Integrated piloting of a thermophilic bioleaching process for the treatment of a low-grade nickel-copper sulphide concentrate

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Synopsis

Mintek was a leading participant in the BioMinE project between 2004 and 2008. This project, which was funded in part by the European Commission, was aimed at the development of biotechnology for the minerals industry in Europe. Mintek's research programme focused mainly on the development of integrated bioleach-based processes for the recovery of base metals from complex, low-grade sulphide concentrates. Specific European mineral resources were targeted and used in integrated piloting campaigns involving bioleaching, solution purification, and metals recovery.

This paper describes the use of thermophilic bioleaching for the recovery of nickel and copper from a low-grade nickel-copper concentrate produced at the Aguablanca Mine in southern Spain. Currently, the Aguablanca Mine produces a bulk nickel-copper concentrate for sale to a smelter, and the proposition is to increase the profitability of the operation by the on-site production of metal or metal intermediate.

Initially, bench-scale bioleach tests were conducted to determine the bioleach operating conditions. These tests included an evaluation of mesophilic, moderately thermophilic and thermophilic microorganisms. In order to achieve sufficiently high levels of both copper and nickel extraction, a thermophilic process was selected—this was necessary for leaching of the refractory chalcopyrite that occurs in this concentrate. Additional bench-scale test work was carried out to derive a conceptual process flowsheet for the solution purification and metals recovery circuit.

The results of the bench-scale tests were used to design, construct and commission an integrated pilot plant, which was subsequently operated at Mintek for over seven months. During this time, the solution purification and metals recovery processes were optimized, and all recycle loops were closed. The final process flowsheet included the following unit operations: concentrate regrinding, thermophilic bioleaching at 70°C, primary iron removal using limestone, copper solvent extraction and electrowinning, secondary iron removal, nickel hydroxide precipitation using magnesia, and final solution purification using lime. Where applicable, process solutions were recycled to preserve water.

The process design data derived from this pilot-plant campaign formed the basis for a conceptual engineering study for the developed process. In the study, mass and energy balances were derived, and a process flowsheet was developed and used as the basis for estimating the capital and operating costs of the process. This enabled a preliminary economic analysis of the process to be undertaken. The findings of this study are discussed.

Introduction

The BioMinE project

The BioMinE project[1,2] was a four-year long integrated project under the Sixth Framework Programme of research supported by the European Commission, which ran from November 2004 to October 2008. The project was aimed at the evaluation of biohydrometallurgy to improve the exploitation of the European non-ferrous metal resources in a sustainable way. At the time of its completion, the project had 37 participating partners covering a diversity of interests: research organizations, academic institutions, and industrial and mining companies. South Africa was well represented: besides Mintek, two academic institutions and one mining company from South Africa were members of the BioMinE consortium.

Mintek's role in BioMinE

Mintek's Biotechnology Division played a leading role in the project, both as coordinator of the bioleaching 'Work Package' and by active participation in key aspects of the research and technology development (RTD). One of the main focuses of Mintek's research programme was the development of integrated bioleach-based processes for the recovery of base metals from complex, low-grade sulphide concentrates. Specific European mineral resource types were targeted and used in integrated piloting campaigns involving bioleaching, solution purification, and metals recovery. One such resource is copper polymetallic concentrates, and a nickel-copper concentrate, which is currently produced at the Aguablanca Mine in southern Spain, was selected as one of the target resources in this

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project. Mintek conducted extensive bench-scale bioleach testing on this material, the results of which formed the basis for an integrated bioleaching pilot-plant campaign that was conducted at Mintek.

Bioleaching of base metal concentrates

The bioleaching of concentrates of base metal sulphides has been the subject of extensive research and development over several decades[3]. The use of mesophilic bacteria to oxidize sulphide minerals such as cobaltiferous pyrite, pentlandite and millerite (nickel), sphalerite (zinc), galena (lead), chalcocite, covellite, cubanite and bornite (all copper) has been demonstrated, and in some cases the bioleaching process has successfully been integrated with downstream metals recovery processes[4,5,6].

In the early 1990s, BHP Billiton demonstrated the bioleaching of a pentlandite concentrate using mesophiles and moderate thermophiles, in a demonstration plant with a design capacity of 300 kg/d of concentrate. This plant produced more than 700 kg of Class 1 nickel in a six-month long campaign. Subsequently, an integrated pilot plant situated at the Yabula Nickel Refinery in Queensland, Australia, was successfully operated as part of a feasibility study[4].

In the late 1990s, the world's first commercial bioleaching plant treating a base metal concentrate was established at the site of the defunct Kilembe Copper Mine in Uganda. The plant, owned an operated by the Kasese Cobalt Company, was designed to produce about 1 000 t/a of cobalt cathode from a cobaltiferous pyrite concentrate. The technology for this plant was supplied by the French company, BRGM[5].

Bioleaching of the primary copper sulphide, chalcopyrite, has presented more of a challenge, since the surface of the chalcopyrite tends to passivate at the temperatures and redox potentials found in typical mesophilic bioleach processes, effectively stopping the leaching of the mineral. This behaviour can be overcome by a variety of approaches, often applied in combination, which include increasing the operating temperature (requiring the use of thermophilic microorganisms), finer grinding of the concentrate, the addition of catalysts (such as silver), and control of the redox potential[3,6,7,8,9].

In 2001, Mintek and its partners, Industrias Peñoles S.A. de C.V. of Mexico and BacTech, successfully demonstrated the bioleaching of chalcopyrite using moderate thermophiles at an operating temperature of 45°C, in an integrated demonstration plant that was commissioned and operated in Monterrey, Mexico. The plant, with a design capacity of 1 t/d of copper, produced more than 40 t of LME A-grade cathode copper from a complex polymetallic concentrate in which the main copper mineral was chalcopyrite[6].

In 2003, Alliance Copper, a joint venture between BHP Billiton and Codelco, demonstrated the thermophilic bioleaching of chalcopyrite concentrate, at an operating temperature of 78°C in a prototype plant situated at Chuquicamata in Chile, which was designed to produce 20 000 t/a of copper cathode[9]. A unique feature of this prototype plant is that the air supply is supplemented with pure oxygen, which necessitates the use of an automated system to control the dissolved oxygen concentration in the process.

In the past 20 years, the bioleaching of a wide variety of base metal concentrates has been demonstrated. In the past decade, the development has focused mainly on the bioleaching of chalcopyrite, and the investigations have been aimed either at complex polymetallic concentrates containing several sulphide minerals (the Mintek/BacTech/Peñoles project), or copper concentrates that contain impurities such as arsenic (the Alliance Copper operation).

The Aguablanca Mine

The Aguablanca Mine is an open-pit nickel-copper sulphide mine, located in the province of Badajoz, about 80 km north of Seville in southern Spain (Figure 1a). The mine currently produces a bulk, low-grade copper-nickel concentrate for sale to a smelter. Aguablanca was the first nickel sulphide mine in production in Western Europe. Construction of the on-site treatment plant (Figure 1b) was completed in December 2004, commissioning and production commenced early in 2005, and production parameters were reached in early 2006. The project has an initial projected mine life of 10½ years.

Figure 1a—Location of the Aguablanca Mine in southern Spain

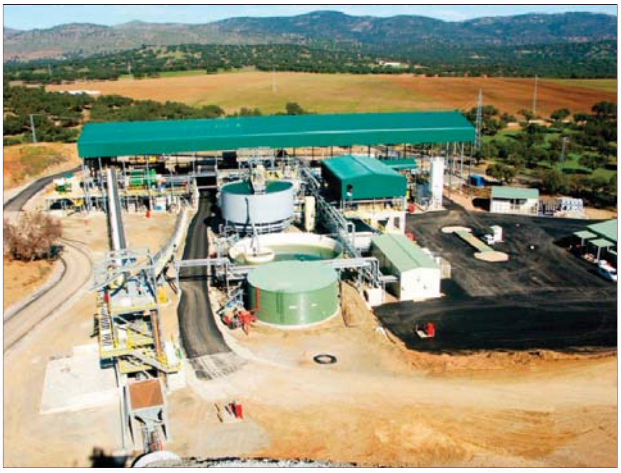


Figure 1b—The Aguablanca treatment plant

The on-site flotation plant was designed to treat 1.5 million tonnes of ore per year, with an additional milling capacity of 0.3 million tonnes per year. The plant produces a bulk copper-nickel-PGM concentrate, but also has the flexibility to produce separate copper and nickel concentrates. The process flowsheet comprises primary jaw crushing, primary semi-autogenous grinding, secondary ball milling, gangue preflotation, flotation, concentrate thickening and filtration. Production data for the first two full years of operation are summarized in Table I.

The concentrate is transported either to Glencore's Huelva smelter, located 140 km away on the Spanish coast, or to other smelters, for treatment. Cash costs (including by-product credits) in 2007 amounted to US$7.23 per pound of nickel sold[10].

The mineralization of the Aguablanca deposit is similar in type to both the Voisey's Bay deposit in eastern Canada and the Norilsk deposit in Russia. Nickel, copper, platinum and palladium mineralization occurs within magmatic breccia bodies that form gossans at the surface. Pyrrhotite, pentlandite and chalcopyrite comprise the dominant sulphide mineralization.

Following the acquisition of the operation by Lundin Mining Corporation (Lundin) in 2007, encouraging exploration results were announced, indicating the existence of a deep orebody below the existing one. The most recent exploration results have confirmed the potential for a significant increase in the mine's resource base and for an expansion of the current open pit production to include underground reserves. In March 2008, Lundin announced that a feasibility study was being conducted on undertaking underground mining, which was expected to be completed towards the end of 2008. If approved, a plant expansion project will be finalized to cater for the processing of underground ore.

Objectives and scope of work

The proposition underpinning this work is to extend the Aguablanca treatment plant by the addition of a bioleaching and metals-refining facility, allowing the on-site production of metal or metal intermediate products, and thus substantially increasing the realized metal value.

Table I

Aguablanca Mine production data

	2006	2007
Ore mined (tonnes)	1 550 437	1 707 330
Ore milled (tonnes)	1 486 800	1 668 959
Ore grade		
Nickel (%)	0.6	0.5
Copper (%)	0.5	0.4
Flotation recovery		
Nickel (%)	72	76
Copper (%)	90	92
Concentrate grade		
Nickel (%)	6.6	7.3
Copper (%)	6.8	6.9
Production (metal contained)		
Nickel (tonnes)	6 398	6 630
Copper (tonnes)	6 616	6 281

The selection of bioleaching for treatment of Aguablanca concentrate may have certain advantages over alternative hydrometallurgical processes such as pressure leaching:

- ➤ The Aguablanca operation is relatively small, and bioleaching may be more suitable for this scale of operation than pressure leaching, which has relatively high capital and operating costs and has been shown to be more cost-effective at larger scales[11].
- ➤ Bioleaching is also readily able to treat low-grade concentrates, and one option that could be considered if bioleaching was to be implemented is to produce an even lower-grade concentrate than is presently the case, in order to improve the overall recovery of metal (particularly nickel) in the concentrate.
- ➤ If plant expansions were implemented at Aguablanca in the future, the modular nature of bioleach tanks would make it possible to expand the plant with relative ease.

