

The recovery of nickel from high-pressure acid leach solutions using mixed hydroxide product – LIX[®]84-INS technology

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Received 12 December 2005; accepted 6 January 2006

Available online 3 March 2006

Abstract

This route for the recovery of nickel from HPAL laterite solutions was successfully demonstrated by the Centaur Mining Cawse Nickel Operation and has been selected by BHPBilliton for their Ravensthorpe Nickel–Yabulu Refinery circuit. The mixed hydroxide route is also well suited to treatment of solutions by other leach processes such as Activox[®].

The technology can use an acid strip coupled to nickel electrowinning or an ammonia strip and basic nickel carbonate production. The solvent extraction technology involved in these two circuits is reviewed and compared. The ancillary operations of pregnant leach solution clarification, reagent remediation, reductive stripping of cobalt, zinc and copper transfers are discussed. The treatment of ammoniacal leach solutions containing higher copper concentrations than those normally seen in laterite processing is also addressed.

Possible routes for recovery of cobalt and nickel from the solvent extraction circuit are also discussed.

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Keywords: Solvent extraction; Hydrometallurgy; Non-ferrous metallic ores; Electrowinning; Ion exchange

1. Introduction

At the 1996 ALTA Nickel Conference, Cognis introduced the concept of treating nickel laterites (Mackenzie et al., 1996) using a high-pressure acid leach (HPAL) followed by mixed hydroxide precipitation (MHP), ammonia leach of nickel and cobalt and the refining of nickel using LIX[®]84-INS based solvent extraction (SX). In the intervening years this process has received significant attention and acceptance.

The nickel refinery part of the circuit can use either acid strip plus electrowinning (EW) by the Cognis–Cawse route, or a strong ammonia strip coupled with steam stripping to produce a basic nickel carbonate (Queensland Nickel Route).

Both routes have their merits.

The general process of producing a MHP followed by ammonia releaching of nickel and SX of the nickel using LIX[®]84-INS, a 2-hydroxy-5-nonylacetophenone oxime specially produced for nickel extraction from ammoniacal leach solutions, was first used by Centaur Mining and Exploration's Cawse Nickel Operations. Although no longer producing nickel metal the Cawse operation has come to be recognized as technically the most successful of the first generation West Australian HPAL nickel laterite plants. Subsequently BHPBilliton has chosen this route for their 30,000 tons per annum Ravensthorpe Nickel project. From laterite ore to loaded organic, the Ravensthorpe–Yabula circuit is essentially the same as the Cawse circuit. Companhia Vale do Rio Doce (CVRD) has also selected the Cognis–Cawse Nickel process route for their feasibility study for the Vermelho Nickel project.

Since their original publication in 1996, Cognis has published a number of papers on aspects of the chemistry of

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the SX circuit involved in the Cawse and Queensland Nickel operations (Virnig et al., 2001; Mackenzie et al., 1998; Mackenzie and Virnig, 2004; Mackenzie et al., 2005). The MHP–ammonia re-leach–nickel SX–EW circuit described in these papers should not be considered as an SX circuit in isolation. There are several serendipitous aspects to the overall circuit which make it probably the most effective circuit for the recovery of nickel metal and the associated cobalt by product from laterites. These aspects, as well as the potential use of the MHP–ammonia re-leach–LIX[®]84–INS route in the recovery of nickel from sulphide ores will be discussed in this paper.

The LIX[®]84–INS SX circuit involves chemical reactions and metallurgical processes which are not commonly seen in SX circuits recovering other metals and are certainly absent from conventional copper SX circuits and these issues merit special mention.

This paper will

- Summarize some of the advantages of the MHP–ammonia re-leach–LIX[®]84–INS SX–EW route for the recovery of nickel metal from laterites.
- Discuss how the SX–EW circuit is integrated into the overall circuit.
- Discuss the SX chemistry which is specific to nickel recovery using LIX[®]84–INS.
- Discuss the EW of nickel from the acid strip solutions.
- Review some of the possible impurity control strategies for the nickel SX circuits.
- Discuss the application of the MHP–ammonia re-leach–SX–EW process for the treatment of nickel sulphide ores.

2. Advantages of the MHP route for the recovery of nickel from HPAL laterite leach solutions

The leach solutions exiting typical laterite HPAL plants are characterized by high residual acid concentrations of about 10–60 g/l, iron concentrations of about 2–8 g/l, aluminium concentrations of 1–2 g/l and magnesium concentrations of about 10–20 g/l. The nickel and cobalt concentrations are about 3–6 g/l and 0.2–0.8 g/l respectively. In addition, there are manganese concentrations of about 1–3 g/l together with the presence of small, but problematic concentrations of other metals including copper, zinc and chrome.

While removal of the Fe(III) and neutralization of the acid with limestone provides an initial partial purification step, separation of the remaining constituents to produce a saleable nickel product is not easy and cannot currently be done in a single metallurgical process.

Some of the possible processes used to treat a HPAL leach solution include:

1. Precipitation of a mixed sulphide product followed by re-leaching of the transition metals and application of the Sherrit process. Examples include Murrin Murrin and Moa Bay.
2. Precipitation of a mixed sulphide precipitate and re-leaching in sulphate media followed by sequential SX circuits to recover Cu, Co and Ni.
3. Precipitation of a mixed sulphide product followed by the Falconbridge chloride leach route.
4. Direct SX on the neutralized solution using two SX circuits in series. Goro and Bulong are examples of this approach although Goro also uses ion exchange (IX) to remove some impurity metals. The Western Minerals Technology process (Kloiber et al., 2005) downstream of the Activox[®] leach used for treating nickel sulphide concentrates would also fall into this category.
5. Precipitation of a MHP followed by ammonia re-leaching and SX–EW.

All of these processes and some others can be made to work. An intriguing aspect of nickel laterite hydrometallurgy is that there are almost as many possible treatment routes as there are research hydrometallurgists. The selection of the optimum process will depend on economic issues and on risk minimization.

The MHP–ammonia re-leach route has a number of advantages:

1. The MHP can be transported from the mine site to an existing nickel refinery. This is the route followed by BHP Billiton with their Ravensthorpe operation. Ammonia re-leaching of aged MHP, while not quite as straightforward as leaching of fresh hydroxide precipitate is quite feasible.
2. Where the nickel deposit is located in a country with some political risk the refinery can be located in an alternative country.
3. The MHP can be stored in slurry form within the circuit to provide several days operating time for the downstream refinery. This ensures continuous operation of the refinery should there be disruptions to the HPAL operation.
4. This storage can also be used to provide a feed of uniform quality and quantity to the nickel refinery.
5. The MHP and ammonia re-leach stage increases the purity and concentration of the nickel product prior to refining. In this way the MHP stage is analogous to a concentrator in a nickel sulphide operation. It is this purification step which has contributed greatly to the acceptance of the LIX[®]84–INS solvent extraction route for nickel.
6. The MHP can be sold as an intermediate product.

3. The overall nickel recovery circuit

The LIX[®]84–INS SX used in nickel laterite circuits is only one part of a much larger metallurgical flowsheet which employs a number of steps to concentrate and purify the nickel solution derived by HPAL leaching of a laterite.

Without the contribution of these other steps, the LIX[®]84-INS SX would not be effective. These steps and the overall advantages of the MHP–ammonia re-leach–SX circuit will now be considered in more detail.

