EVALUATING PROCESS OPTIONS FOR TREATING SOME REFRACTORY ORES

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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.

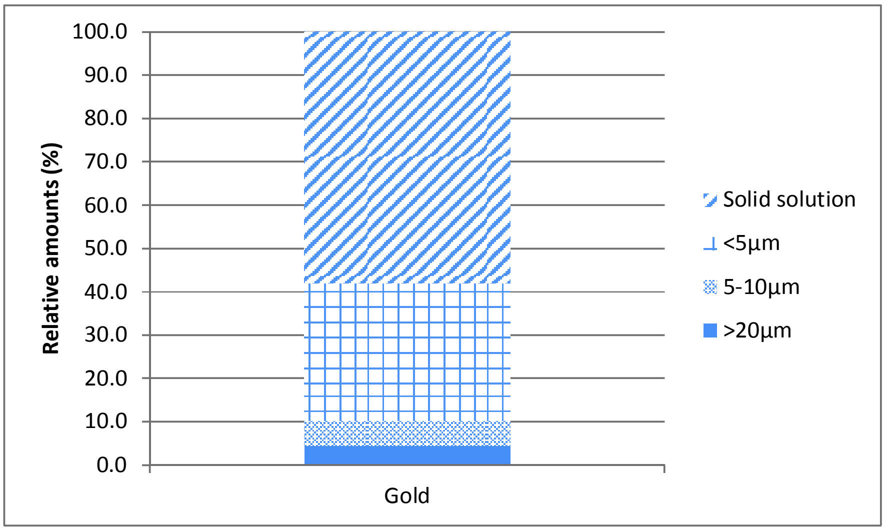


Figure 1: Gold deportment in Ore A

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.

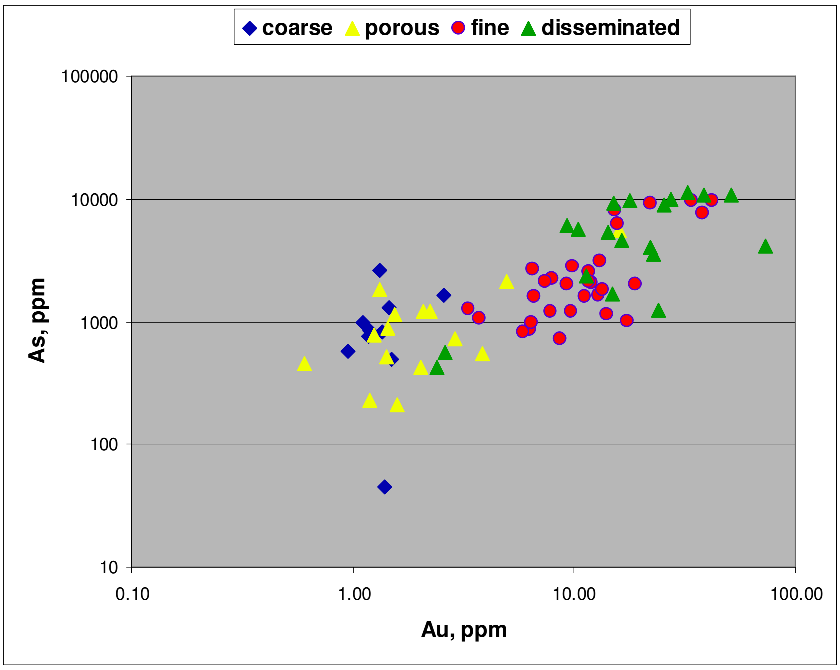


Figure 2: Correlation between measured concentrations of sub-microscopic gold and arsenic in different morphological types of pyrite (from 74 point analyses)

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.

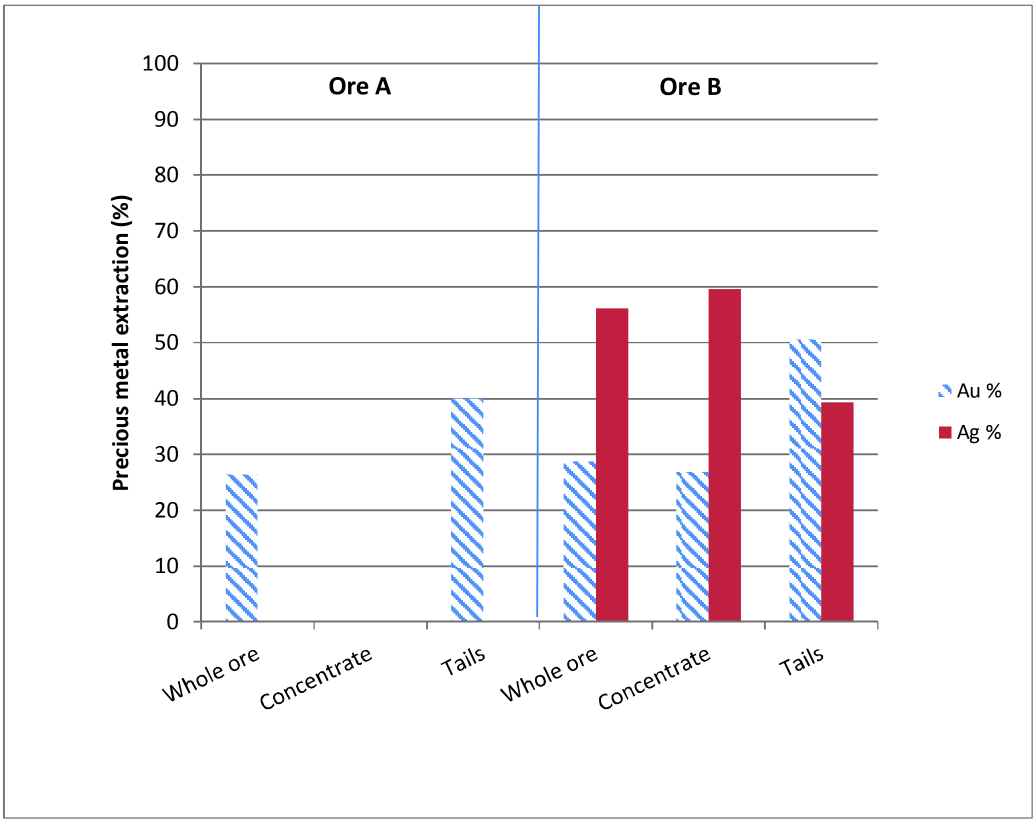


Figure 3: Precious metal extraction following cyanidation on both Ore A and Ore B samples

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.

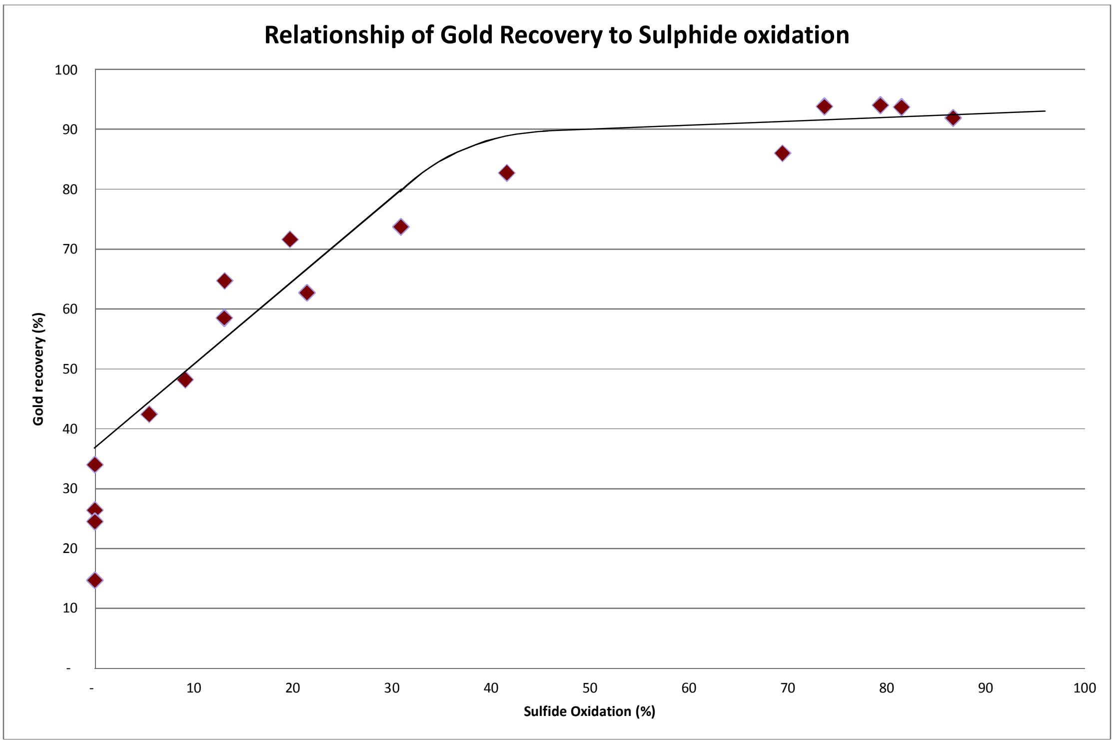


Figure 4: Relationship of gold recovery to sulfide oxidation required for Ore A and concentrate to be amendable to cyanidation

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.

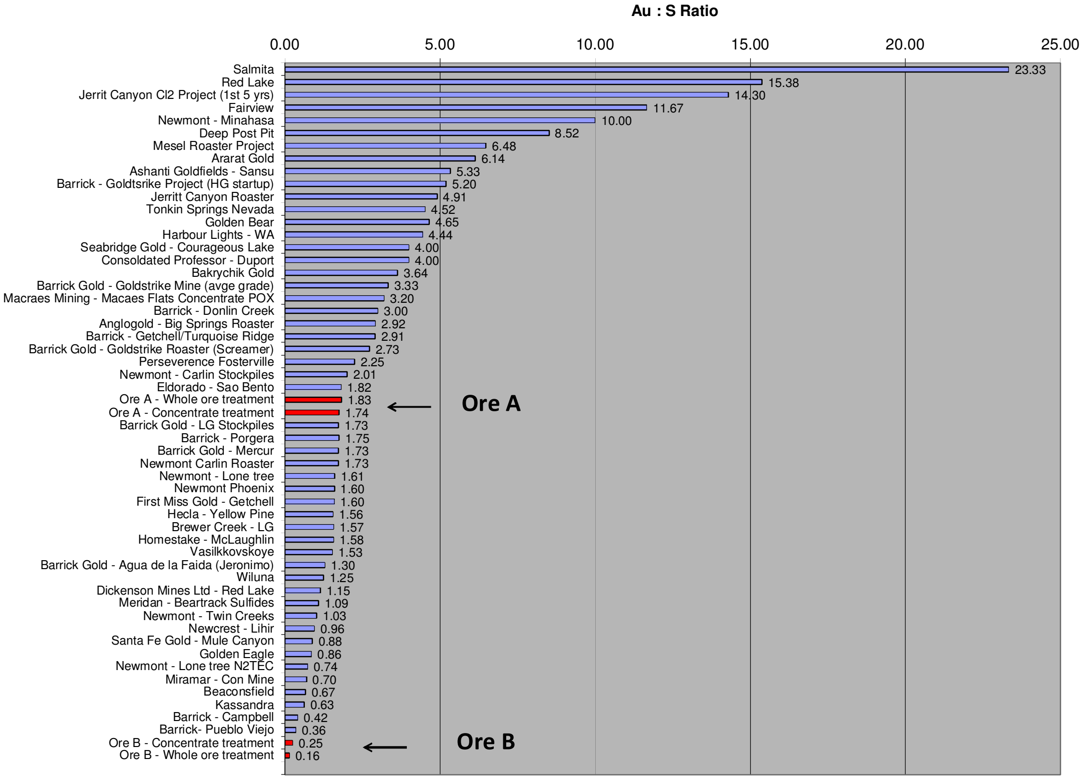


Figure 5: Comparison of gold to sulfur ratio for the two ore types and various existing and previous process operations around the world

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. Ulfrafine milling the ore or concentrate was considered as a low capital option even though gold recovery would not be expected to be high. Historical test work data ensured that biooxidation heap leach option be considered. The Albion process was considered as a possible viable approach given its development in recent years towards full commercial scale.