Extensive bench-scale amenability and pilot-plant testing of the Aguablanca concentrate was performed at Mintek. Three phases of testing were completed.

In phase one, the concentrate sample used in the test work was subjected to chemical, physical and mineralogical characterization. Amenability testing comprised open-circuit bioleach testing in small-scale, continuously operated bioleach reactors, with process volumes of a few litres. The results of these tests were used to define the bioleach operating parameters for the next two phases of test work.

In the second phase, a continuous bioleach miniplant was operated in open circuit, using the bioleach operating parameters defined in phase one. The product from this miniplant was used to carry out bench-scale tests aimed at assessing the operating parameters for solution purification and metals recovery. The results of these tests were used to define the initial operating parameters for the final phase of test work.

In the third and final phase, an integrated pilot plant comprising bioleaching, solution purification and metals recovery was operated for a period of over seven months. During this integrated piloting campaign, seven cycles of downstream processing were completed, and integration was achieved by recycling of the purified process solution. The primary aims of the integrated piloting campaign were to assess the impact, if any, of recycling of the purified process solution on the bioleach process performance, to optimize the operating parameters for the purification and metals recovery circuit, and to define the process design criteria to be used as the basis for a conceptual engineering study and preliminary economic evaluation of the process. In this phase, additional optimization test work was also conducted to assess the effect of bioleach process parameters such as redox potential, concentrate grind size and feed solids concentration on copper and nickel extractions and leach kinetics. The results of these additional optimization tests were included in the engineering and economic study.

The outcomes of this comprehensive bench-scale and pilot-plant test work programme are discussed here, together with the results of the engineering and economic study.

Conceptual flowsheet development

At the outset, a conceptual process flowsheet for bioleaching and the recovery of metals from the Aguablanca concentrate was envisaged. The conceptual flowsheet aimed to address the following key issues:

- Solid-liquid separation and washing of the bioresidue, wastes and metal products
- Removal of iron from the leach liquor
- Purification of the leach liquor
- Recovery of copper from the leach liquor
- Recovery of nickel from the leach liquor
- Waste products and their disposal
- Re-use of water
- The acid balance.

Solid-liquid separation and washing of the residues, wastes and products is usually achieved either by counter-current decantation (CCD) using a series of thickeners, or by filtration. The choice depends on many factors, including the settling and filtration characteristics of the solid materials, the required washing efficiency, and the availability of water.

The removal of iron from leach liquors such as these is usually achieved by precipitation using limestone, at a pH level of around 3.0. The aim of this process step is to remove iron prior to copper solvent extraction, without coprecipitating or entraining the valuable metals (nickel and especially copper), and in so doing to produce a waste product which has good filtration or settling characteristics. The factors that govern this process include the operating temperature, the overall residence time, the number of precipitation stages and the rate at which the pH level is raised, and the final pH level.

The recovery of copper from bioleach liquors is usually achieved by the well-established method of solvent extraction (SX) using a copper-selective organic reagent. The SX process selectively removes copper from the pregnant solution, and produces a purified copper solution from which copper can be extracted as a cathode using electrowinning (EW). Important factors that govern this process are the number of extraction and stripping stages, the organic reagent that is applied, and the pH level and copper tenor of the pregnant solution.

A few variations for the residue washing, iron removal and copper SX-EW operations were considered, including:

- Precipitation of the iron directly after bioleaching, without first separating the pregnant leach liquor from the bioresidue. This would remove one solid-liquid separation and washing step from the process, but it would have to be established that the resultant solids, comprising the bioleach residue, iron precipitates and gypsum, could be thickened or filtered easily.
- Performing the copper SX step prior to removing the iron from the leach liquor. This would eliminate the possibility of encapsulation or coprecipitation of copper in the iron precipitate, but it would have to be established that the iron in the SX feed did not affect the extraction of copper.

Purification of the SX raffinate is usually required prior to nickel recovery. The SX process reduces the pH level of the liquor, and this purification is usually achieved by further neutralization of the liquor by limestone, to a pH level of about 5.0. The objective is to remove all impurities that may contaminate the nickel product.

There are several options that could be considered for the recovery of nickel from the leach liquor. In the nickel laterite industry, which has seen the implementation of hydrometallurgical nickel extraction and recovery processes in a number of projects in the past decade, several different approaches have been followed. These include precipitation to form a metal salt, precipitation to produce a metal sulphide, and SX-EW, usually after the formation and redissolution of an intermediate product (either a salt or a sulphide), to produce metal.

After a careful evaluation of the various alternatives, it was decided that the integrated pilot plant would incorporate the metal salt precipitation route, for the following reasons:

- It is conceptually the simplest route to follow.
- Nickel SX-EW is not as well established as copper SX-EW, and the process circuits are complex, difficult to operate, with high capital and operating costs.
- The production of an intermediate metal salt does not preclude the option of producing metal by SX-EW. The metal salt could be redissolved and the solution used as a feed to a refinery including nickel SX-EW. Such a refinery could be added at a later stage; this would reduce the initial technical risk associated with the project.

One of the objectives of the bench-scale test work programme conducted prior to the integrated piloting campaign was to evaluate which metal salt should be produced as a nickel intermediate.

The final step in this process (in which a nickel salt is produced) would involve a partial or total neutralization of the barren liquor to remove any residual impurities that remain or that may have been added during the nickel precipitation process, prior to recycling of the liquor to the front end of the circuit.

This conceptual flowsheet is shown in Figure 2. In this case, magnesia is used as the alkali in the nickel precipitation process, and the product is nickel hydroxide.

For this process flowsheet, the soluble magnesium sulphate produced in the nickel hydroxide precipitation process would need to be removed by the addition of lime in the final neutralization step.

Concentrate characterization and bioleach amenability testing

Characterization of the Aguablanca concentrate

The Aguablanca concentrate was characterized both physically and chemically.

The dry density of the Aguablanca concentrate was determined to be 3 292 kg/m^3.

In all of the test work, the concentrate was reground prior to bioleach testing, and a particle-size distribution with a d_{90} of 10–12 μm was targeted. The size distributions of the as-received and reground concentrates are illustrated in Figure 3.

The typical chemical analysis of the Aguablanca concentrate is summarized in Table II.

A scanning electron microscope was used to perform a quantitative modal analysis of the Aguablanca concentrate, and the results are summarized in Table III. This analysis shows that the bulk of the concentrate comprises the sulphide minerals pyrite, pyrrhotite, chalcopyrite and pentlandite, with

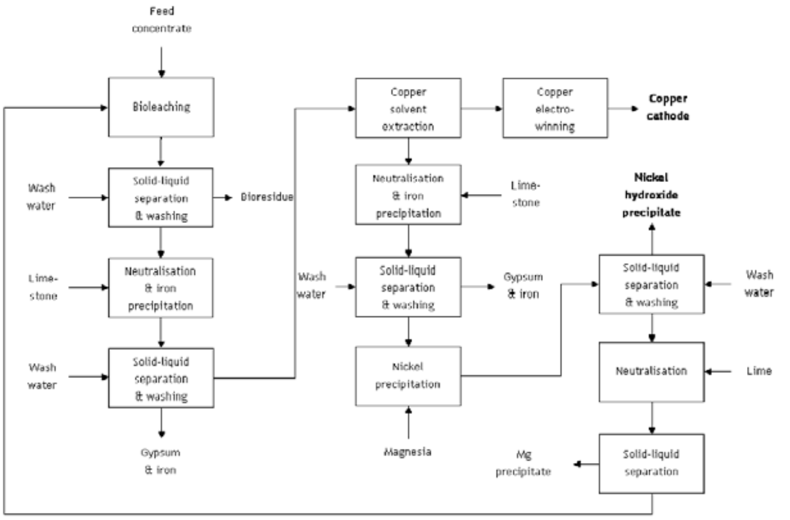


Figure 2—Conceptual flowsheet for the Aguablanca nickel-copper concentrate

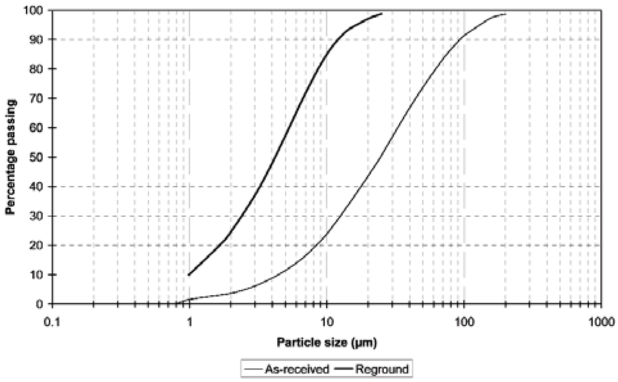


Figure 3—Typical particle-size distributions for the as-received and reground Aguablanca concentrates

Table II

Typical chemical analysis of the Aguablanca concentrate

Constituent	Concentration (%)
Cu	6.29
Ni	5.24
S_{TOT}	24.40
S^{2-}	24.36
Fe	28.90
Si	6.28
Mg	2.74
Al	1.16
Ca	1.00
Co	0.18

a variety of silicates such as amphibole, feldspar, talc and chlorite being the predominant gangue minerals.

Bioleach amenability test work

The results obtained from amenability test work performed in

Table III

Typical modal analysis of the Aguablanca concentrate

Mineral groups	Mineral name	Chemical formula	Mass % *	
Sulphides	Pyrite	FeS_2	23.7	77.4
	Pyrrhotite[1]	$Fe1_{-x}(Ni)S$	20.7	
	Chalcopyrite	$CuFeS_2$	18.5	
	Pentlandite	$(Fe,Ni)_9S_8$	14.5	
	Gersdorffite	NiAsS	0.04	
	Cu-sulphides[2]	CuS	0.03	
	Sphalerite	Cu_2S	0.02	
	Galena	ZnS	0.02	
Silicates	Amphibole	$NaCa_2(Mg,Fe)_4Al(Si_6Al_2)O_{22}(OH)_2$	9.8	21.6
	Feldspar	$CaAl_2Si_2O_8$; $KAlSi_3O_8$	4.2	
	Talc	$Mg_3Si_4O_{10}(OH)_2$	3.2	
	Chlorite	$(Mg,Fe,Al)_6(Si,Al)_4O_{10}(OH)_8$	2.2	
	Ni-bearing silicates	Fe(K,Al,Mg,Ni)-silicate	1.8	
	Quartz	SiO_2	0.3	
Oxides	Fe-oxides	Fe_2O_3; α-FeOOH; Fe_3O_4	0.6	0.6
	Ilmenite	$FeTiO_3$	0.04	
Sulphates	Cu-bearing sulphate	Ca(Fe,Cu)-sulphate	0.2	0.4
	Gypsum	$CaSO_4.2H_2O$	0.1	
Phosphate	Apatite	$Ca_5(PO_4)_3(OH,F,Cl)$	0.02	<0.5
PGM species	-	PtPd-BiTe and PtRh-S	<0.01	<0.5

1 Including pyrrhotite with and without detectable nickel
2 Mostly covellite but includes chalcocite
* Where relevant, two decimal figures are given merely to demonstrate proportional differences between trace minerals

continuously operated bench-scale reactors using mesophile, moderate thermophile and thermophile cultures indicated that nickel extractions of over 98% could be obtained using moderately thermophilic and thermophilic cultures operating at 45 and 70°C, respectively. For copper, which occurs as chalcopyrite, extractions of over 95% could only be achieved using the thermophile culture at 70°C (Table IV). Based on these results, the thermophile culture was selected for further optimization test work.