A schematic flowsheet of a Cognis–Cawse version of the circuit is shown in Fig. 1. In Table 1 the typical analyses of the nickel-containing aqueous solution at various stages of the circuit is presented. A schematic of the SX circuit is shown in Fig. 2.

The leach solutions produced by HPAL of nickel laterites contain a wide range of metals, some of them at significant concentrations. Currently available SX technology cannot purify these solutions in a single SX circuit.

A feature of the MHP–ammonia re-leach–LIX[®]84-INS circuit is the gradual purification of the leach solution as

it progresses from the HPAL leach to the SX feed. The more important purification steps are discussed in Sections 3.1–3.4.

3.1. Fe(III) removal and acid neutralization in primary neutralization

The addition of ground limestone to produce a pH of around 2.5 precipitates the Fe(III) and neutralizes the acid.

3.2. Fe(II) removal and partial removal of Mn, Cr and Cu in secondary neutralization

The addition of limestone and air at a pH of approximately 4.5–4.7 removes almost all of the Fe, Al and Cr.

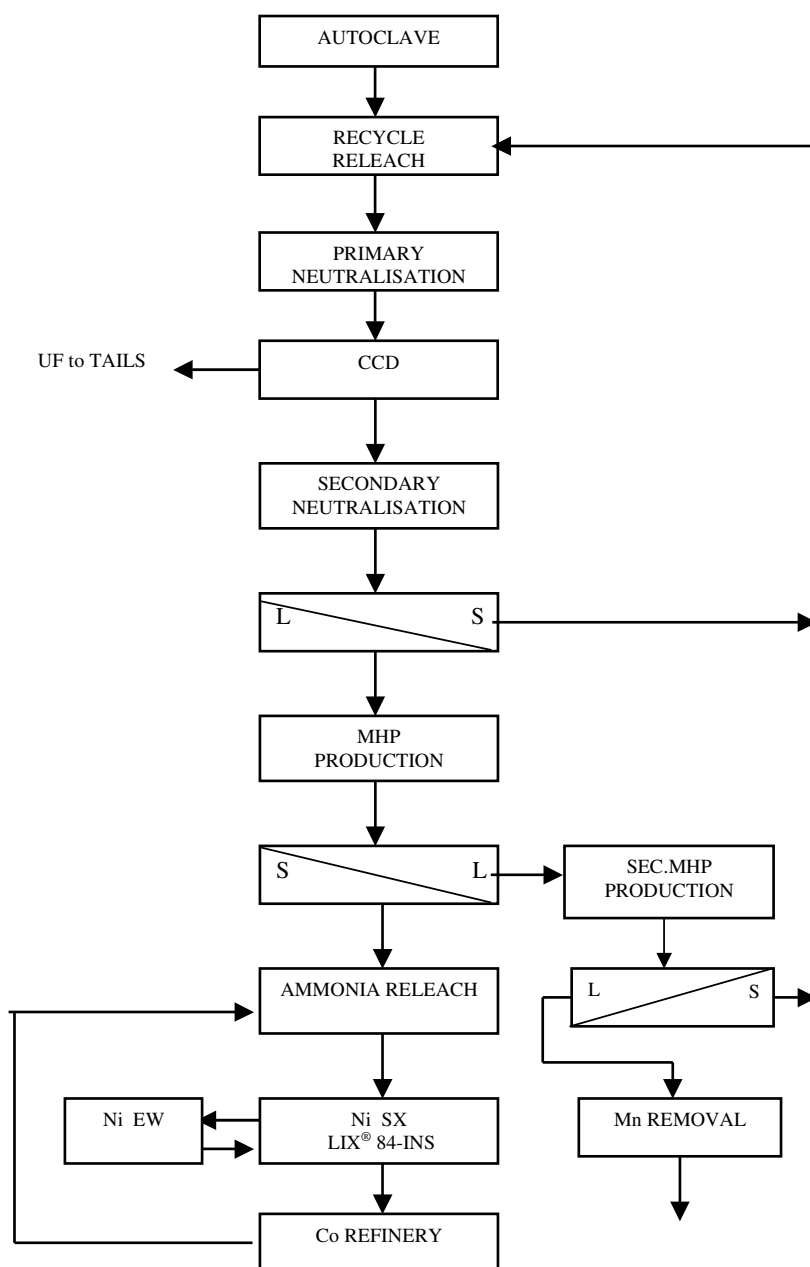


Fig. 1. Simplified Cognis–Cawse nickel laterite circuit.

Table 1
Metal concentrations at various stages in the MHP–ammonia leach–LIX®84-INS nickel circuit

Stage	Metal concentration, mg/l							
	Ni	Co	Fe	Al	Mn	Cu	Zn	Cr
Primary neutralization feed–HPAL discharge plus recycle releach	8000–12,000	300–500	10,000–16,000	3000–8000	1000–3000	100–500	100–300	100–500
Secondary neutralization thickener O/F	3000–4500	100–200	2–10	1–10	500–1000	5–20	50–150	1–5
Ammonia releach solution–SX feed	12,000–15,000	600–1000	1–2	1–2	1–2	30–70	300–400	1
Metal content, %								
MHP	38–45	1.2–2.5	0.1–0.3	0.05–0.1	3–5	0.05–0.2	0.6–1.8	0.01–0.02

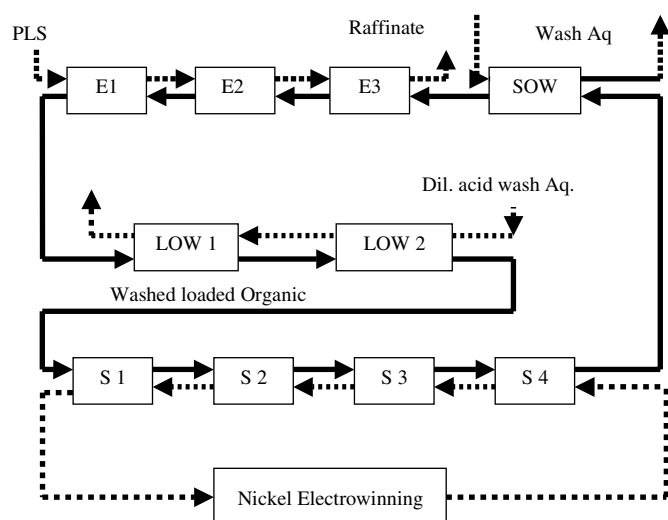


Fig. 2. Schematic diagram of solvent-extraction circuit.

At the same time, Mn and Cu are partially removed. Details of the chemical reactions involved in primary neutralization, secondary neutralization and manganese removal have been previously published (Dry and Miller, 2001).

3.3. Selective releaching of Ni, Co, Cu, Zn and Mn in the ammonia–ammonium carbonate releach of the MHP

MHP is releached in an ammonia–ammonium carbonate solution containing approximately 100 g/l NH_3 and 90 g/l CO_2 . This leach dissolves only those metals which form stable amine complexes and is selective for the transition metals, Ni, Co, Zn and Cu. Only a small amount of the Mn is leached. Typical ammonia releach efficiencies are summarized in Table 2.