For Ore B only the three conventional processes were considered. Notably the high silver to gold ratio in the proposed feed favours the recovery of precious metals via the Merrill Crowe process discussed below.

Table 3: Process Options Considered for ore A

Processing options	**Crushing and grinding**	**Flotation**	**Pre-treatment**	**Cyanidation and recovery**	**Expected Au Recoveries (Overall)**
Bacterial Oxidation Heap leach – Whole ore	Tertiary crushing, stacking	None	Heap bio oxidised and then reclaimed	Product processed through a conventional CIL plant	70%
Ultra fine milling – Whole Ore	Ore is crushed and milled to 75µm	None	Ore fined milled to 10µm		50%
Ultra fine milling – Concentrate		Floated for a 9% mass pull	Concentrate fined milled to 10µm		45%
Roasting – Concentrate					
Albion - Concentrate			Fine milled to 10µm Atmospheric oxygenated leach		83%
Bacterial Tank Leach – Concentrate			Bacteria oxidation Residue neutralised and washed		83%
Pressure Oxidation - Concentrate			High pressure oxidation Residue neutralised and washed		85%

Table 4: Process Options Considered for ore B

Processing options	Crushing and grinding	Flotation	Pre-treatment	Cyanidation and recovery	Expected Au & Ag Recoveries (Overall)
Roasting – Concentrate	Ore is ground to 75µm	floated for a 30% mass pull	Roasted in air and residue quenched	Oxidised concentrate combined with flotation tails and cyanide leached. Solid/liquid separation. Au & Ag in PLS recovered by the Merrill Crowe process.	81.5% Au 62% Ag
Bacterial Tank Leach – Concentrate			Bacteria oxidation Residue neutralised and washed		81.5% Au 62% Ag
Pressure Oxidation - Concentrate			High pressure oxidation Residue neutralised and washed		81.5% Au 62% Ag

DESIGN PARAMETERS

The overall process design based on input from clients and derived from average values obtained in the test work programme are summarized in Table 5.

Expected throughputs, battery limits, expected available process infrastructure and unit costs associated with power and reagent supply were supplied by clients or derived from other work carried out by Bateman.

A basic mass balance for all the process options was generated to determine the stream flow rates and compositions. Sizing of key equipment was based on the technical specifications derived from the mass balance. Pricing of equipment was derived from vendor quotations, together with in-house database costs.

Table 5: Physical and Plant Data

Ore		A	B	
Time Data				
	Life of Mine	8	10	yrs
	Overall Plant availability	91.3	91.3	%
Plant Data				
	Resource	-	50	Mt
	Basis for Design	600,000	3,650,000	t_{ore}/yr
	Flotation Plant production	54,000	1.095,000	t_{concs}/yr
Ore Mineralogical Composition				
	Arsenopyrite FeAsS	2.0	0.01	%
	Pyrite FeS_2	1.75	12.3	%
	Copper Sulphides (Covellite, chalcopyrite,tetrahedrite)	-	0.07	
	Carbonates	20.0	-	%
	Gangue	76.25	86.84	%
	Au	2.25	1.04	g/t
	Ag	na	32	g/t
Concentrate Mineralogical Composition				
	Arsenopyrite FeAsS	20.0		%
	Pyrite FeS_2	18.5	34.7	%
	Copper Sulphides (Covellite, chalcopyrite,tetrahedrite)	-	0.06	
	Carbonate	3.3	-	%
	Gangue	58.2	87.5	%
	Au	22.5	2.5	g/t
	Ag	-	45	g/t
Bond work data				
	Ore	17	15	kWh/t
	Concentrate	30	20	kWh/t

A description of the ore preparation, cyanidation and tails detoxification processes which are applicable for the majority of the process treatment options are presented in the following sections. Details of the pre-treatment processes are described later. The design of tailings deposition facility has not been considered in the study.

Crushing and Grinding

For Ore A crushing will comprise a primary jaw crusher only, the product from which will feed the SAG mill.

For Ore B a primary and secondary crusher system will have been previously installed for treating oxide ore for a heap leach operation. For the purpose of this study the crushed feed from the secondary crusher is transported by conveyor to be screened and cone crushed to yield a P80 of 2000 micron particles.

For both ores the grinding circuit were configured as a SAG Mill / Ball Mill circuit to grind the material to 80% passing 75 microns. The SAG mill will be operated in open circuit and the ball mill in closed circuit with hydro-cyclones. The cyclone overflow will feed either into the flotation circuit or thickener for whole ore processing.

For the whole of ore heap bioleach option for treating Ore A the ore is crushed to 100% passing 9 mm prior to bioleaching. The residue reclaimed from the bioleaching is ground to 80% passing 75 microns in a 5 x 5.5m ball mill.

Flotation Circuit for Concentrate Options

The trash screen underflow from the cyclone overflow gravitates to the conditioning tank where lime, collector and fresh water, if required, are added. The slurry in the conditioning tank flows to rougher flotation tank cells in series with an assumed total retention time of 24 minutes. Flotation air to each tank cell is provided by a dedicated external blower which ensures each cell receives a controlled air flow. The concentrate from the rougher cells is pumped to a high rate concentrate thickener. The thickened flotation concentrate is then pumped to one of the refractory treatment processes.

For Ore A the flotation circuit has been considered to be a rougher-scavenger circuit with a single cleaner which will upgrade the concentrate by a ratio of 10:1. The cleaner tailing will be recycled back to the scavenger feed. The overall recovery is estimated at 90% and the weight recovery 9% based on the results of the flotation test work. The flotation tails with the high residual carbonate content will be used as neutralising material in acidic oxidation processes before being disposed in a tailings dam.

For Ore B the gold and silver recovery is estimated at 70% and 45% respectively, with the weight recovery 30% based on the results of the flotation test work. The thickened tails is pumped to the cyanidation circuit to be treated along with the oxidized residue from the refractory process plant to extract the gold and silver.

Cyanidation and Gold Recovery Circuits

Ore A residue

The gold cyanide leaching circuits were designed on a 24 hour residence time. The same circuit configuration will be used for either whole or concentrate, with the whole ore sized for treating 75 t/h, whereas the concentrate is sized for treating 7 t/h.

Feed from the refractory pre-treatment plant is pumped to a slurry neutralising tank where applicable and on to the Carbon in leach (CIL) circuit for gold recovery to carbon. Gold will be recovered from the leach pulp using granular carbon before stripping in a spilt AARL elution circuit.

Gold loaded to carbon will be recovered in batches from CIL Tank #1 before acid washing with dilute hydrochloric acid and rinsing with water. The loaded carbon will be stripped using the AARL elution process. The pregnant strip solution flows from the elution column into the pregnant solution tank from where it is pumped at a controlled rate to the EW circuit.

The stripped carbon is thermally reactivated in a regeneration kiln at a temperature of 700°C. The reactivated carbon is screened to remove carbon fines before being returned to the last CIL tank to replace the forwarded carbon. The fine carbon is forwarded to a settling pond and periodically recovered and bagged for sale.

Gold is plated onto the electrowinning cell stainless steel wool cathodes. The barren electrolyte flows to a collection tank and pumped to the barren solution storage tank for recycling to the elution circuit. Cells are periodically cleaned of gold sludge. The sludge will be calcined and smelted to produce gold doré bullion for transport.

Ore B Residue

The carbon in leach or carbon pulp process conventionally used in recovering precious metals is not conducive to recovering high silver content in ore and concentrate. Ores with high silver to gold content generally favour Merrill-Crowe recovery process. This is largely because of the very large carbon stripping and electro winning systems required for processing large quantities of silver in a conventional CIL plant(12). Examples of operations in Peru where the Merrill Crowe approach has been used are Newmont's Minera Yanacocha mine and Barrick's Pierina Mine. Ore B has a silver to gold ratio of 30:1 and consequently the Merrill Crowe process is favoured.

The gold and silver cyanide leaching circuit is designed on a 24 hour residence time. The circuit configuration has been sized to treat the oxidised concentrated as well as the tails from the flotation circuit.

Oxidised feed from the refractory pre-treatment plant is pumped and combined with the flotation tails in a conditioning tank where lime is added, if required, to bring the pH up to 10.5, and cyanide added prior to being pumped to a series of six cascading tanks.

Following leaching the pregnant cyanide liquor solution is separated and washed from the solids via a three stage CCD thickener circuit consisting of 35m diameter high rate carbon steel thickeners. The pregnant cyanide liquor overflow from the final CCD thickener is treated in the Merrill Crowe plant to recover silver and gold.

The 940 m^3/hr Modular Merrill Crowe and refinery plant consists of the following steps following cyanidation:

- Solid liquid separation using three 35 m high rate counter current decantation thickeners to remove the pregnant precious metal solution as described in the above section.
- The pregnant solution is clarified in a polish leaf filter to reduce the suspended solids to approximately 1 mg/L.
- The clarified solution is de-aerated using packed tower (Crowe) under vacuum.
- Powdered zinc is added to the de-aerated clarified solution inline or tank to precipitate precious metals from solution.
- The resultant precipitate is pumped to a filter press and washed.
- The filtered precipitated is collected and acid washed to remove the zinc.
- The acid washed precipitated product is refiltered, washed and air dried.
- The product is smelted to form silver/gold dore.

Detoxification Circuits

Following the requirements of the International Cyanide Management Code tailings will be detoxified before storage in a lined tailings facility. Weak acid dissociable (WAD) cyanide in the leach tailings will be neutralised to below 0.5 ppm CN using an Inco system. The resultant detoxified slurry residue is pumped to a tailings storage facilities.

TREATMENT OPTIONS FOR ORE A

Ultra Fine Milling – Whole Ore or Concentrate

The general concept flow sheet for whole ore and concentrate is shown in Figure 6. Concentrate or whole ore feed from the grinding circuit is further ground to a particle size of less than 10 µm to physically liberate gold using either a vertical or horizontal stirred mill consisting of rotating stirrers within a stationary mill shell to impart kinetic energy to a fine media charge (usually sand). Xstrata Technology owns the rights for the commercialisation of the IsaMill™, a horizontal stirred mill. The mill product is then treated in a conventional cyanidation CIL plant to recover the gold. The presence of very fine inclusions of gold in the order of 10 µm and the possibility of some surface oxidation of the sulphide minerals may improve in gold recovery from coarser ground feed.

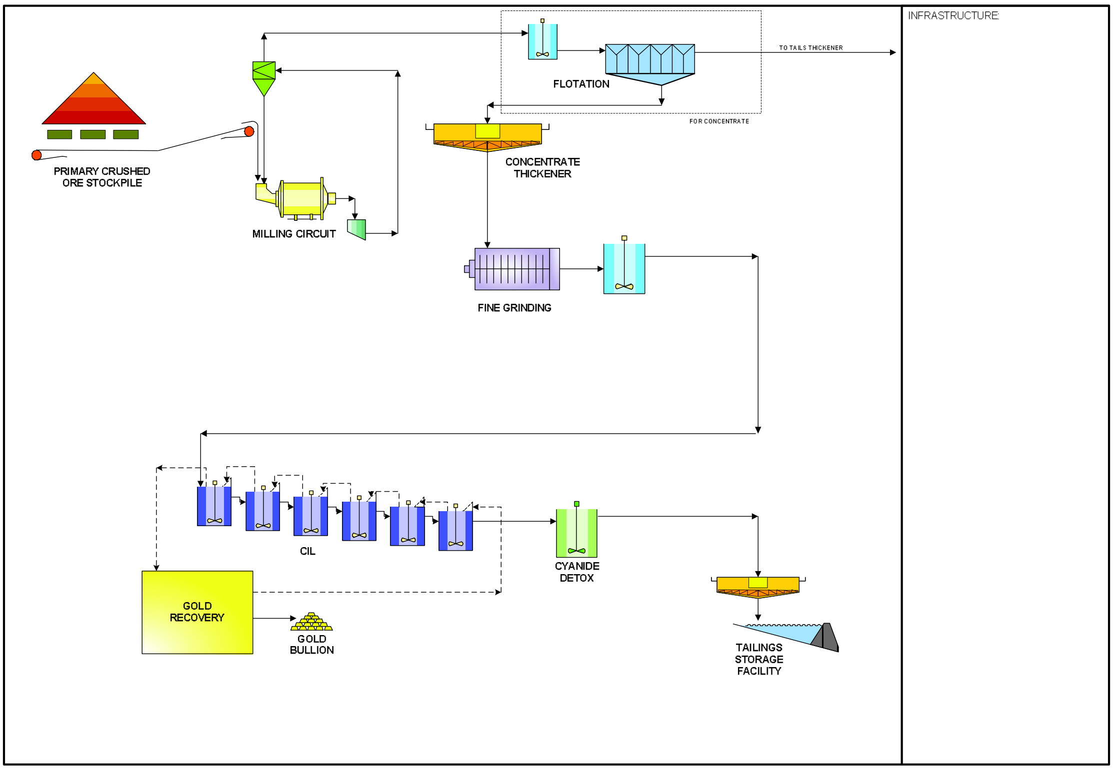


Figure 6: Ultra fine grinding circuit for whole ore and concentration (flotation circuit inclusive)

Bacterial Heap Oxidation/Leaching

Biological heap leach technology can be used to oxidise whole ore sulphides in open stockpiles before treatment in a conventional oxide gold treatment plant. Biological heap leaching is often used for base metal recovery but in that case the solutions are recovered rather than the solids. The most successful biological heap refractory leaching has been conducted by Newmont in their operations in Nevada(13). Heaps of low grade ore are treated at a rate of 3.5 Mtpa. The schematic diagram of the bacteria whole ore heap is shown in Figure 7.