Further optimization of the process focused on determining the effects of key process parameters such as grind size, residence time and feed solids concentrations on the leach kinetics, metal extractions and performance of the thermophile culture.

It was shown that the performance of the thermophile culture was negatively affected when operating at feed solids concentrations above 12%, and so a conservative feed solids concentration of 10% was selected.

Bioleach tests performed at 70°C at grind sizes ranging between 10 and 35 μm indicated that a nickel extraction of over 97% could be obtained over the range of grind sizes. There was, however, a reduction in copper extraction with increasing grind size, and copper extractions of 95% could only be achieved at a particle size of d_{90} = 10 μm and an overall six-day residence time (Table V).

Based on the results of the amenability test work, a set of bioleach operating parameters was set for the integrated piloting campaign. These are summarized in Table VI.

Bench-scale purification and metals recovery test work

Phase two of the overall metallurgical test work programme comprised bench-scale testing that evaluated the downstream unit operations: iron removal, copper SX, nickel precipitation, and solution purification. The results of these tests were used to set the initial operating conditions for the third phase of testing—the integrated pilot plant run.

In order to provide sufficient samples for the bench-scale test work, a continuous bioleach miniplant was operated in open circuit treating a sample of Aguablanca concentrate. The product from this plant was used to undertake a series of batch bench-scale tests. These tests aimed to assess each of the proposed unit operations, and in particular to evaluate various options for the precipitation of iron and nickel, and to carry out an initial assessment of the solid-liquid separation characteristics of the products from each process.

It was recognized that optimization of the precipitation processes can be undertaken only during continuous testing, where the impact of steady process operating conditions, and the use of recycle streams for seeding of the precipitation reactions, can be assessed. These factors are likely to have a direct and beneficial impact on the rate and extent of metal precipitation that can be achieved in practice, and on the settling and/or filtration characteristics of the resulting precipitates. The aim of these tests, therefore, was merely to provide the initial operating parameters for the integrated piloting campaign.

Table VII contains a summary of the initial operating parameters that were selected for each unit operation.

Integrated piloting test work

Integrated pilot plant

The pilot plant that was used for the integrated piloting tests is comprised of three main units: a bioleach plant, a precipitation plant, and a solvent extraction plant. The bioleach plant was operated continuously for the duration of the campaign. However, it was not possible to operate the precipitation and solvent extraction plants continuously throughout the campaign, because of the wide disparity between the

Table IV

Metal extractions achieved in three-stage continuously operated reactor systems, 6-day residence time, 10 % feed solids concentration, particle size of d_{90} = 10 μm

	Metal extraction (%)	
	Cu	Ni
Mesophiles (35°C)*	30	76
Moderate thermophiles (45°C)	65	99
Thermophiles (70°C)	95	99

* Single-stage reactor at 3-day residence time

Table V

Summary of bioleach results showing the effect of grind size on metal extractions

Operating conditions	Cu extraction (%)			Ni extraction (%)		
	R1	R2	R3	R1	R2	R3
3-stage, 70°C, 10 µm, 6 days	83.5	92.0	95.1	98.7	99.3	99.4
3-stage, 70°C, 20 µm, 6 days	72.4	82.0	90.7	96.3	97.4	97.4
1-stage, 70°C, 35 µm, 3 days	52.9			87.7		
Batch, 70°C, 35 µm, 6 days	72.8			98.1		

R1/2/3 = Reactor 1/2/3

Table VI

Bioleach operating parameters for Aguablanca nickel-copper concentrate

Operating conditions	
Microbial culture	Thermophiles
Operating temperature (°C)	70.0
Feed solids concentration (%)	10.0
Feed grind size (d_{90}) (μm)	10
Overall residence time (d)	6.0
Operating Eh level (mV vs. Ag|AgCl)	600
Operating pH level	1.3–1.1
Oxidation level and metal extractions	
Sulphide oxidation (%)	99.0
Copper extraction (%)	95.1
Nickel extraction (%)	99.4
Solution metal tenors	
Iron concentration (g/ℓ)	26.0
Copper concentration (g/ℓ)	8.0
Nickel concentration (g/ℓ)	6.5

Table VII

Initial operating parameters for integrated piloting of Aguablanca concentrate

Primary iron precipitation	
Temperature (°C)	60.0
Final pH level	3.0
Final iron tenor (g/ℓ)	0.45
Iron removal (%)	99.4
Limestone consumption (kg/t liquor)	25.0
Precipitate settling rate (m/h)	0.0382
Precipitate underflow density (%)	27.6
Calculated thickener area (m^2/(t/h))	69.1
Copper solvent extraction	
Organic extractant	LIX®984N-C
Diluent	ShellSol 2325
Extractant concentration (%)	25
Organic-to-aqueous ratio	0.8:1
Final copper tenor (g/ℓ)	0.163
Copper extraction (%)	97.5
Secondary iron precipitation	
Temperature (°C)	60.0
Final pH level	5.0
Final iron tenor (g/ℓ)	0.05
Iron removal (%)	98.3
Limestone consumption (kg/t liquor)	13.7
Precipitate settling rate (m/h)	6.0-7.8
Precipitate underflow density (%)	20-24
Calculated thickener area (m^2/(t/h))	2.0-4.3
Nickel precipitation	
Temperature (°C)	60.0
Final pH level	7.8
Final nickel tenor (g/ℓ)	~0.2
Nickel removal (%)	~97.1
Product nickel grade (%)	24.2
Magnesia consumption (kg/t liquor)	18.4
Precipitate settling rate (m/h)	9.3
Precipitate underflow density (%)	12.8
Calculated thickener area (m^2/(t/h))	4.6
Final (magnesium) precipitation	
Temperature (°C)	60.0
Final pH level	10.0
Final magnesium tenor (g/l)	0.0062
Magnesium removal (%)	99.5
Product magnesium grade (%)	12.7
Lime consumption (kg/t liquor)	4.65
Precipitate settling rate (m/h)	3.0
Precipitate underflow density (%)	7.0
Calculated thickener area (m^2/(t/h))	22.7

retention time in the bioleach plant and in the subsequent purification and metals recovery processes. To overcome this, the product from the bioleach plant was collected over a period of several weeks, and then used to feed the precipitation and solvent extraction units—which were then operated continuously, but for short periods. The same precipitation plant, comprising five equal-volume reactors arranged in series, was used for each of the successive precipitation processes.

The integrated pilot plant is illustrated in Figures 4 and 5.

The integrated pilot plant was operated for a period of 220 days, during which seven downstream solution purification and metals recovery cycles were completed. Each downstream cycle consisted of five additional unit operations, carried out in sequence. The product from each unit operation was stored, and used as the feed for the next unit operation in the sequence. The five unit operations were:

Figure 4—The integrated pilot plant: the solvent extraction plant is on the left, the bioleach pilot plant is in the centre, and the precipitation plant is on the right

Figure 5a—The integrated plant's bioleach reactors, with the solvent extraction plant in the background: on the right is the feed tank, the dark grey vessel is the 60-litre primary reactor, with three 20-litre reactors arranged in series to the left of that

Figure 5b—The integrated plant's neutralization reactors, comprising a feed tank (hidden) and five 20-litre reactors arranged in series; in the foreground are various settlers that were used during the operation of this plant

Figure 5c—The integrated pilot plant's solvent extraction units; this versatile plant has eight mixer-settlers that can easily be configured to provide the required number of extraction, washing and stripping stages

- Primary iron precipitation, in which the bioleach pulp was neutralized with limestone.
- Copper SX, treating the pregnant leach solution after primary iron removal.
- Secondary iron precipitation, in which the SX raffinate was neutralized with limestone.
- Nickel precipitation, in which nickel hydroxide was precipitated from the purified solution.
- Magnesium precipitation, to remove excess magnesium from the circuit.

At the completion of each cycle, the barren solution was recycled to the bioleach plant, by using it as the make-up water in the bioleach feed tank. In this way, full integration of the process was achieved. The use of recycled barren process solution for bioleach feed dilution was the only connection between the bioleach process and the downstream solution purification and metals recovery cycles.

Bioleach plant performance

The operation of the bioleach pilot plant is illustrated in Figures 6 to 12, which contain graphs of the routine data for several of the measured operating parameters.

In the early stages of the pilot-plant campaign, it was observed that the measured feed solids concentration was below the target value of 10%, as shown in Figure 6. This was despite the fact that the feed constituents (concentrate, nutrients and dilution water) were carefully metered to target this feed solids concentration. At first, it was suspected that this may be due to inadequate agitation in the feed tank, but even after this was addressed (by installing a larger impeller in the feed tank), the measured feed solids concentration remained below the target value. Inspection of the feed material revealed that, during the drying process after regrinding, the concentrate had formed persistent agglomerates that that did not break up in the feed tank. It is possible that residual flotation chemicals associated with the concentrate may have caused this. These large particles remained in the lower region of the feed tank, and were preferentially pumped into the primary bioleach reactor.

