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A TECHNO-ECONOMIC COMPARISON OF PYROMETALLURGICAL AND HYDROMETALLURGICAL OPTIONS FOR TREATING HIGH-ARSENIC COPPER CONCENTRATES

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ABSTRACT

As the high-grade copper deposits in the world become depleted, there is an increasing focus on developing lower-grade ores which contain problematic impurities such as arsenic. Indeed, many high-arsenic ore bodies remain undeveloped because of their high arsenic levels. Separation of arsenic from copper at the beneficiation stage is difficult using the current processing techniques and invariably copper concentrates produced from these ores contain high arsenic levels, sometimes so high as to make them unacceptable to copper smelters. New, sustainable approaches are needed to allow development of high-arsenic copper ores while at the same time minimizing arsenic dispersion into the biosphere.

The most promising route for separating arsenic from copper at the flotation stage centers on the use of pulp potential control. In recent years CSIRO has demonstrated at laboratory scale, for a number of high-arsenic copper ores, that crude flotation separation between arsenic-bearing minerals and copper-bearing minerals can be made. The differences in specific minerals responses to changes in pulp potential and pulp pH are exploited for this separation. Such separations nominally produce a low-arsenic high-copper concentrate and a low-copper high-arsenic concentrate. The former can be smelted without penalty, while the latter requires further treatment to capture and stabilize the arsenic and to recover the residual copper.

Consideration has now been given to treatment options for the high-arsenic low-copper product. A pyrometallurgical option, involving a low temperature roast to selectively fume off the arsenic and leave a high-copper calcine, coupled with an arsenic stabilization step, has been demonstrated in the laboratory, and a preliminary techno-economic evaluation shows the process to be economically favorable. These results were recently published. CSIRO has now investigated an alternate hydrometallurgical option, involving atmospheric alkaline sulphide leaching of the high-arsenic product, where the arsenic is selectively dissolved leaving a copper-rich leach residue. The leaching testwork was carried at on the same high-arsenic low-copper material used in the roasting studies, enabling a techno-economic comparison of the alternate roasting and leaching treatment options to be made. The alkaline leaching work is briefly described and the results of the process option comparison presented and discussed.

Keywords: arsenic, copper, economics, enargite, ore flotation, tennantite

INTRODUCTION

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About 90% of world copper production is produced pyrometallurgically from sulphide-based copper minerals, with the remainder from oxide-based copper minerals (Ayres *et al,* 2002, Davenport *et al*, 2002). Among the copper sulphide minerals, the main one is chalcopyrite ($CuFeS_2$) comprising 50% of total copper production (CDAA, 2012). However, often present with chalcopyrite in copper orebodies are sulphide-based copper minerals which also contain arsenic. These include enargite (Cu_3AsS_4) and tennantite ($Cu_{12}As_4S_{13}$). Their presence is problematic as these minerals are generally concentrated with the sulphide copper minerals during flotation leading to high-arsenic copper concentrates, destined for smelting. The current general industry practice is that copper smelters only accept concentrates with less than 0.2% (i.e. 2000 ppm) arsenic without payment of any penalty by the seller of the concentrate (Bruckard *et al*, 2010). If the arsenic content is between 0.2 to 0.5%, some penalty is imposed on the price received for the concentrate. Concentrates with more than 0.5% arsenic content are not generally accepted by the smelters. A simplified typical flowsheet for the current treatment of high-arsenic copper ores is given in Figure 1.

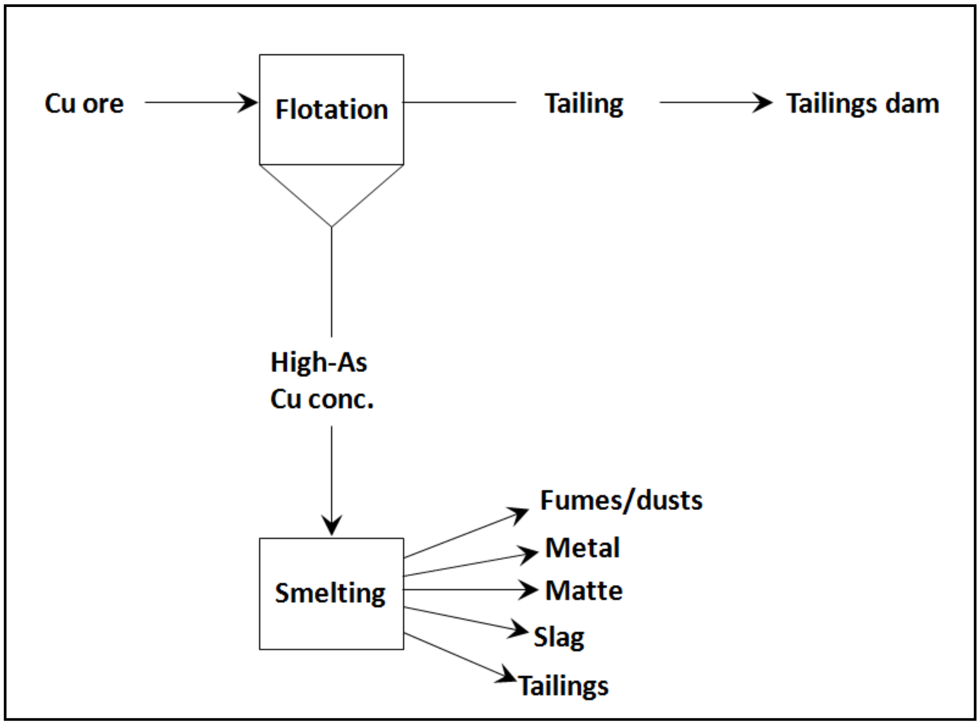


Figure 1. Basecase flowsheet for treatment of high-arsenic copper ores

Recently, a new early arsenic removal process for treating high-arsenic copper ores was developed by CSIRO (Jahanshahi *et al*, 2006; Senior *et al*, 2006). It comprises three main stages: a flotation stage where arsenic and copper minerals are separated into high-arsenic low-copper concentrate and a low-arsenic high-copper concentrate, based on exploitation of differences in the specific minerals responses to changes in pulp potential and pulp pH; a roasting stage where the high-arsenic low-copper concentrate produced in the flotation stage is treated to remove arsenic to a fume leaving a high-copper calcine; and a stabilization stage where the high-arsenic fume is treated to produce a stable low-volume stream that can safely be disposed of. It was demonstrated that removed arsenic in the form of As_2O_3 can be immobilized in high pH low temperature ceramics (LTC). X-ray diffraction data for the cured ceramics indicated the presence of anhydrous sodium arsenite ($NaAsO_2$), meaning the arsenic did not move into the LTC microstructure but persisted as aqueous sodium arsenite in the pore-water of the matrix before and after drying (Brew and Vance, 2008). A simplified flowsheet for the early removal process with roasting option is shown in Figure 2. A techno-economic evaluation of this new process has been published (Haque and Norgate, 2008; Haque *et al*, 2010). A detail description of the processing and stabilization methods have been given elsewhere (Jorgensen et al, 2007; Brew and Vance, 2008; Bruckard *et al*, 2010).

