TREATMENT OF NICKEL LATERITES BY CHLORIDE AND HYBRID CHLORIDE-SULPHATE PROCESSES

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ABSTRACT

With laterites representing about 70% of the world's nickel reserves, there continues to be an incentive to develop an economically and technically viable process for their treatment. A number of sulphate-based and chloride-based processes have either been built or have been proposed over the past fifteen years, but none has really been successful. PAL and HPAL (high pressure acid leaching) continue to be the favoured routes, but as long as the prices of sulphur and sulphuric acid remain high, it is doubtful whether such processes can be truly economic. A number of chloride-based processes have also been proposed, but these depend very much on the economic viability of both regenerating and recycling hydrochloric acid and controlling magnesium.

The present paper briefly looks at the merits of both sulphate and chloride processing, and at the potential of hybrid chloride-sulphate circuits, nominally combining the best aspects of each. A number of potential scenarios are presented, with order of magnitude capital and operating costs for each. It is concluded that flowsheets which offer additional marketable by-products, such as hematite (iron ore) and magnesia, have a greater chance of being successful, provided that by-product values are included in the overall economic assessment.

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INTRODUCTION

Laterites comprise ~70% of the world's nickel reserves, and despite the very severe economic turndown experienced at the end of 2008, nickel will nevertheless remain a very important commodity as the economies of Asia, South America and Africa continue to develop. It is essential, therefore, that a viable extraction process, both technical and economic, be found for the recovery of nickel from laterite deposits. Conventionally, the traditional approach has been:

- Consider only one medium, either sulphate, chloride, or, occasionally, ammonia;
- Focus only the valuable metals (nickel and cobalt) in the ore, and consider the economic assessment only on nickel;
- Treat iron and magnesium (and aluminium) as impurities, with only nuisance value.

Given that it is almost impossible to significantly upgrade laterites, the whole ore has to be processed, resulting in the size of the front end leaching process being much larger than in conventional hydrometallurgical plants. Nevertheless, attempts have been made to build plants that treat the whole ore, and over the past decade, there has been some acceptance of the High Pressure Acid Leach (HPAL or PAL) processes as a viable process technology, albeit one wherein the economics are marginal at best.

Unfortunately, laterites do not easily lend themselves to heap leaching, the normally accepted method of treating low-grade oxidic ores, although there are notable exceptions such as European Nickel in Turkey [1]. For heap leaching to be successful, the ore has to have certain physical characteristics which allow it to be agglomerated and have the necessary integrity to form heaps, whilst at same time permitting percolation of the lixiviant through the heap. Most tropical laterites do not have these characteristics, and therefore the heap leach approach is not possible.

Sulphuric acid leaching, because of its very nature, is essentially a once-through process, with whatever acid is added to effect dissolution of nickel and cobalt exiting the process as gypsum and/or magnesium sulphate. Attempts have been made to consider calcining the magnesium sulphate to magnesia, and recycling the sulphur dioxide to an acid plant. However, the calcination process is difficult, inefficient and really requires the monohydrate as feed compared to the more normal higher hydrates.

A number of chloride-based processes have also been proposed, notably by the present authors. Inherent in these processes is the absolute necessity to recycle hydrochloric acid and minimise chloride losses. The conventional approach employing pyrohydrolysis is energy-intensive, and can generate only 18% acid at best, thus placing a load on the water balance of the circuit. As with the sulphate approach, magnesium is leached, and the end product of the circuit is magnesium chloride which has to be dealt with.

Most of the processes either in operation or contemplated do not take into account by-products. One notable exception is Sherritt's Ambatovy Project in Madagascar [2], where considerable amounts of ammonium sulphate will be produced and sold as fertilizer. However, all laterites contain significant amounts of iron, magnesium and aluminium. At the time of writing, the demand for iron, especially from China, is again picking up, and therefore if the nickel extraction process can generate a good quality hematite, this would be an added bonus to the project. At the very worst, the product is benign and can be returned to the mine, but potentially, it is an additional source of revenue. The same applies to magnesium, and to a lesser extent aluminium. Generating a high-quality magnesia product from magnesium chloride, in particular, is not difficult.

The situation with aluminium is somewhat different. In order to be able to market hematite, much of the aluminium has to be removed from the iron. The unusual solubility properties of $AlCl_3$ suggest that it might be possible to control aluminium dissolution during leaching or possibly it can be salted out of solution [3] and treated separately.