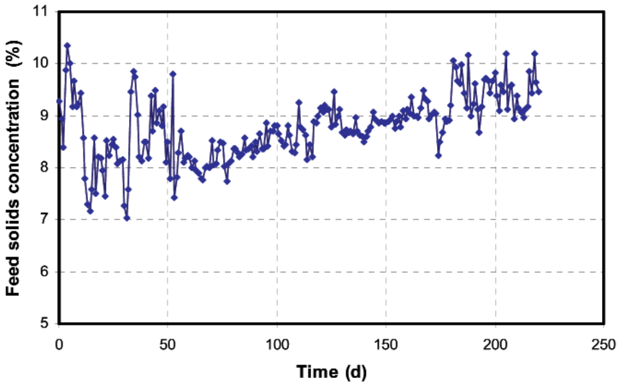


Figure 6—Bioleach feed solids concentrations during the integrated piloting campaign

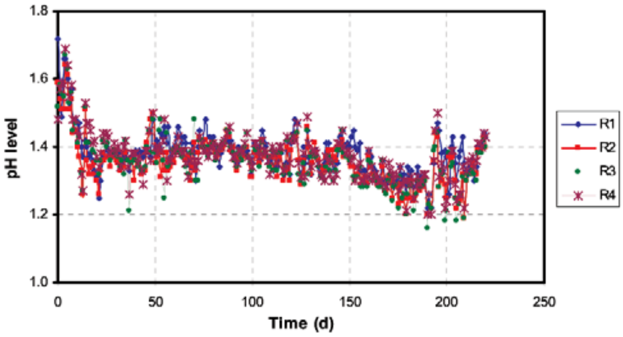


Figure 7—Measured pH levels during the integrated piloting campaign

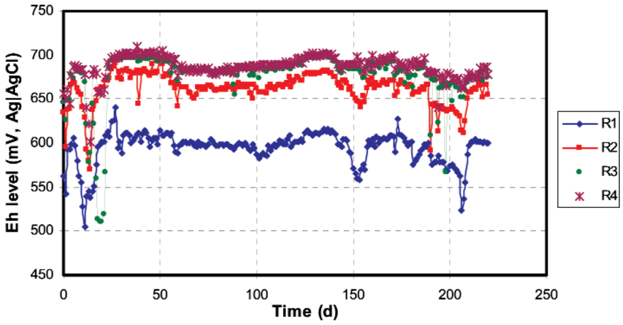


Figure 8—Measured Eh levels during the integrated piloting campaign

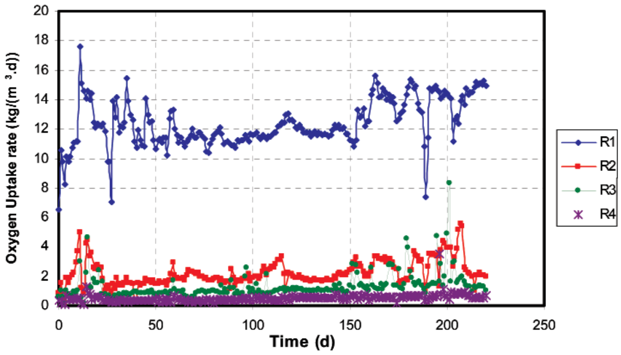


Figure 9—Measured oxygen uptake rates levels during the integrated piloting campaign

Thus, when the solids concentration in the feed tank was measured just before refilling the tank, the measured value was below the target value. To overcome this, the concentrate was given a light regrind prior to use, in order to break up the coarse agglomerates.

The rate of evaporation from the bioleach reactors was high, since they were being operated at a temperature of 70°C. At this scale of operation, the ratios of surface area to tank volume and of aeration rate to tank volume are very high. This results in much higher rates of evaporation than would be experienced in a production-scale process. To counter this, dilution water was pumped into each reactor to replace the water lost through evaporation. The rate of evaporation was not constant, and was affected by seasonal variations in the ambient humidity. Consequently, step changes were occasionally made to the rate of dilution water addition, which explains why the soluble metal concentrations shown in Figures 10 to 12 display several sudden changes during the course of the campaign. This had no material impact on the process performance, as evidenced by the stable redox potential levels (Figure 8) and oxygen uptake rates (Figure 9) for most of the campaign.

The bioleach plant operation was found to be stable and robust, and capable of withstanding upsets in plant operation. At various times, power, heater and compressed air failures were experienced, but the process recovered swiftly from these disturbances.

Four mass balance sampling campaigns were performed over the bioleach section of the plant during the integrated pilot-plant trial, allowing a detailed set of performance parameters to be calculated. The results of these mass balances were broadly similar. The results of one of these, performed on day 92, are summarized in Table VIII.

The results of the mass balances indicated that the target nickel and copper extractions of 99.4% and 95.1% were consistently achieved or exceeded. The target sulphide oxidation level of 99.0% was also consistently achieved.

Microbial identification

The microbial populations present in the bioleach reactors were identified and quantified using Q-PCR and T-RFLP techniques, respectively (Tables IX and X). *Acidianus sp.* was identified as the most abundant organism in all four stages, with *Metallosphaera sp.* and *Sulfolobus sp.* present in much lower numbers.

Additional bioleach optimization test work

Additional bench-scale bioleach test work was conducted to study the effect of redox potential, grind size and feed solids concentration on copper and nickel extractions and bioleach kinetics.

Although the most common means of overcoming the slow and incomplete extraction of copper from chalcopyrite is to operate the process at high temperature using thermophiles, an alternative approach is to operate at controlled redox potential levels. Previous test work has indicated that faster copper leach kinetics can be achieved by maintaining the redox potential at levels between 410 and 440 mV vs. Ag|AgCl, although a reduction in the rate of nickel extraction was observed at redox potential levels below 460 mV vs Ag|AgCl.

Based on these results, a three-stage continuously operated reactor system was operated at 70°C and an overall 6-day residence time, using Aguablanca concentrate milled to a particle size of d_{90} = 10 μm. The redox potential in the first-stage reactor was maintained at 430 mV vs. Ag|AgCl, while the redox levels in the secondary stages were not controlled. Redox potentials of 550 and 580 mV vs. Ag|AgCl were measured in the second- and third-stage reactors, respectively.

The results of this test are summarized in Table XI. This test confirms that, by controlling the redox potential in the first-stage reactor, considerably faster copper leach kinetics can be achieved. Around 96% of the copper was extracted in the first-stage reactor, compared to an extraction of 78% without redox control. The overall copper extraction improved from 95 to 98%. Furthermore, the high redox potential levels

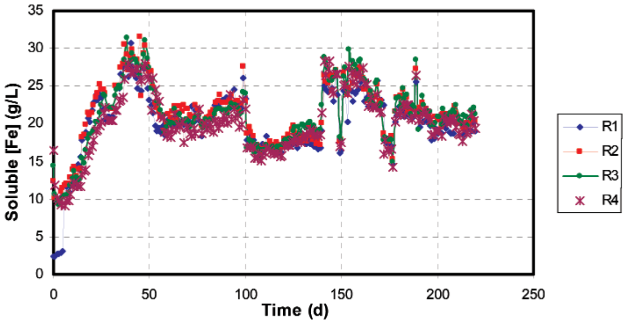


Figure 10—Measured soluble iron concentrations during the integrated piloting campaign

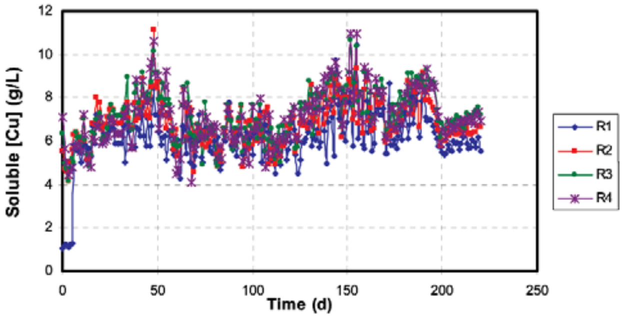


Figure 11—Measured soluble copper concentrations during the integrated piloting campaign

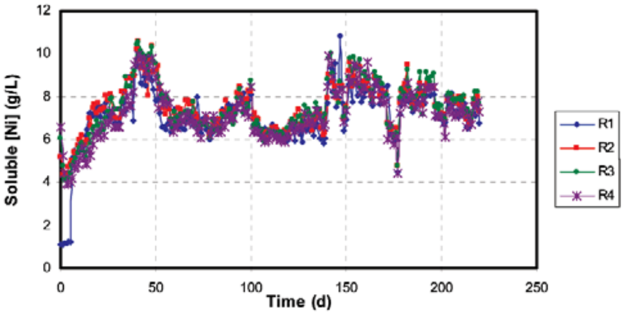


Figure 12—Measured soluble nickel concentrations during the integrated piloting campaign

Table VIII

Results of bioleach mass balance conducted on day 92

Reactor number	**Soluble analysis**				**Total (HCl-washed) analysis**			
	1	**2**	**3**	**4**	**1**	**2**	**3**	**4**
Solids concentration %	8.6							
Feed flow rate (ℓ/d)	20.2							
Product flow rate (ℓ/d)	19.8							
Residence time (d)	3.03	4.04	5.06	6.07				
Temperature (°C)	70	70	70	70				
pH level	1.35	1.33	1.34	1.35				
Redox (mV, Ag/AgCl)	599	667	685	687				
[Fe] (g/ℓ)	19.4	19.8	19.3	17.9				
[Ni] (g/ℓ)	6.2	6.3	6.2	5.8				
[Cu] (g/ℓ)	5.2	5.6	5.8	5.6				
Fe extraction (%)	65.6	64.7	66.7	65.9	92.5	92.6	94.3	95.8
Fe precipitation (%)					29.1	30.1	29.3	30.4
Ni extraction (%)	97.0	97.7	99.6	99.3	97.5	99.4	99.7	99.8
Ni precipitation (%)					0.46	1.69	0.17	0.21
Cu extraction (%)	78.4	86.7	92.7	95.0	81.7	87.8	93.0	96.5
Cu precipitation (%)					3.94	1.24	0.31	3.92
S^{2-} extraction (%)	93.8	98.2	98.9	99.4				
[S°] (%)	0.2	0.3	0.26	0.23				

Table IX

Quantification of the microorganisms using Q-PCR

Microorganism	**Cell concentrations (cells/ml)**			
	Reactor 1	**Reactor 2**	**Reactor 3**	**Reactor 4**
Acidianus sp.	2.7 x 10^{9}	6.1 x 10^{9}	4.5 x 10^{9}	3.2 x 10^{9}
Metallosphaera sp.	5.4 x 10^{7}	1.3 x 10^{8}	8.4 x 10^{7}	5.4 x 10^{7}
Sulfolobus sp.	7.6 x 10^{4}	9.9 x 10^{4}	6.4 x 10^{4}	ND

Table X

Relative abundance of the microbial populations occurring in the bioleach reactors

Sample	***Acidianus sp.* (%)**		***Metallosphaera sp.* (%)**		***Sulfolobus sp.* (%)**	
	Q-PCR	**T-RFLP**	**Q-PCR**	**T-RFLP**	**Q-PCR**	**T-RFLP**
Reactor 1	98.0	78	1.9	22	0.003	ND
Reactor 2	97.9	85	2.0	15	0.002	ND
Reactor 3	98.2	89	1.8	11	0.001	ND
Reactor 4	98.3	92	1.7	8	0	ND

*ND: not detected

Table XI

Effect of controlled low redox potential on bioleach process performance

	Feed	**Reactor 1**	**Reactor 2**	**Reactor 3**
Microbial culture	Thermophiles			
Temperature (°C)		70	70	70
pH level		1.6	1.3	1.25
Redox (mV, Ag|AgCl))		430	550	580
Cumulative residence time (d)		3	4.5	6
Feed solids concentration (%)	9.8			
Grind—d_{90} (µm)	10			
[Fe] (g/ℓ)		14.5	17.4	18.0
[Cu] (g/ℓ)		7.1	7.0	7.0
[Ni] (g/ℓ)		6.7	6.9	7.0
Fe extraction (%)		48.4	68.6	75.6
Cu extraction (%)		96.2	97.8	98.1
Ni extraction (%)		93.8	98.4	98.7
S^{2-} extraction (%)		67.3	94.2	98.8

maintained in the secondary reactors ensured that a nickel extraction of over 98% was achieved in the second-stage reactor. This result indicates that operating the first-stage reactor at a low redox potential level allows a reduction in the overall bioleach residence time from six to four days.