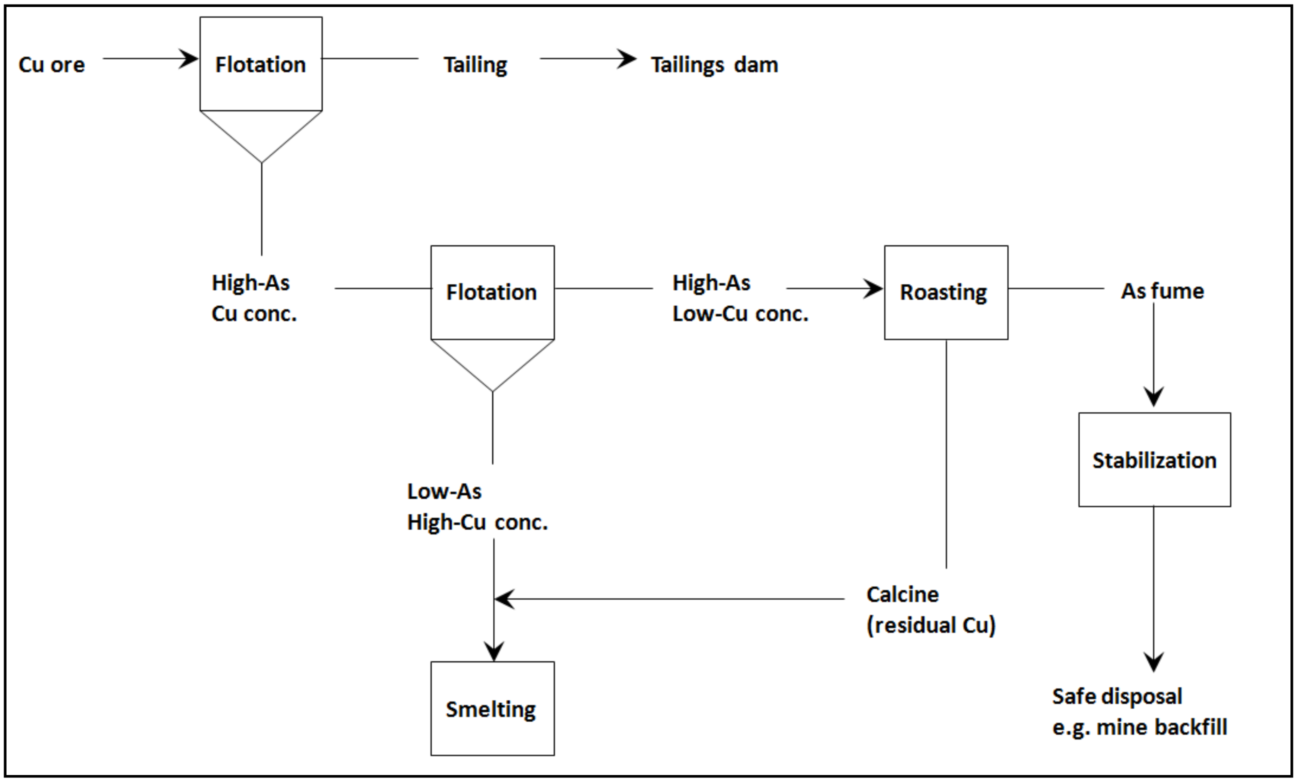


Figure 2. Early arsenic removal flowsheet for treatment of high-arsenic copper ores (roasting scenario)

An alternative option to the roasting step in the early arsenic removal process is a scenario where the roasting step is replaced by a hydrometallurgical step. In this case, the high-arsenic low-copper concentrate is leached with alkaline sulphide (a mixture of NaOH and Na_2S) to selectively dissolve the arsenic from the arsenic-bearing mineral matrix, leaving behind a copper-rich leach residue (Cuevas *et al*, 2011). A simplified flowsheet is shown in Figure 3. The alkaline sulphide leaching system was selected for treatment after the merits of various leaching systems for high-arsenic concentrates were determined (Cuevas *et al,* 2011). With regard to this leaching system:

- Test work was carried out on same high-arsenic copper concentrate as used in roasting option.
- The best leach condition was found to be 100 g/L Na_2S, 50 g/L NaOH, 100°C, for 1 h, and with 34% solids in the feed - this yielded an arsenic extraction of 98% with only 0.5% copper dissolved.
- The reaction in leaching tank is as follows

  $2\ Cu_{10}Fe_2As_4S_{13} = 4\ Cu_5FeS_4 + 8\ As + 5\ S_2$
- The reaction efficiency in removing arsenic was assumed to be 98% based on the laboratory experimental observation referred to above.

A techno-economic evaluation has been undertaken on the leaching-based early arsenic removal process in the present study, with an update of the costs for the previous conventional and pyrometallurgical options. This paper reports the results of this techno-economic comparison of the pyrometallurgical process using roasting with the new alkaline leaching process along with this update of the conventional base case.

