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TECHNICAL & COST COMPARISON OF LATERITE TREATMENT PROCESSES
PART 4 – HYDROXIDE TO METAL

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ABSTRACT

This is the fourth and final paper in a series examining the economics of processing a hypothetical nickel-cobalt laterite. Part 1 outlined the processes examined and their applicability to limonite and saprolite. Part 2 presented the results of process modelling done to quantify reagent and utility requirements and to calculate the variable portion of the operating costs. Part 3 extended the comparison to the fixed operating and capital costs and used simple financial modelling to compare the different processes. In these papers, the assumption was that the nickel and cobalt are recovered as an intermediate oxide or hydroxide product that would be processed further elsewhere.

This paper examines processing the intermediate product to nickel and cobalt metal. The question addressed is whether or not on-site processing of the intermediate product would enhance the economics of the project. Three possible routes for this onward processing are examined:

- Leaching the intermediate product in an ammonia-ammonium sulphate system, with solvent extraction and electrowinning of the nickel, precipitation of cobalt sulphide from the resulting raffinate, oxidative leaching of the cobalt sulphide and electrowinning of cobalt from the resulting solution.
- Leaching the intermediate product in spent electrolyte from nickel electrowinning, solvent extraction and electrowinning of cobalt from the resulting solution and electrowinning nickel from the raffinate.
- Eliminating production of the intermediate hydroxide by using synergistic solvent extraction to recover nickel directly from the solution from which the intermediate product would otherwise be produced, stripping the organic phase with spent electrolyte and electrowinning nickel from the loaded strip liquor. Then using the same synergistic solvent extraction system to extract cobalt from the nickel raffinate, stripping and electrowinning cobalt in the same way as the nickel.

While the previous papers examined several processing routes for making the intermediate product, this paper assumes just the conventional HPAL route, but the analysis would apply equally to the other routes making a mixed hydroxide.

The three onward processing options are ranked. The impact on the process economics of metal and reagent prices and the percentage of the metal prices for which the intermediate hydroxide is sold are examined.

INTRODUCTION

Increasingly, future nickel supplies will have to be produced from lateritic deposits, including the lower grade limonite zones which are not suitable for pyrometallurgical processing. This has led to pressure acid leaching and to other processes operating at atmospheric pressure, including heap leaching and agitated tank leaching with sulphuric acid. Processes using hydrochloric acid and nitric acid are also under development.

The technical status of several processes was presented in the Part 1 of this series[1]. Part 2 presented a review of the reagent and utility costs associated with the established and the developing processes[2]. Part 3 extended the comparison to include capital and fixed operating costs[3].

The processes covered in Parts 1 to 3 of this series are listed in Table 1. The commercial category contains processes that are currently applied to laterites on a stand-alone basis. Heap leaching and atmospheric tank leaching with sulphuric acid have been commercially applied[1]. The developing processes are the Neomet and Direct Nickel processes.

Table 1 - Processes & Products

Processes	Product
Commercially Applied	
Pressure Acid Leaching	Nickel-cobalt hydroxide
Enhanced Pressure Acid Leaching	Nickel-cobalt hydroxide
Heap Leaching with sulphuric acid	Nickel-cobalt hydroxide
Caron Reduction Roast-Ammonia Leach	Nickel oxide and nickel-cobalt sulphide
Rotary Kiln Electric Furnace Smelting	Ferronickel
Sintering/Blast Furnace Smelting	Nickel pig iron
Sintering/Submerged Arc Smelting	Nickel pig iron
Commercially Applied as Satellite Operations	
Atmospheric tank teaching with sulphuric acid	Nickel-cobalt hydroxide
Developing Technologies	
Neomet chloride leach	Nickel-cobalt oxide
Direct Nickel nitric acid leach	Nickel-cobalt hydroxide

The results presented in Part 3 indicated that, in the absence of by-product revenue, the HPAL technology appears to offer the strongest overall economics for processing laterite, at least to the intermediate product. A caveat is that the two developing processes, Neomet and Direct Nickel, both reject the iron as pure hematite, and if this can be sold for appreciably more than as iron ore, both Neomet and Direct Nickel could well offer distinctly superior economics.

HPAL technology was chosen as the baseline for the exercise presented in this paper because it is the dominant currently proven option. While similar exercises could certainly be done for the other laterite processing routes, covering them all in a single paper would be excessive.

FEED AND HPAL

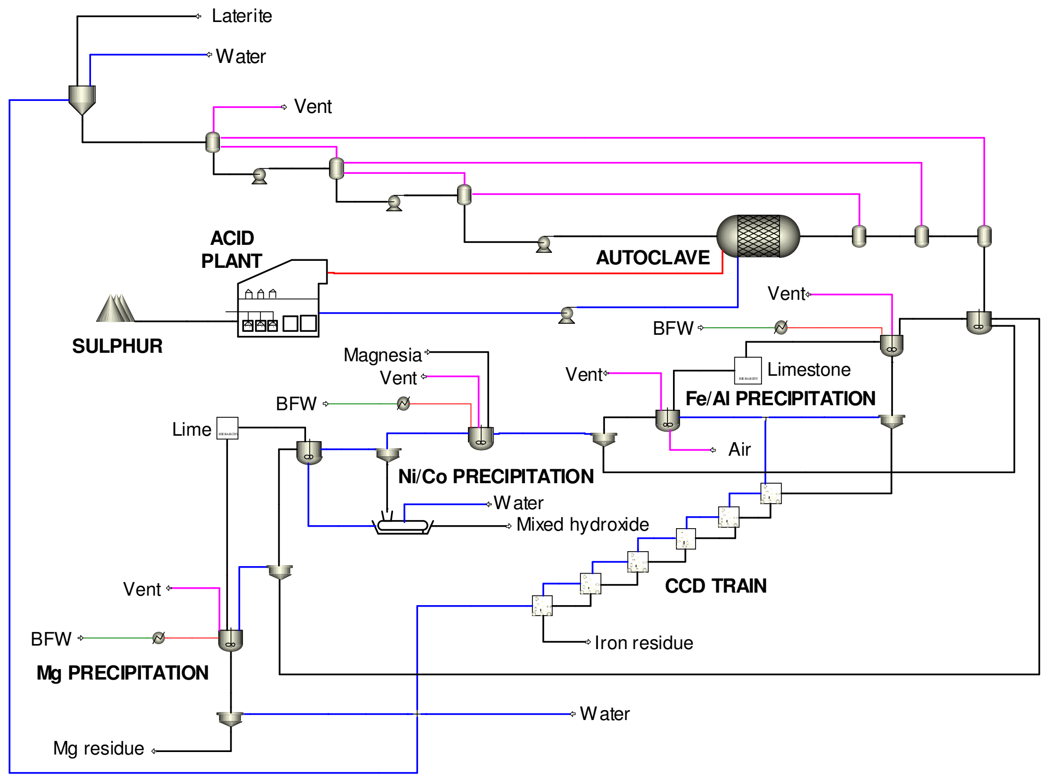
Table 2 lists the composition of the feed limonite used in this exercise. The rationale for the choice of this laterite is discussed in the previous papers of this series[1,2,3].