In the process refractory ore is crushed to -9 mm before conveyor stacking on a heap leach type leaching pad constructed on re-usable lined leach pads. A bio-solution (containing bacterial inoculum and nutrients) is then added to the ore on the stacker conveyor and recirculated through the heap. Air is percolated through the heaps over a nominal total cycle time of 400 days, including 165 days bio oxidation. A mixture of bacteria consisting of mesophiles, moderate thermophiles and thermophiles are propagated with the heap with heap temperatures reaching up to 80 C. The bio-solution is recycled to the heap and a bleed from the bio-solution recycle is neutralised using limestone before discharge or re-use to maintain moderate pH and neutralise the acid generated during oxidation. Arsenic is precipitated as stable ferric arsenate during neutralisation. At the completion of the oxidation cycle, the ore is washed and allowed to drain before reclaiming.

The oxidised ore is reclaimed from the oxidation heaps and processed through a conventional cyanide leaching plant. The process has a low capital cost but has a relatively high technology risk. Notwithstanding this, it can be a cheap simple technology.

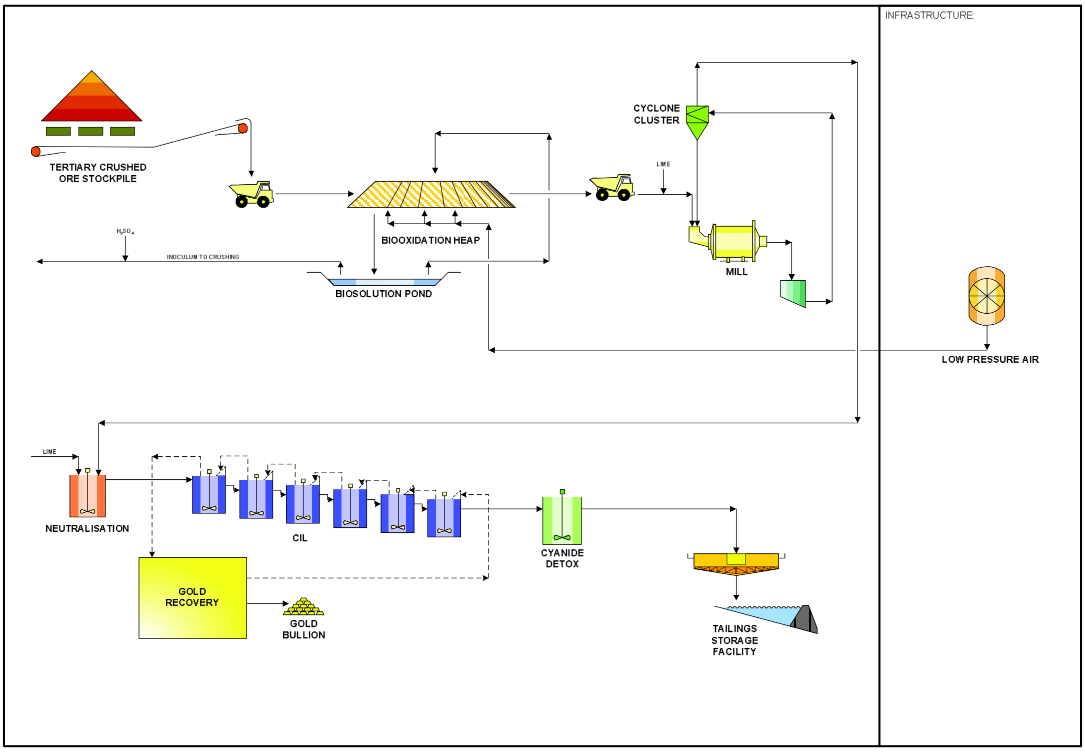


Figure 7: Biological Oxidation Heap Leaching – Whole Ore

BIOX® Agitated Tank Reactors

The Biological oxidation (BIOX®) process uses bacteria to oxidise sulphides in large agitated tanks before treatment in a conventional oxide gold treatment plant. The technology is well proven in particular for concentrates and in use in a number of similar operations(14). The process is illustrated in Figure 8.

The concentrate with a P_{80} of 75 µm is fed to the biological oxidation circuit comprising a series of primary bioleach reactors in parallel. Partially oxidised product exiting the primary reactors are recombined and enter a series of secondary reactors. The configuration (normally three primary reactors in parallel and three secondary in series) allows for at least double the residence time in the primary reactors to promote growth and division of the bacteria and prevent washout. The temperature is controlled to between 40-45 C at ambient pressure. Cooling and aeration are significant contributors to cost and power demand.

The pulp pH is maintained at less than 2 with a total residence time of 3 to 5 days. A CCD circuit separates solids and liquids. Iron and arsenic and any base metals report to solution and are neutralised with limestone followed by lime. Arsenic is precipitated as the stable ferric arsenate phase. The residue is washed and neutralised for cyanidation. Gold and silver are leached from the residue, recovered on carbon and electrowon in the conventional way.

Biological oxidative leaching is easily used to control oxidation levels of the ore for optimum recovery. BIOX® technology requires significant capital although the technology is relatively simple and well understood. This technology has the registered trade mark BIOX® and is now owned by Gold Fields of South Africa. Similar technology developed by Mintek of South Africa and BacTech Mining Company has been installed commercially.

Biooxidation testwork has been conducted on Ore A. Gold recovery from the oxidised residue was 92% at an 80% oxidation level. It is probable that there was some preferential oxidation of the arsenopyrite / arsenian pyrite to achieve the recovery at the lower oxidation level. Tank biooxidation is particularly suited to partial oxidation of the sulphides. Hence some costs associated with power and reagent consumption can be reduced compared with full sulphide oxidation processing.

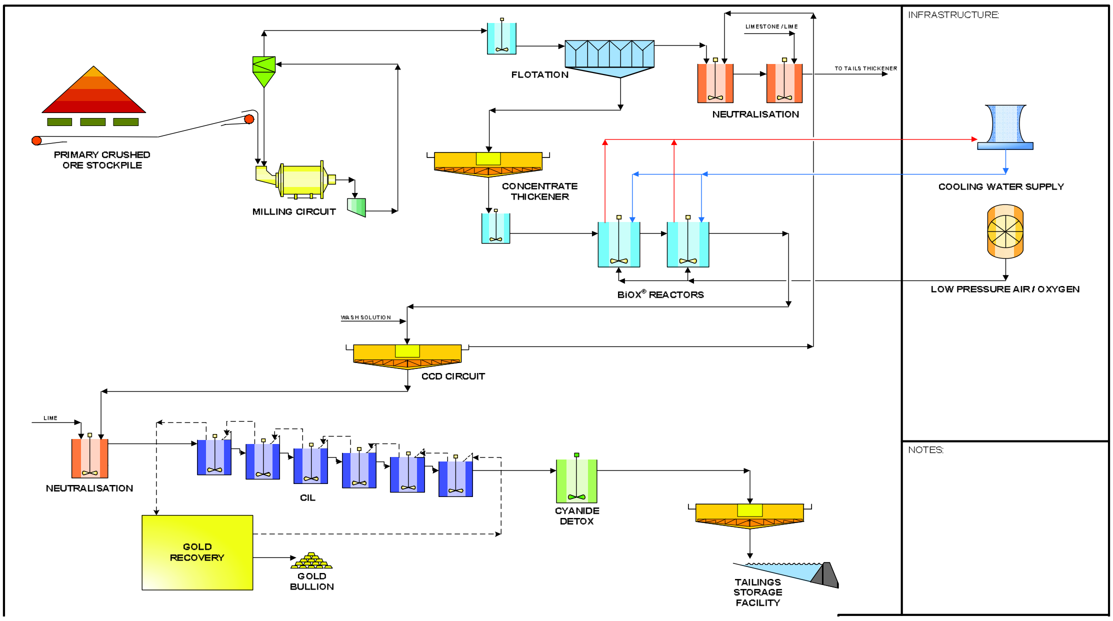


Figure 8: Agitated Tank Bioleaching – Concentrate

Albion Process

The Albion process uses ulfrafine grinding in stirred mills to produce a highly activated, finely ground concentrate at low energy inputs. The finely ground concentrate is then leached with oxygen at atmospheric pressure in conventional agitated tanks. The flow sheet of the process is illustrated in Figure 5.

The Albion Process was developed by a MIM subsidiary, Hydrometallurgical Research Laboratories in the mid 1990s, and a US patent was obtained in late 1999. Xstrata acquired control of MIM in June 2003. Numerous piloting campaigns at the Hydromet Research Laboratories have been run since 1995, on copper-gold and gold applications at up to 50 kg/day. In recent years development to full commercialisation is proceeding. The Certej Project owned by European Goldfields is under development in Romania. This project uses an Albion Leach to leach gold from a flotation concentrate produced from 3 Mtpa ore. The process has been subjected to bench scale and pilot testing at Xstrata's Hydrometallurgical Research Laboratories in Australia. A total of 57 operations have tested the Albion process at HRL and 11 have proceeded to full pilot plant trials.

The sulphide concentrate produced from flotation will be fed to a fine grinding M1000 Isamill to yield a product P_{80} of 10 µm. The ground ore is then transferred to three Albion oxidation reactors where low pressure oxygen is sparged into the slurry. In the process the sulphide matrix containing the gold is broken down to release ferric ions and sulphuric acid. The temperature is not controlled and rises to close to 100°C. The breakdown of the sulphides liberates the gold. The ferric ion and acid are neutralised in situ by the continual addition of limestone slurry. The limestone is dosed to maintain the leach pH within the range of 5 to 5.5. Arsenic is precipitated as the stable ferric arsenate phase. The neutralised iron oxides (mainly goethite) and gypsum formed in the process is then processed through the CIL circuit.