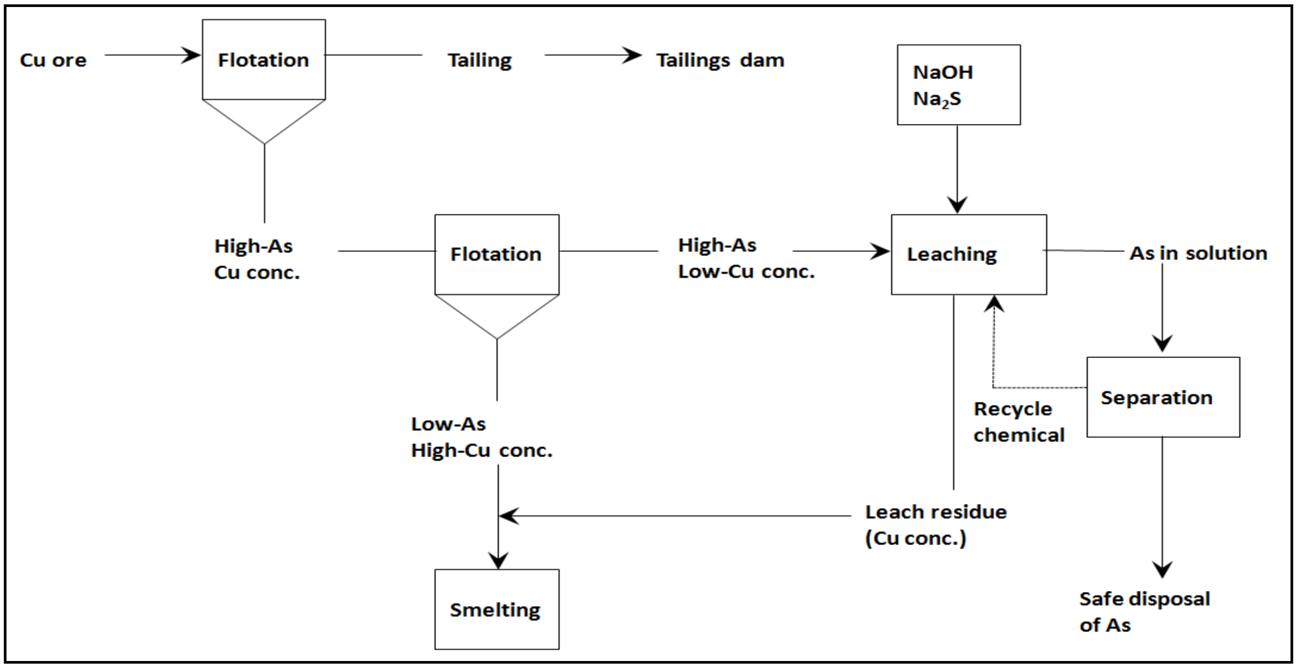
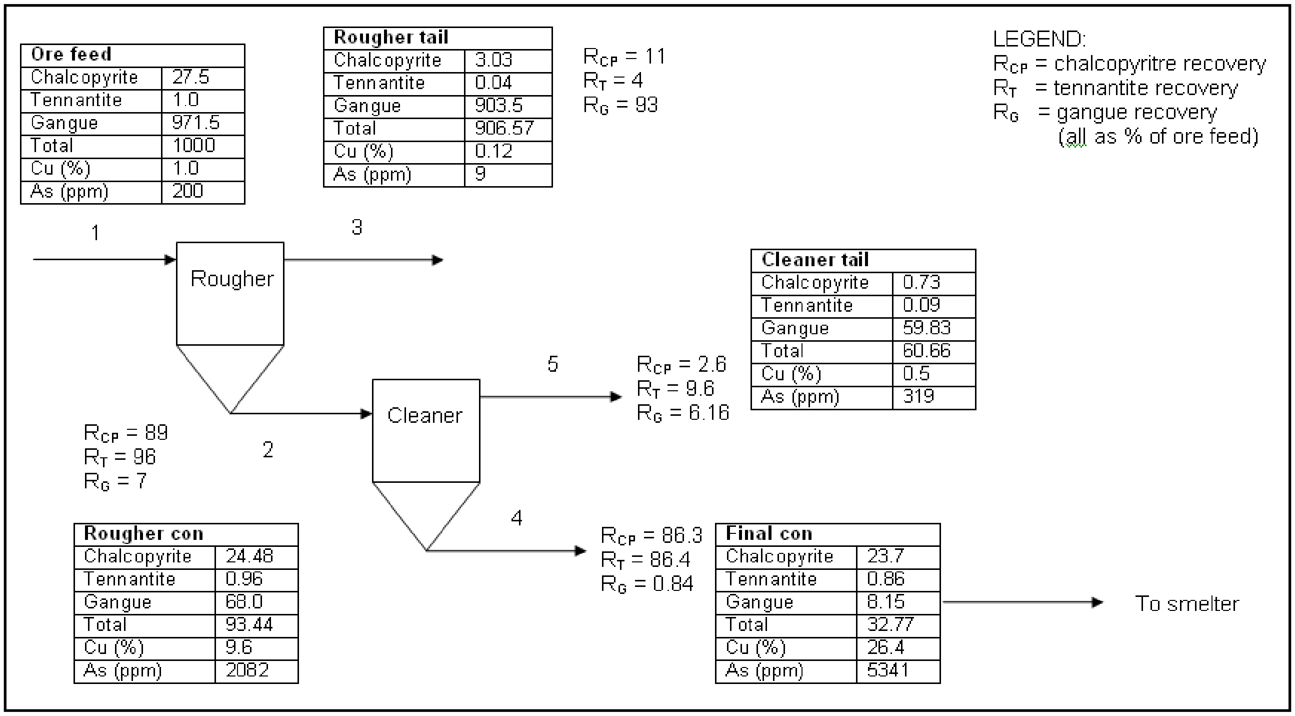


Figure 3. Early arsenic removal flowsheet for treatment of high-arsenic copper ores (leaching scenario with recycling of chemicals)

PROCESS MODELLING

Process modelling of the conventional (base case) flowsheet and the new proposed flowsheets was undertaken using the METSIM process simulation software using the specified mass flow rates and compositions of each stream. The base case process flowsheet showing component mass flowrates is given in Figure 4.



Ore feed	
Chalcopyrite	27.5
Tennantite	1.0
Gangue	971.5
Total	1000
Cu (%)	1.0
As (ppm)	200

Rougher tail	
Chalcopyrite	3.03
Tennantite	0.04
Gangue	903.5
Total	906.57
Cu (%)	0.12
As (ppm)	9

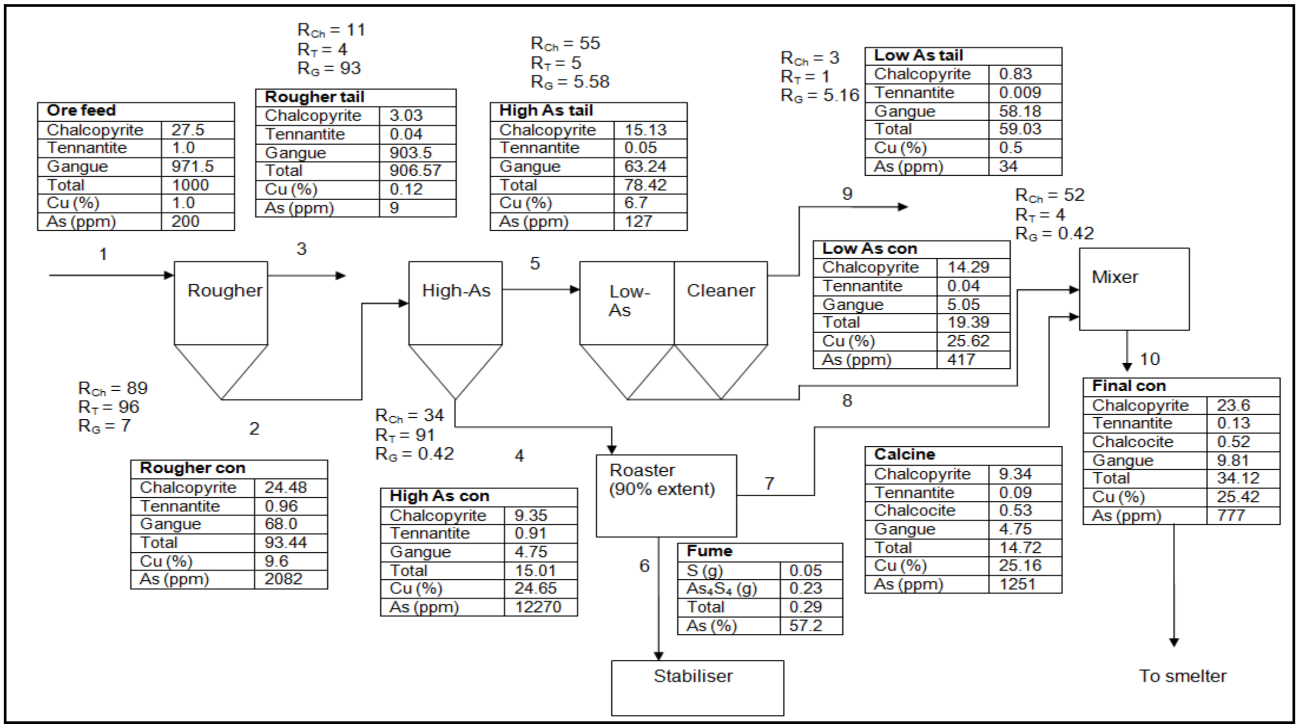
Cleaner tail	
Chalcopyrite	0.73
Tennantite	0.09
Gangue	59.83
Total	60.66
Cu (%)	0.5
As (ppm)	319

