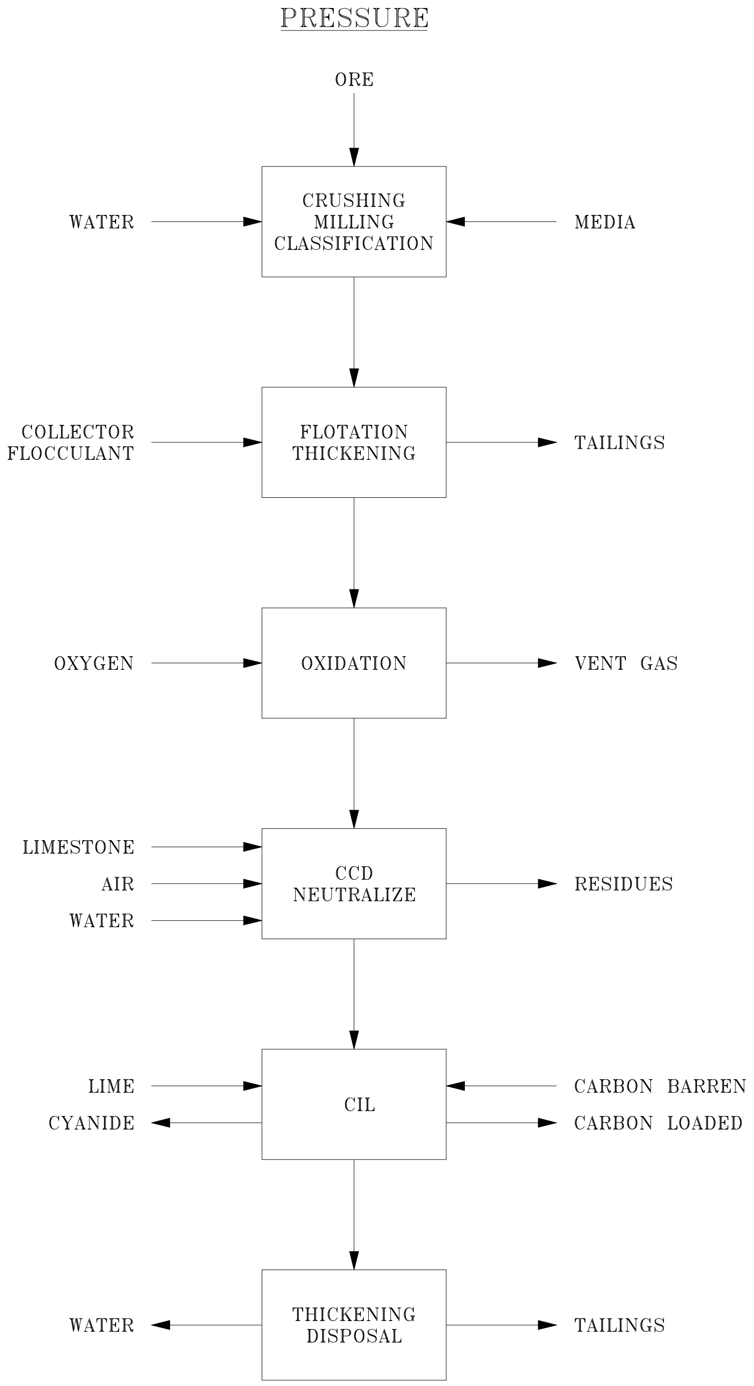
4.4.1 Albion Process

4.4.1.1 Block flow diagram



4.4.2 Pressure oxidation

4.4.2.1 Block flow diagram



4.4.3 Bacterial oxidation

4.4.3.1 Block flow diagram

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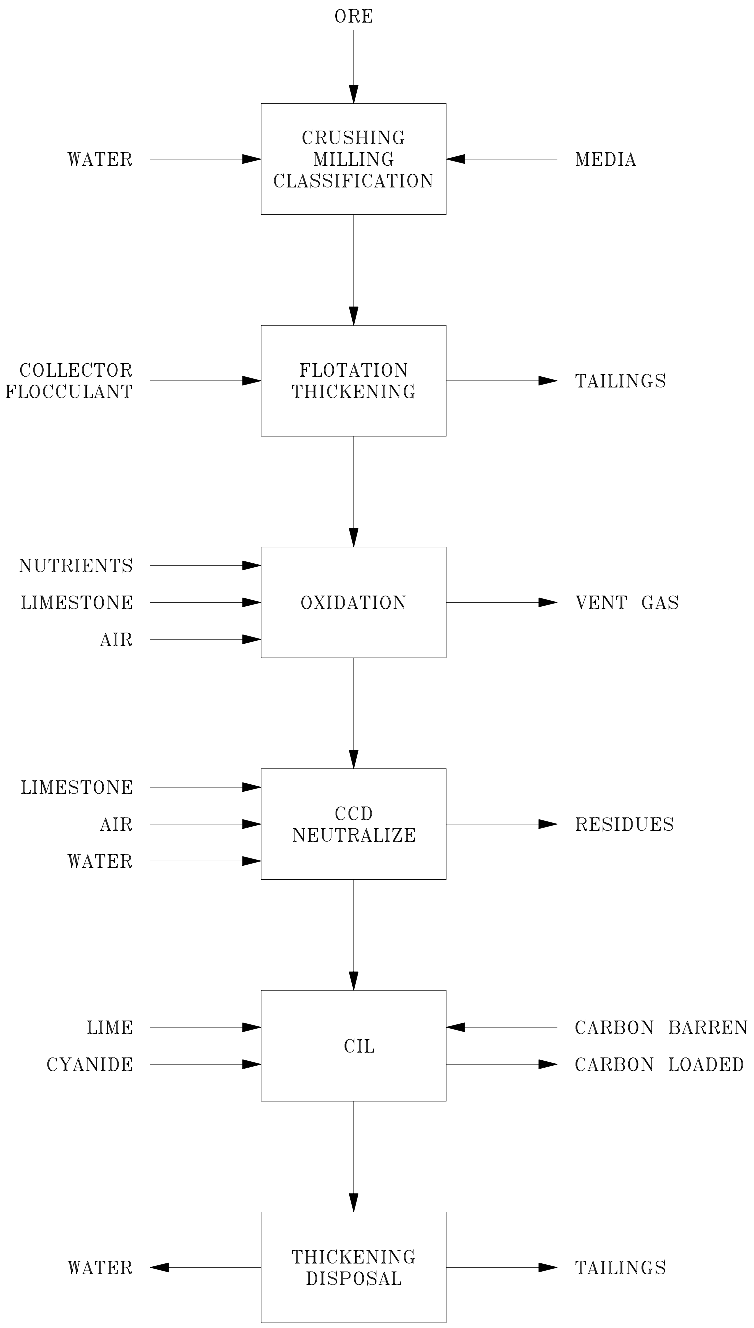

4.5 Capital Cost Estimate

4.5.1 Basis of Estimate

This order of magnitude estimate has been prepared to a “Class 1 – Conceptual” standard as defined by AKA, which is considered to be -30%/+50%. Capital costs are estimated at September 2004 and to not capture 2005 escalation of labour and materials. The scope of the direct cost estimate for the three options includes capital for the process facilities only.

The process plant estimates were developed by factoring from similar areas in other recent gold project estimates, and adjusted for size, flow or capacity. Currency exchange rates at September 2004 were assumed for this comparison of the metallurgical treatment options.

The estimates reflect installed costs in a green-fields location with access to regional infrastructure, illustrative therefore of North Queensland and of similar locations. Equipment costs cover instrumentation and control, electrical equipments, structural and concrete, piping and fabrication costs and installation. A contingency allowance of 10% was applied to direct costs to reflect only the conceptual scope uncertainty.

The summary of the capital costs is shown in Table 1. Costs are presented in Australian dollar (A$) values.