Table 2 - Hypothetical ore analysis, mass %

Ni	Co	Al_2O_3	Cr_2O_3	Fe_2O_3	CaO	MgO	MnO	Na_2O	K_2O	SiO_2
1.22	0.2	5.61	3.36	64.31	0.06	1.54	1.51	0.17	0.01	7.62

Figure 1 illustrates the process model used to represent the HPAL technology. It includes a standard sulphur-burning acid plant that produces the required amount of sulphuric acid and the steam needed for heating the autoclave.

Figure 1 – HPAL circuit



The incoming laterite is mixed with recycled process water and pumped through three pre-heating steps in which it is contacted with steam from three flash-down steps after the autoclave. The pre-heated slurry and concentrated sulphuric acid are pumped into the autoclave. Steam from the acid plant is injected into the autoclave to heat the leach. After leaching in the autoclave, the pressure of the slurry is reduced to atmospheric in three stages. The steam flashed off is used in the pre-heating sequence.

The leached slurry is first contacted with recycled hydroxide precipitates from downstream to re-dissolve co-precipitated nickel and cobalt and to partially consume the free acid left after the leach. The partly neutralised slurry is neutralized further with limestone to precipitate the bulk of the iron and aluminium while co-precipitating essentially no nickel or cobalt. The resulting iron-aluminium slurry is thickened and the thickener underflow is washed with recycled process water in a six-stage counter-current decantation train. The washed iron-aluminium residue leaves the circuit.

The Fe-Al thickener overflow is combined with the supernatant from the counter-current decantation train and neutralized further with more limestone, to precipitate essentially all of the remaining iron and aluminium. Some nickel and cobalt are co-precipitated in this step, and the underflow from the subsequent thickening step is recycled. The remaining solution is neutralized further with magnesium oxide to precipitate the bulk of the nickel and cobalt as hydroxides. The resulting slurry is thickened, the underflow is filtered and the filter cake is washed with water. The washed filter cake is the intermediate hydroxide product. The supernatant and filtrate are combined and neutralized again, this time using lime, to precipitate the remaining nickel and cobalt as hydroxide and gypsum that are thickened, the underflow recycled and the supernatant contacted with lime to precipitate the magnesium and manganese into a hydroxide-gypsum residue that leaves the circuit after thickening. The supernatant from the final thickening step is recycled as process water.

PROCESSING THE INTERMEDIATE PRODUCT

Three routes for taking the intermediate product to cathode nickel and cobalt were chosen for this exercise:

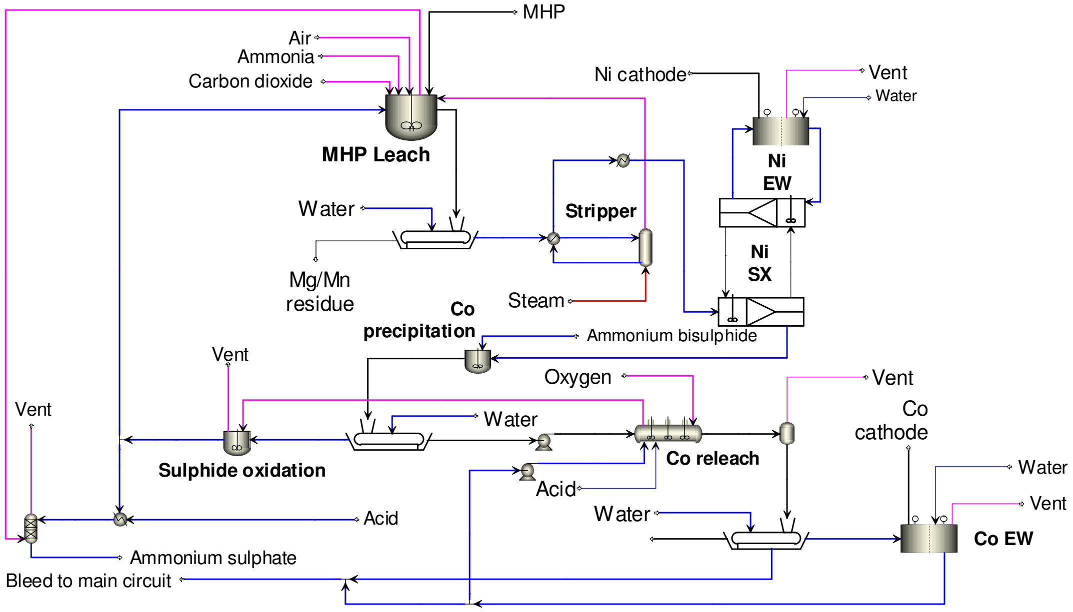
- Leaching the intermediate product in an ammonia-ammonium sulphate system, with solvent extraction and electrowinning nickel, sulphide precipitation of cobalt from the resulting raffinate, oxidative leaching of the cobalt sulphide and electrowinning cobalt from the resulting solution.

- Leaching the intermediate product in spent electrolyte from nickel electrowinning, solvent extraction and electrowinning cobalt from the resulting solution and electrowinning nickel from the raffinate.
- Eliminating production of the intermediate product and using synergistic solvent extraction to recover nickel selectively from the solution from which the intermediate product would otherwise have been produced, stripping the organic phase with spent electrolyte and electrowinning nickel from the loaded strip liquor. Then using the same synergistic solvent extraction system to selectively extract cobalt from the nickel raffinate, stripping and electrowinning cobalt in the same way as nickel.

Ammonia route

Figure 2 illustrates the ammonia-ammonium sulphate route, which is based on the circuits used by Queensland Nickel and Cawse Nickel[4,5]. The intermediate hydroxide product (MHP) is leached in recycled solution with air and ammonia/ammonium sulphate. The nickel and cobalt hydroxides dissolve and the cobalt is oxidized. The stoichiometry used to represent the leach chemistry is shown in Table 3. Any manganese present is oxidised and precipitated as manganese dioxide. The slurry (the solids being essentially magnesium oxide/hydroxide and a minor amount of manganese dioxide) is filtered and washed with water. The washed filter cake is discarded and the filtrate is stripped with steam to remove the excess ammonia for recycle to the leach.