Rougher con	
Chalcopyrite	24.48
Tennantite	0.96
Gangue	68.0
Total	93.44
Cu (%)	9.6
As (ppm)	2082

Final con	
Chalcopyrite	23.7
Tennantite	0.86
Gangue	8.15
Total	32.77
Cu (%)	26.4
As (ppm)	5341

Figure 4. Base case flowsheet showing component mass flow rates (t/h)

The corresponding flowsheets for the proposed roasting-based flowsheet (pyrometallurgical) and the alternative leaching-based (hydrometallurgical) flowsheet are shown in Figure 5 and 6, respectively.



Ore feed	
Chalcopyrite	27.5
Tennantite	1.0
Gangue	971.5
Total	1000
Cu (%)	1.0
As (ppm)	200

Rougher tail	
Chalcopyrite	3.03
Tennantite	0.04
Gangue	903.5
Total	906.57
Cu (%)	0.12
As (ppm)	9

High As tail	
Chalcopyrite	15.13
Tennantite	0.05
Gangue	63.24
Total	78.42
Cu (%)	6.7
As (ppm)	127

Low As tail	
Chalcopyrite	0.83
Tennantite	0.009
Gangue	58.18
Total	59.03
Cu (%)	0.5
As (ppm)	34

Low As con	
Chalcopyrite	14.29
Tennantite	0.04
Gangue	5.05
Total	19.39
Cu (%)	25.62
As (ppm)	417

Rougher con	
Chalcopyrite	24.48
Tennantite	0.96
Gangue	68.0
Total	93.44
Cu (%)	9.6
As (ppm)	2082

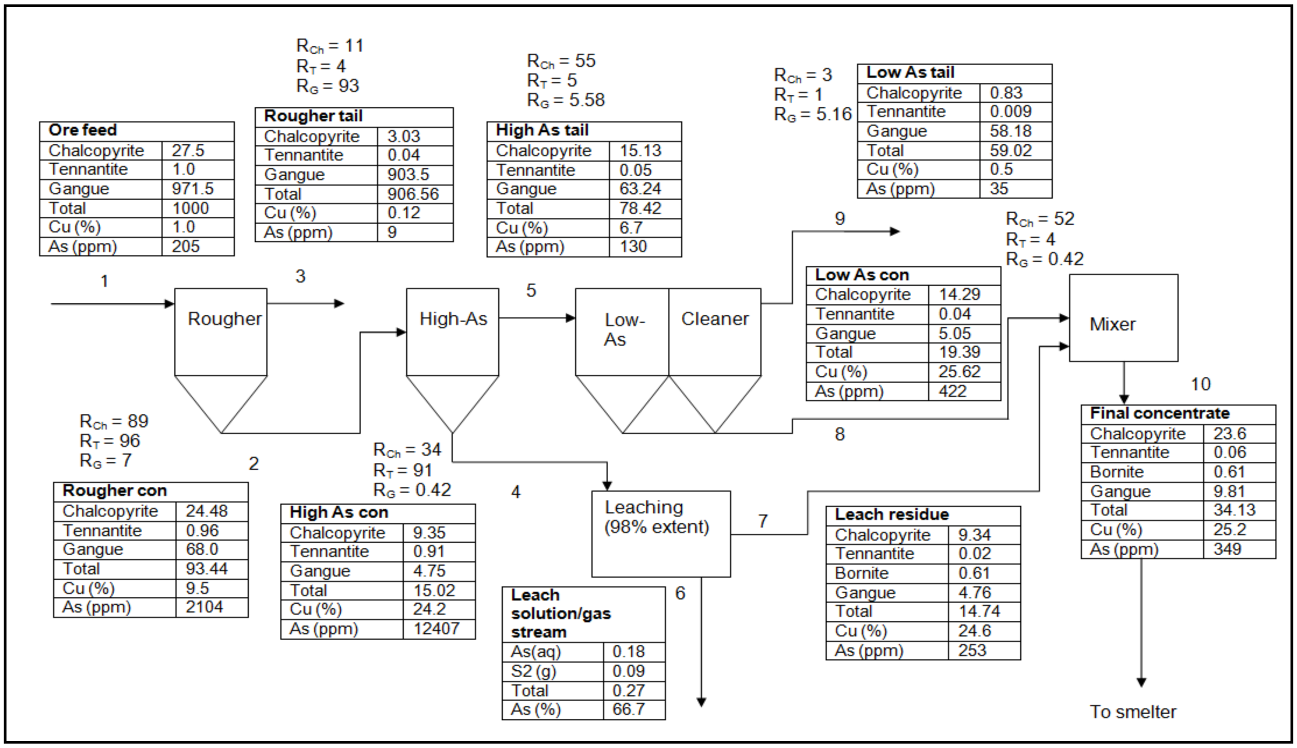
High As con	
Chalcopyrite	9.35
Tennantite	0.91
Gangue	4.75
Total	15.01
Cu (%)	24.65
As (ppm)	12270

Fume	
S (g)	0.05
As_4S_4 (g)	0.23
Total	0.29
As (%)	57.2

Calcine	
Chalcopyrite	9.34
Tennantite	0.09
Chalcocite	0.53
Gangue	4.75
Total	14.72
Cu (%)	25.16
As (ppm)	1251

Final con	
Chalcopyrite	23.6
Tennantite	0.13
Chalcocite	0.52
Gangue	9.81
Total	34.12
Cu (%)	25.42
As (ppm)	777

Figure 5. Roasting-based flowsheet showing component mass flow rates (t/h)



Ore feed	
Chalcopyrite	27.5
Tennantite	1.0
Gangue	971.5
Total	1000
Cu (%)	1.0
As (ppm)	205

Rougher tail	
Chalcopyrite	3.03
Tennantite	0.04
Gangue	903.5
Total	906.56
Cu (%)	0.12
As (ppm)	9

High As tail	
Chalcopyrite	15.13
Tennantite	0.05
Gangue	63.24
Total	78.42
Cu (%)	6.7
As (ppm)	130

Low As tail	
Chalcopyrite	0.83
Tennantite	0.009
Gangue	58.18
Total	59.02
Cu (%)	0.5
As (ppm)	35