The original objective of this paper was to evaluate some hybrid circuit, combining the best aspects of both chloride and sulphate, and to see whether there is any merit in doing so. Both processing circuits have advantages and disadvantages, and the aim was to attempt to combine the best of both into one circuit. Several different scenarios are proposed, notably varying the feed nickel concentration to determine if there is a level below which processing will never be economic.

SULPHATE-BASED CIRCUITS

Sulphate-based circuits essentially fall into two categories, those under pressure and those attempting atmospheric leaching only. The fundamentals of pressure leaching are such that twice the amount of acid as needed stoichiometrically for the valuable metals has to be added because sulphuric acid is monoprotic at the operating temperature of the autoclave (240-270°C). However, it is still less than in atmospheric sulphate leaching, because the iron and aluminium re-precipitate in the pressure leach, releasing acid, whereas in the atmospheric case it does not. The first, and probably most successful, laterite leach plant is that at Moa Bay in Cuba, which first commenced operations in 1959 [4]. The process is based on a sulphuric acid pressure leach at around 240°C in pachuca autoclaves, and precipitation of a mixed nickel/cobalt sulphide intermediate for further refining. In the mid-1990s, three major pressure leach projects were commissioned in Western Australia, Murrin Murrin (now Minara Resources), Bulong and Cawse [5]. Of these three, only Cawse has had any true success [6]. Recently, the Rio Tuba project in the Philippines was successfully commissioned [7-9], although it has to be said that this project benefited greatly from the existing infrastructure and port facilities at the site. Other recent projects include BHP Billiton’s Ravensthorpe Project [10], Dynatec’s Ambatovy Project in Madagascar [2], Vale Inco’s Goro Nickel Project in New Caledonia [11], Eramet’s Weda Bay Project [12], and the Gladstone Pacific Nickel Project in Queensland, Australia [13-15]. Of these, BHP Billiton has announced the closure of Ravensthorpe, but the others are believed to be proceeding.

All of these projects are based on HPAL. An advantage of sulphate circuit is that a dedicated acid plant can be built, which not only supplies the acid but also much of the power needed to operate the process plant. Being able to generate this power minimises, or eliminates altogether, the need for imported fossil fuels, and therefore significantly reduces the carbon footprint of the plant. The major disadvantage is that the acid cannot be recycled, and results in a gypsum or magnesium sulphate effluent. A further disadvantage is that the sulphuric acid route is largely limited to processing just the limonite (low magnesium) horizon of the laterite. Attempts have been made both to eliminate the use of autoclaves and to neutralize a portion of the acid using saprolite (high magnesium) laterite [16, 17], but this still leaves a substantial magnesium sulphate effluent to dispose of.

CHLORIDE-BASED CIRCUITS

Although several chloride-based circuits have been proposed over the past few years, none has been built, or has yet been piloted [18-26]. The advantages of chloride circuits are atmospheric operation, better leach residue filtration, and the ability to generate valuable by-products. The disadvantages are the need to recycle the acid and more expensive materials of construction. Compared to sulphate processing, relatively little effort has been applied to developing chloride circuits, so that any comparative evaluation is necessarily limited.

Chloride circuits make use of the high proton activity afforded by a strong chloride matrix to effect efficient leaching of the ore, and maximise pay-metal recovery. This results in significant co-dissolution of iron, and to a lesser extent magnesium and aluminium, which then have to be dealt with. Of the circuits mentioned above, only the one conceived by Intec [24] has attempted to combine chloride with sulphate. However, the iron and magnesium residues from the flowsheet were not considered to be marketable, although the calcium sulphate product was. Additionally, the sulphuric acid consumption appeared to be on the high side, and nothing further of this development has been published.

CIRCUITS CONSIDERED FOR STUDY

Basis

All of the circuits considered were based on the chloride leaching approach developed in-house, which has extensively been modeled, and of which the key unit operation is the low-temperature hydrolysis of iron using water as the only reagent [23, 27-31]. The hematite produced by this technique is extremely pure, generally with <0.1% chloride, and with no detectable base metals. Similarly, the circuit allows for the recovery of a pure magnesia product, and both of these can have a significant impact on the overall project costs.