In a second test, the effect of concentrate grind size on the copper and nickel extractions was evaluated using concentrate milled to a d_{90} of 20 µm. As before, the redox potential in the first-stage reactor was maintained at a level of 430 mV vs. Ag|AgCl, while the redox potential levels in the secondary reactors were not controlled. As before, redox potential levels of 550 and 570 mV vs. Ag|AgCl were measured in the second- and third-stage reactors, respectively.

The results of the second test are summarized in Table XII. Despite the coarser feed material, a copper extraction of around 92% was achieved in the first-stage reactor, and an overall copper extraction of 96% was achieved. The high redox potential levels in the secondary reactors ensured that a nickel extraction of almost 98% was achieved. This result suggests that, with a coarser feed material and by controlling the redox potential in the first-stage reactor, the overall bioleach residence time can be reduced from six to five days.

Subsequent testing was conducted in which the feed solids concentration was raised to 12%, and it was found that stable operation could be maintained at this solids concentration.

Integrated purification and metals recovery tests

As mentioned previously, seven cycles of operation of the integrated purification and metals recovery circuit were completed during the integrated pilot plant campaign.

Primary iron precipitation

During the commissioning of the plant, it was established that it was possible to conduct the primary iron precipitation on the bioleach product pulp, without first having to separate and wash the bioresidue. Since this removed a solid-liquid separation and washing operation from the flowsheet, it was decided to proceed on this basis for the remainder of the integrated piloting campaign.

The primary iron precipitation process was operated at a variety of temperatures between 45 and 70°C. The pH level in the first-stage reactor was maintained at a level of between 1.8 and 2.9, and the final pH was maintained at a level of between 3.0 and 3.5. The neutralizing agent was limestone. Recycling of the product was used to promote seeding in the first reactor. Various recycle ratios were used.

It was found that the product pulp from the primary iron precipitation process was very difficult to settle, irrespective of the operating temperature and the recycle ratio. It was also found that, as the operating temperature varied, the settling characteristics changed to such an extent that different flocculants were required to achieve the best result. Although some flocculants achieved good settling rates, it was observed that the product liquor was turbid and unsuitable as a feed to the copper solvent extraction plant. The clarity of the settler overflow was improved considerably by the addition of a coagulant. It was shown that the coagulant could be used on its own or in conjunction with a neutrally-charged flocculant. However, large coagulant doses of over 1 kg/t were required, compared with around 3–30 g/t for flocculants.

The best set of results resulted in a batch settling rate of 282 mm/h. In this run, the settler overflow was clear and relatively small doses of coagulant and flocculant were required.

Copper solvent extraction

The copper SX plant was a mixer-settler unit operated using a mixture of 25% LIX®984N-C in ShellSol 2325 as the organic extractant. LIX®984N-C, which is manufactured by Cognis, is a reagent that has been developed especially for use with pregnant liquors containing high copper concentrations, and ShellSol (made by Shell Chemicals) is a paraffin-based hydrocarbon solvent. The sulphuric acid concentration in the strip liquor was maintained at 180 g/ℓ.

The flow rate through the SX plant is determined by a required minimum residence time of three minutes in the mixer, taking account of both the organic and aqueous phases. The second parameter is the loading capacity of the

Table XII

Effect of grind size and controlled low redox potential on bioleach process performance

	Feed	Reactor 1	Reactor 2	Reactor 3
Microbial culture	Thermophiles			
Temperature (°C)		70	70	70
pH level		1.7	1.4	1.3
Redox (mV, Ag|AgCl))		430	550	570
Cumulative residence time (d)		3	4.5	6
Feed solids concentration (%)	9.4			
Grind—d_{90} (µm)	20			
[Fe] (g/ℓ)		13.6	16.0	19.2
[Cu] (g/ℓ)		7.3	7.4	7.6
[Ni] (g/ℓ)		6.7	7.0	7.5
Fe extraction (%)		40.7	63.1	70.1
Cu extraction (%)		92.7	95.8	96.0
Ni extraction (%)		85.1	96.9	97.9
S^{2-} extraction (%)		63.7	84.7	94.2

organic, specified by the manufacturer as being 0.3 g/% or 7.5 g/ℓ copper for a 25% solution of LIX®984N-C. This, together with the copper concentration in the pregnant leach solution (PLS) determines the organic-to-aqueous (O:A) ratio. The strip liquor flow rate is determined by the electrowinning section, where a reduction of 15 g/ℓ copper is required. In the stripping section of the SX plant, this quantity of copper must be replaced, thus setting the organic-to-strip (O:S) ratio. In this instance an O:S ratio of two is typical.

Provided that the O:A ratio was maintained above about 0.7:1, copper extractions between 98.2 and 98.9% were achieved using three extraction stages.

Copper electrowinning

At the completion of the pilot plant campaign, a sample of the strip liquor (which had built up to a copper concentration of around 50 g/ℓ) was used to perform an electrowinning test, using a standard set of operating conditions. Two small copper cathodes (each with an area of 0.0315 m^2) were produced in this test. They were characterized as being smooth, compact and bright. One cathode was drilled and submitted for chemical analysis, to determine whether the copper cathode meets the requirements for LME A-grade copper. Table XIII summarizes the results.

The detection limits for the analyses of the elements P and S are higher than the threshold values. The measured values of the other elements are well below the LME A-grade threshold values, and the sum (excluding P and S) of 16.07 ppm for these elements is well below the threshold value of 65 ppm. The LME A-grade specification also provides upper limits for several elements in combination, and once again the measured values (excluding P) are well below the threshold limits.

These results indicate that, unless the cathode contains over 15 ppm of either S or P, it will easily satisfy the requirements for LME A-grade classification.

Secondary iron precipitation

The purpose of the secondary iron precipitation process was to reduce the concentration of iron and other metals in the pregnant liquor as far as possible prior to the precipitation of nickel hydroxide, in order to reduce the level of contamination of the nickel hydroxide precipitate by these elements. This was achieved by raising the pH of the liquor to a level of around 5.0 using limestone, at a temperature of 60°C. Recycling of the product was used to promote seeding in the first reactor. Various recycle ratios were used.

The product from this process was mostly gypsum, and the product pulp had a very low solids concentration (of between about 2 and 4 %). The gypsum particles in the product solids were large, and were easily settled.

Copper and nickel scavenging

There was some loss of both copper and nickel to the precipitated solids in the secondary iron removal process. Gravity separation of the product was investigated as a means of upgrading the copper and nickel. Three options were evaluated: a shaking table, a Falcon concentrator, and a cyclone.

The shaking table was ineffectual, with around 90% of the copper and nickel reporting to the slimes, which comprised over 60% of the mass. The same was true of the Falcon concentrator, where over 99% of the copper and nickel reported to the tailings, which comprised over 90% of the mass.

The cyclone proved effective in upgrading the copper and nickel. Over 75% of the copper and over 80% of the nickel could be recovered in the cyclone overflow, at a mass recovery of 18.4%. It is likely that further optimization of this process, with a slightly coarser cut and with underflow recycle to scavenge the tails, would further improve this result.

The cyclone overflow would be recycled back to the bioleach process to redissolve the copper and nickel.

Nickel precipitation

Nickel precipitation to produce nickel hydroxide was achieved by raising the pH of the liquor to a level of around 7.8 using magnesia, at a temperature of 60°C. Recycling of the product was used to promote seeding in the first reactor. Various recycle ratios were used.

The objective of this process was to produce a nickel hydroxide product of the highest possible purity, with reasonable filtration characteristics, and to ensure a high recovery of nickel. This presents a challenge: a good nickel recovery is obtained at pH levels in the region of 7.7 to 8.0, but as the pH increases above a level of about 7.5, the product becomes increasingly contaminated with magnesium—either in the form of unreacted magnesia or precipitated magnesium hydroxide.

One approach that has been adopted is to conduct a two-stage nickel hydroxide precipitation. The first-stage process is operated at a target pH level of around 7.2, and aims to

Table XIII

Aguablanca copper cathode analysis

Element(s)	LME A grade threshold (ppm)	Measured value (ppm)
Ag	25	0.17
As	5	0.00
Bi	2	0.53
Cd	-	0.00
Co	-	0.00
Cr	-	0.00
Fe	10	0.00
Mn	-	0.50
Ni	-	0.00
P	-	<20
Pb	5	0.00
S	15	<100
Sb	4	0.00
Se	2	1.17
Si	-	12.67
Sn	-	0.50
Te	2	nd
Zn	-	0.53
Sum	65	16.07
As+Cd+Cr+Mn+P+Sb	15	0.50
Bi+Se+Te	3	1.70
Se+Te	3	1.17
Co+Fe+Ni+Si+Sn+Zn	20	13.70

recover around 90% of the nickel and produce a high-grade nickel hydroxide product. The second-stage process is operated at a pH level of between 7.5 and 8.0, and aims to recover the remainder of the nickel in a lower-grade nickel hydroxide precipitate. This product can either be recycled to the leach process or sold at a lower cost.

A second approach is to conduct the precipitation in multiple stages with careful control of the pH level throughout the process, to ensure that both targets—recovery and product purity—are met. This approach, if achievable, has the advantage of reducing the number of unit operations required, and producing a single high-grade nickel hydroxide product. It was therefore decided to aim for a single-stage nickel hydroxide precipitation process.

In the piloting campaign, the solids were settled in order to provide a recycle stream for seeding purposes. In practice, the nickel hydroxide product will be filtered to minimize the water content. It was observed that nickel hydroxide precipitation required a considerable amount of seeding to initiate the precipitation process—more so than the iron removal steps. The product was observed to flocculate well, although the overflow remained slightly turbid due to the presence of very small particles in suspension. When part of the settler underflow was recycled for seeding, nickel recoveries of well over 99% were achieved.

It was also noted that a fresh magnesia slurry was required for the best performance—it was found that when an older magnesia slurry was used, the product nickel grade was lower.

The nickel content of the nickel hydroxide precipitate varied between 31.6 and 47.5%. This variance reflects the fact that the operating conditions in the nickel recovery process were changed in an effort to understand the impact of the operating conditions on the process. Based on the results, it is considered that a target nickel grade of between 40 and 45% can be achieved using a single-stage precipitation process.

Magnesium precipitation

The final step in the process flowsheet is magnesium precipitation, which was achieved by raising the pH of the liquor to a level of around 9.5 using slaked lime, at a temperature of 60°C. Recycling of the product was used to promote seeding in the first reactor. Various recycle ratios were used. In some cases, the product was settled prior to recycling, whereas in others the recycle stream comprised the product pulp.

The objective of this process step was to remove magnesium and most other contaminants from the stream, to enable it to be recycled to the front of the process. The treatment of a bleed stream could be considered, provided that the presence of some magnesium could be tolerated in the bioleach process. Bioleach amenability tests had shown that up to 10 g/ℓ of magnesium could be tolerated, but there were concerns that the presence of high concentrations of magnesium in the iron removal, copper solvent-extraction and nickel hydroxide processes would be undesirable. Therefore, it was decided to process the entire stream through the magnesium precipitation process.