Low As con	
Chalcopyrite	14.29
Tennantite	0.04
Gangue	5.05
Total	19.39
Cu (%)	25.62
As (ppm)	422

Rougher con	
Chalcopyrite	24.48
Tennantite	0.96
Gangue	68.0
Total	93.44
Cu (%)	9.5
As (ppm)	2104

High As con	
Chalcopyrite	9.35
Tennantite	0.91
Gangue	4.75
Total	15.02
Cu (%)	24.2
As (ppm)	12407

Leach solution/gas stream	
As(aq)	0.18
S2 (g)	0.09
Total	0.27
As (%)	66.7

Leach residue	
Chalcopyrite	9.34
Tennantite	0.02
Bornite	0.61
Gangue	4.76
Total	14.74
Cu (%)	24.6
As (ppm)	253

Final concentrate	
Chalcopyrite	23.6
Tennantite	0.06
Bornite	0.61
Gangue	9.81
Total	34.13
Cu (%)	25.2
As (ppm)	349

Figure 6. Leaching-based flowsheet showing component mass flow rates (t/h)

In carrying out the process modelling, the ore feed rate was assumed to be 1000 t/h (Stream 1 in Figures 4-6) with the mineral composition of this stream given in Table 1. This typical ore composition corresponds to a copper grade of 1.0% and an arsenic content of 0.02% (200 ppm). Assumptions made for all flowsheets are shown in Table 2.

Table 1. Mineral composition of ore feed (1% Cu, 200 ppm As)

Mineral	Weight %
Quartz (SiO_2)	50.6
Pyrite (FeS_2)	46.6
Chalcopyrite ($CuFeS_2$)	2.8
Tennantite ($Cu_{12}As_4S_{13}$) or ($Cu_{10}Fe_2As_4S_{13}$)	0.1
Total	100

Table 2. Assumptions for base case and proposed flowsheets

Stage	Base case flowsheet	Proposed flowsheet
Stream 1	1000 t/h	1000 t/h
Rougher flotation stage	Recoveries: chalcopyrite 89% tennantite 96% gangue 7%	Recoveries: chalcopyrite 89% tennantite 96% gangue 7%
High-As flotation stage	Not applicable	Recoveries: chalcopyrite 34% tennantite 91% gangue 1%
Low-As/cleaner flotation stage	Recoveries: chalcopyrite 86% tennantite 86% gangue 6%	Recoveries: chalcopyrite 52% tennantite 4% gangue 1%
Roasting stage (refer Figure 5)	Not applicable	Reaction (Padilla *et al*, 2001) $2\ Cu_{12}As_4S_{13} =$ $12\ Cu_2S + 2\ As_4S_4 + 3\ S_2$ Reaction extent is 90%
Leaching (refer Figure 6)	Not applicable	Reaction (Cuevas *et al*, 2011) $2\ Cu_{10}Fe_2As_4S_{13} =$ $4\ Cu_5FeS_4 + 8\ As + 5\ S_2$ Reaction extent is 98%
Mixing stage	Not applicable	Calcine product from roaster or leach residue is mixed with low-As high-Cu concentrate

Modelling results

The mass flowrates (including the main mineral components) together with the copper and arsenic contents of all process streams are given in Figure 4 for the conventional (base case) flowsheet and Figures 5 and 6 for new roasting and leaching flowsheets, respectively. Figure 4 shows the copper concentrate produced (Stream 4) had a copper grade of 26.4% and an arsenic content of 5341 ppm, which for the purpose of the study was considered to be just within acceptable limits for smelters. Figure 5 shows that the primary chalcopyrite concentrate (Stream 8) had a copper grade of 25.6% and an arsenic content of 417 ppm. After mixing with the calcine product from the roaster (Stream 7), the final concentrate (Stream 10) had a copper grade of 25.4% and an arsenic content of 777 ppm, well below the limit (2000 ppm) for the imposition of an arsenic penalty. Similarly, for leaching option Figure 6 shows that the primary chalcopyrite concentrate (Stream 8) had a copper grade of 25.6% and an arsenic content of 422 ppm. The final concentrate (stream 10) in Figure 6 had a copper grade of 25.2% with only 349 ppm of arsenic.

Copper and arsenic balances for all three flowsheets are given in Tables 3, 4 and 5, along with overall copper and arsenic recovery to the final concentrate.

Table 3. Copper and arsenic balances for base case flowsheet

IN (t/h)		**OUT (t/h)**	
Copper Ore feed (1)	10.037	Rougher tail (3)	1.085
		Cleaner tail (5)	0.302
		Final concentrate (4)	8.650
			10.037
Overall copper recovery to concentrate = 86.2%			
Arsenic Ore feed (1)	0.203	Rougher tail (3)	0.008
		Cleaner tail (5)	0.019
		Final concentrate (4)	0.176
			0.203
Overall arsenic recovery to concentrate = 86.2%			

Table 4. Copper and arsenic balances for proposed flowsheet (roasting scenario)

IN (t/h)		**OUT (t/h)**	
Copper Ore feed (1)	10.037	Rougher tail (3)	1.068
		Low As tail (9)	0.293
		Fume (6)	0.000
		Final concentrate (12)	8.676
			10.037
Overall copper recovery to concentrate = 86.4%			
Arsenic Ore feed (1)	0.203	Rougher tail (3)	0.008
		Low As tail (9)	0.002
		Fume (6)	0.166
		Final concentrate (12)	0.027
			0.203
Overall arsenic recovery to concentrate = 13.3% Arsenic recovery in roaster (from high-As concentrate to fume) = 90.2%			

Table 5. Copper and arsenic balances for proposed flowsheet (leaching-based scenario)

IN (t/h)		**OUT (t/h)**	
Copper Ore feed (1)	10.037	Rougher tail (3) Low As tail (9) Leach residue (7) Low As con (10)	1.064 0.292 3.632 4.966 10.037
Overall copper recovery to concentrate = 86.3%			
Arsenic Ore feed (1)	0.203	Rougher tail (3) Low As tail (9) Leach solution (6) Final concentrate (10)	0.008 0.002 0.182 0.011 0.203
Overall arsenic recovery to concentrate = 5.4% Arsenic recovery in leach tank (from high-As concentrate to leaching solution) = 98%			

The overall recovery of copper from ore to final concentrate was marginally higher with the new flowsheet (86.4% (roasting) and 86.3% (leaching) compared to 86.2% (basecase)), but the overall arsenic recovery was significantly lower (13.3% (roasting), 5.4% (leaching) compared with 86.2% (basecase)). In carrying out the simulations to make the comparison between the flowsheets, it should be noted that in order to produce a saleable final concentrate from the base case, it was necessary to downgrade slightly the overall tennantite recovery (R_T) of the base case (R_T = 86.4%) relative to the proposed processes (R_T = 91 + 4 = 95%). If this was not done, the arsenic content of the final concentrate for the conventional base case was too high and the concentrate would not be acceptable to smelters.