Four case studies were examined by generating and examining detailed mass/energy balances for each case, and varying the grade of base metals (nickel, cobalt and copper) in the feed laterite. Table 1 lists the analyses of the feed laterite for each of the four cases, which is based on a mixture of the various profiles of a low-grade laterite deposit. The feed also contains aluminium, chromium, manganese, zinc, cadmium and calcium. An initial study, called Case 0, was also carried out, where Case 0 assumes the same feed as Case 1, but the final recovery of hydrochloric acid is achieved by first converting the barren solution from magnesium chloride to calcium chloride by adding lime, which also generates magnesium hydroxide. The concentrated solution of calcium chloride is then contacted with concentrated sulphuric acid, generating calcium sulphate and volatile hydrochloric acid. As noted above, the development of Case 0 had been the original impetus for carrying out this study.

Table 1 – Laterite analyses, mass %

Assay	Case 1	Case 2	Case 3	Case 4
Nickel	0.75	1.00	1.25	1.50
Cobalt	0.08	0.11	0.13	0.16
Copper	0.01	0.01	0.01	0.01
Iron	20.3	20.2	20.1	20.0
Magnesium	8.0	7.93	7.89	7.86

The fundamental flowsheet contains the following unit operations, which are described below.

Leach

The incoming laterite is leached with recycled strong hydrochloric acid, with the acid being almost totally consumed due to the high proton activity resulting because of the high chloride concentration. Equations 1, 2 and 3 list the main leaching reactions, where the symbol Me denotes the divalent metals (nickel, cobalt, copper, manganese, zinc, cadmium, calcium and magnesium) and M denotes the trivalent metals (iron, aluminium and chromium). The divalent metals are distributed between oxide and silicate minerals.

$$MeO + 2H^+ \rightarrow Me^{2+} + H_2O \tag{1}$$

$$MeSiO_3 + 2H^+ \rightarrow Me^{2+} + 2H^+ + SiO_2 \tag{2}$$

$$M_2O_3 + 6H^+ \rightarrow 2M^{3+} + 3H_2O \tag{3}$$

Part of the leached slurry is recycled to mix with incoming ore, in order to give a high chloride background to the solution. The balance is settled and the overflow is boiled to drive off any residual hydrochloric acid, which returns to the leach. The settler underflow is filtered and washed, and the washed residue (predominantly unleached silicate gangue) leaves the circuit. The wash filtrate is evaporated in a two-stage sequence to concentrate it to the same level as the primary filtrate. However, this is expensive, and future flowsheets will look at more effective and efficient methods of handling this wash. Steam from the second effect is condensed in the first effect, which is operated under reduced pressure. The steam from the first effect is condensed with cooling water and becomes process water.

Evaporation and Hydrolysis

The concentrated wash filtrate, primary filtrate and the boiled settler overflow are combined and evaporated to a chloride concentration of about 18 molar. The steam and the resulting highly concentrated liquid are heated further in the hydrolysis section, causing the ferric chloride to decompose to gaseous hydrochloric acid and steam, with the iron precipitating as dense, highly crystalline hematite. Equation 4 shows the stoichiometry used to represent this. Aluminium and chromium, also being trivalent, behave similarly to iron.

$$2FeCl_3 + 3H_2O \rightarrow Fe_2O_{3\downarrow} + 6HCl^{\uparrow} \tag{4}$$

A major aspect of the chemistry around the hydrolysis step is that sufficient H_2O must be present, both to react with the trivalent metal chlorides and to remove the HCl from the gas above the melt. Steam is added to ensure this. Experimental work to date has achieved HCl/steam mixtures containing well over 30% HCl leaving the reactor. Steam is added, rather than water, because firstly it is available, and secondly, the overall hydrolysis process is strongly endothermic and adding water would unnecessarily increase the heat transfer duty required of the process equipment. The indirect energy is supplied as condensing medium pressure (12 Bar) steam from the energy recovery section. The liquid phase leaving the hydrolysis train, at about 30 molar chloride, is more a molten salt containing some water than an aqueous solution – i.e. a molten salt hydrate. (The thermodynamics around the hydrolysis section will be described in more detail in a future publication.)

The hematite slurry leaving the hydrolysis train is filtered hot. A characteristic of the hematite produced is that the hot filter cake contains about 90 percent solids; testwork has shown as low as 4% moisture after washing, but 10% moisture has been used in the present study. The filtrate is returned to the start of the hydrolysis train. The per-pass conversion of ferric chloride to HCl and hematite has been taken as 75%, although the experimental work indicates that whilst this is easily achieved, it is a simplification of the actual processes involved. However, for the purposes of this study, the mass/energy balances assume this value, but the overall conversion, taking into account the recycled hot filtrate, is about 98%. The liquid in the unwashed hot filter cake constitutes the base metals advancing from the hydrolysis section.