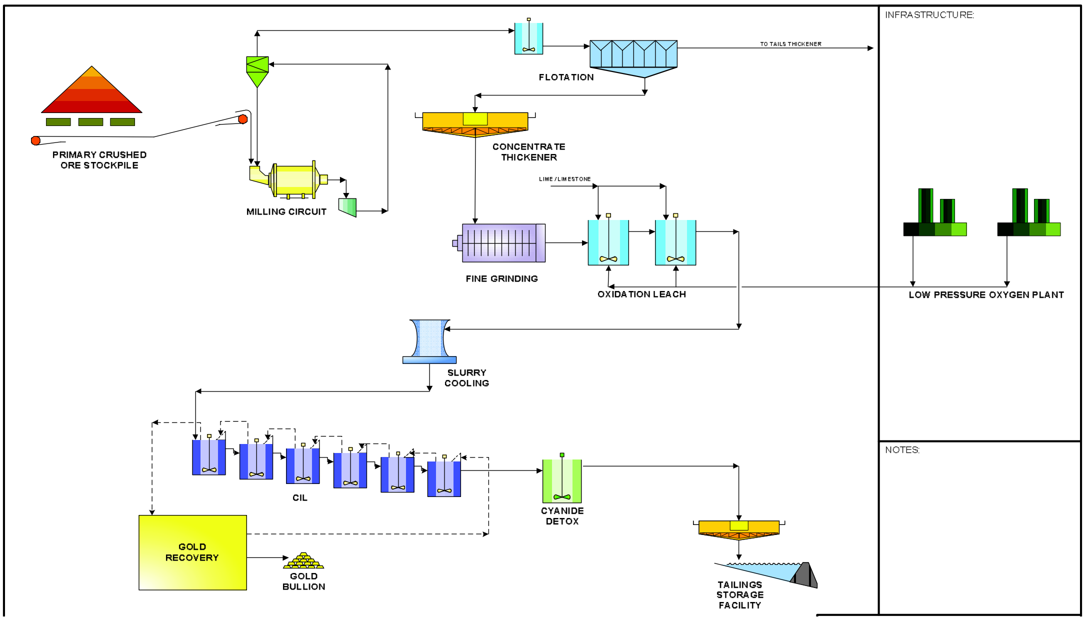


Figure 9: Albion Process – Concentrate

Pressure Oxidation

Pressure Oxidation is a high intensity process to oxidise sulphides in autoclaves before treatment in a conventional oxide gold treatment plant. Over the last 20 years around 15 pressure oxidation plants were commercialised specifically for the gold industry. Of these, eight were designed for treating concentrate. Original processes followed the designs developed in the Sherritt Gordon patents. Treatment approaches for overcoming operational issues have been patented largely by Newmont (15).

The plant layout is illustrated in Figure 10. Flotation concentrate is fed to the autoclave feed tank and then into the autoclave. The autoclave circuit includes a high pressure slurry feed system, autoclave vessel and agitators, flash vessel, and gas handling system. The autoclave will operate at 200ºC and 3,100 kPa and a retention time of approximately 30 – 60 min. Close to full oxidation of the sulphide is expected as partial oxidation is not likely to be viable. The process is expected to be autothermal operation not require any pre heating of slurry to achieve operating temperature. The sulphide minerals are oxidised to haematite with the arsenic precipitated as stable ferric arsenate.

The quenching of the excess flash steam and autoclave vent gas is accomplished with CCD overflow solution. The acidic CCD overflow solution is neutralised with the tails from the flotation circuit which contain high carbonate content. By using the tails to neutralise the effluent will enable a reduction in limestone consumption and costs.

Slurry from the let down tank is treated in a three-stage CCD circuit. Each thickener will be constructed from 316 L stainless steel walls, floor and rakes. The nominal wash ratio in the CCD circuit is maintained with wash efficiency of 99.0%.

The washed oxidation product will be pumped to the CIL tanks for conventional gold recovery. No test work on the performance of pressure oxidation on the concentrate has been carried out. However, from experience it is expected that gold recovery may be slightly better (~2%) than bio-oxidation.

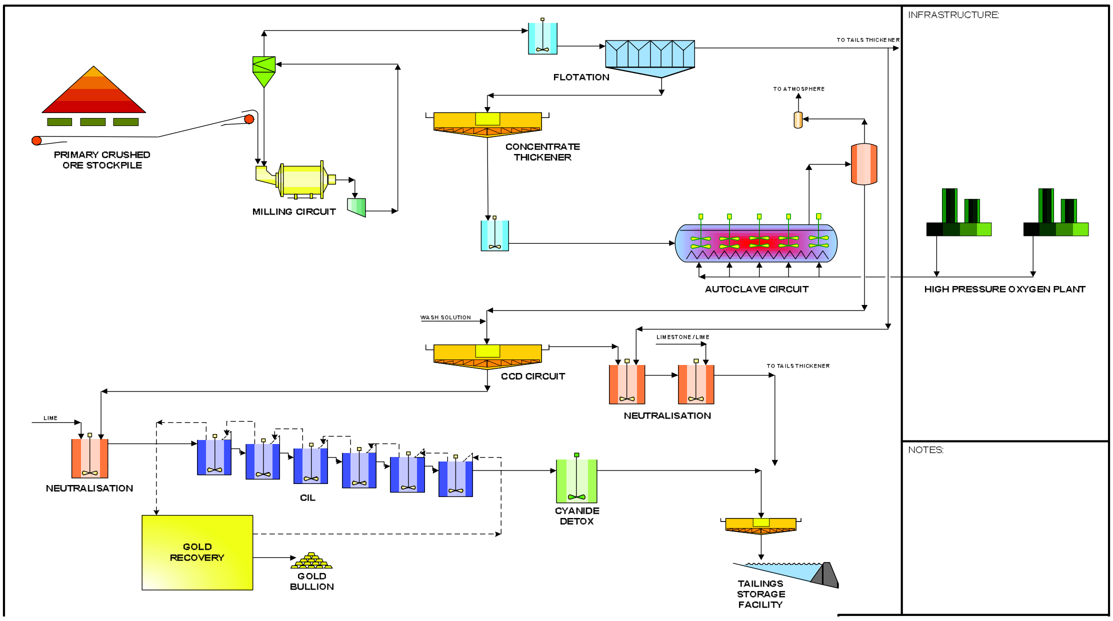


Figure 10: High Temperature Pressure Oxidation

TREATMENT OPTIONS FOR ORE B

BIOX® Agitated Tank Reactors – Concentrate

The process is illustrated in Figure 11. The concentrate with a P_{80} of 75 µm is fed to the biological oxidation circuit comprising a series of primary bioleach reactors in parallel. Partially oxidised product exiting the primary reactors are recombined and enter a series of secondary reactors. The configuration (5 modules each containing four primary reactors in parallel and four secondary in series) allows for at least double the residence time in the primary reactors to promote growth and division of the bacteria and prevent washout. The temperature is controlled to between 40 and 45 ºC at ambient pressure. Cooling and aeration are significant contributors to cost and power demand.

The pulp pH is maintained at less than 2 with a total residence time of 3 to 5 days. A CCD circuit separates solids and liquids. Iron and arsenic and any base metals report to solution and are neutralised with limestone followed by lime. Arsenic is precipitated as the stable ferric arsenate phase for disposal as tailings.

The washed oxidation product is pumped to the precious metals recovery plant. The oxidised residue is treated in the Lime boil circuit where silver containing jarosite formed during the oxidation process is decomposed to liberate the silver. The lime boil circuit consists of four cascading agitated tanks with the temperature maintained at 90˚C using steam generated a boiler or refractory process. Lime is added to obtain a pH of 10.5 (at ~25˚C). To conserve heat in the process a heat exchanger transfer heat from the outgoing slurry to preheat fresh oxidation product feeding into the lime boil circuit.

To recover silver and gold, the neutralised oxidised residue, combined with the flotation tails, is cyanide leached to extract the precious metals. The pregnant liquor is separated from the solids via counter current decantation to recover the precious metals by a Merrill Crowe plant.

No biooxidation testwork has been conducted on ore B concentrate. Tank biooxidation is particularly suited to partial oxidation of the sulphides. Hence some costs associated with power and reagent consumption can be reduced compared with full sulphide oxidation processing, if test work was to show the precious metals could be recovered from a partial sulphide oxidation product.

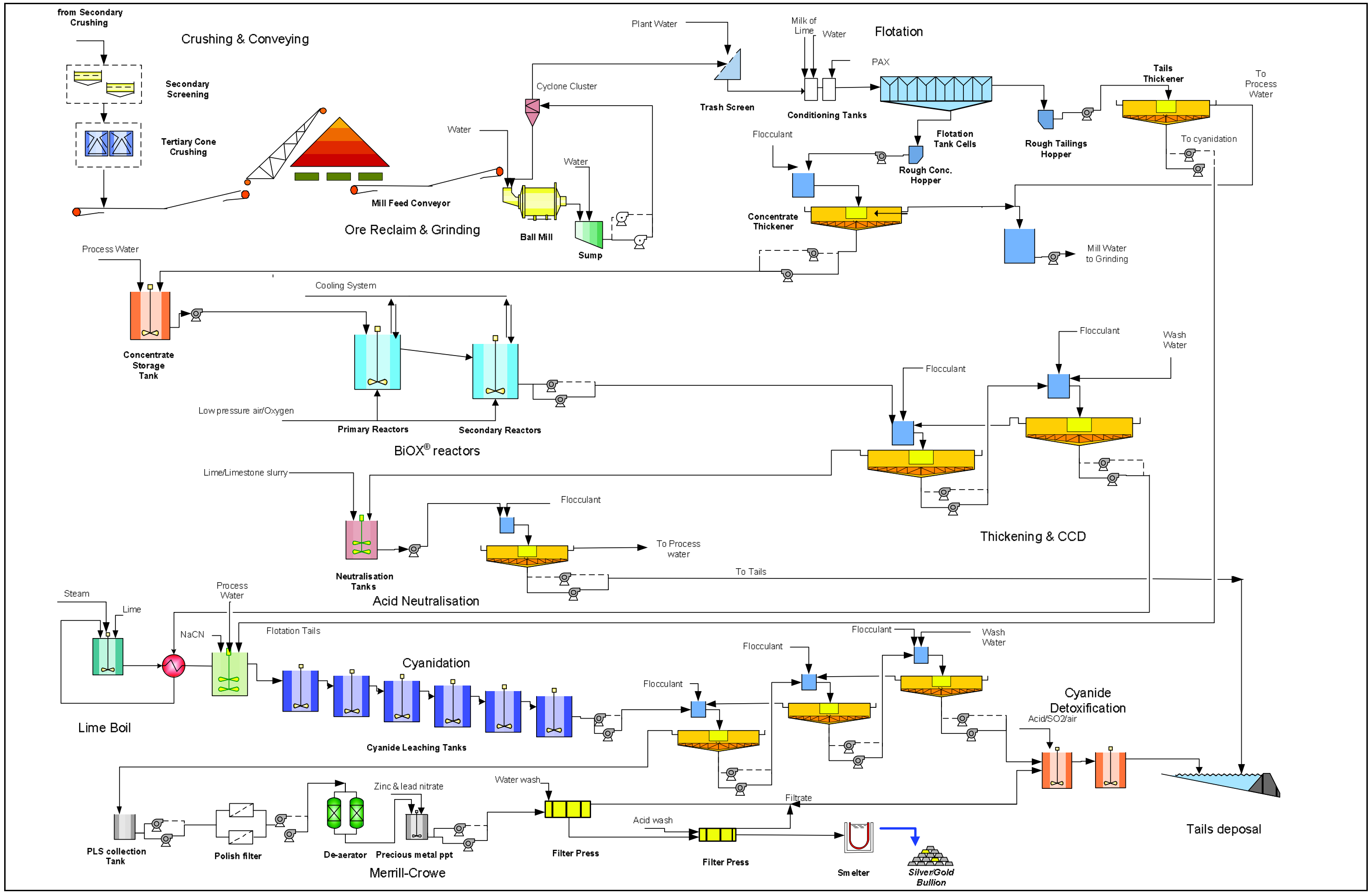


Figure 11: Agitated Tank Bioleaching – Concentrate

Pressure Oxidation – Concentrate

The Pressure Oxidation plant layout is illustrated in Figure 12. Flotation concentrate is fed to the autoclave feed tank, diluted and then fed into one of two autoclaves. The autoclave circuit includes a high pressure slurry feed system, autoclave vessels and agitators, flash vessels, and gas handling system. The autoclave will operate at 210-220ºC and 3,100kPa-3,300KPa and a retention time of approximately 30 – 60 min. Full oxidation of the sulphide is expected as partial oxidation is not likely to be viable. The process is expected to be autothermal and not requiring any pre heating of slurry to achieve operating temperature during operation. The sulphide minerals are oxidised to haematite with the arsenic precipitated as stable ferric arsenate. Slurry from the let-down tank is treated through a hot cure stage consisting of four tanks in series. This process is required to convert all basic ferric sulphate formed during the pressure oxidation process to ferric sulphate so as to reduce lime consumption during neutralisation.