ECONOMIC CALCULATIONS

The mass and energy balances obtained from the METSIM simulations of the respective proposed flowsheets were used to estimate the operating costs of both processes. Unless specified otherwise, all costs are reported in Australian dollars (2010). The following assumptions were made in estimating these costs:

- Flotation reagent consumption in rougher – 100 g/t ore sodium hydrosulphide (NaHS), 10 g/t ore xanthate, 30 g/t AP208, 50 g/t ore frother (Smith and Bruckard, 2007);
- Flotation reagent consumption in cleaner stages is approximately 25% of that in rougher stages, although no NaHS is used in the cleaner stages;
- Prices of flotation reagents – sodium hydrosulphide \$577/t, xanthate (sodium isopropyl) \$1,298/t, AP208 \$1,923/t, frother \$4,377 (quotes from bulk chemical companies in China, 2010);
- Number of flotation cells for the proposed process is 50% greater than for the base case;
- Manning levels for the flotation circuit in the base case were estimated at 84 personnel over 4 shifts (0.021 manhours/t ore);
- Manning levels for the flotation circuit and roaster for the new process were estimated at 96 personnel over 4 shifts (0.024 manhours/t ore);
- Labour hourly rates were estimated to be \$81.91/manhour based on Australian Workers Ordinary Time Earnings and other factors (Australian Bureau of Statistics website, 2011; Haque *et al*, 2010);
- Electricity consumption for the flotation circuit of the base case was estimated to be 6 kWh/t ore, and 6.9 kWh/t ore for the new process;
- Price of electricity \$0.08/kWh;

- Consumables and maintenance are 0.9% and 5% of plant capital cost respectively – a very approximate first estimate of the capital cost of the flotation circuit for the base case is \$12 M, and for the new process (including roasting stage but not stabilization stage) is \$17 M;
- Fresh water consumption is 0.37 m^3/t ore for the base case and 0.56 m^3/t ore for the new process with roasting, 0.58 m^3/t ore for leaching and 0.57 m^3/t ore for leaching with 50% recovery;
- Price of water is \$1.0/kL;
- Price of natural gas (roaster and fuel for heating of leaching solution) is \$5/GJ;
- Cost of nitrogen used in the roaster has not been included at this stage;
- Operating cost of arsenic stabilization stages is \$350/t fume from roaster (or \$0.10/t ore), assumed to be the same for As stabilization from leaching for this study - based on data provided by Environment Australia (2007) for a number of arsenic stabilization technologies;
- For the leaching scenario, the Na_2S consumption was 0.47 kg/kg of feed in the leaching tank and the NaOH consumption was 0.1 kg/kg based on laboratory experimental data. This corresponds to 11.3 kg Na_2S/t of ore and 3 kg NaOH/t of ore processed assuming the purity specification provided by the chemical suppliers. The leaching temperature was assumed to be at 100ºC. Natural gas fuel consumption in leaching has been accounted for this study.

Operating cost

The estimated operating costs (both as per tonne of ore and per tonne of concentrate) are given in Tables 6, 7 and 8 for the conventional and new processes with roasting and leaching scenarios, respectively. The operating cost for the leaching scenario with 50% recovery (recycling) of leaching chemicals is shown in Table 9, assuming there is no significant cost associated with recycling these chemicals. There are three additional stages in the new flowsheets. One is a flotation stage, one is a roasting or leaching stage, and the other is an arsenic stabilization stage. Detailed information about the stabilization and roasting stages/equipment are given in earlier paper (Bruckard *et al*, 2010). The natural gas consumption as fuel for heating the leaching solution to 100°C has been estimated using the METSIM flowsheeting software and is shown in Tables 8 and 9. For the recycled chemicals case, it has been assumed that the leaching solutions can be recycled without any significant heat loss.

Table 6. Operating cost of base case flowsheet (32.8 t/h concentrate)

Item	Consumption	Unit	Price	Unit	Operating cost	
					\$/t ore	**\$/t conc.**
Reagents						
Modifier (NaHS)	100	g/t ore	577	\$/t	0.06	1.76
Frother	62.5	g/t ore	4377	\$/t	0.27	8.21
Xanthate	12.5	g/t ore	1298	\$/t	0.02	0.52
AP208	37.5	g/t ore	1923	\$/t	0.14	2.26
Electricity	6	kWh/t ore	0.08	\$/kWh	0.48	14.65
Water	0.37	kL/t ore	1.0	\$/kL	0.37	10.36
Labour	0.021	manh/t ore	81.91	\$/manh	1.72	52.42
Consumables & maintenance					0.00	0.09
Total					**2.99**	**90.27**

Table 7. Operating cost of proposed flowsheet (34.1 t/h concentrate) – roasting scenario

Item	**Consumption**	**Unit**	**Price**	**Unit**	**Operating cost**	
					$/t ore	**$/t conc.**
Reagents						
Modifier (NaHS)	100	g/t ore	577	$/t	0.06	1.96
Frother	75	g/t ore	4377	$/t	0.33	9.47
Xanthate	15	g/t ore	1298	$/t	0.02	0.54
AP 208	45	g/t ore	1923	$/t	0.09	2.61
Electricity	6.9	kWh/t ore	0.08	$/kWh	0.55	16.24
Water	0.56	kL/t ore	1	$/kL	0.56	15.31
Labour	0.024	manh/t ore	81.91	$/manh	1.97	57.81
Fuel (natural gas)	0.01	GJ/t ore	5	$/GJ	0.05	1.32
Consumables & maintenance					0.00	0.11
Arsenic stabilization					0.10	2.97
Total					3.72	108.07

Table 8. Operating cost of proposed flowsheet (34.1 t/h concentrate) – leaching scenario without recycling of leaching chemical

Item	**Consumption**	**Unit**	**Price**	**Unit**	**Operating cost**	
					$/t ore	**$/t conc.**
Reagents						
Modifier (NaHS)	100	g/t ore	577	$/t	0.06	1.7
Frother	75	g/t ore	4377	$/t	0.33	9.5
Xanthate	15	g/t ore	1298	$/t	0.02	0.5
AP 208	45	g/t ore	1923	$/t	0.09	2.6
Electricity	6.9	kWh/t ore	0.08	$/kWh	0.55	16.2
Water	0.56	kL/t ore	1	$/kL	0.56	15.3
Labour	0.024	manh/t ore	81.91	$/manh	1.97	57.8
Leaching reagents						
Na_2S	11.8	kg/t ore	450	$/t	5.31	155.53
NaOH	3	kg/t ore	725	$/t	2.18	63.71
Water	0.02	kL/ore	1	$/kL	0.02	0.58
Fuel (natural gas)	0.01	GJ/t ore	5	$/GJ	0.05	1.39
Consumables & maintenance					0.00	0.11
Arsenic stabilization					0.1	2.97
Total					11.22	328.0

Table 9. Operating cost of proposed flowsheet (34.1 t/h concentrate) – leaching scenario with 50% recovery of leaching chemicals