It was observed that the magnesium product flocculated relatively easily, but that the overflow contained some fine particles that did not settle.

Upgrading of magnesium precipitate

An attempt was made to upgrade the magnesium precipitate using a cyclone, in order to assess whether the upgraded product might be used in the nickel hydroxide precipitation process. This test was only partially successful, in that around 40% of the magnesium was recovered in the cyclone overflow, which had a mass recovery of 11.5%. The upgraded product had a magnesium grade of 22.2%, but also contained 12.5% calcium, indicating that gypsum had also reported to the overflow. If this was to be used in the nickel precipitation process, the gypsum would contaminate the nickel hydroxide product, and it is therefore not proposed to utilize this upgraded product in the nickel precipitation process.

Summary of integrated piloting tests

The results of the integrated piloting tests were used to establish the basic process design criteria for the proposed bioleaching and metals recovery plant for the treatment of the Aguablanca concentrate. The design criteria are summarized in Table XIV.

The piloting programme demonstrated that integration of the bioleaching process with the downstream purification and metals recovery processes had no impact on the bioleach process performance. Therefore, the design criteria for the bioleaching process remained unchanged from those obtained in the amenability tests.

These basic process design criteria were used as the basis for the process design and costing of the commercial-scale plant.

Process design and costing

The flowsheet and data derived from the integrated piloting of the Aguablanca nickel-copper concentrate were used to develop a process design and costing model for a hydrometallurgical plant consisting of bioleaching and associated purification unit operations. The data from the integrated piloting test work were used as inputs for the model. In some instances, most notably in sizing the settling and filtration equipment, in-house database information was used to obtain realistic settling and filtration rates. The settling and filtration tests that were performed during the integrated pilot plant campaign were limited in scope and extent, and were not optimized. It is therefore considered that the in-house database information is more reliable than the test work data.

Process description

The proposed hydrometallurgical plant will be situated at the site of the existing Aguablanca concentrator in southern Spain. The plant will be designed to treat 96 000 t/a of Aguablanca concentrate, producing 4 857 t/a of nickel in an intermediate precipitate and 5 400 t/a of copper cathode. The design criteria for the plant are summarized in Table XV.

Milling

The concentrate is repulped to between 50 and 60% solids, and fed to a vertical bead mill, where it is reground to achieve a particle size distribution with a d_{90} of 10 μm. Regrinding of the concentrate to this extent is required to achieve suitable

Table XIV

Basic process design criteria based on integrated piloting results

Bioleaching	
Bacterial culture	Thermophiles
Operating temperature (°C)	70.0
Feed solids concentration (%)	10.0
Feed grind size (d_{90}) (μm)	12
Overall residence time (d)	6.0
Operating Eh level (mV vs. Ag|AgCl)	650
Operating pH level	1.4–1.2
Sulphide oxidation (%)	99.0
Copper extraction (%)	95.1
Nickel extraction (%)	99.4
Primary iron precipitation	
Temperature (°C)	60.0
Number of stages	5
Residence time per stage (h)	1
Final pH level	3.0
Final iron tenor (g/ℓ)	0.09
Iron removal (%)	99.5
Limestone consumption (kg/m^3 slurry)	60
Thickener underflow density (%)	30–40
Calculated thickener area (m^2/(t/h))	10-20
Copper solvent extraction	
Temperature (°C)	35
Organic extractant	LIX®984N-C
Diluent	ShellSol 2325
Extractant concentration (%)	25
Organic-to-aqueous ratio	1:1
Number of extraction stages	3
Number of scrubbing stages	2
Number of stripping stages	3
Final pH level	1.1–1.2
Final copper tenor (g/ℓ)	0.1
Copper extraction (%)	97.0
Secondary iron precipitation	
Temperature (°C)	60.0
Number of stages	5
Residence time per stage (h)	1
Final pH level	5.0
Final iron tenor (g/ℓ)	0.00
Iron removal (%)	95.0
Limestone consumption (kg/m^3 slurry)	12.5
Filter cake moisture (%)	46.5
Filtration rate (m^2/(t/h))	62.4
Nickel precipitation	
Temperature (°C)	60.0
Number of stages	5
Residence time per stage (h)	1
Final pH level	7.8
Nickel precipitation	
Final nickel tenor (g/ℓ)	0.02
Nickel removal (%)	99.5
Product nickel grade (%)	40.0–45.0
Magnesia consumption (kg/m^3 slurry)	4.0
Thickener underflow density (%)	20–25
Calculated thickener area (m^2/(t/h))	2.0–3.0
Filter cake moisture (%)	55.2
Filtration rate (m^2/(t/h))	126
Final (magnesium) precipitation	
Temperature (°C)	60.0
Number of stages	5
Residence time per stage (h)	1
Final pH level	9.5
Final magnesium tenor (g/ℓ)	0.00
Magnesium removal (%)	98.8
Product magnesium grade (%)	10.2
Lime consumption (kg/m^3 slurry)	10.0
Filter cake moisture (%)	64.0
Filtration rate (m^2/(t/h))	166

Table XV

Design criteria used in the costing model

Plant Capacity	t/a Ni	4,857		
	t/a Cu	5,401		
	t/h concentrate	12.00		
	t/d concentrate	288.0		
	t/h S^{2-} oxidized	2.75		
Metal dissolution				
Ni	%	99.5		
Cu	%	95.0		
Fe (leach - ppt)	%	62.0		
S	%	93.9		
Leach				
Residence time, primaries	d	3.0		
Residence time, secondaries	d	3.0		
Total residence time	d	6.0		
Temperature	°C	70		
Feed solids concentration	%	10		
O_2 consumption	kg/kg S	1.96		
$CaCO_3$ addition to leach	kg/t conc.	15.4		
Primary bioleach reactors				
Number of modules	#	2		
Number of reactors per module	#	3		
Tank material		Steel or Stebbins		
Tank volume	m^3	1625		
Motor power per tank	kW	81		
Oxygen utilization	%	39		
Secondary bioleach reactors				
Number of modules	#	2		
Number of reactors per module	#	3		
Tank material		Steel or Stebbins		
Tank volume	m^3	1,625		
Motor power per tank	kW	34		
Oxygen utilization	%	26		
Air blowers				
Number	#	12		
Blower capacity	m^3/h	4,341		
Total air requirement, bioleach + iron removal	Nm^3/h	52,096		
Total air requirement, bioleach	Nm^3/h	51,916		
or	$Nm^3/t\ S^{2-}$	18,911		
Cooling tower				
Number	#	1.00		
Cooling water flow rate	m^3/h	363		
Heat load bioleach reactors	kW	-8,392		
Fe removal/CCD				
Temperature	°C	60		
Number of stages	#	5		
Residence time per stage	h	1		
Tank volume	m^3	147.6		
Solids rate	t/h	24.38		
Settling rate	$(t/h)/m^2$	0.24		
Settler area per CCD	m^2	11.37		
Number of CCD stages	#	4		
$CaCO_3$ consumption	kg/t conc.	559		
Cu SX				
Reagent		LIX 984		
Reagent concentration	%	11		
		Extraction	Scrubbing	Stripping
O/A ratio		1.00	45.0	4.94
Mixer residence time	min	3	3	3
Number of stages	#	2	1	2
Settling rate for phase separation	$m^3/(m^2.h)$	5.00		
Cu recovery	%	97		
EW power	kWh/t Cu	2,200		
Secondary Fe removal				
Temperature	°C	60		
Number of stages	#	5		
Residence time per stage	h	1		
Tank volume	m^3	241.4		
Solids rate	t/h	1.66		
Filtration rate	$(t/h)/m^2$	0.48		
Filtration area	m^2	3.46		
$CaCO_3$ consumption	kg/t conc.	102		

Table XV (continued)

Design criteria used in the costing model

Ni hydroxide precipitation				
Temperature	°C	60		
Number of stages	#	5		
Residence time per stage	h	1		
Tank volume	m^3	256.4		
Solids rate	t/h	1.32		
Filtration rate	(t/h)/m^2	0.09		
Filtration area	m^2	14.65		
Settling rate	m^2/(t/h)	2.08		
Settler diameter	m	1.87		
MgO consumption	kg/t conc.	55.1		
Mg precipitation				
Temperature	°C	60		
Number of stages	#	5		
Residence time per stage	h	1		
Tank volume	m^3	269.0		
Solids rate	t/h	2.48		
Filtration rate	(t/h)/m^2	0.48		
Filtration area	m^2	5.17		
$Ca(OH)_2$ consumption	kg/t conc.	71.3		
Alternative Ni recovery option 1				
S^{2-} precipitation				
Temperature	°C	90		
Pressure	kPa	600		
Number of stages	#	1		
Number of modules	#	2		
Residence time per stage	h	0.5		
Autoclave volume	m^3	123.9		
Precipitate	t/h	0.95		
Settling rate	m^2/(t/h)	3.00		
Settler diameter	m	3.82		
Filtration rate	(t/h)/m^2	0.48		
Filtration area	m^2	1.99		
H_2S consumption	kg/t conc.	29.8		
Post-S2- precipitation neutralization				
Temperature	°C	70		
Number of stages	#	5		
Number of modules	#	1		
Residence time per stage	h	1		
Volume per tank	m^3	243.4		
Precipitate	t/h	1.43		
Settling rate	m^2/(t/h)	3.00		
Settler diameter	m	2.34		
$Ca(OH)_2$ consumption	kg/t conc.	65.0		
Alternative Ni recovery option 2:				
Ni SX				
Extractant		Versatic acid in n-paraffin		
Extractant	%	6.7		
		Extraction	**Scrubbing**	**Stripping**
O/A ratio		1.25	5.50	5.50
Mixer residence time	min	4.00	6.00	6.00
Number of stages	#	4.00	3.00	3.00
Settling rate for phase separation	m^3/(m^2.h)	5.00		
Extraction pH		6.25-6.40		
NaOH consumption	kg/t conc.	69.0		
Ni recovery	%	100		
	kWh/t Ni	3,700		

copper recoveries in the bioleaching process. For the design capacity, it is considered that the regrinding will be achieved in a mill similar to a Deswik 1000 vertical bead mill. The Deswik 1000 is fitted with up to 24 polyurethane impellers on a vertical shaft. The milling medium comprises 1 mm zirconia beads. This mill is similar in concept to the IsaMill, with the principal difference being the orientation of the mill: the IsaMill is a horizontal bead mill, whereas the Deswik has a vertical orientation. Bead mills have been found to be more energy efficient than stirred-media mills such as the Svedala VertiMill and the Metso Minerals Stirred Media Detritor (SMD), particularly for ultra-fine grinding to particle sizes below a d_{90} of around 20 μm. Based on information obtained from the Deswik mill supplier in South Africa, it is estimated that the power consumption required to achieve the target particle-size distribution in a Deswik 1000 vertical bead mill is about 14.3 kWh/t.

The reground pulp is delivered to a pulp storage tank, with a sufficient capacity to allow the bioleach plant to be fed for two days.