The quenching of the excess flash steam and autoclave vent gas is accomplished with CCD overflow solution. Slurry from the hot cure stage is treated in a three-stage CCD circuit. Each thickener will be constructed from 316 L stainless steel walls, floor and rakes. The nominal wash ratio in the CCD circuit is maintained with wash efficiency of 99.0%. The overflow wash liquor from the CCD circuit is collected and neutralised in a series of eight tanks. The neutralised water is stored and reused in the process plant.

The washed oxidation product is pumped to the lime boil circuit where silver containing jarosite formed during the pressure oxidation process is decomposed to liberate the silver. Similar to the Biox plant the lime boil circuit consists of four cascading agitated tanks with the temperature maintained at 90˚C using steam generated by a boiler. Lime is added to obtain a pH of 10.5 (at ~25˚C). To conserve heat in the process a heat exchanger transfers heat from the outgoing slurry to preheat fresh oxidation product feeding into the lime boil circuit. The lime boiled slurry is pumped and combined with the flotation tails in a conditioning tank for cyanidation to extract gold and silver. Following liquid-solid separation step silver and gold are recovered from the pregnant cyanide liquor solution via the Merrill Crowe plant.

No test work on the performance of pressure oxidation on the concentrate has been carried out.

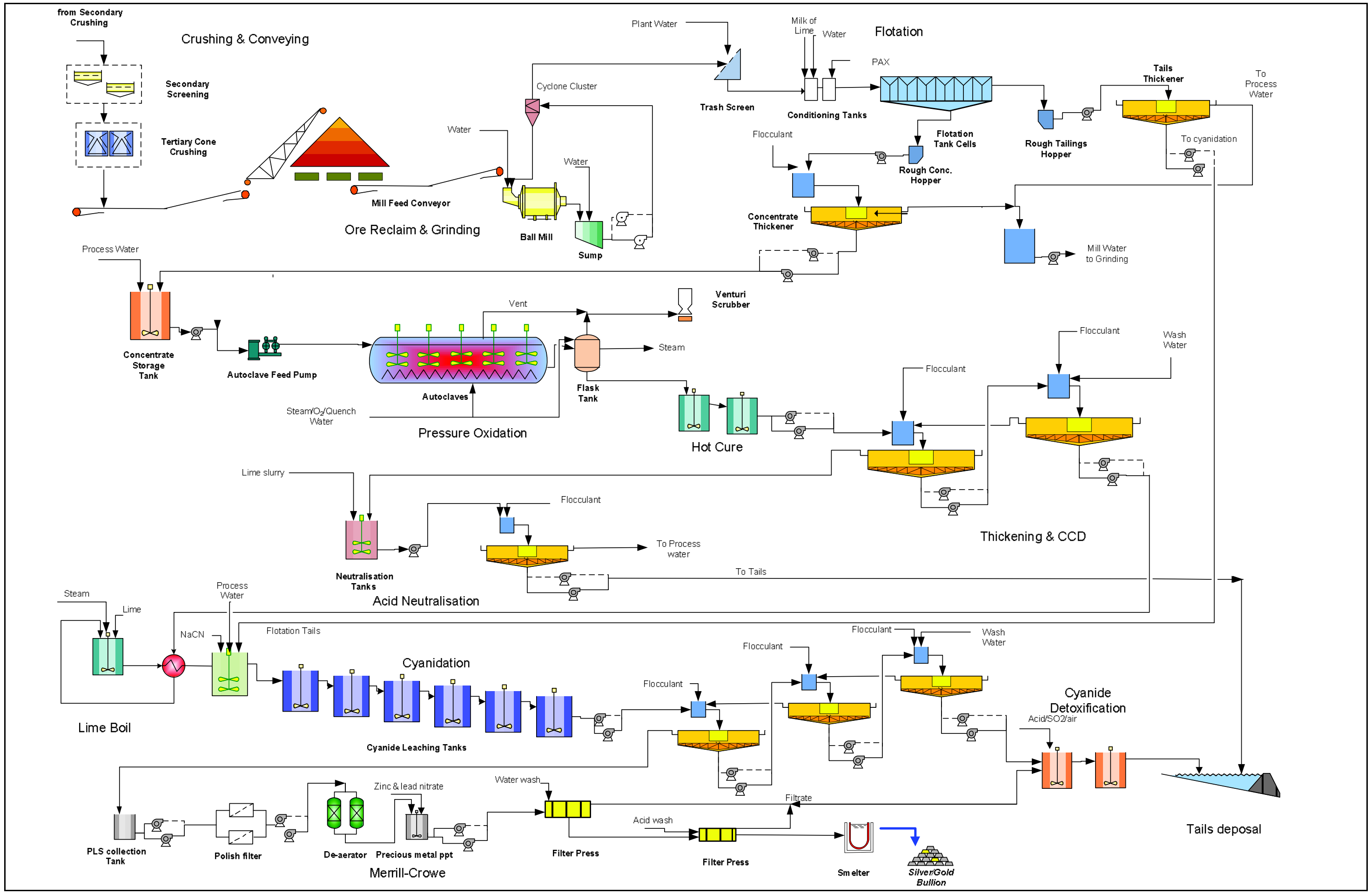


Figure 12: High Temperature Pressure Oxidation – Concentrate

Air Roasting – Concentrate

Roasting in an oxidising atmosphere, to produce a cyanide leach amenable porous calcine, has traditionally been the most commonly applied oxidative pre-treatment process for refractory gold ores. The concentrate from the flotation circuit will be thickened and filtered to produce a feed to the roasters. Oxidation by roasting uses air or oxygen enriched combustion to oxidise sulphides before quenching and treatment in a conventional oxide gold treatment plant. Three roasting technologies are available; circulating fluid bed (CFB) roasting under air and oxygen enriched atmosphere and conventional fluidised (or bubbling bed) technology. A single stage roasting step should be sufficient to fix arsenic at the levels encountered in the ore. This is subject to confirmation by testwork.

The concentrate is fed to a rotary drier, where it contacts hot roaster flue gas to reduce the moisture content to 5%. The dried concentrate is then distributed by the roaster rotary valve to the fluidised roasting bed chambers where the sulphide minerals react with the fluidising air. The bed temperature will be maintained at around 600 to 650ºC, to achieve 95% oxidation of total sulphide. For the treatment of concentrate the process design is based on two roasters supplied by Sinopec Nanjing Design Institute (SDNI) with one acid plant supplied by Noram.

The off-gases will be passed through a waste heat boiler at approximately 10% SO_2, and 6-7% O_2, before passing through the electrostatic precipitator (ESP). The ESP will reduce particulates in the gas to ~0.2g/Nm^3 before entering the gas purification section of the sulphuric acid plant. Arsenic trioxide will be captured in the ESP and returned to the calcine.

The gas is diluted with air to give the optimum SO_2 concentration for conversion in the double conversion and double absorption process. Nominally 99.75% of the SO_2 is removed and converted to acid, with a tail gas exhausted to atmosphere at an SO_2 concentration of 250 ppm. The acid is adsorbed in a circulating acid solution and 98-98.5% sulphuric acid is cooled in the product acid cooler before going to the acid storage tank for storage and export.

A water-tube boiler will produce saturated steam for use in electricity generation. It is expected that a nominal amount of electricity will be generated in the roaster/acid plant for potential use in the process. Excess LP saturated steam will be available for heating duties. For the purposes of this study, the benefit arising from potential power generation has not been included as a credit in the overall cost of the process.

The quenched roaster calcine will be slurried and pumped to the cyanide leach circuit and gold and silver recovered from the pregnant liquor by the Merrill Crowe plant.

In the absence of tests, the Au and Ag recoveries across the roasting process have been chosen from a range that would be typically expected for pyritic feedstocks. While roasting can produce a relatively inert calcine leading to lower cyanide consumption by complete oxidation of sulphur, the control of emissions of As and SO_2 is difficult and adds to the capital cost and complexity of the operation. Fixing the arsenic in calcine product is possible but would reduce gold recovery and increase reagent consumption in the subsequent cyanidation stage. Furthermore issues such as sintering and lock-up in hematite may also lower recoveries. Test work would be required to determine viability of an approach.

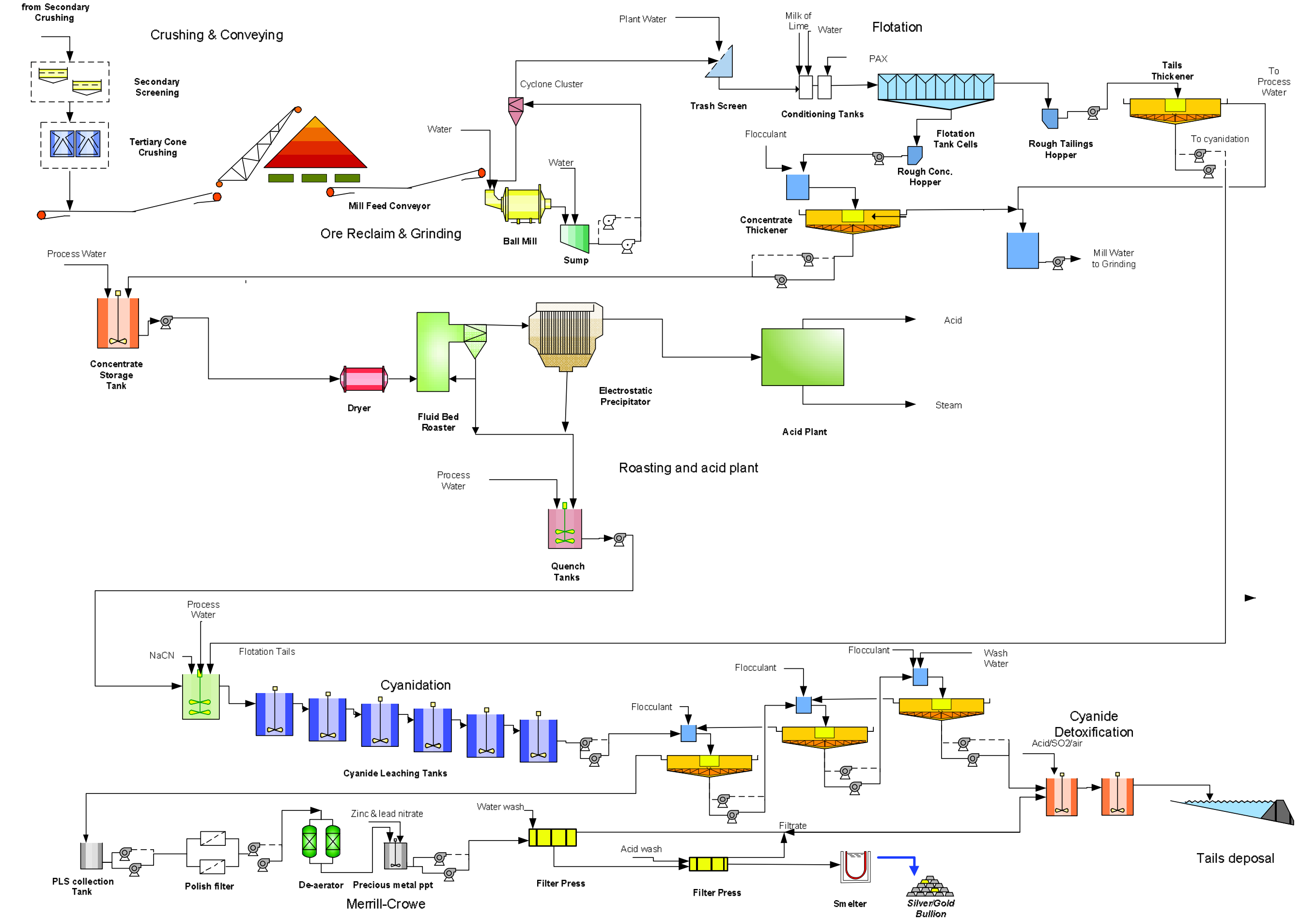


Figure 13: Roaster – Concentrate

CAPITAL AND OPERATING COST ESTIMATES FOR EACH PROCESS OPTION

Basis for Estimate

Specific factors that were considered in the economics of the process include the following:

- Concentrate grade
- Mineralogical composition
- Process route adopted
- Power costs
- Labour rates
- Transport rates
- Accessibility to existing infrastructure and land

These are broken down and taken into account in the capital and operating cost estimates, with similar methodology taken for both Ore A and B.