Purification and Base Metal Recovery

The hot filter cake is quenched and diluted with recycled solution from the following step, in which magnesium oxide and oxygen are used to oxidize any ferrous iron to ferric iron and precipitate the last of the iron onto the hematite. The chemistry of the circuit is managed such that the chloride concentration remains sufficiently high for the copper, cobalt, zinc and cadmium to be present as anionic chloride complexes, enabling the use of anionic ion exchange and/or solvent extraction in the subsequent recovery steps. Nickel and manganese either do not form anionic chloride complexes, or form very weak ones such that they effectively behave as cations. The resulting slurry is settled, the supernatant being used to quench and dilute the hot filter cake from hydrolysis. The underflow is filtered and washed with water. The washed hematite leaves the circuit. The combined primary and wash filtrate passes through an anion exchange step to remove zinc and cadmium impurities. The purified solution is subjected to anionic solvent extraction to separate out the copper and cobalt, leaving the nickel and manganese in the raffinate. In both of these steps, water is used to elute/strip the resin/solvent, causing the anionic chloride complexes to dissociate and be released. The copper and cobalt removed from the anion exchange solvent are separated and recovered by fairly conventional cationic solvent extraction and electrolysis. The nickel in the raffinate from the anionic solvent extraction step is also recovered by fairly conventional cationic solvent extraction and electrolysis. Prior to nickel recovery, the manganese is removed – this exercise assumed electrolysis to manganese dioxide, but other methods such as chlorination could also be used.

Acid Recovery

Case 0

The barren solutions leaving the cationic solvent extraction steps are partially recycled to the quench/dilution step after the hydrolysis section and the balance is neutralized with calcium oxide (lime) to precipitate any remaining divalent metals and magnesium. The resulting hydroxide slurry is filtered and washed, the washed solids leaving the circuit. The filtrate is evaporated under reduced pressure, using condensing steam from the leach section, to concentrate it ahead of the next step. The concentrated solution of calcium chloride is contacted with concentrated sulphuric acid, forming calcium sulphate and releasing volatile hydrochloric acid that is separated from the slurry by boiling. The resulting gaseous hydrochloric acid and steam return to the leach. The calcium sulphate is highly crystalline, and hence readily filterable, and leaves the circuit.

Cases 1 Through 4

The barren solutions leaving the cationic solvent extraction steps are partially recycled to the quench/dilution step after the hydrolysis section and the balance is neutralized with magnesium oxide to precipitate any remaining divalent metals other than magnesium (and a little calcium). The resulting hydroxide slurry is filtered and washed, the washed solids leaving the circuit. The filtrate is evaporated under reduced pressure, using condensing steam from the leach section, to concentrate it ahead of a step that is similar to conventional pyrohydrolysis, in which the magnesium chloride is thermally converted into magnesium oxide and gaseous hydrochloric acid. The resulting gaseous acid/steam returns to the leach section instead of being cooled and scrubbed with process water (as is done in conventional pyrohydrolysis). Leaching with HCl gas is surprisingly efficient and effective, and results in a better-filtering leach slurry. The magnesium oxide is partially recycled to the various neutralization steps in the circuit, but most simply leaves the circuit, carrying out the calcium.

Energy Balance

The evaporation and hydrolysis section of the circuit is strongly endothermic and steam at the appropriate pressure is required to drive the chemistry, a requirement which is common to all cases. The steam leaving the hydrolysis and acid recovery sections is used to heat the leach section, but the available latent heat in the returning steam/acid is substantially more than is required merely to heat the leach. The excess latent heat is captured as low pressure steam in the leach section. The present assumption is that steam coils in the leach reactors can be used to keep the leach reactors at about 110°C, converting boiler feed water to atmospheric pressure steam. Another option would be partially condensing the steam/acid in a heat exchanger, just ahead of the leach train. Some of this steam is sent to the hydrolysis section to supplement the steam from the evaporation stage. Some of the remaining steam is compressed to 12 Bar and used to drive the hydrolysis chemistry. Some is compressed to 6 Bar and used to drive the evaporation step ahead of the hydrolysis train. The balance is used to drive the reduced-pressure evaporation step after the recovery of the base metals.