Item	Consumption	Unit	Price	Unit	Operating cost	
					$/t ore	$/t conc.
Reagents						
Modifier (NaHS)	100	g/t ore	577	$/t	0.06	1.7
Frother	75	g/t ore	4377	$/t	0.33	9.5
Xanthate	15	g/t ore	1298	$/t	0.02	0.5
AP 208	45	g/t ore	1923	$/t	0.09	2.6
Electricity	6.9	kWh/t ore	0.08	$/kWh	0.55	16.2
Water	0.56	kL/t ore	1	$/kL	0.56	15.3
Labour	0.024	manh/t ore	81.91	$/manh	1.97	57.8
Leaching reagents						
Na_2S	5.9	kg/t ore	450	$/t	2.66	77.76
NaOH	1.5	kg/t ore	725	$/t	1.09	31.85
Water	0.01	kL/ore	1	$/kL	0.01	0.29
Fuel (natural gas)	0.004	GJ/t ore	5	$/GJ	0.03	0.70
Consumables & maintenance					0.00	0.11
Arsenic stabilization					0.10	2.97
Total					7.45	217.4

The operating cost of the new process using the roasting option was approximately 20% higher ($108.1/t concentrate compared with $90.3/t concentrate) than the conventional process. However, it was found that with the leaching option, the operating cost is at present over three and a half times higher than the conventional process. The cost of chemical reagents accounts for about 67% of the operating cost for this option. If about 50% recovery of these chemicals is possible by recycling, the leaching option operating cost may be reduced by 34% ($217.4/t conc. compared with 328.0/t conc.). The contributions to the operating cost for all the processes are compared in Figure 7. Excluding leaching chemicals, the cost of labour was the major contributor followed by electricity and flotation reagents for all cases. In the case of the leaching option, the cost of sodium sulphide and sodium hydroxide is significant. Although the leaching process works very well in the laboratory, unless a reduction in chemical consumption during leaching is achieved by further optimization of the process, the high operating cost will significantly affect its financial viability. However, the new processes may still prove to be financially viable when based on the ratio of operating cost to the value of copper in the saleable concentrate product, using the current copper metal price from the London Metal Exchange (A$8,645/t). If the ore cannot be processed using the conventional process to produce a product with an acceptable arsenic content or the arsenic specification cannot be achieved by blending, then the proposed process with either roasting or leaching becomes the only option for processing this ore. Furthermore, if the penalty for arsenic increases over time, which is a possible scenario, then the new proposed process using roasting could become even more economically attractive.

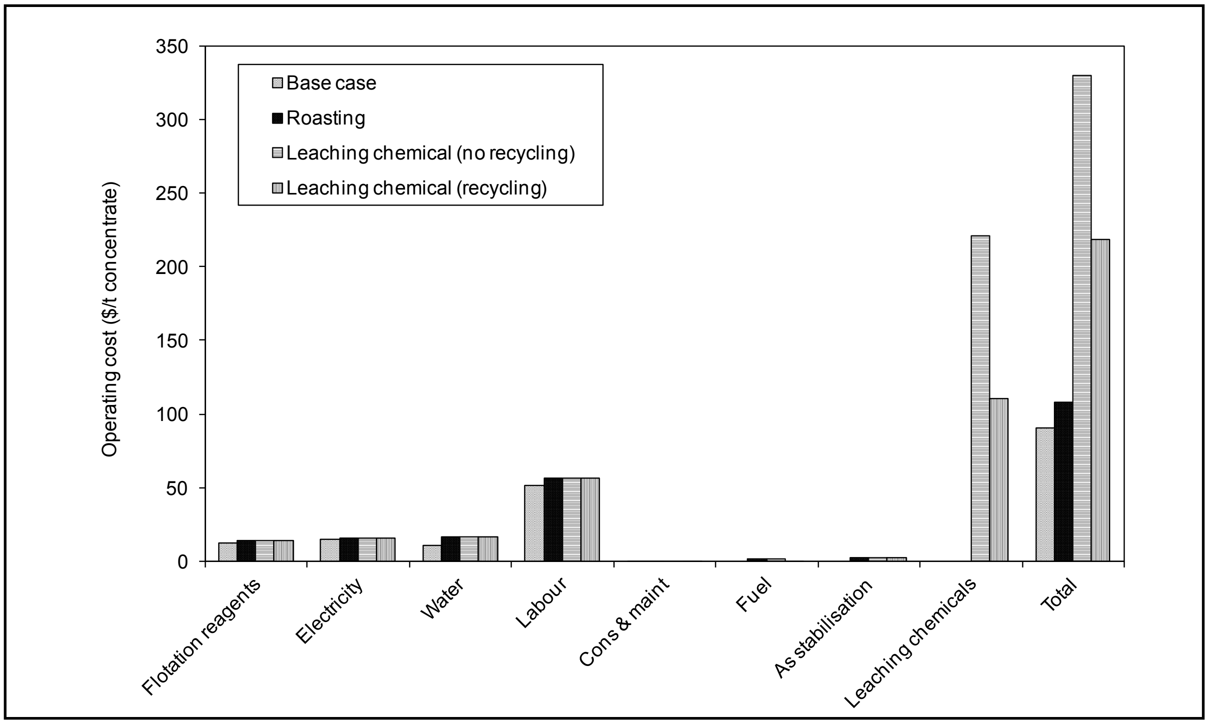


Figure 7. Comparison of operating cost contributions for base case and new processes with roasting and leaching scenarios (with cost of leaching reagents)

Revenue estimation

Arsenic penalty calculation

Based on the amount of arsenic present in the concentrate, the seller must pay a penalty on the price received for the concentrate. The arsenic penalty values were obtained from three different data sources and are shown plotted in Figure 8 (Haque *et al,* 2010). Although not varying dramatically, the middle data set in this figure was used in the subsequent analysis of the three flowsheets. According to this data source, the arsenic penalties are:

< 0.2% (2000 ppm)	No penalty
0.2% - 1.0%	US$3 per 0.1% As

The final concentrate from the base case flowsheet contained 5341 ppm (0.53%) As and thus attracted a penalty of US$3 × (0.53-0.20)/0.10 = US$9.90/t concentrate (or A$9.9/t at an exchange rate of A$1 = US$1.0). The final concentrate produced from the proposed flowsheet had an arsenic content of 777 ppm (0.08%) for the roasting and 349 ppm (0.03%) for the leaching scenarios, with neither attracting any penalty.