Figure 2 – Ammonia-ammonium sulphate circuit



Nickel is removed from the steam-stripped solution by solvent extraction with an organic reagent such as LIX-84I. The loaded organic phase is stripped with spent electrolyte from the subsequent nickel electrowinning step and the nickel is recovered from the resulting advance electrolyte by electrowinning. The stoichiometry used to represent the relevant chemistry is shown in Table 4. The raffinate from the nickel solvent extraction section is contacted with ammonium bisulfide (or H_2S and NH_3) to precipitate the cobalt as cobalt sulphide that is filtered from the resulting slurry and washed with water. The filtrate returns to the MHP leach via an oxidation step to convert residual sulphide to sulphate. The filter cake is pressure leached with oxygen in spent electrolyte from the subsequent cobalt electrowinning step, converting the cobalt sulphide to dissolved cobalt sulphate and solid elemental sulphur. The resulting slurry is filtered, the filter cake is washed with water and the washed elemental sulphur is returned to the acid plant in the laterite circuit. The wash filtrate is returned to the laterite circuit to purge impurities. Cobalt is recovered from the primary filtrate by electrowinning. The spent electrolyte, minus a small bleed to the main laterite circuit to purge impurities and the balance supplemented with fresh sulphuric acid, is returned to the pressure oxidation step. The applicable chemistry is represented by the stoichiometry shown in Table 5.

Table 3 – Ammonia leach stoichiometry

$Ni(OH)_2 + 2NH_3 + 2NH_4^+ \rightarrow Ni(NH_3)_4^{2+} + 2H_2O$
$Ni_4(OH)_6SO_4 + 8NH_3 + 6NH_4^+ \rightarrow 4Ni(NH_3)_4^{2+} + SO_4^{2-} + H_2O$
$2Co(OH)_2 + 6NH_3 + 6NH_4^+ + ½O_2 \rightarrow 2Co(NH_3)_6^{3+} + 5H_2O$
$Co_4(OH)_6SO_4 + 14NH_3 + 10NH_4^+ + O_2 \rightarrow 4Co(NH_3)_6^{3+} + SO_4^{2-} + 8H_2O$

Table 4 – Nickel SX and EW chemistry

Loading $Ni(NH_3)_4^{2+} + 2RH \rightarrow R_2Ni + 2NH_3 + 2NH_4^+$
Stripping $R_2Ni + 2H_3O^+ \rightarrow Ni^{2+} + 2RH + 2H_2O$
Ni electrowinning $Ni^{2+} + 3H_2O \rightarrow Ni + ½O_2 + 2H_3O^+$

Table 5 – Cobalt precipitation, oxidation and EW chemistry

Sulphide precipitation $3NH_4HS \rightarrow 3NH_4^+ + 3HS^-$ $Co(NH_3)_6^{3+} + 2HS^- \rightarrow CoS + S + 2NH_4^+ + 4HN_3$
Residual sulphide oxidation $H_2S + 2O_2 \rightarrow H_2SO_4$
Oxidative dissolution $CoS + 2H_3O^+ + ½O_2 \rightarrow Co^{2+} + 3H_2O + S$
Electrowinning $2Co^{2+} + 6H_2O \rightarrow 2Co + O_2 + 4H_3O^+$

Sulphuric acid route

This option is based on work published by Iliev *et al*[6] on the solvent extraction of cobalt with Cyanex 272 from nickel sulphate solution. This process is used commercially by Níquel Tocantins at their São Paolo Refinery in Brazil. Figure 3 illustrates the circuit and Table 6 shows the stoichiometry used to represent the leach chemistry. Table 7 shows the stoichiometry used to represent the electrowinning of nickel and the solvent extraction and electrowinning of cobalt.

Figure 3 – Sulphuric acid leach circuit

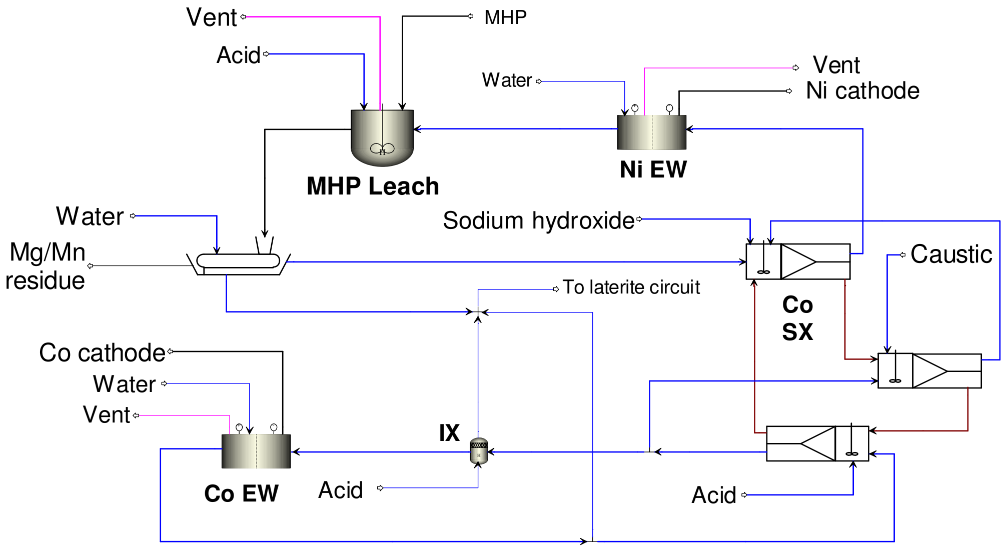


Table 6 – Sulphuric acid leach stoichiometry

$Ni(OH)_2 + 2H_3O^+ \rightarrow Ni^{2+} + 4H_2O$
$Ni_4(OH)_6SO_4 + 6H_3O^+ \rightarrow 4Ni^{2+} + SO_4^{2-} + 12H_2O$
$Co(OH)_2 + 2H_3O^+ \rightarrow Co^{2+} + 4H_2O$
$Co_4(OH)_6SO_4 + 6H_3O^+ \rightarrow 4Co^{2+} + SO_4^{2-} + 12H_2O$
$Mn(OH)_2 + 2H_3O^+ \rightarrow Mn^{2+} + 4H_2O$
$Mg(OH)_2 + 2H_3O^+ \rightarrow Mg^{2+} + 4H_2O$