3.4. Partial steam strip of the ammonia releach solution

The ammonia releach solution is subjected to a partial steam strip in a column. This reduces the ammonia concentration of the ammonia releach solution to around 40 g/l. At this lower total ammonia concentration, some of the ammine complexes dissociate and the associated metal is precipitated. This process is particularly effective in removing

Table 2
Ammonia releach efficiencies for various metals

Metal	Ammonia releach efficiency
Ni	99
Co	92–95
Cu	95
Zn	95
Fe	<1
Mn	3–5
Al	<1
Mg	<1

ing manganese from the leach solution. A small percentage of the Ni and Co together with some Cu and Zn are also precipitated and the partial steam strip solids discharge must be subjected to a further leaching process to recover the Ni and Co.

The selectivity of the steam strip for Ni and Co over Mn is a result of the stability constants for the respective ammine complexes. These are shown in Table 3. The reaction considered is



$$K_n = [\text{M}(\text{NH}_3)_n] / [\text{M}][\text{NH}_3]^n \quad (2)$$

The manganese ammine complexes are fortunately much less stable than the nickel and cobalt ammine complexes. Dissociation of the manganese complexes and precipitation of the manganese results in an SX feed typically containing less than 1 mg/l Mn. This is most significant as the presence of Mn ammine complexes in the SX feed would have very adverse effects on the SX circuit operation. Unfortunately the stability complexes for Cu and Zn are similar to those for Ni and Co making removal of Cu and Zn from the solution in the partial steam strip impossible. The solution exiting the partial steam strip with the composition given in Table 1 is an excellent feed solution however for Ni SX–EW using LIX®84-INS.

4. Special chemical aspects of the extraction of nickel from ammoniacal leach solutions produced from a MHP

A PLS containing the metal concentrations given in Table 1 together with 20–40 g/l free ammonia is a very

Table 3

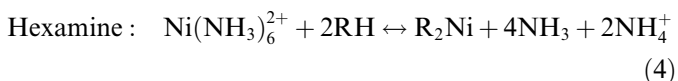
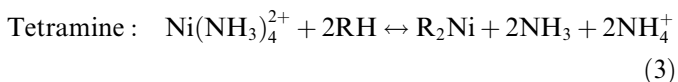
Stability constants for metal ammonia complexes which are important in partial steam stripping of nickel laterite ammoniacal leach solutions

Metal ion	N	Temperature, °C	Log K_n
Mn^{2+}	1	20	1.00
	2		1.54
	3		1.70
	4		1.30
Cu^{2+}	1	25	4.24
	2		7.83
	3		10.80
	4		13.00
	5		12.43
Fe^{2+}	1	20–30	1.40
	2		2.20
Co^{2+}	1	30	2.10
	2		3.67
	3		4.78
	4		5.53
	5		5.75
	6		5.14
Co^{3+}	1	30	7.30
	2		14.00
	3		21.10
	4		25.70
	5		30.75
	6		35.20
Ni^{2+}	1	30	2.81
	2		5.08
	3		6.85
	4		8.12
	5		8.93
	6		9.08
Zn^{2+}	1	25	2.33
	2	25	4.83
	3	25	7.46
	4	25	9.34

acceptable feed to a LIX[®]84-INS solvent extraction circuit. However the chemistry involved is complex. Some of the main aspects of this chemistry will be discussed. Differences between the chemistry of acid and strong ammonia strip circuits will be examined.

4.1. Extraction of nickel by oximes

The extraction proceeds as follows:



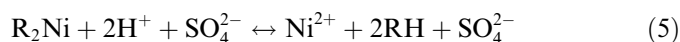
The extraction is an equilibrium reaction and can be driven in either direction depending on the concentrations of the species involved. During extraction of nickel, ammonia and ammonium ions are produced. These species report to the raffinate and can be recovered by steam stripping for recycle to leach. This is a positive aspect of the extrac-

tion process. However, if the leach solution contains significant concentrations of nickel, e.g. 50 g/l, then extraction of the tetrammine will yield about 60 g/l ammonia/ammonium ion while extraction of the hexamine will yield about 90 g/l of these species. The extraction isotherm for LIX[®]84-INS (see Fig. 3) shows that at total ammonia concentrations above 60 g/l, the extraction of nickel is reduced by the ammonia/ammonium ion concentration in the raffinate.

The equilibrium equation imposes restraints on the maximum tenor of nickel in the PLS that can be treated. In practice, ammonia is steam stripped from the PLS in order that the feed to SX contains about 20–40 g/l ammonia. In order to limit the NH_3/NH_4 produced during extraction, the nickel concentration in the PLS is usually restricted to about 12–18 g/l although it would be possible to produce a nickel concentration in the PLS much higher than this (50–60 g/l).

The extraction and stripping circuits operate at 50–60 °C. This reduces the viscosity of the organic and LIX[®]84-INS concentrations of 30–35 v/v% may be used. The maximum nickel loading on the organic will be about 15.0–18 g/l. In practice, this is reduced by some Co loading and 12–15 g/l is probably a more realistic figure. Either the ammonia strip or the acid strip will deliver a very low nickel concentration on the stripped organic and a nickel transfer of 12–15 g/l is achievable.

Following nickel extraction by LIX[®]84-INS, the loaded organic can be stripped by high (≈ 280 –290 g/l) NH_3 solutions or sulphuric acid plus nickel sulphate solutions. The ammonia strip follows the extraction equilibrium relationship given above. In order to improve the stripping of nickel with ammonia, Cognis has developed a modified version of LIX[®]84-I called LIX[®]87QN. This reagent uses tridecyl alcohol as an equilibrium modifier. The acid strip proceeds as follows:



The acid concentration required to strip LIX[®]84-I are relatively low. Using multi-stage stripping, it is possible to produce a strong nickel electrolyte containing 70–100 g/l nickel at a pH of approximately 3.0 which is suitable for EW.

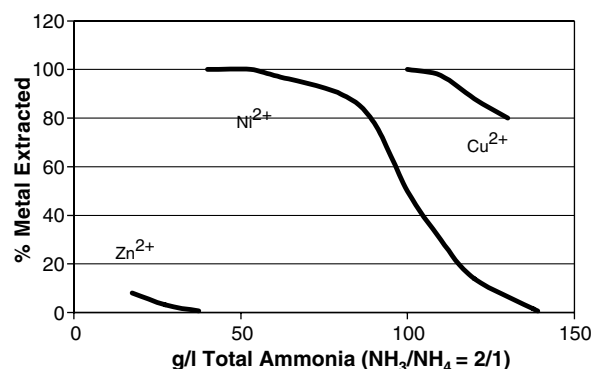


Fig. 3. Extraction isotherms from ammoniacal solutions for LIX[®]84-INS.

Let us now consider some of the special aspects of the SX circuit.