Case 0

In this case, it is assumed that the mechanical energy required for steam compression comes from high-pressure steam generated in a sulphuric acid plant with the shortfall coming from a gas-fired turbine. The hot exhaust gas from this turbine is used to generate more 12 Bar and 6 Bar steam, minimizing the amount of steam requiring compression.

The sulphuric acid required for the conversion of calcium chloride to hydrochloric acid and calcium sulphate is generated on site by burning elemental sulphur in air and converting the resulting sulphur dioxide to concentrated sulphuric acid. The energy released by the sulphuric acid plant is captured as high-pressure steam that is used to drive an expansion turbine, producing some of the mechanical energy that is used to compress low-pressure steam from the leach section.

Cases 1 Through 4

The mechanical energy required for steam compression was assumed to come entirely from a gas-fired turbine. The hot exhaust gas from this turbine is used to generate more 12 Bar and 6 Bar steam, minimizing the amounts of steam requiring compression.

Results

The mass/energy balances are based on the following assumptions:

- An annual nickel production of 50,000 tonnes.
- A plant availability of 350 days per year, 24 hours per day.
- Use of natural gas (57.7 GJ/t) as a fuel source.

The mass/energy balances generated the results summarized in Table 2. These numbers were used to size the process equipment and to generate estimates of the capital cost associated with each case.

Table 2 – Main Process Flows, t/h

Process Flow	Case 0	Case 1	Case 2	Case 3	Case 4
Incoming ore, dry basis	872	864	646	517	431
Fe_2O_3 ex hydrolysis,	257	259	193	153	127
Acid ex hydrolysis, as 100% HCl	355	345	256	204	169
Acid ex $MgCl_2$ conversion, as 100% HCl	167	170	128	104	88
Total evaporation load, H_2O	1741	765	573	460	384
Total steam in regenerated acid	851	814	618	504	427
Total steam raised from the leach	1441	890	640	493	394
Steam compressed to 6 Bar	776	221	189	170	157
Steam compressed to 12 Bar	765	303	213	160	125
Natural gas consumed as fuel	25	76	58	47	40

ECONOMIC ANALYSIS

The predominant aim of this exercise was to conduct an economic analysis of the chloride and hybrid processing routes for treating nickel laterites. As with all such exercises, it has been broken down into the capital component and the operating component.

Capital Costs

The capital costs are summarised in Table 3. Essentially, Cases 0 and 1 have been treated the same, with the cost of the pyrohydrolysis equipment for magnesium chloride being equated to that of an acid plant plus sulphur burner. It is recognised that this is a gross over-simplification, but it is sufficient for the purposes of the present exercise. These costs have order of magnitude accuracy, and have been derived from a similar project using accepted standard engineering factors.

The costs reflect the ore treatment rate required for a plant to produce the baseline tonnage of 50,000 tpa of nickel metal. They are consistent with costs published for other large nickel laterite plants such as Ravensthorpe, Ambatovy and Gladstone Pacific Nickel, especially for Cases 0 and 1. It is clear that as the nickel grade of the ore improves, then the capital expenditure comes down for an equivalent tonnage of nickel, as would be expected.

Table 3 – Capital Cost Summary

	Cost, US$ million				
	Case 0	**Case 1**	**Case 2**	**Case 3**	**Case 4**
Mining Equipment	60	60	47	41	37
Ore Preparation Storage and Handling	177	177	121	99	84
Leach	90	90	70	61	55
Evaporation/Hydrolysis	127	127	100	86	86
Purification	15	15	12	10	9
IX/SX	171	171	135	113.5	97.5
EW	72	72	70	68	67
Evaporation	12	12	9	8	7
Pyrohydrolysis	-	162	126	108	90
Sulphur Burner/Acid Plant	162	-	-	-	-
Power Plant	270	270	210	184	164
Reagents Handling	10	10	8	7	6
Utilities and Auxiliaries	30	30	23	20	18
Direct Equipment Costs (DEC)	1196	1196	931	807	713
Plant Factor (2 times)					
Sub-Total Direct (Installed) Costs	**2392**	**2392**	**1862**	**1614**	**1427**
EPCM (12% Installed Cost)	287	287	223	194	171
Owner Costs (4% Installed Cost)	96	96	74	65	57
Spares and Maintenance (5% DEC)	60	60	47	40	36
Transportation (12% DEC)	144	144	112	97	86
Chemicals Initial Fill (3% DEC)	36	36	28	24	21
Sub-Total Indirect Costs	**622**	**622**	**484**	**420**	**371**
Total	3015	3015	2346	2034	1798
Contingency (15%)	452	452	352	305	270
Total Capital Investment	**3467**	**3467**	**2698**	**2339**	**2067**