Table 7 – Ni EW, Co SX, IX and EW stoichiometry

Ni EW
$Ni^{2+} + 3H_2O \rightarrow Ni + ½O_2 + 2H_3O^+$
Co SX
$(Co,Mn,Ni)^{2+} + 2LH + 2H_2O \rightarrow L_2(Co,Mn,Ni) + 2H_3O^+$
$L_2Ni + Co^{2+} \rightarrow L_2Co + Ni^{2+}$
$L_2(Co,Mn) + 2H_3O^+ \rightarrow (Co,Mn)^{2+} + 2LH + 2H_2O$
Co EW
$Co^{2+} + 3H_2O \rightarrow Co + ½O_2 + 2H_3O^+$

The MHP is dissolved using fresh sulphuric acid and spent electrolyte from the nickel electrowinning section. The residual solids are filtered from the resulting slurry, washed with water and discarded. The wash filtrate returns to the laterite circuit to purge magnesium and manganese and to recycle any contained nickel and cobalt to the hydroxide precipitation step. The primary filtrate goes to the cobalt solvent extraction section, which uses Cyanex 272 to extract the cobalt, the manganese and a part of the nickel. The loaded organic phase is scrubbed with some of the loaded strip liquor from the subsequent stripping section to remove the co-extracted nickel, the spent scrub solution returning to the feed to the extraction section. The scrubbed organic phase proceeds to the stripping section where it is stripped with spent electrolyte from the cobalt electrowinning section. The loaded strip liquor, less the part used as scrub solution, goes to the cobalt electrowinning section via an ion exchange stage to remove any residual nickel. The loaded resin is stripped with sulphuric acid, the spent eluate returning to the laterite circuit. Cobalt is recovered from the purified solution by electrowinning. The spent electrolyte, minus a small bleed to the laterite circuit to purge manganese, is replenished with fresh sulphuric acid and returned to the solvent extraction section.

Synergistic solvent extraction route

This option arises from the assumption that eliminating the precipitation and re-dissolution of nickel and cobalt hydroxide could reduce the cost of recovering metallic nickel and cobalt from solution produced by leaching laterite. Work has been published on synergistic systems that appear to be applicable. CSIRO has developed a synergistic solvent for just this application[7], but the information required to effectively model that particular system was not available in time for this paper.

Work was previously published by Du Preez and Kotze[8] on solvent extraction using versatic acid and a synergist called Nicksyn™ to extract nickel and cobalt from laterite leach liquor. Figure 4 shows data published by Du Preez and Kotze and the fit to their data achieved using an Aspen Plus® model reproducing that experimental work. This model fitted equilibrium constants for the reactions listed in Table 8, in which H_2A_2 is the versatic acid dimer, L is the synergist, HA.L is a versatic acid-synergist complex and NiL_2A_2, CoL_2A_2, etc. are the metal species extracted into the organic phase. The process by which this model was built and fitted to the data will be presented as a separate paper at an appropriate future occasion.

Once built and fitted to the experimental data, the model was used to investigate the application of this synergistic solvent extraction system to the solution going to MHP precipitation in the HPAL circuit assumed for this paper. Figure 5 illustrates the circuit developed, which would replace the MHP precipitation section in the main laterite circuit.

Figure 4 – SSX data and model fit

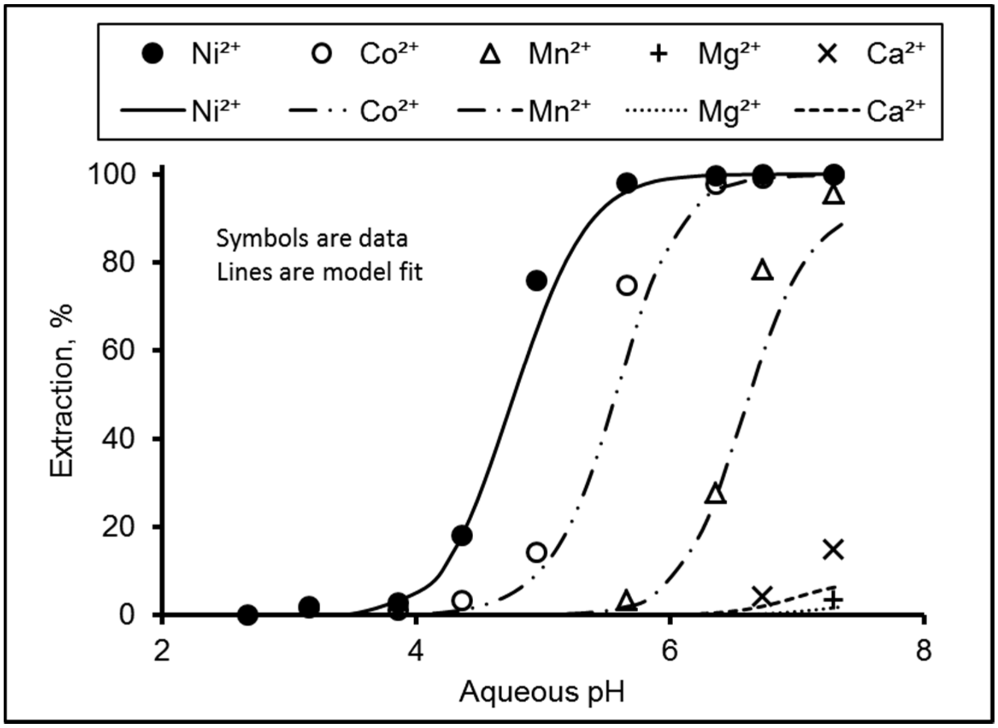


Table 8 – SSX reaction stoichiometry

$Ni^{2+} + H_2A_2 + 2L \leftrightarrow NiA_2L_2 + 2H^+$
$Co^{2+} + H_2A_2 + 2L \leftrightarrow CoA_2L_2 + 2H^+$
$Mn^{2+} + H_2A_2 + 2L \leftrightarrow MnA_2L_2 + 2H^+$
$Mg^{2+} + H_2A_2 + 2L \leftrightarrow MgA_2L_2 + 2H^+$
$Ca^{2+} + 3H_2A_2 \leftrightarrow CaA_6H_4 + 2H^+$
$H_2A_2 + 2L \leftrightarrow 2HA.L$