4.2. Cobalt extraction

Cobalt is the most problematic issue in the use of Cognis technology to recover nickel. Although the detailed chemistry of how cobalt influences the technology is not clear, Cognis and plant operators do understand how to control the problems caused by cobalt:

1. PLS solutions produced by HPAL of laterites, MHP production and ammonia leaching will, depending on the laterite analysis, contain cobalt concentrations of about 1/10–1/15 of the nickel concentration or about 1–2 g/l Co.
2. Co(II) will co-extract with nickel as an R_2Co chelate and will, quite rapidly, oxidize on the organic to an R_3Co chelate. In this form the stripping kinetics of cobalt in the nickel strip (either acid or strong ammonia) are so slow as to be virtually non-existent. Cobalt will therefore accumulate on the organic and will tie up oxime. One gram per liter Co on the organic as Co(III) will reduce the nickel transfer capacity of the organic by approximately 1.5 g/l.
3. Unfortunately not all the Co(II) extracted will have oxidized to Co(III) by the time the loaded organic reaches the strip circuit. A small percentage of it will still be present as Co(II) and will strip with the nickel. This cobalt will plate with nickel during electrowinning.
4. An even more adverse effect of the cobalt extraction is the damage that it causes to the organic. Firstly, it increases the viscosity of the organic leading to physical problems in SX. It is thought that the viscosity increase is caused by cobalt catalysed formation of polymeric species in the organic. The exact mechanism whereby cobalt promotes the degradation of the organic is not clear.
5. By far the best way to overcome the cobalt problem is to oxidize all the cobalt to Co(III) prior to SX. In the ammonia leach stage, aeration of the leach tanks can help to oxidize cobalt. Following this, the steam strip used to reduce the ammonia content of the PLS can aid in cobalt oxidation. Hydrogen peroxide may be added to the PLS prior to SX in order to complete the cobalt oxidation. It has always been thought that this peroxide addition would lead to degradation of the organic but unless excessive peroxide is added, this does not seem to be the case.
6. Oxidation of cobalt is facilitated by the fact that while the E0 for oxidation of Co(II) to Co(III) is 1.8 V in acid sulphate, it is only 0.17 V in alkali solution.
7. In some Ni SX circuits the raffinate goes to a cobalt sulphide precipitation stage where the cobalt sulphide is thickened and sent to a cobalt refinery. The sul-

phide thickener overflow may return to the nickel recovery circuit and fine sulphide particles in the thickener overflow can make oxidation of the cobalt in the Ni SX difficult.

8. Unfortunately there is no direct chemical analytical procedure for analysing the Co(III)/Co(II) ratio in the Ni SX feed. The usual method is an indirect one using oxime extraction and to report the Co extracted as Co(II).
9. It is useful to add extra peroxide to the aqueous entering E2 to ensure that any Co(II) produced in E1 is oxidized.
10. Unfortunately, it is inevitable that over time some cobalt will be extracted onto the organic and will remain on the circulating organic. This cobalt will have to be removed by a reductive strip.

4.3. Reductive stripping of cobalt from circuit organic

Cobalt and manganese, which have been oxidized to a higher valency state on the organic, can be stripped using acid plus a solid reductant. The main features of the reductive strip are

1. The solid reductants available are iron and zinc powder. Zinc is the more electronegative metal and is more effective than iron in stripping Co(III) from the organic. However, the strip produces zinc sulphate. Since the strip aqueous is often recirculated back to the leach, this introduces soluble zinc into the circuit. Because there is no simple way to remove zinc from the circuit water, it will build up and will find its way into the SX circuit and possibly into the nickel or cobalt final product. For this reason, iron powder is the preferred reductant in an operating plant. Nickel powder and gaseous hydrogen are not effective as cobalt reductants. HCl in the absence of any reductant will also strip cobalt from LIX®84-INS. This is sometimes used as an analytical procedure in the lab. The use of high concentration HCl also results in a significant increase in the degradation by hydrolysis of the oxime.
2. The strip uses sulphuric acid in the range of 60–120 g/l. The kinetics of stripping increase with the acid concentration but the degradation of the oxime also increases with the acid strength and 80–100 g/l is usually the upper concentration employed.
3. The strip kinetics increase markedly with an increase in strip temperature. The strip is normally operated at 50–60 °C. Higher temperatures can lead to excessive degradation of the organic.
4. The strip kinetics increase with an increase in concentration and surface area of the reductant. Iron powder concentrations of 25–50 g/l in the organic are typical.
5. The strip mixer residence time should be 30–60 min. The iron should be introduced to the organic phase so that it is organic wetted. The exact chemistry of the reductive

strip is not clear but it involves some reaction at the iron–organic interface.

The cobalt strip will usually be operated as a continual bleed stream circuit and should have sufficient capacity to maintain the cobalt loading on the nickel circuit organic below 1–2 g/l. In general the effect of cobalt extraction on nickel solvent extraction with LIX®84-INS is similar for the Cognis–Cawse and QNI circuits although the lower leach efficiency of the RRAL process means that these circuits generally have lower cobalt concentrations in the PLS than the HPAL circuits.

4.4. The effect of copper extraction

Although much of the copper will be precipitated in the secondary neutralization stage, up to 100 mg/l may be present in the Ni SX feed and will be quantitatively extracted. In the acid strip circuit a little of the copper may strip in S4 where the acid concentration is around 40 g/l. However as the aqueous progresses to S1 where the pH is about 3.0–4.0 it will be re-extracted onto the organic and the pregnant strip solution will contain only a few mg/l copper. The stripped organic returning to nickel extraction will typically contain a few hundred mg/l of copper. This can be stripped in a bleed stream copper strip using 180–200 g/l sulphuric acid. The copper will also be stripped in the cobalt strip bleed stream. For an ammonia strip circuit, about 10–20% of the copper loaded on the organic will be stripped by the 290 g/l ammonia and will report to final product. The copper loading on the stripped organic will be about 200–500 mg/l. Copper can be removed from the ammonia strip final product by precipitation with sodium hydrosulphide or by using a combination of nickel powder and sodium hydrosulphide.

4.5. Zinc extraction

As shown in Table 1, almost all the zinc leached in HPAL will report to Ni SX. In a LIX®84-INS circuit most of the zinc will report to the nickel SX raffinate. If the cobalt is recovered from the raffinate by sulphide addition, the zinc will precipitate with the cobalt as a sulphide. The mixed sulphide will have to be treated in a cobalt refinery. The zinc/cobalt separation can be done with organo-phosphorus acid extractants, such as D2EHPA.

With fresh LIX®84-INS, the nickel to zinc selectivity is high. The Cognis quality control specification is 50,000:1. However for LIX®84-INS circuit organics, the selectivity drops to 250–500:1. Zinc is crowded from the organic by nickel (Table 4).

The kinetics of nickel extraction are lower for the sulphate media. With fresh LIX®84-INS, some zinc will entrain in the loaded organic. If the nickel concentration on the loaded organic is not close to the maximum loading, a small amount of zinc may be co-extracted with the nickel. An acid scrub of the loaded organic at pH 5–5.5 will

Table 4
Nickel and zinc extraction kinetics^a

PLS media	Metal	Time, s				
		0	30	60	90	300
Carbonate	Ni, g/l	0	13.1	13.8	14.1	14.3
	Zn, mg/l	0	5.6	2.2	1.7	1.2
Sulphate	Ni, g/l	0	12.4	12.4	14.1	14.6
	Zn, mg/l	0	13.5	13.5	2.2	1.0

^a PLS = 15 g/l Ni, 1 g/l Zn, 32.5 g/l NH₃, 25 g/l ammonium sulphate or carbonate; O/A = 1.0, organic = 30 v/v% LIX 84-INS in aliphatic diluent.

remove any small traces of zinc extracted on the organic. With aged circuit organic however this scrub does not completely remove the zinc from the loaded organic. Some zinc (approximately 50% of the zinc on the E1 organic) will report to the nickel final product. Similarly the use of a dilute ammonia scrub of the loaded organic in a circuit employing an ammonia strip will remove only about 50% of the zinc from the nickel loaded organic. Typically the zinc transfer to the strip circuit will be 10–20 mg/l. The exact mechanism for this zinc transfer is not well understood but an oxime degradation product may be involved. One of the degradation products of oximes which are observed in SX circuits treating ammoniacal leach solutions are ketimines. Cognis has studied the effect of ketimines on Ni/Zn selectivity and has found that they do not contribute to the transfer of zinc to the strip circuit.