Operating Costs

Table 4 lists the amounts of reagents and utilities consumed per tonne of base metal (nickel plus copper plus cobalt) produced. The consumptions of hydrochloric and sulphuric acid are to replenish losses, except for Case 0, where sulphuric acid is required for HCl regeneration. The water is mainly cooling water make-up. Table 5 lists the unit prices used and the calculated contributions of each reagent or utility to the variable cost. By far the dominant contribution throughout is the cost for natural gas, which is used to drive the gas turbine used to compress steam for the hydrolysis and evaporation sections. However, the cost of sulphur is also highly significant in Case 0.

Table 4 – Reagent and Utility Consumptions, per tonne of base metal produced

Item	Case 0	Case 1	Case 2	Case 3	Case 4
HCl (100%), t	0.26	0.54	0.40	0.32	0.27
H_2SO_4 (100%), t	33.8	0.09	0.09	0.09	0.09
Natural gas, GJ	394	659	500	406	344
Water, t	228	154	122	102	88
Power, MWh	14	14	11	10	9

Table 6 lists the capital and operating costs estimated from the numbers emanating from the mass/energy balances. As noted elsewhere, the capital costs, in particular, are indicative only as they were derived by extrapolation from information on a similar but slightly different circuit. The operating costs are also shown as per pound of nickel produced. This does not include any co-credit whatsoever, not even for cobalt. Table 7 lists the selling prices assumed for this exercise and the various quantities produced.

Table 5 – Reagent and utility costs, US$ per tonne of base metal

Item	Unit price	Case 0	Case 1	Case 2	Case 3	Case 4
HCl (100% basis)	$857/t	0.10	0.21	0.12	0.09	0.08
H_2SO_4 (100%)	$200/ton	3.08	0.01	0.01	0.01	0.01
Oxygen	$750/t	0.02	0	0	0	0
Natural gas	$6.50/GJ	0.02	1.95	1.45	1.18	1.00
Electricity	$11.50/GJ	1.17	0.26	0.21	0.18	0.16
Water	$0.25/ton	0.03	0.02	0.01	0.01	0.01

Table 6 – Estimated Capital and Operating Costs

Item	Case 0	Case 1	Case 2	Case 3	Case 4
Estimated capital cost, $ million	3467	3467	2698	2339	2067
Operating costs, US$ million/year:					
Mining cost ($10/t mined)	73	73	54	43	36
Chemicals and raw materials	394	27	15	12	10
Consumables	2	2	2	2	2
Manpower	9	9	9	9	9
Utilities	149	273	205	168	144
Total Operating Costs, US$ million/year	**627**	**384**	**285**	**234**	**201**
Cost, US$/lb Ni	**5.69**	**3.48**	**2.59**	**2.12**	**1.82**

Table 7 – Prices and Quantities of Metal Produced

Metal	Selling price	Quantity produced, kt/year				
		Case 0	Case 1	Case 2	Case 3	Case 4
Nickel	$5/lb	50.0	50.0	50.0	50.0	50.0
Cobalt	$20/lb	5.6	5.6	5.6	5.6	5.6
Copper	$2/lb	0.4	0.4	0.4	0.4	0.4
Hematite	$50/t	2191	2211	1641	1306	1082
Magnesia	$100/t	500	847	641	519	438
Gypsum	$50/t	2624	-	-	-	-

Financial Models

Table 8 shows the gross revenues from each of the cases, where it can be seen that both hematite and magnesia make a significant contribution, although nickel and cobalt remain predominant. These numbers were used to generate a simple cash flow analysis for each of the four cases examined. The assumptions going into these cash flow analyses are as follows:

- The capital expenditure occurs over the two years preceding the first year of production.

- The production achieved is 25 percent of design in year 1, 50 percent in year two, 75% in year 3 and 100% subsequently.
- The full operating cost is incurred from year 1.
- The applicable tax rate was assumed to be 25%.
- The project life was assumed to be 20 years.