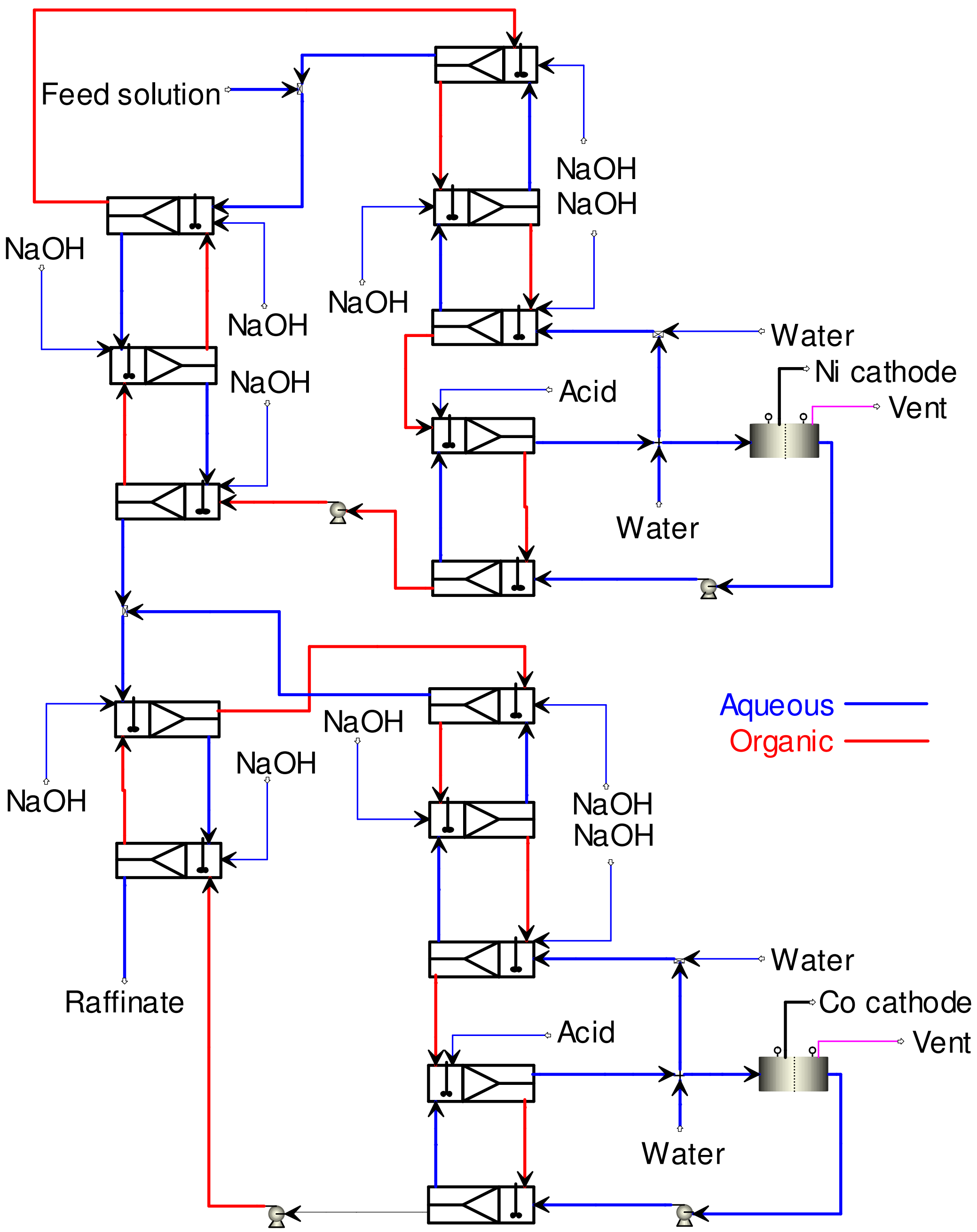
The feed to the previous MHP precipitation section combines with spent aqueous liquor from the scrub section and passes through three equilibrium stages of nickel extraction, extracting essentially all of the nickel, some of the cobalt and a little of the manganese. The raffinate from the third nickel extraction stage proceeds to the cobalt extraction section.

The loaded organic phase from the first nickel extraction stage passes through three equilibrium stages of scrubbing with advance nickel electrolyte from the subsequent stripping section, in which essentially all of the manganese and cobalt are removed from the organic phase and replaced with nickel. The spent scrub liquor leaves the first scrub stage and returns to the nickel extraction section.

The scrubbed organic from the third scrub stage is stripped in two equilibrium stages with spent electrolyte from the nickel electrowinning section, plus make-up sulphuric acid replenishing the acid removed in the portion of resulting advance electrolyte used in the scrubbing section. The resulting advance electrolyte, less the part used in the scrubbing section, goes to the nickel electrowinning section, where cathode nickel is produced. The spent electrolyte returns to the stripping section.

The raffinate from the nickel extraction section is combined with the spent strip solution from the cobalt scrubbing section and essentially all of the cobalt, some of the manganese and very minor amounts of magnesium and calcium are extracted into organic phase of the same composition as is used for the nickel, in two equilibrium stages. The resulting raffinate returns to the laterite circuit, where the remaining magnesium, manganese and calcium are rejected.

Figure 5 – SSX circuit



The loaded organic phase from the extraction section is scrubbed in three equilibrium stages with part of the advance electrolyte produced in the stripping section, removing essentially all the manganese, magnesium and calcium and replacing them with cobalt. The spent scrub solution recycles to the cobalt extraction section, and the scrubbed organic is stripped in two equilibrium stages with spent electrolyte from the cobalt electrowinning section and make-up sulphuric acid. The stripped organic phase returns from the second stripping stage to the cobalt extraction section. The advance electrolyte from the first stripping stage, less the part used in the scrubbing section, goes to the cobalt electrowinning section where cathode cobalt is produced. The spent electrolyte returns to the stripping section.

The raffinate from the cobalt extraction section returns to the main circuit. The sodium hydroxide and sulphuric acid added in the nickel and cobalt solvent extraction sections become sodium sulphate that ends up in the cobalt raffinate, and ultimately as sodium jarosite in the iron-aluminium residue leaving the main laterite circuit. Table 9 lists the numbers calculated for this route.