Capital Cost Estimate

The capital cost estimates have been defined as a "Scoping/conceptual Level Study". This corresponds to a Class 4 estimate in the Bateman guidelines and is deemed to have accuracy in the range of +/- 30%.

Vendor estimates were obtained for proprietary technologies. The bio-oxidation tank processes were sourced through Gold Fields Ltd and the Albion process through Xstrata. Capital costs for crushing, grinding, pressure autoclave, CCD, CIL, gold refinery, thickeners and tails detoxification plants were based on recently completed works of similar operational throughputs. Price escalation was factored in where applicable to allow for current costs. The Merrill Crowe plant was sourced from FLSmidth Salt Lake City.

Indirect costs were calculated based on erection of Temporary Facilities, First Fills and Lubricants Commissioning and Start-Up Spares, Engineering, Procurement and Construction Management and Contingency.

No allowance has been made for Infrastructure, Environmental studies, investigations, permits or liabilities, Costs of permits. Legal fees, Land acquisition or rights of way and any resettlement Costs.

Operating Cost Estimate

The operating cost estimates have been prepared to an accuracy of ± 40%. The operating labour costs were developed from the schedule of required labour prepared by Bateman, with costs for each labour resource based on typical wages expected for the regions.

The consumption rates of reagents and consumables were based on assumptions from test work data and predicted calculated values from mass balance models. Pricings including power were supplied by the clients. Reagent and power costs are listed in Table 6.

Table 6: Local Reagent and Power Costs Assumed

Local Reagents	Ore A	Ore B	Unit
Lime	130	121	$/t
Limestone	25	75	$/t
Sulphuric acid	390	125	$/t
Sodium cyanide	2660	2250	$/t
Power			
Power Cost	0.08	0.10	$/kWh

Power consumption was calculated based on unit processes. Maintenance costs for each plant area were estimated as a percentage of the total direct capital cost of mechanical equipment for that area. A contingency of 10% was included to cover unforeseen operating costs and is meant to cover the normal level of unknowns that are inherent in early design.

Comparative Cost Estimates for Ore A

A summary of plant design data, operating and capital costs for all six process treatment options are presented in Table 7. The capital and operating cost estimates for each of the process options is shown graphically in the following figures.

Table 7: Summary Comparison of selected Refractory Gold processes for Ore A

Summary Comparison of Selected Refractory Gold Processes for Ore A

OPTION		**1**	**2**	**3**	**4**	**5**	**6**
PROCESS		**Bacterial Heap Oxidation Whole Ore**	**UFM Con Concentrate**	**UFM Ore Whole Ore**	**Albion Concentrate**	**Bacterial Tank Leaching (BIOX) Concentrate**	**POX Concentrate**
Plant Design Data							
Ore Processed	t / yr	600,000	600,000	600,000	600,000	600,000	600,000
Concentrate Production	t / yr	0	54,000	0	54,000	54,000	54,000
Au in Feed	oz / yr	43,403	39,062	43,403	39,062	39,062	43,403
S in Feed	t / yr	7,401	6,942	7,401	6,942	6,942	6,942
CO3 in Process Feed	t / yr	71,946	1,080	71,946	1,080	1,080	1,080
Final Grind Size (P_{80})	µm	150	10	10	10	75	75
Sulphur Oxidation	%	60	2	2	80	80	95
Overall Recovery of Au	%	70	45	50	83	83	85
Acid produced	t/yr	13,599				1,057	1,057
Net Acid required by process	t/yr	56,859			1,200		
Power	kWh/t ore	43.9	31.6	67.5	41.7	74.8	96.6
Annual Operating Cost							
Labour	US $ million / annum	2.6	2.9	2.2	2.5	2.5	2.7
Reagents	US $ million / annum	28.2	3.9	5.4	3.8	3.7	4.5
Power	US $ million / annum	2.1	1.5	3.2	2.0	3.6	4.6
Sundry Other	US $ million / annum	3.2	1.5	1.7	2.0	2.0	3.0
Total	***US $ million / annum***	***36.1***	***9.8***	***12.5***	***10.3***	***11.9***	***14.9***
Contingency	US $ million / annum	3.6	1.0	1.2	1.0	1.2	1.5
Total Operating Cost	***US $ million / annum***	**39.7**	**10.8**	**13.7**	**11.3**	**13.0**	**16.4**
Capital Cost							
Direct Capital Cost - Front End and Gold Recovery	US $ million	20.32	28.45	29.38	31.67	27.00	32.17
Direct Capital Cost - Refractory	US $ million	15.42	3.50	7.80	5.00	12.67	14.91
Total Direct Capital Cost	**US $ million**	**35.74**	**31.95**	**37.18**	**36.67**	**39.67**	**47.07**
Total Indirect Capital Cost	**US $ million**	**15.55**	**13.90**	**16.17**	**15.95**	**17.26**	**20.48**
Total	**US $ million**	**51.3**	**45.9**	**53.3**	**52.6**	**56.9**	**67.6**
Cash Flow							
Gold Produced	oz / annum	30,382	17,578	21,701	32,344	32,344	36,719
Operating Cost Relative to Ounces of Au	**US $ / oz**	1,308	612	633	350	403	447

The whole ore Bacteria heap leach and ultrafine milling ore processes will treat 600,000 tpa feed whereas the other processes will treat 54,000 tpa flotation concentrate. The mass pull from the flotation circuit will be 9% by weight recovering 90% of the gold in the ore.

The final particle size of the bacteria heap will be coarser than the other processes for cyanidation. The ultrafine milling processes, including Albion, will be ground to 10 micrometre, while for bacteria tank and pressure oxidation a grind to 75 micrometres is considered.

For heap leaching only 60% of sulphide minerals is assumed to be oxidised due to percolation and surface passivation issues through the heap. While the ultrafine milling process is considered a physical reduction in particle size process some allowance has been made for surface oxidation. Both the Albion and bacteria tank leach processes sulphide oxidation can be controlled to 80% to allow for maximum gold recovery without excess oxidation. In pressure oxidation partial sulphide oxidation is not easy to control and 95% of the sulphide has been considered in this study.

Gold recoveries are estimated based on the limited test work on bio-oxidation work. No gold recovery in tails from the concentrate treatment processes have been considered in this study. The tails are expected to be too low grade and refractory to be treated economically. Hence overall gold recovery is based on the 90% recovered in the flotation circuit for processes treating concentrate. The gold recovered from pressure oxidation is expected to be 2% higher than bio-oxidation tank process.

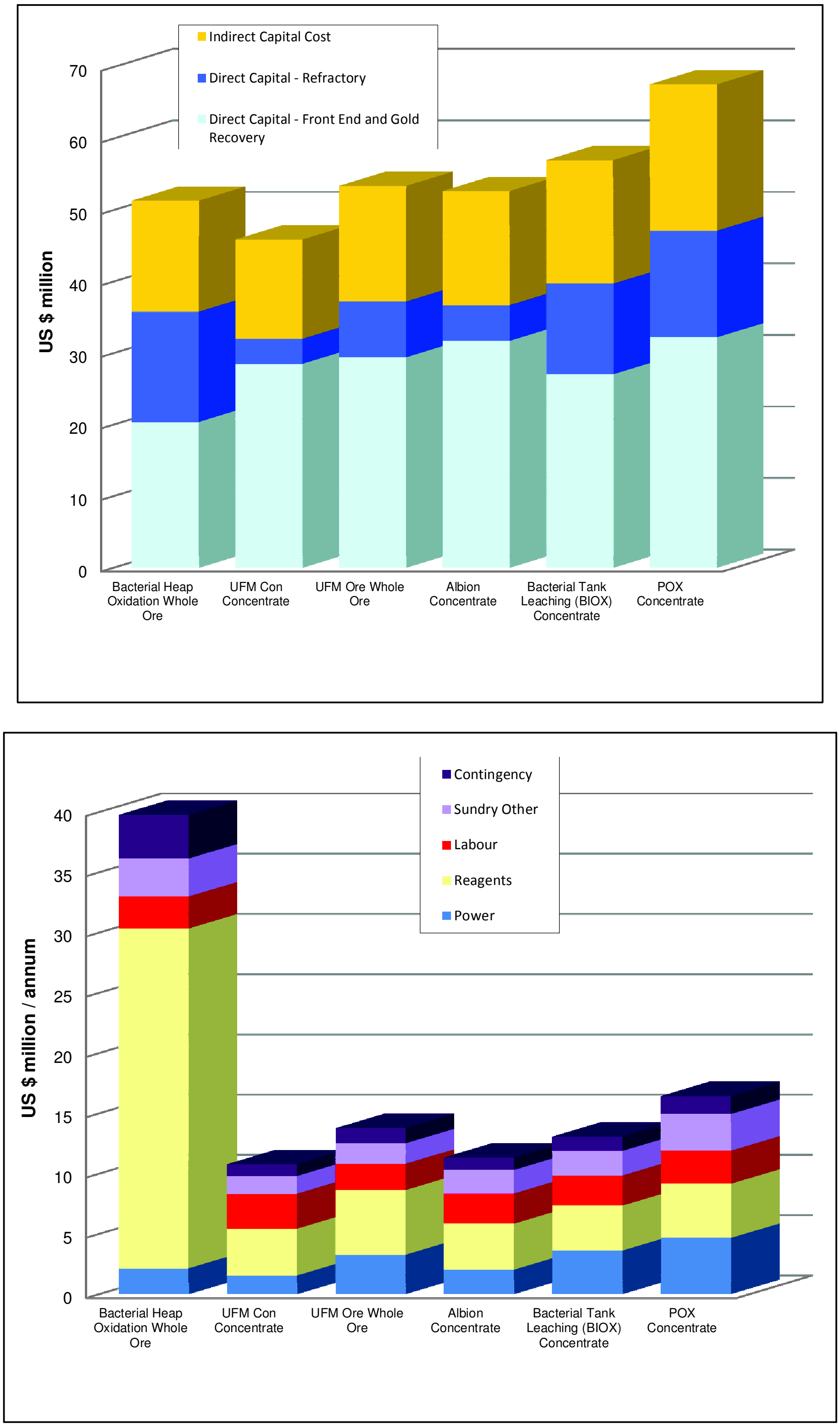


Figure 14: Capital (a) and Operating (b) costs shown graphically for Ore A

For Ore A the high operating costs for the bacteria heap leach is associated with high acid demand for neutralisation of the carbonate in the ore. Early test work data alluded to the requirement of either acid or excess pyrite being added to the heap to neutralise the carbonate component of the ore. Utilisation of the flotation tails (containing carbonate) as a neutralising material reduces the costs of limestone in the concentrate treatment processes. The bacteria tank leaching and pressure oxidation are slightly more energy intensive than the other processes.

A comparison of the operating cost per ounce of gold recovered versus process capital costs is presented in Figure 15. For the ultrafine milling processes to be cost effective in line with the other concentrate process options a gold recovery of 81.2% for concentrate or 78.4% for whole ore is required. Test work however will be required to validate actual gold recoveries.