Bioleaching

The bioleach feed pulp is diluted to the required solids concentration of 10% (m/m), using fresh water or water recycled from the downstream process. Appropriate control systems are utilized to maintain the desired pulp feed rate and solids concentration.

The bioleach reactors are operated at a temperature of 70°C, and consist of two modules, each comprising six reactors. Each module comprises three primary reactors operated in parallel, and three secondary reactors operated in series. The feed pulp is evenly distributed to the primary reactors to ensure a uniform feed to each primary reactor.

The overall residence time of the bioleaching process is maintained at six days, and each individual reactor has a volume of approximately 1 620 m^3. The bioleach reactors are mechanically-agitated, aerated vessels, made from an appropriate material—either a duplex stainless steel or a ceramic-lined concrete—to enable them to withstand the corrosive, acidic process conditions. The reactors are agitated using specialized impellers that promote the dispersion of large volumes of air and maintain the solids in suspension. Aeration is via specially-designed aerators that are positioned in the proximity of the impellers. The oxidation reactions taking place in the reactors are exothermic, and the temperatures of the reactors are controlled at 70°C using cooling water that is circulated through a cooling tower. In some of the secondary reactors, where the reaction rates may be lower, some heating may be required, which is achieved by live steam injection.

The product from the two bioleaching modules, comprising a pulp with a pH level of around 1.6, containing unleached gangue minerals, some iron precipitates, and a liquor containing sulphuric acid, ferric sulphate, copper sulphate and nickel sulphate, is combined and delivered to the primary iron removal plant.

Primary iron removal

The primary iron precipitation takes place in a series of five neutralization reactors, in which the temperature is maintained at 60°C. Limestone slurry is used as the neutralizing agent. The pH is raised to a level of 3.0, and the overall residence time is 5 hours. The main products from the neutralization reaction are gypsum (from acid neutralization) and ferric hydroxide (from iron precipitation).

The product pulp from the primary iron removal reactors is fed to a CCD circuit, comprising four thickeners. In this circuit, wash water is used to wash the product solids. The liquor from the CCD circuit, containing copper sulphate and nickel sulphate, is fed to the copper SX plant.

Copper solvent extraction and electrowinning

The SX plant receives the pregnant leach solution (PLS) from the primary iron removal circuit, and produces a rich electrolyte that is sent to the electrowinning plant. Copper SX is performed in two extraction, one scrubbing, and two stripping stages. In the extraction stages, copper is extracted into the organic phase. The organic phase comprises a mixture of 11% LIX®984N-C in ShellSol 2325 as the organic extractant (based on a specification of 0.29 g/ℓ per volume % in the organic phase). Although a concentration of 25% organic was used in the pilot-plant test work, the process model takes into account dilution from the CCD wash, which results in a lower copper tenor and organic concentration. An organic-to-aqueous ratio of 1:1 is maintained in the extraction stages. The tailing (raffinate) from the extraction stages, which contains nickel sulphate, forms the feed to the secondary iron removal circuit.

The loaded organic is fed to the stripping section, where it is contacted with spent electrolyte from the electrowinning section, typically containing around 180 g/ℓ of sulphuric acid. The copper is stripped from the organic phase into the electrolyte, which forms the feed to the electrowinning plant. The barren organic phase is returned to the extraction stages.

In the electrowinning section, the rich electrolyte from the copper solvent extraction plant is fed to a series of conventional electrowinning cells, where the copper is plated out on stainless steel cathodes to produce LME A-grade copper. The electrowinning plant comprises a series of cells, with non-soluble lead anodes and 316L stainless steel cathodes, a rectifier, an overhead crane to handle the cathodes, an automatic cathode stripping machine and accessory equipment. The spent electrolyte is returned to the stripping section of the copper SX plant.

Secondary iron removal

The copper SX raffinate is fed to the secondary iron removal circuit, where the pH is raised to a level of 5.0 using a limestone slurry in order to purify the leach solution prior to nickel recovery. The plant comprises five reactors in series, operated at a temperature of 60°C. In this process, the acid produced in the copper SX process is neutralized, and any residual iron is precipitated. Additional elements that are removed in this process are aluminium and chromium, with partial removal of silicon and zinc. Any copper remaining in the raffinate is also precipitated, and there may be a small loss of nickel sulphate. The main products from the neutralization reaction are gypsum (from acid neutralization) and ferric hydroxide (from iron precipitation).

The product pulp from the secondary iron removal reactors is fed to a continuous filter, where wash water is added. The filtrate, containing nickel sulphate, is fed to the nickel precipitation plant. The filter cake is repulped and fed to a cyclone, with the objective of concentrating and recovering most of the precipitated copper and nickel. The cyclone overflow, containing 20% of the mass and about 75–80% of the copper and nickel, is recycled to the bioleach circuit. The cyclone underflow is disposed of.

Nickel precipitation

Three downstream nickel recovery options were investigated.

- ➤ *Hydroxide precipitation*—this is the base case, and is the one that was piloted. Nickel is precipitated as an hydroxide by the addition of magnesia. This the least sophisticated option, as it generates a lower-quality product, which is difficult to dewater and therefore more expensive to transport. It is environmentally more benign than the other routes, provided inexpensive supplies of lime and magnesia can be obtained.
- ➤ *Sulphide precipitation*—through the addition of hydrogen sulphide. This process involves a higher degree of complexity, requiring pressurized reactors and a low cost-source of hydrocarbon or H_2S. This route also has environmental concerns, owing to the use of pressurized reactors and the production of toxic

and flammable gases, and it requires a more sophisticated workforce. However, the product has a lower moisture content than the hydroxide product, and is therefore less expensive to transport.

- *Solvent extraction and electrowinning*—of nickel cathode is the most sophisticated and complex route, requiring a skilled workforce. A reliable power supply is required. This route is suitable if refining charges are high and the additional capital expenditure can be justified.

Nickel hydroxide precipitation

In this route, the filtrate from the secondary iron removal circuit is fed to the nickel precipitation plant, which comprises five reactors operated in series, each at a temperature of 60°C. The pH is raised to a level of 7.8 using a magnesia slurry, and the overall residence time is five hours. The nickel precipitates as nickel hydroxide, and some unreacted magnesia and magnesium hydroxide also report to the solids. The aim is to produce a nickel hydroxide product containing at least 45% nickel and less than 10% magnesium (measured on a dry basis). The product pulp is fed to a settler, and the settler underflow is further dewatered with a continuous Larox-type filter, where the aim is to reduce the moisture content of the nickel hydroxide product to below 80%.

The barren filtrate, containing magnesium hydroxide that was formed in the nickel precipitation process, is fed to the magnesium removal circuit in order to purify the process solution prior to recycling. Magnesium precipitation is performed in a plant comprising five reactors operated in series, each at a temperature of 60°C. The pH is raised to a level of about 9.5 using a slaked lime slurry, and the overall residence time is five hours. The product pulp from the magnesium removal reactors is filtered, the barren filtrate is recycled, and the filter cake, containing mainly magnesium hydroxide and gypsum, is disposed of.

Nickel sulphide precipitation

Nickel sulphide precipitation is carried out in two autoclaves operated in parallel, with a 30-minute residence time in each. The feed solution is heated to 90°C, and H_2S gas is introduced at a pressure of 600 kPa. The product is fed to a settler, and the underflow is filtered to produce a dewatered nickel sulphide product. The settler overflow is fed to a neutralization plant, consisting of five reactors in series, with a total residence time of five hours. The solution is neutralized with slaked lime to a pH level of 9.5 in order to neutralize the free acid and precipitate impurities. The gypsum product is fed to a settler, and the settler overflow is recycled to the bioleach process.

Nickel solvent extraction and electrowinning

Nickel SX consists of four extraction, three scrubbing, and three stripping stages. The organic extractant is 7% Versatic acid in a paraffin diluent. The O:A ratio is 1.25:1 for extraction, 5.5:1 for scrubbing and 5.5:1 for stripping. Extraction is carried out at a pH level of 6.5, using sodium hydroxide (NaOH) as the neutralizing agent. For the purposes of this study, it was assumed that nickel SX is performed in closed circuit with the leaching circuit. However, this approach may not be technically feasible, owing to the build-up of the SX neutralizing agent (NaOH or NH_4OH). It may be necessary to precipitate and redissolve the nickel, followed by SX in a separate purification circuit, as is done in most laterite pressure leaching circuits. Such an option would add significantly to the capital and operating costs, and was not considered in this study.

Process costing and financial analyses

Mass and energy balances were developed for the overall hydrometallurgical circuit, followed by the design, sizing and costing of major equipment items, using in-house equipment cost indices. The equipment costs were factored to estimate the installed capital costs. Capital costs were calculated from installed equipment costs and factored indirect costs. Operating costs were calculated from stoichiometric reagent consumptions, estimated labour rates, and calculated power consumptions.

Capital cost estimate

Capital costs (basis: 2007/2008) were calculated for the three proposed process flowsheets, covering the three different nickel recovery options. Based on the mass balance and design specifications, major equipment items such as tanks, thickeners, filters and compressors were sized and costed using an in-house Mintek equipment cost database. Installation and auxiliary items were estimated by multiplying the equipment costs by the appropriate factors.

Indirect capital costs were calculated by multiplying the total installed capital cost by appropriate factors. The total project cost is the sum of the total installed capital cost and the indirect costs, plus a contingency. The breakdown of capital costs is given in Table XVI for the three purification options. Capital costs were approximately US$5.7/lb annual nickel for the hydroxide and sulphide precipitation options, and approximately US$6.1/lb annual nickel for the SX-EW option.

Operating cost estimate

Operating costs were calculated from the stoichiometric reagent consumptions, estimated reagent costs, calculated power consumption and labour rates. The cost for maintenance was included in the labour element, and an additional 6% of the total installed capital cost was provided for maintenance materials and consumables. The breakdown of operating costs for the various components is shown in Figure 13.

Figure 14 expresses the operating cost elements for four different refining options: concentrate sale, nickel hydroxide production, nickel sulphide production and nickel cathode production. In each case, the value of refining charges and by-product credits was included in the operating cost breakdown. The cost of mining and concentrate production of US$3.8/lb nickel was obtained from the most recent published data for the Aguablanca operation (2007 data).