Table 9 – SSX parameters

Nickel section						
Stream	pH	Concentration, g/L				
		Ni	Co	Mn	Mg	Ca
Aqueous to extraction	5.2	6.06	1.19	2.51	1.98	0.30
Aqueous ex extraction	5.8	0.00	0.44	2.42	1.92	0.29
Organic ex extraction		13.52	1.64	0.02	0.000	0.000
Organic to extraction		0.001	0.000	0.000	0.000	0.000
Aqueous to scrubbing	4.1	56.25	0.03	0.00	0.00	0.00
Aqueous ex scrubbing	5.8	45.56	10.56	0.13	0.00	0.00
Organic to scrubbing		13.52	1.64	0.02	0.00	0.00
Organic ex scrubbing		15.17	0.01	0.000	0.000	0.000
Aqueous to stripping	1.3	51.16	0.02	0.00	0.00	0.00
Aqueous ex stripping	4.0	100.58	0.05	0.00	0.00	0.00
Organic to stripping		15.17	0.01	0.00	0.00	0.00
Organic ex stripping		0.00	0.00	0.00	0.00	0.00
Extraction efficiency (O:A=0.46)		99.98%	61.9%	0.4%	0.00%	0.01%
Scrub efficiency (O:A=6.45)		-	99.6%	100%	100%	100%
Strip efficiency (O:A=3.18)		100%	100%	-	-	-
Overall recovery to Ni cathode		99.95%	0.30%	99.95% Ni in cathodes		

Cobalt section						
Stream	pH	Concentration, g/L				
		Ni	Co	Mn	Mg	Ca
Aqueous to stage EX1	5.8	0.001	0.438	2.416	1.916	0.288
Aqueous ex stage EX2	6.0	0.000	0.002	2.348	1.862	0.279
Organic ex stage EX1		0.017	2.381	2.860	0.004	0.004
Organic to stage EX2		0.000	0.000	0.000	0.000	0.000
Aqueous ex stage SC3	4.1	0.199	11.661	44.097	0.066	0.065
Aqueous to stage SC1	4.1	0.198	60.001	0.000	0.000	0.000
Organic to stage SC3		0.017	2.381	2.860	0.004	0.004
Organic ex stage SC1		0.018	5.466	0.000	0.000	0.000
Aqueous to stage ST1	1.5	0.154	45.880	0.000	0.000	0.000
Aqueous ex stage ST2	4.0	0.286	86.811	0.000	0.000	0.000
Organic to stage ST2		0.018	5.466	0.000	0.000	0.000
Organic ex stage ST1		0.000	0.000	0.000	0.000	0.000
Extraction efficiency (O:A=0.27)		100%	99.7%	24.1%	0.06%	0.4%
Scrub efficiency (O:A=15.62)		-	-	100%	100%	100%
Strip efficiency (O:A=7.61)		100%	100%	-	-	-
Overall recovery to Co cathode		0.05%	99.3%	99.67% Co in cathodes		

CAPITAL AND OPERATING COSTS

Commercially available process simulation software known as Aspen Plus® (AP) was used to build process models (numerically rigorous mass/energy balances) for the three process examined. The reagent and utility consumptions predicted by the process models were combined with estimates of the various unit costs to calculate variable operating costs for each process.

Estimates of the capital and fixed operating costs were generated using commercially available cost estimation software known as Aspen Process Economic Analyser® (APEA). For each process, the mass-energy balance generated using the AP software was electronically exported into the APEA software, the process equipment, material of construction and residence time were specified for the various unit operations and the APEA software was used to estimate the capital cost of the installed process equipment and the fixed operating cost, for each process. In Part 3 of this series, the results obtained for a number of different processes were checked against published information, as a validation exercise[3]. The methodology[3] is described in more detail in Part 3.

Table 10 lists the capital costs calculated for the for the installed process equipment required to produce cathode metal from the hydroxide intermediate. Table 11 lists the calculated additional fixed costs.

Table 10 – Capital costs for add-on processing, $ million

Account	SSX	Ammonia	Acid
Equipment	50.1	35.7	22.1
Piping	63.9	55.1	40.5
Civil	5.4	3.4	2.0
Steel	32.6	32.5	31.9
Instruments	29.2	32.9	24.2
Electrical	20.5	19.2	17.8
Insulation	1.9	1.9	0.3
Paint	2.0	1.9	1.5
Total Direct Field Costs	205.7	182.8	140.5
Indirect Field Costs	23.5	23.1	16.6
Total Field Costs	229.2	206.0	157.1
Freight	26.0	22.8	17.7
Taxes and Permits	13.0	11.4	8.8
Engineering and HO	9.4	9.1	5.0
Other Project Costs	16.9	15.5	11.7
Contingency	53.0	47.7	36.1
Total Non-Field Costs	118.3	106.5	79.4
Project Total Costs	347.5	312.5	236.4

Table 11 – Additional fixed costs, $ million/year

	SSX	Ammonia	Acid
Manpower	0.8	1.3	0.8
Maintenance	2.7	2.0	1.1

Table 12 gives a breakdown of the additional equipment costs by process section. The estimated equipment costs for the nickel and cobalt electrolysis sections are very similar for all three options because the quantities of cathode nickel and cobalt are determined by the amounts of nickel and cobalt in the feed, which is the same in all three cases.

Table 12 – Equipment cost breakdown, $ million

Section	SSX	Ammonia	Acid
MHP leach	-	6.0	4.7
Nickel solvent extraction	41.2	35.8	-
Nickel electrowinning	101.4	101.0	101.2
Cobalt sulphide precipitation & leach	-	7.6	-
Cobalt solvent extraction	24.3	-	5.7
Cobalt electrowinning	22.5	22.9	22.4
Sub-total	189.4	173.3	133.9

The economic benefit associated with onward processing arises from the extra revenue accruing from the sale of nickel and cobalt for the full market price of these metals instead of the discounted price realized by selling the intermediate hydroxide. Processing the intermediate product entails additional operating costs, thus the economic evaluation has to consider the incremental revenue less the incremental operating costs against the incremental capital cost associated with each option. Table 13 lists the reagent and utility costs calculated for the HPAL process producing the mixed hydroxide intermediate and for the HPAL circuit plus each of the three additional processing options. The reagent costs are long term averages calculated from data from the open literature[3]. In Table 13, the reagent costs for each route are for the production of MHP from laterite plus, in the three cases of further processing, the costs for converting the hydroxide to cathode metals.

Table 13 – Variable costs, laterite to product

Reagent	Cost	Reagent & utility costs, $/lb Ni+Co			
		MHP	SSX	Ammonia	Acid
Laterite	$5/t	0.170	0.170	0.170	0.170
Sulphur	$187/t	0.726	0.927	0.738	0.767
$CaCO_3$	$60/t	0.375	0.452	0.446	0.442
MgO	$616/t	0.170	-	0.171	0.270
CaO	$119/t	0.202	0.134	0.182	0.214
NaOH	$148/t	-	0.250	-	0.037
NH_3	$650/t	-	-	0.067	-
H_2S	$800/t	-	-	0.046	-
O_2	$3000/t	-	-	0.067	-
Steam	$10/t	-	-	0.001	-
Water	$1/t	0.002	0.016	0.059	0.059
Power	$11/GJ	0.073	0.086	0.090	0.091
Variable cost, $/lb		1.788	2.035	2.201	2.050

The incremental benefit depends on the metal price and the percentage of the metal value realized from the intermediate hydroxide. In the exercise presented in Part 3 of this series the intermediate hydroxide was assumed to fetch 85 percent of the market value of the contained nickel and cobalt. The long term average prices, inflation-adjusted to US dollars (year 2013), for nickel and cobalt are $7.21/lb and $26.07/lb, respectively, with standard deviations of $2.76/lb for nickel and $13.15/lb for cobalt[3].