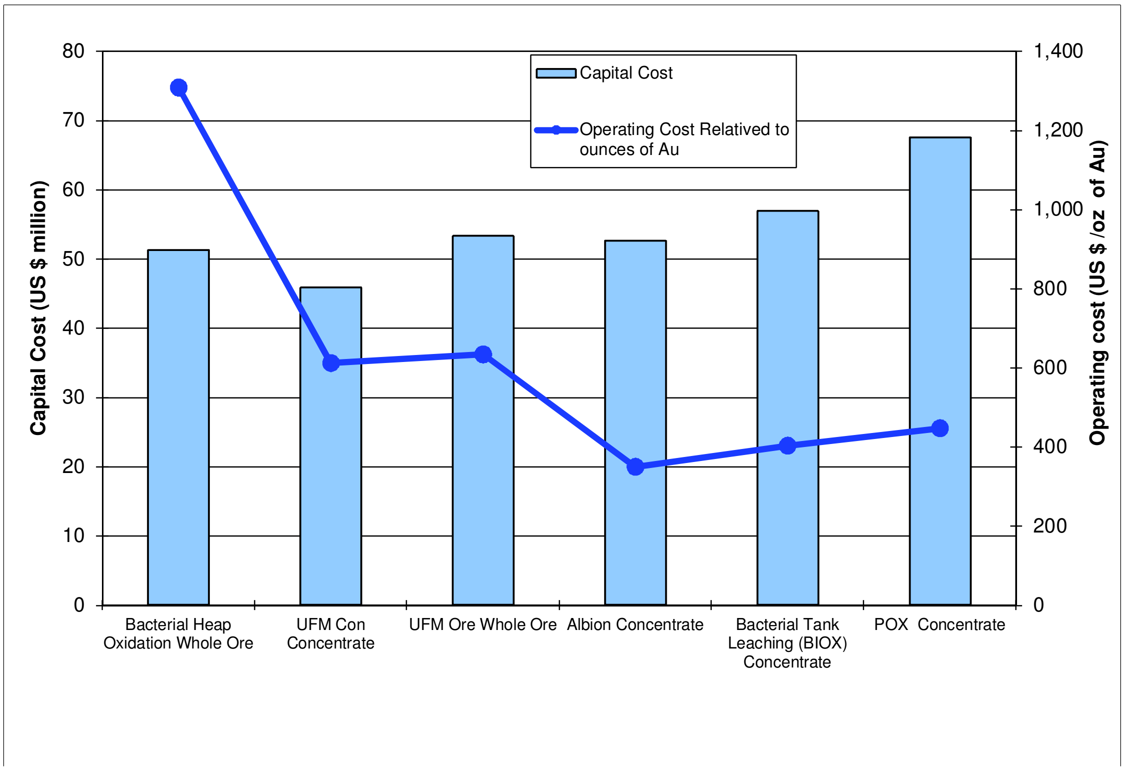


Figure 15: Comparison of conceptual Operating cost to ounces of gold produced versus capital costs for the six process options

Comparative Cost Estimates for Ore B

A summary of plant design data, operating and capital costs for all six process treatment options are presented in Table 8. The capital and operating cost estimates for each of the process options is shown graphically in the following figures.

Table 8: Summary Comparison of selected Refractory Gold processes for Ore B

Summary Comparison of Refractory Gold Processes for Ore B				
OPTION		**1**	**2**	**3**
PROCESS		**Air Roasting Concentrate**	**Bacterial Tank Leaching Concentrate**	**POX Concentrate**
Ore Processed	t / yr	3,650,000	3,650,000	3,650,000
Au in ore	oz / yr	124,388	124,388	124,388
Ag in ore	oz / yr	3,755,120	3,755,120	3,755,120
Concentrate Production	t / yr	1,095,000	1,095,000	1,095,000
Au in Conc	oz / yr	87,072	87,072	87,072
Ag in Conc	oz / yr	1,689,804	1,689,804	1,689,804
S in Conc	t / yr	202,985	202,985	202,985
Final Grind Size (P_{80})	µm	75	75	75
Sulphur Oxidation	%	95	95	100
Overall Recovery of Au	%	81.5	81.5	81.5
Overall Recovery of Ag	%	62.0	62.0	62.0
Acid produced	t/yr	590,558	590,558	621,640
Power	kWh/t ore	56.4	115.1	84.7
Annual Operating Cost				
Labour	US $ million / annum	4.5	4.0	4.1
Reagents	US $ million / annum	12.5	79.2	75.9
Power	US $ million / annum	19.9	42.0	30.9
Sundry Other	US $ million / annum	36.3	33.6	38.2
Total	***US $ million / annum***	***73.1***	***158.8***	***149.2***
Contingency	US $ million / annum	7.6	15.9	14.9
Total Operating Cost	***US $ million / annum***	**80.8**	**174.7**	**164.1**
Capital Cost				
Direct Capital Cost - Front & End	US $ million	115.7	119.4	135.8
Direct Capital Cost - Refractory	US $ million	124.4	136.0	136.6
Direct Capital Cost	**US $ million**	**240.0**	**255.4**	**272.4**
Indirect Capital Cost	**US $ million**	**108.3**	**115.4**	**122.8**
Total	**US $ million**	**348.3**	**370.8**	**395.2**
Revenue				
Gold Produced	oz / annum	101,377	101,377	101,377
Silver Produced	oz / annum	2,326,297	2,326,297	2,326,297
Operating Cost Relatived to Ounces of Precious Metals Produced #				
	US $ /oz of Au equivalent (with Ag)	**580**	**1,205**	**1,132**
#Silver value converted to Au equivalent based on ratio of current price of Au and Ag				

All three process options considered will treat 1,095,000 tpa of flotation concentrate. Based on the limited test work data the mass pull from the flotation circuit will be 30% by weight recovering 70% of the gold and 45% of the silver into a flotation concentrate. The oxidised residues from the refractory circuit will then be combined with the flotation tails for cyanidation.

It is assumed that gold and silver extraction on oxidised concentrates will be 95% and 90% respectively for all three process options. Gold and silver recoveries from the flotation tails will be 50% and 39% respectively as observed in the test work. This yields overall gold and silver recovery of 81.5% and 62.0% respectively. Sulphide oxidation is assumed to be 95% except for pressure oxidation which is set at 100%.

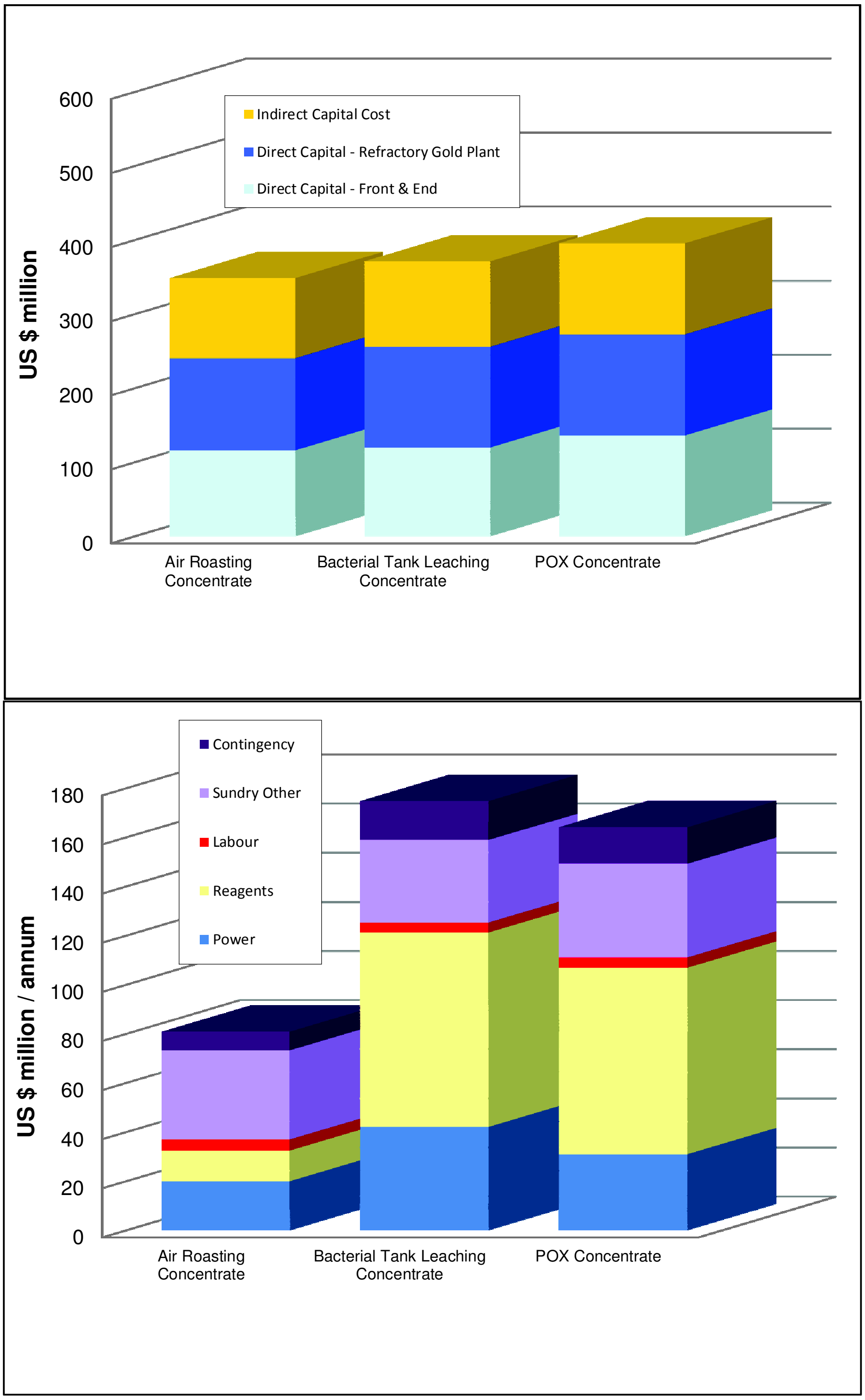
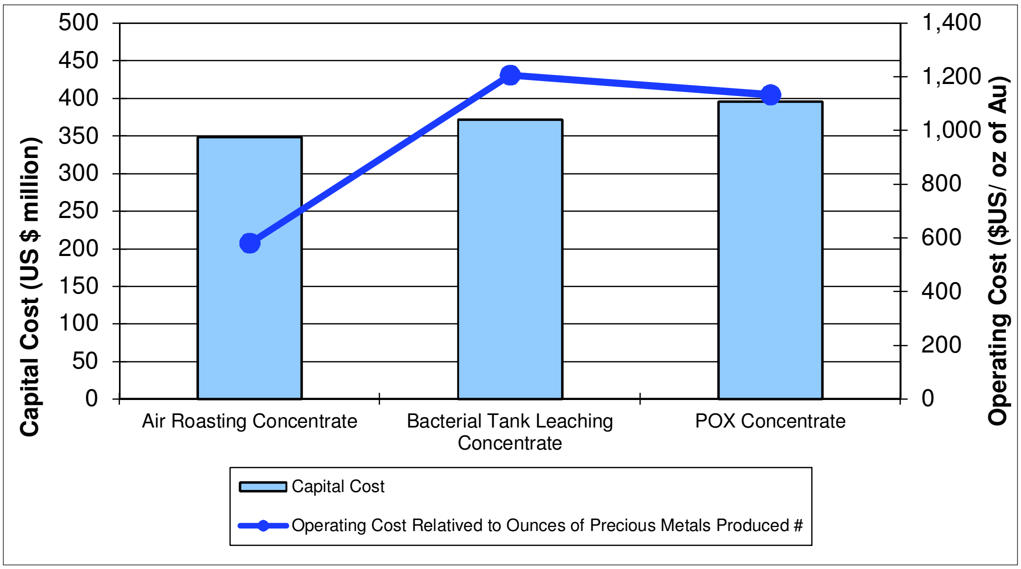


Figure 16: Capital (a) and Operating (b) Cost Estimate for Ore B

For ore B the operating costs for both bacterial tank leaching and pressure oxidation is approximately double the roasting costs. The bacterial tank leaching and pressure oxidation are more energy intensive than roasting and incur high reagent costs associated with limestone use for neutralisation.

To allow comparison between the three refractory process costs the total operating costs were expressed in terms of operating cost per ounce of equivalent gold produced taking into account silver. The ounces of silver produced were converted to ounces of equivalent gold by ratio of the average price of gold at $US 1600/oz and silver at $US 30/oz. A comparison of operating costs per ounce of gold versus process capital costs is presented in Figure 17. The operating costs are significantly high and higher grade ores would appear to be required to make this ore body profitable and sustainable. The capital costs of all three process options are not significantly different with roasting having the lowest cost. The capital cost comparisons are consistent with early studies(6).