4.5.2 Capital Costs Exclusions

The capital cost has been presented to show the relative direct capital cost for the various process facilities. The following items are excluded from the estimate as the level of scope development is conceptual:

- Utilities and reagent supply outside the process plant,
- General infrastructure,
- Capital spares for first year of operation
- EPCM and vendor supervision,
- Heavy cranage, temporary facilities and site distributables,
- First Fill consumables and warehouse stocks,
- Insurances, technology licences and tariffs,
- Construction camp and mobilization,
- Mining and haulage of ore and limestone to the process plant,
- Tailings dams and effluent disposal,
- Sustaining, replacement and deferred capital,
- Assistance during commissioning, ramp up and operation,
- Working capital requirements and owners costs.

4.5.3 Methodology

The capital costs for each process block were generally factored directly from previous AKA projects or recent studies as most appropriate. The reference projects were selected for similarity of equipment and adjusted for flow capacity and currency movements.

The capital cost for the CCD and Neutralisation process block, (not part of the Albion Process flowsheet) was factored from an AKA Study, (May 2003). Differences in the downstream process including gold recovery were also factored on the basis of their respective metallurgical and operational criteria.

Albion Process

The capital cost for the Albion Process block was factored from an AKA study which compared the Albion Process with roasting, (August 2004). The process criteria for oxygen consumption efficiency and reactor retention were applied to obtain the factored costs. The capital cost for an oxygen plant is not included because of through the fence supply allowance captured in the operating costs.

Pressure Oxidation

The capital cost for the Pressure oxidation process block was factored from a recent AKA gold pressure oxidation study (May 2003) allowing for currency movements. The process criteria for oxygen consumption efficiency, feed pumping and reactor retention were applied to obtain the factored costs. The capital cost for an oxygen plant is not included because of through the fence supply allowance captured in the operating costs.

Bacterial Oxidation

The capital cost for the Bacterial oxidation process block was factored from a recent AKA gold Bacterial oxidation study having equivalent tank and agitator configuration. The cost of the air blowers, slurry cooling and other ancillary equipment was calculated from the AKA database. The process criteria for air demand and dispersion, and reactor retention were applied to obtain the factored costs.

4.6 Operating Cost Estimate

4.6.1 Basis of Estimate

This order of magnitude estimate has been prepared to a "Class 1 – Conceptual" standard. Operating costs are estimated at December 2004, and do not capture 2005 escalation of labour and consumables. Comparative operating costs were developed for major expense categories including;

- Labour
- Utilities
- Consumables and reagents
- Maintenance

Each category was broken down by item sufficiently to allocate rates and prices. The costs are based on the "back-end" of the process facilities as shown in the block flow diagrams (see section 4.3). The costs to operate the front end facilities of crushing, milling and flotation are largely identical and not considered in this comparison.

- Concentrate treatment rate; 18 tonnes per hour
- Concentrate grade; 22% sulphur and 22g/t gold
- Operating hours; 8200 per year approximating 94% availability

Prices for reagent utilities were obtained from supplier inquiries or the AKA database. The consumption rates or quantities were derived from mass balances from the testwork or interpreted from studies of similar projects by AKA. The cost of oxygen was based on a dedicated facility for supply 'through the fence' from a supplier quotation.

Labour allocations were nominated for the concentrate treatment facilities including neutralization and CIL requirements based on other recent studies. Salaries or wages were estimated from in-house information for Australian projects.

Maintenance costs were factored against direct equipment costs based on ratios from similar processing facilities. Power and water costs were developed from estimates of area consumer and a power cost of 5 c/kWh was used.

The following items are excluded from the operating cost estimate;

- Management of residue impoundment and rehabilitation
- Regulatory and licence costs
- Head office costs and site security
- Product transport, refining and royalty charges
- Insurances and statutory inspections
- Amortization, depreciation and financing

- Ore comminution and concentrate production

The costs presented for the process facilities are relative and capture direct operational expenses only for the oxidation and downstream operations. The costs are considered at conceptual scoping level.
All costs are presented in Australian dollars with values current at September 2004.

4.7 Summary of Comparative Cost Estimates

The comparative capital and operating costs are presented and discussed below. Preliminary comparisons can be made on a relative basis. Capital costs for the process facilities are shown in Table 2, and include estimates for the concentrate production to show their relative magnitude against concentrate treatment (highlighted in grey). Operating costs are summarised in Table 3 for concentrate treatment including oxidation leaching, neutralisation and washing, followed by CIL treatment of oxidised concentrates.