The long term average metal prices, assuming 85 percent of the metal values for the intermediate hydroxide, the costs listed in Table 13 for the various reagents and utilities and the additional capital and fixed costs listed in Table 10 and Table 11 lead to the cash flow calculations shown in Table 15, Table 16 and Table 17. The baseline cash flow calculation for production of the MHP from laterite is shown in Table 14.

Table 14 – Cash flow analysis, HPAL circuit, selling MHP for 85% of metal prices

Cash flow $ million	Year										
	1	2	3	4	5	6	7	8	9	10	11→
Capital cost	572	572									
Fixed costs			6	6	6	6	6	6	6	6	6
Variable costs			32	65	130	130	130	130	130	130	130
Revenue			151	303	605	605	605	605	605	605	605
Gross margin	-572	-572	113	232	470	470	470	470	470	470	470
Tax	0	0	0	0	0	42	141	141	141	141	141
Net margin	-572	-572	113	232	470	428	329	329	329	329	329
IRR , 10 year	18%										
IRR, 20 year	23%										

Table 15 – Differential cash flow analysis, ammonia leach option, MHP to metal

Cash flow $ million	Year										
	1	2	3	4	5	6	7	8	9	10	11→
Capital cost	156	156									
Fixed costs			3	3	3	3	3	3	3	3	3
Variable costs			4	8	16	16	16	16	16	16	16
Revenue			27	53	107	107	107	107	107	107	107
Gross margin	-156	-156	20	42	88	88	88	88	88	88	88
Tax	0	0	0	0	0	0	4	26	26	26	26
Net margin	-156	-156	20	42	88	88	84	62	62	62	62
IRR , 10 year	10%										
IRR, 20 year	17%										

Table 16 – Differential cash flow analysis, acid leach option, MHP to metal

Cash flow $ million	Year										
	1	2	3	4	5	6	7	8	9	10	11→
Capital cost	118	118									
Fixed costs			2	2	2	2	2	2	2	2	2
Variable costs			5	9	19	19	19	19	19	19	19
Revenue			27	53	107	107	107	107	107	107	107
Gross margin	-118	-118	20	42	86	86	86	86	86	86	86
Tax	0	0	0	0	0	0	25	26	26	26	26
Net margin	-118	-118	20	42	86	86	61	60	60	60	60
IRR , 10 year	15%										
IRR, 20 year	21%										

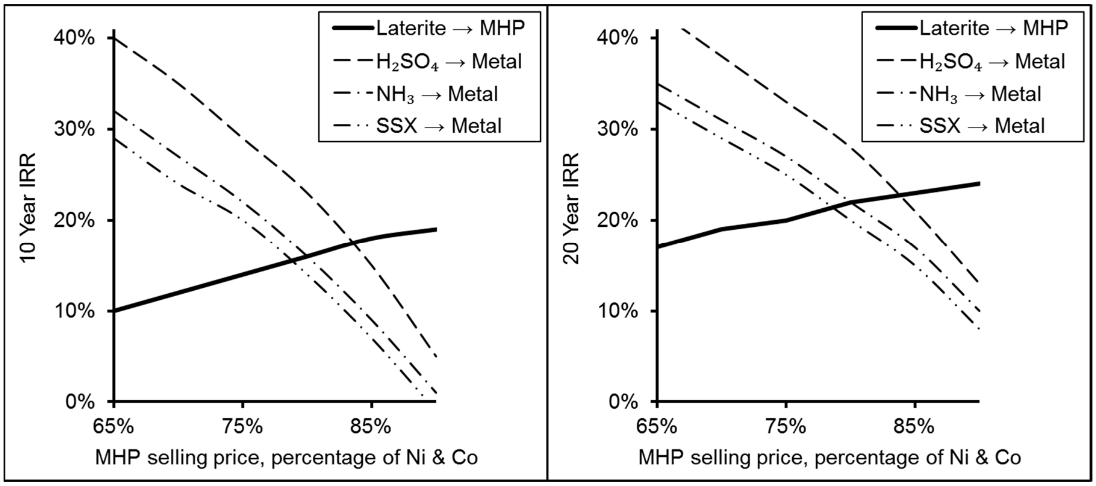
Simplistically, at least, in terms of the calculated internal rates of return for the differential benefit versus the differential cost and the capital required for the extra processing, it would appear that processing the intermediate hydroxide to cathode metal, while not a complete loss proposition, may not be as financially rewarding as producing and selling the intermediate hydroxide. Of the three options for the additional processing, the best would appear to be re-leaching the intermediate hydroxide in spent nickel electrolyte, extracting the cobalt from the resulting solution by solvent extraction and electrowinning nickel and cobalt. The option replacing the precipitation of hydroxide with synergistic solvent extraction and electrowinning appears to be the worst option, economically, even though it seems to have the lowest variable cost, because it appears to have the highest capital cost.

Table 17 – Differential cash flow analysis, SSX option, MHP to metal

Cash flow $ million	Year										
	1	2	3	4	5	6	7	8	9	10	11→
Capital cost	174	174									
Fixed costs			3	3	3	3	3	3	3	3	3
Variable costs			4	9	18	18	18	18	18	18	18
Revenue			27	53	107	107	107	107	107	107	107
Gross margin	-174	-174	19	41	86	86	86	86	86	86	86
Tax	0	0	0	0	0	0	0	16	26	2	29
Net margin	-174	-174	19	41	86	86	86	69	60	60	60
IRR, 10 year	7%										
IRR, 20 year	15%										

Selling the intermediate hydroxide for 85 percent of the value of the contained nickel and cobalt is an assumption. Changing this assumption would change the differential economics associated with the extra processing to convert the intermediate hydroxide to cathode metals. Figure 6 shows the impact of this assumption on the 10-year and 20-year IRR (internal rate of return) numbers.

Figure 6 – Effect of MHP price



As the selling price of the intermediate hydroxide rises, the economics of making and selling it improve, while the incremental benefit associated with processing it to cathode metals decreases. When the IRR for processing the intermediate to cathode metals is higher than the IRR for simply making and selling the intermediate hydroxide, it could be argued that going to the metals would enhance the economics of the overall operation. When the IRR for going from the intermediate hydroxide to metals is at or below the IRR for making and selling the intermediate hydroxide, going from the hydroxide to metal products might not be the best use of the extra capital.