4.6. Potential impurity control strategies

Most base-metal refining circuits will use a combination of SX and ion exchange (IX) to produce a solution of sufficient purity for the recovery of final product. SX is often the technique of choice for removal of high-concentration impurities, while IX is well suited for the removal of low-concentration impurities from the valuable metals in most acid sulphate cobalt/nickel flowsheets. A combination of these techniques would allow base-metal producers to achieve higher metal specifications and consequently add value to their products. Likely applications in a Cognis–Cawse type Ni flowsheet would include:

- Removal of Zn from advance electrolyte to Ni EW or reduction.
- Removal of Cu and/or Zn from Co EW electrolyte.
- Reduction of Ni in Co EW electrolyte.

A wide variety of SX extractants and ion exchangers are available for these applications.

A suitable exchanger will be chosen from chelating, weak acid or even molecular recognition technology (MRT) resins. The choice of contactor type would depend on factors such as size and cost of the resin inventory, and the cost of contacting equipment. Simple fixed-bed columns in a lead-lag configuration are cheaper to construct but require more resin inventory than a counter-current

operation. A counter-current arrangement is more costly to construct and complex to operate.

4.6.1. Removal of Zn by SX from advance electrolyte to Ni or Co EW

If the concentration of zinc in the nickel or cobalt electrolyte is relatively high, an SX unit operation using an organo-phosphorus acid is most likely to be used. Depending on the performance of this SX, an IX polishing step may be required to reduce zinc concentration to acceptable levels.

Murrin–Murrin makes use of Cyanex® 272 to remove zinc from the nickel solution upstream of Co SX and Ni reduction. The selective stripping of Cyanex 272 to separate zinc from cobalt is also feasible. Bulong, (Mayze, 1999) used D2EHPA in a small SX circuit to remove zinc from the cobalt advance electrolyte.

Choice of SX extractant will largely depend on the concentration of other impurities as well as the integration of the zinc SX into the plant circuit. It is always good practice to minimize the number of different reagents used on any particular site.

4.6.2. Removal of Cu and/or Zn from Co EW electrolyte

This has been implemented and reported for the Kakanda Tailings integrated demonstration plant, (Wythe and Kotze, 2000) and the Bulong Ni laterite project using Purolite S950 (Mayze, 1999). In both these cases, the choice of a cheap, chelating resin made the use of fixed-bed lead-lag columns possible.

The use of resins with D2EHPA functionality is well suited to the polishing of Zn from Co and Ni product solutions. A resin such as Lewatit – OC 1026 is reasonably priced and capable of reducing zinc to very low concentrations.

The arrangement and design of the IX unit operation will largely depend on the requirement in terms of solution purity, as well as how it can be integrated into the flow-sheet. A split elution may be required to recover the bulk of the co-extracted valuable cobalt or nickel. If there is an opportunity to recycle solutions to an appropriate point in the circuit, a split elution may not be necessary.

4.6.3. Reduction of Ni in Co EW electrolyte

Removal of small amounts of Ni from a Co solution is extremely difficult due to the similarity of the chemical behaviour of the metals. INCO (Port Colbourne, Canada) and J&W (Chambishi Metals, Zambia) both use Dowex M4195 for the removal of nickel from high-concentration cobalt solution.

The IX process was installed at Chambishi Metals to replace the inefficient precipitation of Ni from Co solution (Rao et al., 1992). Cobalt losses with the previously used precipitation technique were unacceptably high and made the IX circuit attractive. Since using the IX route, the original fixed-bed system has been replaced with an ISEP carousel continuous counter-current system designed to minimize the inventory of the costly resin.

4.6.4. Other impurities

Impurities such as Ca, Fe and Mn can be removed from Co or Ni solutions in order to purify them for final metal production. Although the Cognis–Cawse type flowsheet is largely selective over these impurities, these can be introduced in some cases.

Ca can be largely removed from Ni solution using D2EHPA. This has been described for the removal of Ca from a Ni solution prior to production of high-purity NiSO₄ crystals (Clark et al., 1993). Ca can be reduced to very low levels. The stripping of Ca from D2EHPA is problematic as stripping with sulphuric acid can precipitate gypsum. Alternatives are stripping with large volumes of solution or the use of HCl.

Trace Fe is removed from Ni solution along with Zn by SX using Cyanex 272 at Murrin–Murrin (Mayze, 1999). QNI use the same extractant to remove Fe from the Co solution in their circuit (Gray et al., 1996).

Manganese can be largely removed using D2EHPA from cobalt or nickel solutions (Feather et al., 2000). This can be an attractive alternative if a D2EHPA SX unit operation is already under consideration for the removal of zinc. The zinc and manganese can then be removed together. This technique is useful ahead of metal recovery by EW, where some manganese can be tolerated, either in an undivided-cell operation or divided cells using cathode bags.

Probably the best way to treat the zinc problem is to remove the zinc from either the PLS or the final product liquor by IX. For a Cognis–Cawse circuit this is quite feasible and a D2EHPA based resin Lewatit-OC 1026 from Bayer will remove zinc from the tank house electrolyte. For circuits employing strong ammonia as strip liquor (the Queensland Nickel route), an IX option is not readily available and some other way will have to be found to exclude zinc from the final product.

4.6.5. Reagent interactions

In developing alternative flowsheets involving multiple SX circuits to remove/recover various metals, one has to take into consideration the potential effect of cross-contamination of the circuit organics on the operation. While the most obvious way that cross-contamination is likely to occur is simply by addition of the wrong SX reagent to a circuit, the most likely avenue is by slow transfer of organic as entrainment in the aqueous raffinate from the lead SX circuit to the trailing SX circuit.

One obvious potential effect of cross-contamination is loss of selectivity. Bulong experienced problems with the selectivity of the Versatic acid circuit for Ni over Ca when it became contaminated with Cyanex®272 (Mayze, 2001; O'Callaghan, 2003). In the case of an oxime extractant, contamination of the circuit organic with an acidic extractant such as D2EHPA or Cyanex®272 can result in a significant increase in the rate of hydrolysis of the oxime. The hydrolysis chemistry requires an acid catalyst and water. Contamination of an oxime circuit organic by an

oil soluble acid such as DEHPA provides both acid and water due to solvation of the acid functionality directly in the organic phase. Hydrolysis is no longer limited to the interface but can also occur in the bulk organic. This has been observed for synergistic mixtures of LIX[®]63 with D2EHPA (Inoue et al., 1997), under conditions typical of Ni extraction. While no similar experiments have been carried out with LIX[®]84-INS under conditions approximating nickel extraction, stability experiments carried out in the Cognis laboratories have shown that spiking a strong acid such as a dodecylbenzene sulphonic acid into a typical extractant such as LIX[®]984N at a level of 50 ppm will result in a 66–75% reduction in the half life of the reagent against a standard copper electrolyte at 45 °C. In systems where multiple SX circuits are employed using different reagents, care must be taken to insure that cross-contamination cannot occur.