4.7.1 Capital Cost Comparison

Table 2 - Process Facilities Direct Capital Cost Estimate Summaries

Area Description	Albion Process	Pressure Oxidation	Bacterial Oxidation
Crushing and Grinding	15.4	15.4	15.4
Flotation and Concentrate Thickening	4.0	4.0	4.0
Oxidation	10.5	25.3	29.8
CCD and Neutralisation	-	2.5	3.0
CIL, Gold Recovery and Cyanide Detox	1.7	1.0	1.1
Contingency 10%	3.2	4.8	5.3
PROCESS TOTAL A$M	**34.8**	**53.0**	**58.6**

The Albion Process concentrate treatment facilities are estimated to have substantially lower capital cost that the other process routes, i.e. $12.2M compared to $28.8M for Pressure oxidation and $33.9M for Bacterial oxidation. The direct capital saving are in the order of $16M and $22M respectively (excluding contingency).

These estimated differences in capital costs can be viewed in terms of the total capital needs that may be around $100M-$150M for a typical 100,000 oz/yr green-fields, refractory, gold operation in Australia. Total project costs will be influenced by infrastructure requirements and mine development, but this potential saving is substantial at the conceptual accuracy of the estimate.

The lower cost for Albion Process plant and equipment can be attributed to smaller reaction tanks and agitators compared to Bacterial oxidation, while the pressurized Pressure oxidation operation requires substantially more expensive equipment. The amount of installed exotic piping and complexity of control would also be substantially less for an Albion Process oxidation facility.

4.7.2 Operating Cost Comparison

Table 3 - Concentrate Treatment Operating Cost Estimate Summary

Item	Albion Process	Pressure Oxidation	Bacterial Oxidation
Labour	1.7	3.5	2.1
Utilities	2.2	1.6	3.3
Reagents	9.3	6.3	7.7
Maintenance	0.9	2.1	1.3
Concentrate Treatment Annual Operating Cost A$M	**14.1**	**13.5**	**14.5**
Cost per Ounce Produced A$	**147**	**138**	**178**

Concentrate treatment operating costs for the Albion Process plant, $14.1M pa are similar to the Pressure oxidation facility, $13.5M pa with the conceptual accuracy of the estimates. The slightly lower unit cost for Pressure oxidation is indicative of the incrementally higher gold recovery predicted following this aggressive leaching route. The supply costs for oxygen is the major, variable expense items for the Albion Process and Pressure oxidation process flowsheets.

Slightly higher total operating costs are estimated for the Bacterial oxidation process, $14.5M pa compared to Albion Process concentrate treatment. The major variable cost difference arises from the air supply and dispersion power demand for Bacterial oxidation compared to oxygen consumption in the Albion Process reactors. The lower gold recovery from the testwork results in substantially higher unit costs for operating the Bacterial oxidation flowsheet.

The Albion Process plant is estimated to have the lowest labour fixed cost due to the relative simplicity of operation and maintenance. The Bacterial oxidation plant requires closer monitoring by operators to maintain bacterial activity, and has more ancillary facilities, e.g. neutralization and water cooling. The Pressure oxidation plant is expected to have greater process control requirements and higher maintenance workforce due to the more aggressive operating conditions. The additional cost of these complexities will depend on the remoteness of the plant location from major population centres and the availability of skilled labour.

4.7.3 Total Concentrate Treatment Cost Comparison

Table 4 summarises the total concentrate treatment costs including amortising the capital costs for the various oxidation technologies. The costs used covered the process plant from concentrate oxidation to cyanidation of the leach residue. Capital costs that were amortised included the direct costs for concentrate treatment from Table 2 (highlighted in grey) with 10% contingency and 20% indirect costs added. The total capital costs including contingency and indirect costs for concentrate treatment for the three processes are as follows; Albion Process - $16.1M, Pressure Oxidation - $38.0M and Bacterial Oxidation - $ 44.8M.

Capital costs were amortised at 15% per year.