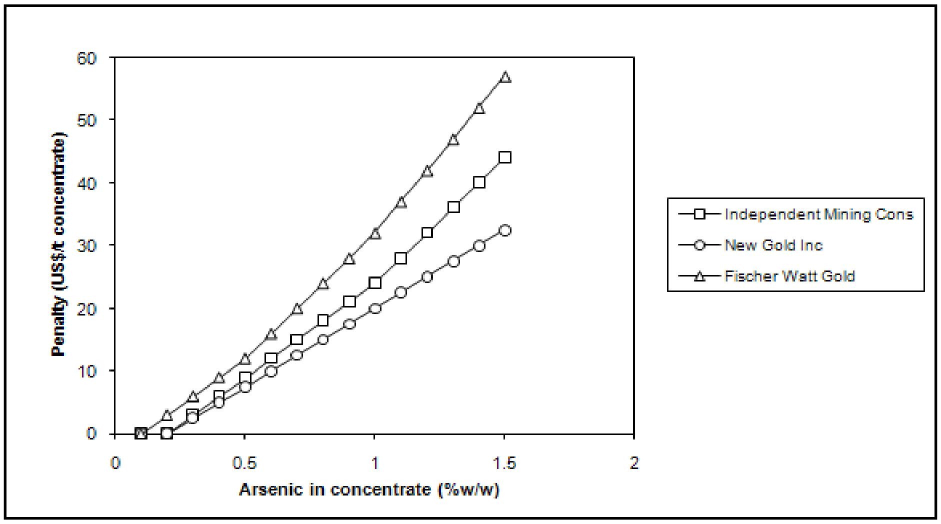


Figure 8. Arsenic penalties for copper concentrates (various sources)

Revenue calculation

The price received for the final copper concentrates from both proposed flowsheets was based on the copper content (96.7% payable) and an average copper price over 2011 of US$8,645/t (LME, 2011) or A$8,645/t at an exchange rate of A$1.0 = US$1.0. Based on these data the price received for the copper concentrates was calculated to be:

Copper concentrate revenue (A$/t) = (%Cu in conc./100) × 0.967 × 8645

= 86.45 × %Cu in conc.

The net revenue received from all flowsheets is given in Table 10. As slightly different amounts of concentrate are produced for the different flowsheets, subsequent calculations discussed below are made on a per tonne of ore basis.

Table 10. Revenue from copper concentrate for all flowsheets

Parameter	**Base case**	**Roasting-based process**	**Leaching-based process**	**Leaching-based process (with 50% recovery of leaching reagents)**
Ore (t/h)	1000	1000	1000	1000
Concentrate (t/h)	32.8	34.1	34.1	34.1
Concentrate grade % Cu % As	26.4 0.53	25.4 0.08	25.3 0.03	25.3 0.03
Cu revenue $/t conc $/t ore	2207 72.4	2123 72.4	2123 72.1	2123 72.1
As penalty $/t conc $/t ore	9.9 0.32	0.00 0.00	0.00 0.00	0.00 0.00
Net revenue after payment for penalty ($/t ore)	72.08	72.40	72.10	72.10

DISCUSSION

Economic results

Based on the economic results (and underlying assumptions) given in Table 10, the net revenue derived from the new proposed processes over the base case conventional process was estimated to be $0.44/t ore for roasting and $0.46/t for leaching. The increased operating cost of the new process using roasting over the conventional process from Tables 6 and 7 was estimated to be $0.73/t ore, while the increased operating cost of the new process using leaching over the conventional process from Tables 6 and 8 was estimated to be $8.23/t ore. However, with recycling of the leaching chemicals with an assumed recovery of 50%, this increase in operating cost is reduced to $4.46/t ore. This analysis does not take into account the increased capital cost (additional flotation cells, roaster and associated equipment) required for the new processes. For the assumed base case nominal feedrate of 1000 t/h ore (approximately 8.0 Mtpa at 92% plant availability), the additional operating cost for the plant would be around $5.8 M/y for roasting and $66 M/y for leaching. With recycling of chemicals with the leaching process, the additional cost is $36 M/y compared with $66 M/y. These results indicate that unless the leaching cost is dramatically reduced, roasting is the preferred option from an economic point of view.

These first estimates of the economic benefits of the new proposed process using roasting are sufficiently encouraging to warrant further development of the flowsheet. It should also be appreciated that while the conventional base case flowsheet was assumed to produce a saleable concentrate for the purpose of the analysis, if this is not the case in practice, the economic benefits of the proposed process increase dramatically. The economic results also indicate that optimization of leaching process is required if the viability of this process is to be improved.

Major uncertainties

As is to be expected in a preliminary evaluation of this nature, there are a number of uncertainties associated with the results obtained. The major uncertainty on the revenue side is the magnitude of the arsenic penalty imposed. On the cost side, the major uncertainties are the arsenic stabilization cost, the additional level of manning required and electricity consumed for the extra equipment, and the additional reagent consumption. The main process uncertainties relate to the separation performance of the high-arsenic and low-arsenic flotation stages and roasting stage, as well as the nature and performance of the arsenic stabilization stage. The associated other penalties for roasting and leaching would be different because of the loss of fuel value in materials in the smelter with concentrate after roasting that has not been considered here. The preliminary investigation of the flow rate of sulphur over the flotation circuit indicates that the difference in sulphur and iron sulphide content of the products of three scenarios is small. However, further detail investigation is necessary. The cost of off-gas handling in roasting scenario has not also been included in this analysis. Thus it is possible that roasting cost may increase by small amount if all these factors are considered in detail but still the cost is much lower than the leaching scenario. As the project moves to larger scale in the coming years, it is anticipated that new or revised data will be obtained to resolve some of these uncertainties. For commercial application of this technology, further studies should be conducted to determine overall environmental impact of both processes for getting approval from the environmental agencies which is beyond of the scope of this study.

CONCLUSIONS

This study is a first attempt to determine whether incorporating additional arsenic separation and stabilization stages into the conventional copper processing flowsheet in order to facilitate the early removal of arsenic from copper ores of high arsenic content would be economically viable. Based on the flowsheet performance data assumed, the analysis showed that the yield of concentrate will be slightly higher with the new proposed process using roasting compared with the conventional (base case) process, although the copper grade is slightly lower, resulting in similar copper revenues from both processes. However, the arsenic content of the concentrate in the new process is significantly lower, with an arsenic penalty no longer being imposed, as was the case with the conventional process. If this roasting process is implemented in a 8Mta plant with 92% availability, the net increase in cost is about $5.8 M/y. Processing options using leaching as an alternative to roasting is significantly more expensive in terms of operating cost compared with the conventional process and also with the roasting option.

However, it should be emphasized that this evaluation was based on the premise that the conventional process was capable of producing a final concentrate with an arsenic content low enough to make it acceptable to smelters. It was against this scenario that the new proposed process was compared. If this is not the case in practice (i.e. no acceptable concentrate can be produced from the ore by conventional processing), meaning a high arsenic orebody cannot be mined and processed, then the economic benefits of the new proposed flowsheet increase dramatically. Furthermore, any improved environmental or sustainability benefits of the proposed flowsheet have not been investigated or quantified. Such benefits, if forthcoming, could further improve the economic credentials of the new process.

Additional capital expenditure would be required to implement the new flowsheet (additional flotation cells, roaster and associated equipment). However, the extent of this increase in capital cost is dependent on the plant throughput, but this is not known at this stage, with a nominal 1000 t/h feed rate of ore assumed for this conceptual study. This factor will be influenced by whether a new roaster is to be purchased or an existing one utilized. Nevertheless, these first estimates of the economic benefits of the new proposed flowsheet for early arsenic removal from copper ores using roasting are sufficiently encouraging to warrant further development of the flowsheet.

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