4.7. Sulphate and ammonia transfer to strip

In a LIX[®]84-INS nickel SX circuit using an acid strip, ammonia transfer to strip may present a problem. Conversely in a LIX[®]84-INS circuit using an ammonia strip, transfer of sulphate to strip may present a problem.

The ammonia loading on fresh 30 vol% LIX84-I loaded organic containing 13–14 g/l Ni is about 0.3–0.5 g/l. If this ammonia transfers to the acid strip circuit and then to the tankhouse, it has the potential to stress the cathode, particularly the starter sheets being plated onto titanium blanks. In practice, nickel tankhouses have been relatively tolerant of ammonium sulphate in the tankhouse. The transfer of ammonia to strip, however, should be minimized by a pH 5–6 dilute sulphuric acid wash of the loaded organic. There is another potential problem created by ammonia transfer to strip. If the ammonium sulphate concentration in the tankhouse electrolyte becomes high enough and if the tankhouse circuit is allowed to cool at times of temporary shutdown, then Nickel Boussingaultite $(\text{NH}_4)_2\text{Ni}(\text{SO}_4)_2 \cdot 6\text{H}_2\text{O}$ may precipitate in the tanks and pipes. Steam tracing may be a useful precaution in these tankhouses.

In LIX[®]84-INS circuits using a strong ammonia strip, sulphur transfer to the final product may be a problem. The sulphur may be present in the PLS as sulphate, or possibly as other sulphur species such as thiols produced in a RRAL circuit. If a MHP–ammonia/ammonium sulphate leach circuit is used, then the SX feed will contain appreciable sulphate and this sulphate may transfer on the loaded organic. The most likely transfer mechanism is via entrained aqueous which is so fine as to be not readily separated by physical means. Karl Fisher analysis on samples of LIX[®]84-INS circuit organic which have been centrifuged to remove gross aqueous entrainment show a water content of 0.1% which is sufficient to transfer appreciable sulphate. Washing of the loaded organic can reduce but not entirely eliminate sulphur transfer.

4.8. Ammonium carbonate and ammonium sulphate media

Either MgO or CaO can be used to precipitate the MHP. If CaO is used, the ammonia releach will take place in an ammonium sulphate media. The kinetics of extraction from an ammonium sulphate medium are slower than from a carbonate medium. Cognis has found that the E60sec/E300sec from the carbonate medium is 96.5% while the corresponding value for the sulphate medium is 85%. One can speculate that this is probably due to the tendency of nickel amines to form nickel ammonium sulphate double salts. No such double salt exists for the carbonate system and in this system the nickel will be present as the readily extracted free amine complex. Where sulphate transfer to strip is an issue, then the use of CaO to form the MHP and the resulting high sulphate concentration in the PLS may increase the sulphate transfer problem. It may be difficult to operate a circuit using a strong ammonia strip and ammonium sulphate releach.

4.9. Diluent types

Because nickel SX circuits using LIX[®]84-INS operate at high temperatures (50–60 °C), a high flash point diluent is mandatory. The diluent type can also influence the chemistry of the extraction process. Both Flett and Melling (Flett and Melling, 1979) and Cognis have found that increasing the aromatic content of the diluent can increase ammonia extraction. This effect is more pronounced for extraction from sulphate media than from carbonate media. From sulphate media, aromatic diluents also show slightly better nickel extraction. It is probable that the better solvency power of aromatics for the extracted species contributes to these results. The diluent type appears to have no effect on the nickel extraction or stripping kinetics.

4.10. Nickel stripping circuits for LIX[®]84-INS

For the acid sulphate strip used in the Cognis–Cawse circuit, the strip kinetics are very slow compared with copper strip kinetics. Mixer residence times of 6 min minimum and temperatures over 50 °C are needed in order to achieve acceptable kinetics. The slow stripping of nickel into acid sulphate mirrors the slow extraction of nickel from acid sulphate solutions. Temperature plays a major role in determining strip kinetics and the time required to achieve 98.4% stripping is 250% longer at 40 °C than it is at 60 °C.

In the acid sulphate strip, the objectives are to almost completely strip the nickel and at the same time virtually completely consume all the acid in the strip aqueous so that the nickel strip liquor leaving S1 (the stage the loaded organic enters) is between pH 3 and 4. This means that the acid level in the strip aqueous entering S4 must be almost stoichiometric with the nickel on the loaded organic. In practice, it is easier to have slightly less than this amount of acid available and to add extra acid with

pH control to strip stages 3 and 4. The typical pH profile in strip will be from 0.6 in S4 to 3.3 in S1. Very little stripping (<5%) takes place in S1. The role of this stage is to neutralize, by stripping nickel, any free acid remaining in the aqueous. Most of the stripping takes place in S2 and S3. It is possible that some copper may strip in the relatively high acid conditions in S4 (30–40 g/l) but as the strip aqueous moves to S1, the pH of the aqueous will increase and this copper will be back extracted onto the organic. The nickel concentration in the anolyte is about 60 g/l and this is built up in strip to 70–100 g/l. This means that the throughput O/A in the acid sulphate strip will be about 1.5–2.5–1.

In the strong ammonia strip circuit, the most important operating variable is the ammonia concentration of the strip aqueous. It should be 280–290 g/l in order to achieve good stripping of the nickel. Even with this concentration of ammonia, the circuit organic will need to contain about 5–10 v/v% tridecyl alcohol and operate at about 55 °C in order to strip the organic down to about 100 mg/l Ni in 3–4 stages.

4.11. EW of nickel from strip solutions

The EW of nickel cathode from LIX®84-INS strip solutions is relatively straightforward. If the strip pH is carefully controlled, then the advance electrolyte to EW will be close to pH 3. The electrolyte will first have any entrained organic removed by coalescers followed by carbon columns. Zinc will be removed by IX as described previously. The pH will then be adjusted to 3–3.5 using sodium hydroxide. This may generate a little sodium sulphate but as this is generally added to nickel tankhouse electrolytes to improve conductivity, this is not a problem.

The nickel EW operation usually involves production of a 24-h starter sheet on a titanium blank followed by a 5–7 day cycle to produce a commercial cathode. A typical electrolyte composition and the main cell parameters are given in Table 5.

In order to maintain the pH in the cathode region at 3.0–3.5 and to reduce hydrogen evolution at the cathode, divided cells are commonly used. In general, older plants used bagged cathodes, while the modern trend is towards

the use of bagged anodes to retain the acid in the anode compartment. The anolyte together with oxygen generated at the anode is removed by suction from the anode compartment. The use of bagged anodes also reduces the formation of mist in the tankhouse which, as a result of the carcinogenic properties of some nickel compounds, is important in nickel EW.

Sodium sulphate is added to increase the conductivity of the electrolyte and to so permit the use of high current densities. Sodium dodecyl sulphate is a surfactant which is stable in acid solutions. It promotes the disengagement of hydrogen bubbles from the surface of the cathode before they can grow large enough to cause pitting and streaking of the cathode.

Hydrogen evolution at the cathode increases the local pH and can lead to nickel hydroxide formation. Boric acid acts as a buffer and prevents the rise in pH. There is also a suggestion that boric acid can enhance the electrodeposition of nickel. Jayasekera and Kyle (1999) and Robinson et al. (2002) have provided a survey of the mechanical aspects of nickel tankhouse construction.