Table 4 - Concentrate Treatment Operating Cost Summary (including Capital charge)

Item	Albion Process	Pressure Oxidation	Bacterial Oxidation
Operating Cost per Ounce Produced A$	147	138	178
Capital Cost Amortised per Ounce Produced A$	25	58	83
Total Cost per Ounce Produced A$	172	196	261*

* Total costs for Bacterial oxidation would reduce to $219/oz, if the gold recovery achieved was the same as the Albion Process (ie 92%)

5. Process Options Comparison

5.1 Operability

The major attractions of the Albion Process compared to other refractory concentrate treatment routes include the simplicity of operation and lower capital cost. All equipment for Albion Process will have conventional and proven designs, and would be smaller than Bacterial oxidation facilities of equivalent concentrate capacity. The process control and monitoring demands for the Albion Process equipment are expected to be simpler than all other competing technologies including Bacterial oxidation, Pressure oxidation, roasting and chlorination.

Start up and shut down are rapid for Albion Process facilities unlike Bacterial oxidation or Pressure oxidation reactors. No equipment or operational risks arise from loss of temperature or pressure that may be experienced in roasters or Pressure oxidation plants. Loss of power, oxygen or air supply presents no major difficulties to the process efficiency or upon a re-start.

The stable nature of the Albion Process allows operators to learn to control the operation quickly and reliably. Personnel with limited process backgrounds can be taught to run the plant optimally in a short time frame. The operational duties are expected to be shared by the CIL operator in most facilities, thus keeping labour costs at a minimum.

5.2 Availability

High run time availability of 94% or better is expected from the Albion Process plant area, which is higher than most milling facilities and other refractory ore treatment plants.

The equipment in the Albion Process plant has conventional design and common materials of construction. The layout and piping configuration can allow for individual equipment bypassing while concentrate treatment continues.

5.3 Flexibility

The Albion Process is able to deal efficiently with all occurrences of refractory concentrates; sulphides, arsenides, or carbonaceous. The ultrafine grinding and effective oxygen uptake allow variable feed grade treatment. The nature of the precipitated iron and sulphur species in the Albion Process leach residue promotes low cyanide consumption in downstream gold recovery.

The presence of base metals and organic contaminants in concentrates are accommodated by the Albion Process. The extent of oxidation achieved in the Albion Process leach is a function of residence time, and is relatively insensitive to mineralogy. As the leach is not bacterial, it does not require the addition of nutrients to the leach, and metals or organic contaminants that may limit bacterial activity will not affect the Albion Process leach.

Volatile metals such as mercury are not volatilised at the operating temperatures of the Albion Process leach, and report to the residue. Toxic metals such as arsenic will report to the leach residue in an environmentally stable form, as ferric arsenate.

5.4 Maintainability

The application of simple proven process steps, conventional equipment and accessible plant layouts contribute to easily maintainable plant. The smaller tank size and the benign slurry conditions used in the Albion Process leach allows improved access and lower maintenance costs compared to Bacterial and Pressure oxidation facilities. The Albion Process flowsheet also avoids CCD and separate neutralization sections completely, and so the overall maintenance workload will be minimal compared to Pressure oxidation facilities.

The generally alkaline slurry for precious metals treatment in the Albion Process largely negates the need for high grade stainless steel throughout the plant. The use of exotic nickel alloys or titanium are not required for the Albion Process facilities. Special surface preparations, frequent inspections and operational contingencies that are prevalent for expensive materials are not required at the Albion Process operating conditions.

5.5 Effluents and Emissions

Slurry discharge from the Albion Process would be further treated through a conventional CIL flowsheet for gold recovery. The final residues can then be handled and stored in impoundments or mine-fill following standard cyanide detoxification. No other liquid effluents arise from the Albion Process flowsheet.

Gaseous emissions from the Albion Process leach tanks are minimal due to the high oxygen utilization. Residual oxygen, and carbon dioxide that arise from in-situ neutralization, will be saturated in moisture. Beside the vapour plume dispersion no gas cleaning will be required to remove acid mist or volatile components.