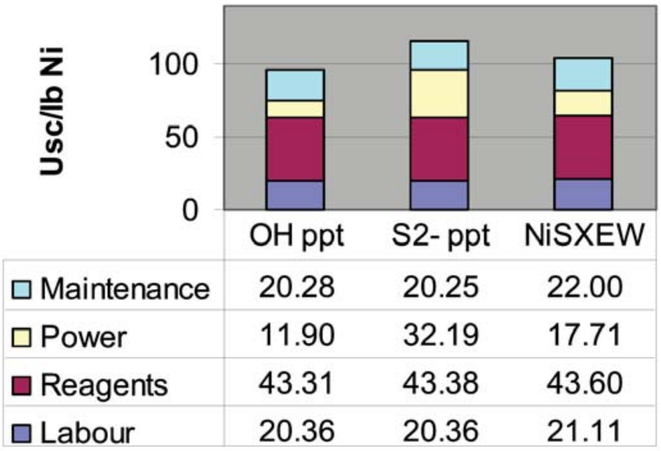
Financial analysis

For the base case financial analysis, it is assumed that capital is spent in the first two years. A mining rate of 1.5 million tonnes per annum is assumed, and a concentrate production rate of 96 000 t/a. Depreciation of 33% per annum and a

Table XVI

Capital cost breakdown

Item $ '000	OH ppt $ '000	S2- ppt $ '000	Ni SXEW $/a lb Ni	OH ppt $/a lb Ni	S2- ppt $/a lb Ni	Ni SXEW $/tpa conc.	OH ppt $/tpa conc.	S2- ppt
Fine grinding	1 920	1 920	1 920	0.18	0.18	0.18	20	20
Bioleach utilities	1 937	1 937	1 937	0.18	0.18	0.18	20	20
Bioleach	11 497	11 497	11 497	1.07	1.07	1.07	120	120
CCD/residue filter	4 702	4 702	4 702	0.44	0.44	0.44	49	49
Fe removal I	620	620	620	0.06	0.06	0.06	6	6
Fe removal II	2 238	2 238	2 238	0.21	0.21	0.21	23	23
Ni-OH precipitation	3 231			0.30			34	
Mg precipitation	2 368			0.22			25	
Ni-S^{2-} precipitation		4,029			0.38			42
Neutral'sn - S^{2-} ppt		1 515			0.14			16
NiSX			3 005			0.28		
NiEW			5 666			0.53		
CuSX	1 778	1 778	1 778	0.17	0.17	0.17	19	19
Cu EW	5 903	5 903	5 903	0.55	0.55	0.55	61	61
Reagents and services	960	960	960	0.09	0.09	0.09	10	10
Tailings neutralization	960	960	960	0.09	0.09	0.09	10	10
Infrastructure	4 235	4 235	4 235	0.40	0.40	0.40	44	44
Total directs (TIC)	**42 349**	**42 287**	**45 761**	**3.95**	**3.95**	**4.27**	**441**	**440**
EPCM	6 352	6 343	6 864	0.59	0.59	0.64	66	66
Site temp facilities	847	846	915	0.08	0/08	0.09	9	9
Technology fees	423	423	458	0.04	0.04	0.04	4	4
First Fill	1 133	1 133	1 133	0.11	0.11	0.11	12	12
Owner's cost	4 235	4 229	4 576	0.40	0.39	0.43	44	44
Total Indirects	**12 991**	**12 974**	**13 946**	**1.21**	**1.21**	**1.30**	**135**	**135**
Contingency	5 534	5 526	5 971	0.52	0.52	0.56	58	58
Total project cost	**60 874**	**60 787**	**65 678**	**5.68**	**5.68**	**6.13**	**634**	**633**



	OH ppt	S2- ppt	NiSXEW
□ Maintenance	20.28	20.25	22.00
□ Power	11.90	32.19	17.71
■ Reagents	43.31	43.38	43.60
■ Labour	20.36	20.36	21.11

Figure 13—Operating cost breakdown (expressed in USc/lb Ni)

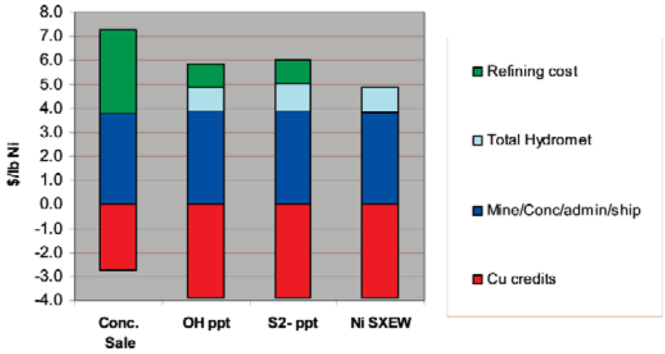


Figure 14—Operating cost breakdown

taxation rate of 35% are assumed. No provision is made for loan financing. The project is assumed to be fully equity funded. The capital and operating costs of the mine and concentrator are included, and these costs were taken from the published Aguablanca prefeasibility study[12] and other published costs, with appropriate escalation. The economic analysis therefore comprises the costs of mining, concentrate production, bioleaching and purification.

In Table XVII, the major outputs from the economic calculations, including a comparison of the internal rate of return (IRR) for each option, are summarized. In this analysis, metal prices prevailing in June/July 2008 were used: these were US$10.00/lb nickel and US$3.50/lb copper. It is noted that these prices have fallen substantially since then. Based on nickel refining charges reported by Lundin in a fourth quarter 2007 press release, an average nickel refining charge of 35% of nickel value was used in this model for direct concentrate sale. Nickel refining charges of between 5 and 15% of value have been quoted[13,14]. An average refining charge of 10% of metal value was used in this model.

An IRR over 10 years of 31% was calculated for concentrate sale, 30% for nickel hydroxide production, 29% for nickel sulphide production, and 33% for nickel cathode production. At a discount rate of 8%, the net present value (NPV) is US$145.7 million for concentrate sale, US$203.5 million for hydroxide production, US$196.4 million for nickel sulphide production, and US$239.7 million for nickel cathode production.

The effect of the low redox bioleach option on the process economics was calculated for the hydroxide precipitation route, and a comparison is summarized in Table XVIII.

As indicated in Table XVIII, the low redox case yielded a higher overall copper extraction of 98% with a 4-day residence time, compared with 95% with a 6-day residence time for the high redox case. The low redox case also tolerated a higher feed solids concentration of 12% (as against 10% for the high redox case). The decrease in residence time and the increase in solids concentration resulted in a decrease in the bioleach tank volume from about 1 600 m^3 to 920 m^3. This reduced the capital cost of the bioleach unit operation by 40%, and the overall hydrometallurgical plant capital cost by 12%. The impact on the overall operating cost for the hydrometallurgical plant was smaller; it decreased by about 4% for the low redox case. The net effect was that the IRR of the project increased from 29.9% to 31.7%.

Table XVII

Capital and operating cost summaries and financial calculations

	Concentrate production	Bioleach + hydroxide precipitate	Bioleach sulphide + precipitate	Bioleach + SX-EW
Mass of ore mined (t/a)	1 500 000	1 500 000	1 500 000	1 500 000
Concentrate production (t/a)	96 000	96 000	96 000	96 000
Nickel price (US$/lb)	10	10	10	10
Copper price (US$/lb)	3.5	3.5	3.5	3.5
Capital cost (US$ million)	119	61	61	66
Operating cost (US$ million/a)	42.0	10.3	12.4	11.2
Metal value paid out (%)	65	90	90	100
NPV, @ 8 % (US$ million)	145.7	203.5	196.4	239.7
IRR (%)	31	30	29	33

Table XVIII

Effect of low-redox bioleach operation on project economics

		High redox case		Low redox case	
Solids concentration in leach	%	10.0%		12.0%	
Leach residence time days		6.0		4.0	
Tank sizes	m^3	1.625		922	
Capex bioleach	M $	11.5		6.9	
Capex bioleach + purification plant	M $	60.9		53.3	
Opex bioleach + purification plant	M $	10.3		9.9	
Opex bioleach	M $	1.6		1.3	
Opex bioleach utilities	M $	0.8		0.7	
Aeration rate	Nm^3/h	51 916		52 004	
IRR, % 10 years	%	29.9%		31.7%	
Oxidation levels—low redox		Stage 1	Stage 2	Stage 3	Stage 4
Redox potential	mV	430	550	560	570
Pentlandite	%	93.0	98.0	98.0	99.5
Chalcopyrite	%	96.0	97.0	98.0	98.0
Pyrite	%	50.0	70.0	80.0	87.0
Pyrrhotite	%	90.2	95.1	99.0	99.5
Oxidation levels—high redox		Stage 1	Stage 2	Stage 3	Stage 4
Redox potential	mV	580	600	640	700
Pentlandite	%	97.0	97.7	99.5	99.5
Chalcopyrite	%	78.0	87.0	93.0	95.0
Pyrite	%	78.5	82.7	86.1	87.0
Pyrrhotite	%	90.2	95.1	99.0	99.5

Discussion and conclusion

Despite extensive research and development work spanning a period of at least 20 years, the uptake by industry of bioleaching technology to treat base metal sulphide concentrates has been slow. This is despite the fact that several projects have proceeded to an advanced commercial demonstration level, and despite a sustained boom in commodity prices. The main reason for this slow uptake is that smelting and refining charges have been on a long-term downward trend for over a decade, since peaking in 1998. There was a brief period (between 2004 and 2006) where these charges rose steeply, but they subsequently fell back to even lower levels. New hydrometallurgical technologies find it difficult to compete with the existing smelting industry, where the capital investment has already been made.

Nevertheless, the indications early in 2009 are that smelting and refining charges are rising steeply (by between 60 and 70% for long-term contracts), and a sustained increase in these charges could result in hydrometallurgical technologies, including bioleaching, being considered for new projects in the future. Further impetus for selecting hydrometallurgical processes such as bioleaching will be gained as environmental considerations become increasingly important, and as the technologies themselves are improved, reducing costs and improving efficiencies.

The present study, using Aguablanca nickel-copper concentrate as an example, has achieved several objectives:

- It has demonstrated the technical feasibility of the technology. The use of thermophilic microorganisms to oxidize chalcopyrite has been demonstrated in an integrated pilot plant over an extended period of time.
- A feature which distinguishes this technology with a previously developed thermophilic bioleach technology for chalcopyrite oxidation is that it is based on the use of air, as opposed to oxygen-enriched air. This obviates the need for sophisticated process control systems to regulate the dissolved-oxygen concentration in the bioleach reactors.
- It has been shown that the economics of this technology is comparable with that of concentrate production and sale.
- Significant advances have been made in applying a redox control strategy to improve chalcopyrite oxidation rates, thereby reducing the capital cost of the bioleach process and improving the overall process economics.

This study has demonstrated that thermophilic bioleach technology is ready for commercial implementation. In Europe, where environmental factors and clean production are strong commercial drivers, and where sulphide deposits often have complex mineralogies and low grades, it could well become the technology of choice. In Africa and the rest of the resource-rich developing world, where large, higher-grade resources are still to be found in abundance, the timeline for acceptance may be longer. However, the first world is consuming most of these resources, and it can be expected that clean production practices will in future be dictated by the consumers and not the producers of raw materials—and then the same drivers currently at play in places like Europe will begin to determine technology choices in the producing countries. Bioleaching stands ready to meet this challenge.

Acknowledgements

This paper is a contribution to the research activities of the BioMinE project (BioMinE – NMP2-CT-2005-500329 – Contract Number 500329), which was co-financed by the European Commission (EC). Mintek received additional financial support for its research activities from the South African Department of Science and Technology (DST). The authors gratefully acknowledge the EC and the DST for their support which made this work possible, and thank all the BioMinE partners who contributed to the joint research effort. In particular, the contributions made by Bioclear and the University of Bangor to the microbial identification and quantification work are noted.

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