4.12. Phase disengagement in LIX®84-INS nickel circuits

A combination of low interfacial tension due to the nature of the ammoniacal leach solutions, high SX reagent concentrations, and the presence of degradation products in the circuit organic results in slow organic continuous phase separation times for these circuits. In addition, the high LIX®84-INS concentrations used and the presence of some polymeric degradation products mean that at 25 °C the circuit organic viscosity will be about 4–5 cst., which is higher than is typically observed in copper SX circuits. Nickel circuits however operate at temperatures in excess of 40 °C. At these temperatures, the nickel circuit organics have viscosities of around 2.5 cst., which is approximately the same as is observed in copper SX circuits at 25 °C. The interfacial tension of LIX®84-INS nickel circuit organics against 40–60 g/l ammonia solutions is between 13 and 18 dynes/cm. This is significantly lower than the 23–26 dynes/cm commonly observed in copper SX circuits. The low interfacial tension in ammoniacal nickel SX circuits will reduce the thermodynamic driving force for phase separation of emulsions.

The phase disengagement times observed in an operating plant will after a period of operation be in the order of

Table 5
Typical Ni EW operating parameters

Ni concentration and pH in EW feed	70–100 g/l, pH 3.0–3.5
Anolyte Ni and acid concentration	Approximately 60 g/l Ni and 35–45 g/l H ₂ SO ₄
Temperature	Approximately 60 °C
Current density	210–250 A/m ²
Cell voltage	3.6–3.9 V
Sodium sulphate	100 g/l
Sodium dodecyl sulphate	10–20 mg/l
Boric acid	5–10 g/l
Anodes	Rolled calcium–tin–lead alloy
Anode thickness	6–8 mm
Electrode spacing	160 mm

Organic continuous – extraction	20–30 min
Aqueous continuous – extraction	45–60 s
Organic continuous – strip	45–150 s
Aqueous continuous – strip	45 s

Not surprisingly the LIX®84-INS circuits are operated aqueous continuous in all stages. The role of degradation products in producing these phase separation times has been discussed in detail elsewhere (Mackenzie et al., 2005).

4.13. Remedial technologies for LIX[®]84-INS nickel SX circuit organic

The organic degradation observed in these circuits has focussed attention on techniques which could be used to remediate the circuit organic. The organic remedial technologies used in copper and uranium SX systems have been tested on LIX[®]84-INS nickel circuit organic. None have been markedly successful although some have shown some promise.

4.13.1. Clay treatment

Conventional clay treatment using acid-activated clay has not been particularly successful in improving either the physical (phase separation and entrainment) or the chemical (Ni/Zn selectivity, ammonia transfer and sulphate transfer) properties of the organic. There are some improvements in phase disengagement properties and in Ni/Zn selectivity and these may justify installation of clay treatment facilities in a nickel SX plant. However the benefits of clay treatment of nickel organics are markedly less than the benefits observed when clay treating copper circuit organics. Recently Cognis has developed and patented a process using chemically modified solid-state ion exchangers. This process has shown greater promise than conventional clay treatment but has not yet been tested in an operating plant.

4.13.2. Washing with alkali plus methanol

In their laboratories Cognis have shown that for some nickel circuit organics this technique significantly improves the performance of the organic, as measured by the performance of the washed and unwashed organic in the Cognis Copper QC test method. The wash aqueous was a 1:1 mixture of methanol plus pH 9.0 Na₂CO₃/NaHCO₃ buffer. The copper extraction and strip kinetics and the Cu/Fe selectivity were all improved. The extraction phase separation time (organic continuous) was however significantly increased.

4.13.3. Wipe film distillation

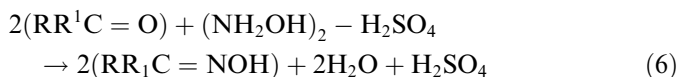
This was exhaustively tested on a nickel circuit organic in a Cognis pilot distillation plant. The feed was a slow phase separating organic which had been removed from the interface zone of a nickel SX plant settler. The wipe film distillation yielded positive results with 37% of the distillation products containing oxime plus ketone which could be re-oximated and returned to the SX circuit. The capital and operating costs of a distillation facility would however be considerable. In addition there are significant safety issues associated with handling oximes at high temperature which must be dealt with.

4.14. Re-oximation of LIX[®]84-INS nickel circuit organics

In nickel SX circuits using LIX[®]84-INS, the degradation rate of the oxime is such that there are economic ben-

efits to re-oximating the ketone degradation product. It should be noted that the degradation rate might not be very high until the organic has loaded some cobalt. Thus for the first months of a new plant operation the need to re-oximate may not be apparent.

The ketone degradation product present in LIX[®]84-INS nickel circuit organic can be re-oximated by the addition of hydroxylamine sulphate under alkaline conditions. The overall reaction is



The organic must be stripped of as much metal as possible prior to re-oximation. Usually this involves a reductive cobalt strip. The organic from the cobalt strip forms the feed to re-oximation. As a result, the capacity of the cobalt strip circuit in terms of cubic meters/hour of organic treated needs to be sufficient to provide the required organic flow rate to the re-oximation stage. Because the re-oximation is carried out under alkaline conditions, entrainment of acid from the cobalt strip to the re-oximation stage should be avoided. A wash stage on the stripped organic may be used. The re-oximation process can be batch or continuous. A batch process gives greater control over the reaction but requires a larger plant. In addition a batch process, being less automated, can involve more labour. Re-oximation is normally commenced once the ketone concentration in the circuit organic reaches a predetermined value. It should be noted that nickel circuits using LIX[®]84-INS cannot rely on re-oximation to completely replace the need to add fresh oxime and diluent. In order to maintain the quality of the circuit organic, fresh oxime and diluent must be added on a regular basis. Re-oximation can, however, provide at least 50% of the additional oxime requirements of an operating plant.

5. Application of the MHP–ammonia re-leach–LIX[®]84-INS SX–EW route to sulphide nickel orebodies

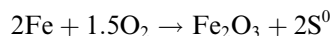
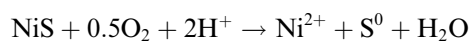
Although the MHP process route was originally developed for laterite orebodies, it may find application in the production of nickel metal from sulphide deposits. Such an application has been proposed before (Jones et al., 1998) in the CESL process. However there may alternative applications to the CESL process.

Nickel sulphide orebodies are typically hosted by Serpentine rocks which contain floatable gangue. Although depression of these floating gangue minerals and the production of smelter grade nickel sulphide concentrates is in some cases possible there are nickel orebodies which do not respond to this approach. For these orebodies the floating gangue may raise the MgO content of the concentrate to a level where conventional smelting is not viable. In addition, some nickel orebodies also contain arsenic bearing sulphide minerals which float and contaminate the concentrate to such an extent that smelting is no longer an option.

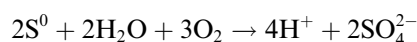
Autoclave leaching of sulphide minerals of copper and nickel is currently receiving considerable attention and both low pressure (LoPOX) and HPAL leach plants are at demonstration plant status (Kloiber et al., 2005).

These pressure leaches can be operated to produce Ni^{2+} and elemental sulphur or the oxidation can be carried further to produce Ni^{2+} and sulphuric acid.