5.6 Gold/Silver Recovery

Albion Process testwork on a variety of ores has achieved gold recoveries generally comparable to Bacterial oxidation, however in this instance it was higher. Pressure oxidation can release incrementally higher levels of gold due to the higher temperature and pressure involved.

A key benefit of the Albion Process is that in every instance tested to date it has achieved significantly higher silver recoveries than Bacterial oxidation and Pressure oxidation.

5.7 Reagent Consumption

The gold is recoverable following Albion Process leaching by conventional CIL/CIP processing at relatively low cyanide and lime consumptions. Limestone consumption will be similar for all oxidation process routes as dictated by the level of sulphur oxidation. Cyanide consumption in the CIL/CIP plant is lower than Bacterial oxidation due to the nature of the leach residue. The Albion Process uses lower levels of neutralising agents for some concentrates as some elemental sulphur is produced.

Nutrient, flocculant or dispersion reagents are not needed for Albion concentrate treatment. Low cost grinding media is needed for fine grinding of concentrate in IsaMills

Oxygen demand by the Albion Process is expected to be similar to that needed on Pressure oxidation reactors. For this study an oxygen utilisation of 80% was used for Pressure oxidation and the Albion Process reactors. The consumption in Pressure oxidation will depend on the concentrate mineralogy, including the carbonate assay.

6. Current Process Options

A plethora of concentrate treatment flowsheets have been promoted over the last decade for both base and precious metals. Novel leaching options for gold, nickel and copper sulphide concentrates may be compared to the Albion Process based on the intensity of process conditions.

A selection of processes that have been developed to piloting or beyond are listed in table 5. Other processes that do not use air or oxygen as the primary oxidant have not been listed. Note that many of these process have emerged specifically to treat copper concentrates but are included as they may be adapted to mixed copper/gold concentrates or gold concentrates

Table 5 - Proposed Routes for Sulphide Concentrate Leaching

Parameter	High	Moderate	Low or Ambient
Temperature, celsius	>155	115-155	<115
	Pressure oxidation	AAC/UBC CESL Dynatec	**Albion Process** Activox Mt Gordon Bacterial oxidation
Pressure, pascal	>1M	150k-1M	<150k
	Pressure oxidation AAC/UBC CESL Dynatec	Activox Mt Gordon	**Albion Process** Bacterial oxidation
Retention, hours	>24	1-24	<1
	Bacterial oxidation	**Albion Process** Mt Gordon Activox	Pressure oxidation AAC/UBC CESL Dynatec
Grind Size, P80 microns	<38	38-75	>75
	Albion Process Activox AAC/UBC	Bacterial oxidation CESL Dynatec	Pressure oxidation Mt Gordon

The Albion Process is designed to operate at ambient pressure and relatively low temperature. This arrangement allows concentrate oxidation to be conducted in conventional agitated tanks which allow the construction costs to be minimized. Bacterial oxidation is the only other mentioned process that operates at these low intensity conditions. The Bacterial oxidation process requires about 5 days retention compared to approximately 1 day for the Albion Process.

The Bacterial oxidation reaction time will decrease marginally as the grind size decreases and the leach temperature rises but is unlikely to drop below 4 days. After concentrate treatment by Bacterial oxidation the metal recovery is likely to be lower while the reagent consumptions will be higher relative to the Albion Process plant. Control of Bacterial oxidation can become problematic at higher slurry temperature due to fluctuating activity of more specialized thermophilic bacteria.

The Albion Process and Pressure oxidation process are the only routes to have been shown to operate in alkaline slurries. Lower cost materials of construction are expected to be installed compared to acid slurries, particularly at the lower temperature of operation.

7. Conclusions

A number of refractory gold processing technologies are currently available. However, the Albion Process now offers the gold mining industry a simpler, robust, and lower cost alternative for treating these ores.

The technology has been extensively tested at large pilot plant scale on a number of different ores and a qualitative comparison of the main process factors shows that the Albion Process conditions are technologically robust and potentially less complex to build and operate.

The study provided some compelling evidence into the cost benefits of the Albion Process. These benefits, along with lower capital exposure and simple commissioning makes the Albion Process worth considering.

Core Resources is currently working with potential partners to bring the first Albion Process plants into production.

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