.# Silver value converted to Au equivalent based on ratio of current price of Au and Ag

Figure 17: Comparison of conceptual Operating cost to ounces of equivalent gold produced (Au + Ag) versus capital costs for the three process options

Roasting is predicted to provide the lowest operating cost per ounce of gold followed by pressure oxidation. In the roasting option if the acid produced is neutralised this increases the overall operating cost from $US80.8M to $US 128.5M. The resultant loss of revenue from the acid selling increases the operating cost per ounce of gold to $909M, which is still predicted to be marginally better than pressure oxidation at $US1132M. However the sulphide sulphur in the concentrate (19%) is considered on the borderline to run a roaster autogenously without some extra heat source. Hence, sulphide grades need to be optimised during flotation to increase sulphide grade. Alternatively to operate the roaster at lower sulphide content the feed will have to be further dried to remove moisture and would incur extra capital and operating costs. Biooxidation option provides the highest expenditure and this is largely due to the high power costs and limestone usage.

Test work is required to determine what actual precious metal recoveries are expected for the different process options, and the degree of sulphide oxidation required to liberate and extract the gold and silver metals from the concentrate.

For the roasting option it is assumed that the arsenic level in the feed is low based on the sample on which flotation tests were conducted and will be captured in the gas and subsequently fixed with the calcine product. Arsenic can be fixed as ferric arsenate in the calcine during roasting but generally results in lower recovery of gold (and silver) during the subsequent cyanidation step. To treat high arsenic content concentrates may require a two stage roasting process with an approximate 25% increase in capital cost. The roasting option assumes the acid can be sold at $US 125/t.

In bacterial tank leach processes sulphide oxidation can be controlled to allow for maximum gold and silver recovery without excess oxidation. Reducing the oxidation requirement would significantly improve the operating costs. In roasting and pressure oxidation partial sulphide oxidation is not easy to control.

Extend of silver being liberated is sensitive to the amount of lime added in the lime boil stage (16). Hence the degree of recovery of silver from pretreatment processes would need to be tested.

Sensitivity

Using the process cost analyses model developed for the study the sensitivities of power, and limestone cost to the overall process costs were examined for Ore B. The base case values (Power $0.10 kWh, Limestone $75/t) were varied separately up to ±50% and the operating costs to ounces of gold produced calculated. The results of these analyses for all three process options are presented in Figure 18. In addition the impact of the variability of gold and silver extraction expected from cyanide leaching the oxidised concentrate separately on operating margin was assessed. The gold and silver extraction from the tails was left fixed at the test work data values. The variability of gold and silver extraction results are presented in Figure 19.

Overall limestone cost has no impact on roasting unless the acid produced is neutralised. In which case the operating costs for the roaster is closer to the pressure oxidation and biooxidation processes. Limestone cost has more of an impact than power cost for pressure oxidation. Whereas for biooxidation power and limestone costs yield similar trends. The availability and cost of limestone will have a major impact on the viability of pressure oxidation and biooxidation process.

As expected variation in the gold extraction in the refractory processes have significant influence on operating cost margin. Roasting has been reported in some studies to yield lower gold recoveries than pressure oxidation and biooxidation(17). Even for allowing a decrease in gold extraction by 10% (-13% from base case) the roaster option still yields lower operating cost per ounce of gold than the other two process options. From experience it may be expected that gold recovery may be slightly better (~2%) for pressure oxidation than bio-oxidation.

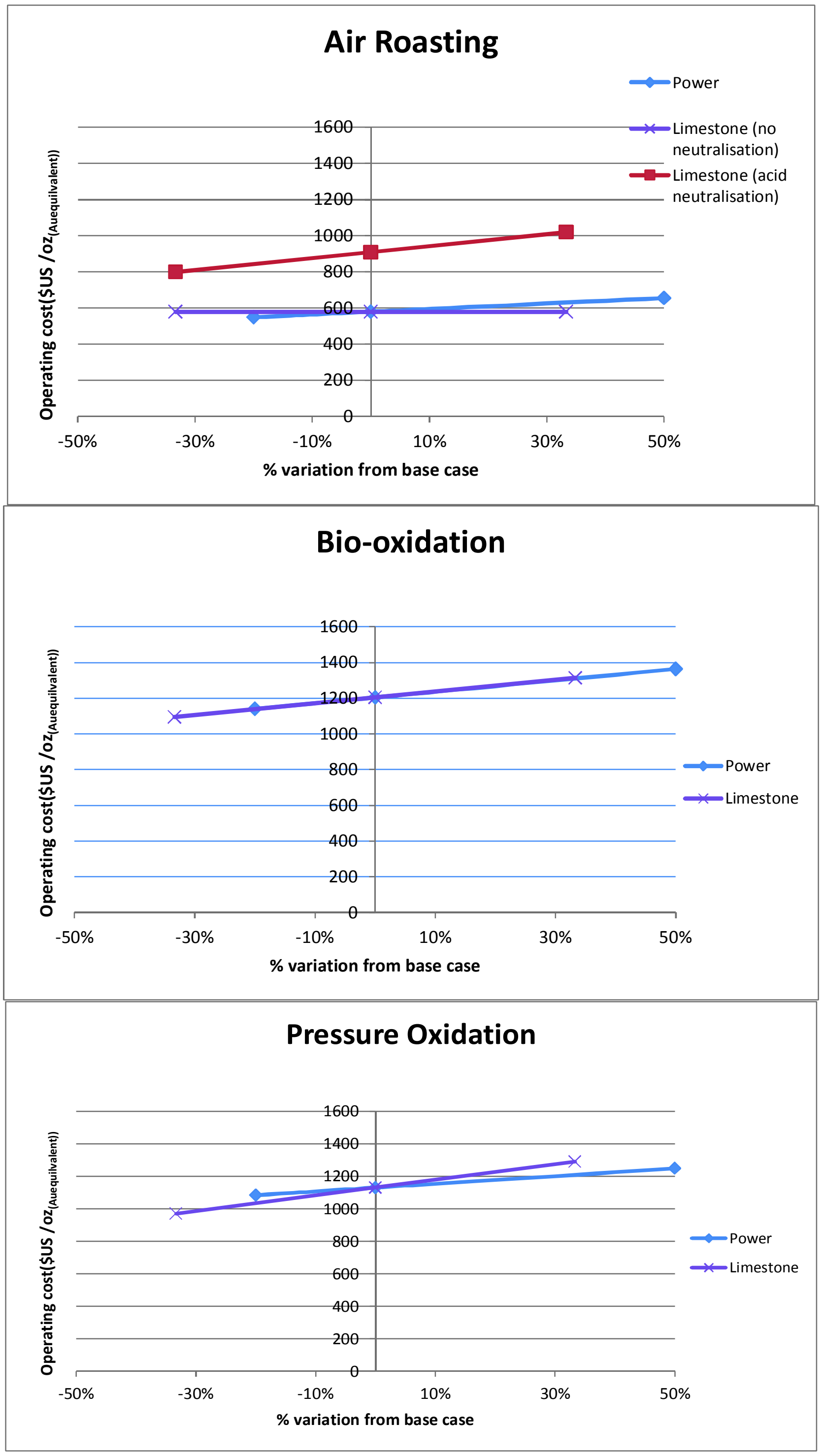


Figure 18: Sensitivity analyses of limestone and power costs for three process options (base case, Power $0.10 kWh, Limestone $75/t)

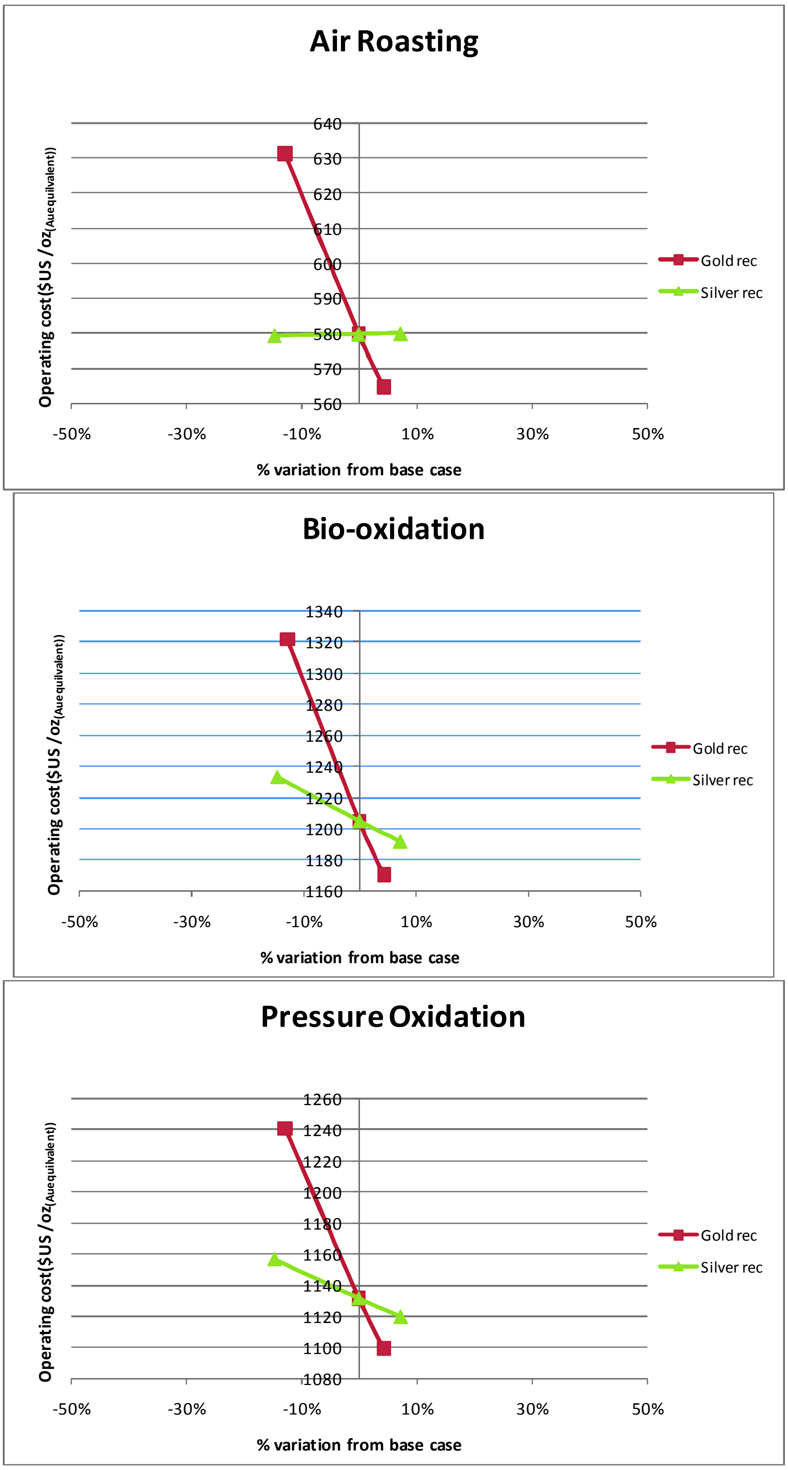


Figure 19: Sensitivity analyses of gold and silver recovery for all three process options (assuming gold and silver extraction variability from oxidised concentrate only, base case Au 95% and Ag 90% in extraction)

SUMMARY AND CONCLUSIONS

Conceptual level studies looking at various process treatment routes were carried out on two different ore bodies. The gold in both ore deposits is locked in arsenic bearing sulfide minerals and require a pre-treatment process prior to recovery.

Five different process options were examined for treating Ore A which has high arsenic and carbonate content at 0.6 Mtpa. The high carbonate content will make heap bio oxidation leach option not viable. The capital cost for alkaline Albion, bacteria tank leaching and POX were not significantly different.

The three conventional technologies were considered for treating the high silver Ore B at 3.65 Mtpa. The significantly high operating costs associated with the low gold to sulphur ratio will make treating Ore B a challenge. The capital costs of all three processes are not significantly different. Roasting operating costs appeared about half that of biooxidation and pressure oxidation. However air roasting assumes acid generated from SO_2 can be sold. In addition it is assumed the arsenic in the concentrate can be contained and stabilised within the roasted calcine.

These estimates were produced to provide the clients with sufficiently detailed options to assess the viability of the project. Further geometallurgical & gold deportment studies are required on both ore bodies. Extensive test work programme on all options is required before further engineering studies can be carried out..

ACKNOWLEDGEMENTS

Permission of two mining companies to present data from studies is greatly appreciated. Information provided by Goldfields and Xstrata is also greatly appreciated.

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