The leaching chemistry for pentlandite $(\text{Ni,Fe})_9\text{S}_8$ can be simplistically described (Habashi, 1997) for the LoPOX and HPAL operations as



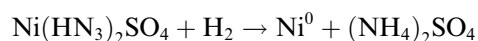
For complete oxidation leaching;



where elemental sulphur is produced in leach operations which are above the melting point of sulphur, surfactants may be added to disperse the sulphur and prevent large balls of fused sulphur forming in the leach vessel. Lignosulphonate is typical of the dispersants used. The lignosulphonate or its degradation products can produce phase separation and crud formation problems in SX operations which directly treat the leach solution. The use of a mixed hydroxide precipitation stage followed by an ammonia leach and LIX[®]84-INS SX will isolate the SX plant from any lignosulphonate additions made to leach.

The nickel sulphate solution produced by LoPOX or HPAL leaching of nickel sulphides can be treated for nickel recovery using the same MHP–ammonia leach–LIX[®]84-INS SX circuit as is employed for laterite ores. In these circuits there will be cobalt present in the leach solution. While it is possible to selectively extract this cobalt using Cyanex[®] 272 ahead of the nickel SX this is not recommended as any Cyanex 272 contamination of the downstream LIX[®]84-INS will result in rapid degradation of the oxime. It is preferable to use one of the processes for cobalt recovery from the nickel raffinate which have been developed for laterite ores.

An alternative route for nickel recovery from nickel sulphide leach solutions would be to add ammonia to the pH 3.0 nickel strip solution produced by SX with LIX[®]84-INS to produce a 40–50 g/l nickel solution with a 2:1 molar $\text{NH}_3/\text{Ni}^{2+}$ solution and to use this as feed to a Sherritt hydrogen-reduction process. The approximately 2:1 molar ratio has been established from Sherritt plant practice, although some Sherritt plants use a 1.9 molar ratio to prevent co-reduction of Co into the Ni product. As the loaded strip solution from SX will be very pure wrt Co, reduction of Ni is maximized. The 2:1 molar ratio also satisfies the requirements for fertilizer grade ammonium sulphate production:



The nickel SX strip solution will be virtually free of copper and this will remove the need for the “copper boil” process

typically used in a Sherritt plant. In addition the strip solution should be free of problematic species such as polythionates, thiosulphates and ammonium sulphamate.

The nickel tenor of the strip solution produced by a LIX[®]84-INS SX circuit can be much higher than what is typically treated in Sherritt plants. This implies that a reduction in size of the Sherritt plant should be possible, with significant effect on reducing capital costs.

6. Selection of contactor equipment for the solvent extraction of nickel from ammoniacal solutions using LIX[®]84-INS

There are several aspects of the process of the nickel SX process described in this paper which make the selection of the contactor equipment and some ancillary equipment a little more critical than it might be for copper SX from acid sulphate using oximes.

These include

- The requirement that all contactor stages operate aqueous continuous. The very slow organic-continuous phase separation prohibits operation in the organic-continuous mode.
- The need for large and positively controlled aqueous recycle loops to ensure aqueous continuous operation.
- The need for a minimum of 6 min residence time in the strip mixers.
- The need to contain and recover ammonia gas liberated in extraction and in the case of QNI used in stripping.
- The need for pH control in the loaded organic wash and strip stages.
- The requirement that the tankhouse advance electrolyte be free of organic entrainment in order to protect the zinc ion exchange and nickel electrowinning operations.

The two plants using LIX[®]84-INS technology employ conventional mixer settlers. Those at the Cawse plant were reverse-flow mixer settlers. Where containment of ammonia gas was required, simple water seals on the settler lids and mixer shafts have proved to be effective.

The use of pulsed columns was piloted at QNI (Parkes et al., 2000) with inconclusive results. Subsequently QNI selected mixer-settler units for their expansion.

Positive control of the aqueous-phase continuity is essential and this is achieved by the use of large aqueous recycles. The pumping turbine should be selected to give adequate feed and aqueous recycle volumes and the recycle lines should be designed to minimize friction losses.

In order to control the O/A ratio in the mixer flow meters on the aqueous recycle lines are recommended. There should also be at least two sample ports on the side of the mixer box to enable rapid determination of the phase ratio in the mixer. Conductivity probes should be fitted to all mixers. These should have read out facilities on the mixer top and in the control room.

Multi-stage mixing is used in both extraction and strip with the strip circuits employing three mixers per stage. The downcomer which transfers the emulsion from one mixer box to the next must be carefully designed to avoid the formation of aqueous locks in the secondary and tertiary mixers.

The requirement of pH control in the strip and loaded organic wash mixers means that the pH of the aqueous in these mixers must be continually measured. The main problem with pH measurement in SX circuits is presenting an organic free aqueous to the pH probe. For nickel circuits an option is to place the pH probe in the settler aqueous weir. This will result in a lag time of a few minutes between the mixer where the acid is added and the pH probe but once a pH profile is established over the four strip stages this lag time can be managed by correct set up of pH controllers. This has also been found to be the case in uranium SX circuits employing a pH profile in strip. The use of pressurized pH electrodes with a continual flow of KCl through the probe would be an advantage. These electrodes ensure that no organic gets into the probe. As the mixers will be operated aqueous continuous, direct pH measurement in the mixer should be possible. As with all pH measurement in SX circuits a vibration free mounting for the probe and the elimination of stray currents is essential.

Effective coalescing, organic de-entrainment and organic removal devices need to be installed on the aqueous exiting strip. Any entrained organic can foul the IX resin for the removal of Zn, and affect cathode quality.

7. Conclusions

The recovery of nickel from laterite sourced ammoniacal leach solutions using the ketoxime, LIX[®]84-INS, has been practiced by two Australian plants. One circuit used a HPAL–MHP route while the other uses a RRAL–SX route. Both plants have demonstrated the technical viability of this recovery route. While current sentiment seems to favour the intermediate product MHP route, the RRAL circuit operated by Queensland Nickel has been successful over 16 years of operation.

Currently two nickel laterite projects under active consideration are evaluating the solvent extraction of nickel using LIX[®]84-INS. One of these, the Vermelho project of Companhia Vale do Rio Doce, may use the HPAL–MHP route combined with SX and acid strip followed by electro-winning while the other, the BHP Billiton Ravensthorpe project, will use the HPAL–MHP solvent extraction–ammonia strip route.

Operating plants using an aged circuit organic show significantly different SX characteristics from pilot circuits using fresh organic. These differences result from the degradation of the organic during operation. This degradation is associated with the extraction of cobalt(II) by the organic and the subsequent removal of this cobalt using a reductive strip.

Plant operators have successfully overcome the operational issues associated with this degradation. Re-oximation of the organic plays an important role in maintaining the health of the circuit organic but cannot completely eliminate the need to add fresh oxime and diluent.

Although the performance of aged circuit organic differs from that of fresh organic, it should be noted that both of the Australian LIX[®]84-INS SX circuits were successfully designed using data obtained in pilot plants operating with fresh organic.

Investigative work by Cognis, by project teams, and by plant operators over the past decade have significantly increased the nickel industries understanding of the technology involved in operating a LIX[®]84-INS nickel solvent extraction circuit. There are many potential circuits available for the recovery of nickel from laterites but the LIX[®]84-INS based circuits have emerged as strong contenders for any new laterite nickel project. Current opinion seems to favour the MHP–ammonia leach–acid strip circuit but the ammonia strip analogue and possibly the RRAL–ASX circuit should also be evaluated for any new laterite project before a final decision is made.

Acknowledgements

The authors acknowledge the permission of Cognis to publish this paper.

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