Table 8. Gross Revenues, US$ million/year

Gross Revenue, US$ million/y					
	Case 0	Case 1	Case 2	Case 3	Case 4
Nickel	550	550	550	550	550
Cobalt	245	245	245	245	245
Copper	1	1	1	1	1
Hematite	110	111	82	65	54
Magnesia	50	85	64	52	44
Gypsum	131	-	-	-	-
Total	1087	991	942	913	894

Table 9 shows the results of these calculations, using the numbers in Table 6 and Table 7.

Table 9 – Cash Flow Analyses, no value for Fe_2O_3, MgO or $CaSO_4$

Calculation	Case 0	Case 1	Case 2	Case 3	Case 4
Internal rate of return, %	0	7	12	15	17
NPV at 10% discount rate, US$ million	-2152	-689	358	852	1201

The experimental work done to date indicates that it is possible to set up the hydrolysis section such that the aluminium and iron are removed separately. (The chromium is a very minor component – its deportment is not yet confirmed, but it is assumed to follow the aluminium.) If the hematite is produced separately from the aluminium, it is likely to be a good source of high grade iron ore. If the hematite can be sold at, say, $50/t instead of being discarded, the cash flow analyses change to those shown in Table 10.

Table 10 – Cash Flow Analyses, $50/tonne for Fe_2O_3

Calculation	Case 0	Case 1	Case 2	Case 3	Case 4
Internal rate of return, %	3	9	14	16	18
NPV at 10% discount rate, US$ million	-1580	-162	739	1151	1449

Similarly, pyrohydrolysis is a standard method of making magnesia, having been used at the Dead Sea in Israel for many years. Hence, the magnesia produced should also be marketable, and if a value of $100/tonne is assigned, then the cash flow analysis changes as shown in Table 11. For the purposes of this study, for Case 0, it has also been assumed that the gypsum produced can be sold at a value of $50/t.

Table 11 – Cash Flow Analyses, $50/tonne for Fe_2O_3, $100/tonne for MgO and $50/tonne for $CaSO_4$

Calculation	Case 0	Case 1	Case 2	Case 3	Case 4
Internal rate of return, %	6	11	15	17	19
NPV at 10% discount rate, $ million	-930	236	1035	1390	1648

DISCUSSION AND CONCLUSIONS

General Considerations

The intent at the beginning of this exercise was to demonstrate that combining the best of chloride processing with that of sulphate could lead to a techno-economically viable flowsheet for treating the whole of the nickel laterite profile, and this circuit was designated as Case 0. However, it is apparent that no matter how the two circuits are combined, a cost-effective hybrid chloride-sulphate flowsheet cannot be conceived, especially for low-grade laterites, even with generating by-product values for iron (hematite), magnesium (magnesia) and gypsum. Consequently, four other cases were considered, with the nickel content being varied from 0.75% to 1.5%, and using a chloride-only flowsheet.

The results of this exercise indicate that the envisaged chloride route could be economically competitive, even for nickel grades as low as 0.75% and currently (April 2009) relatively low nickel prices. Cases 3 and 4 can be justified on nickel and cobalt alone, whereas Case 1 could not, and Case 2 is borderline. These results highlight the importance of not only considering the whole ore, but also taking into account the value of potential by-products. The chloride processing route generates high quality products of iron (hematite) and magnesium (magnesia) which have appreciable value in themselves. These products are generated as a natural consequence of the processing route, and hence incur no additional capital or operating expense. The economic analyses presented show the value of considering these by-products, and demonstrate that what would otherwise be designated as marginal or uneconomic on nickel alone, can have significant merit when taking into account all of the potential products.

The history of nickel laterite ore bodies, except for those producing ferro-nickel from the saprolite horizon, is that they are marginal, and to date, no one process has been capable of processing both horizons, saprolite and limonite. The chloride route is adaptable, being able to treat either or both, and additionally generating valuable by-products.

Future Considerations

The chloride processing route is still in its infancy, and more development work is required. The energy/water balance remains the focus of ongoing work, as well as refining the hydrolysis circuit(s), particularly for aluminium and magnesium. A lower temperature method than pyrohydrolysis of converting magnesium chloride shows promise, but whether it will be competitive with pyrohydrolysis remains to be seen. Separation of iron and aluminium, however, shows considerable promise.

There is considerable scope for improvement of the basic chloride flowsheet presented here, and work is continuing with a view to achieving such goals. The analyses presented in this paper show that the chloride-based approach has sufficient merit to continue its development.

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