A further assumption used so far in this exercise is that the market value of nickel and cobalt is the long term average, as presented in Part 3 of this series. Figure 7 shows the historical prices of nickel and cobalt, inflation-adjusted to 2013 US dollars. The dashed lines are the simple long-term average price, plus or minus one standard deviation. Similar graphs were presented in Part 3 for the prices of the major reagents used in the HPAL circuit.

One way of accommodating price and cost uncertainty in evaluating a potential project is to require the economics to remain acceptable for all combinations of costs and revenue between selected upper and lower bounds. However, examination of the historical price data shows that as nickel and cobalt prices move up and down, the prices of the major reagents used to extract the nickel and cobalt also move up and down in a manner not totally different from the movement of the prices of nickel and cobalt.

Figure 8 is an example of this, showing the inflation-adjusted price of sulphur (in 2013 US dollars) on the left and a dual plot of the nickel price and the sulphur price on the right. Clearly, although the correlation is far from perfect, these two prices follow similar trends. There are similar broad correlations between the nickel price and the prices of the other major reagents used in HPAL processing of laterite.

Figure 7 – Historical nickel and cobalt prices

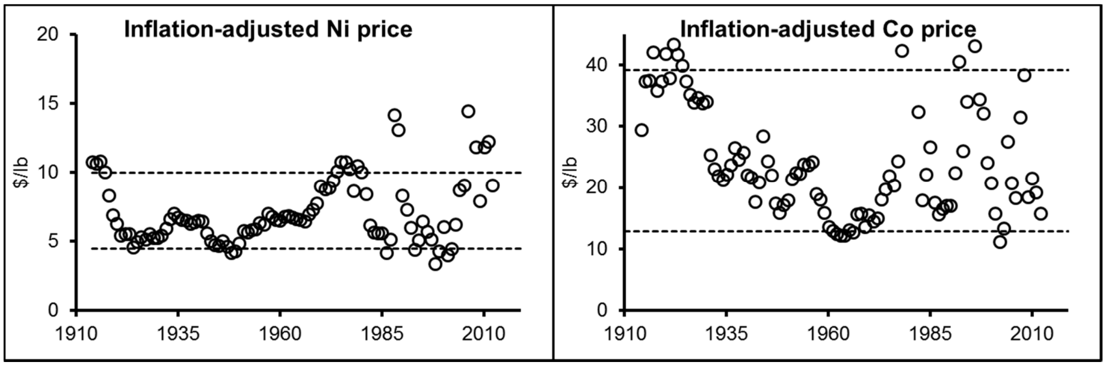
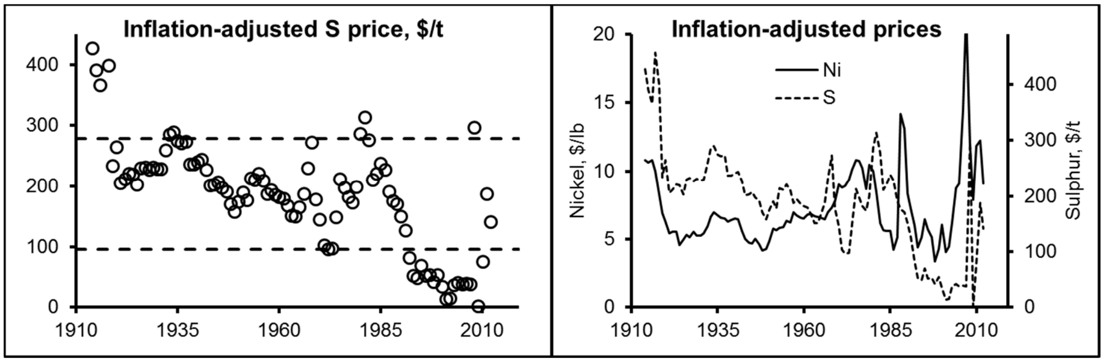
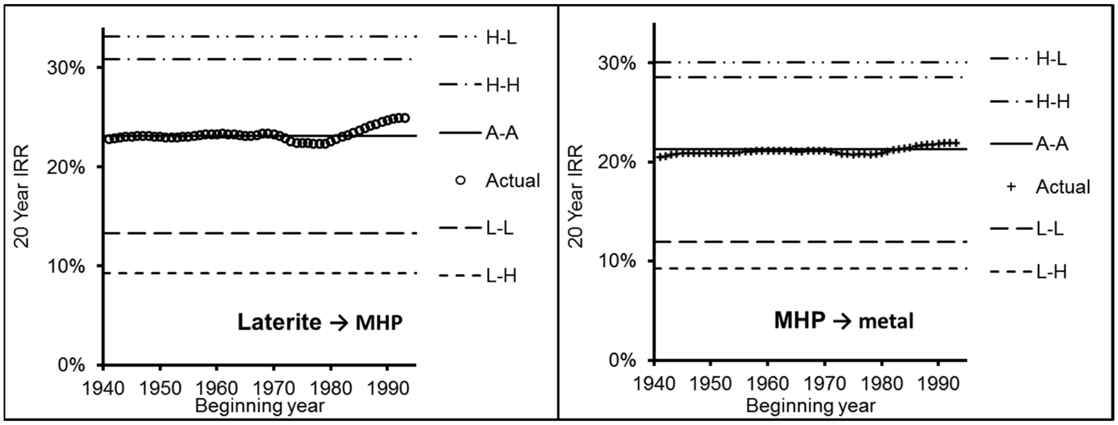


Figure 8 – Sulphur and nickel prices



This implies that the impact of rising reagent prices on the overall economics of the project should, to some extent at least, be countered by the rising price of the metals or intermediate product sold. Similarly, the negative impact of falling metal prices should be countered, at least somewhat, by falling input costs. This hypothesis was tested via a set of cash flow calculations in which the yearly historical prices of nickel, cobalt and the various major reagents were used to calculate the revenue and variable cost numbers over a set of 20 year periods beginning between 1941 and 1992, i.e. 1941 to 1960, 1942 to 1961, etc. Figure 9 shows the results of this exercise.

Figure 9 – IRR calculations using historical price data



The left hand graph in Figure 9 plots the 20 year IRR values calculated for the scenario in which the intermediate hydroxide is sold, and the right hand graph is for processing the intermediate hydroxide to metals by the acid route. The selling price of the hydroxide was assumed to be 85 percent of the contained metal prices. The horizontal lines are the IRR numbers calculated assuming upper-bound metal prices and lower-bound reagent costs (H-L), metal prices and reagent costs at their respective upper bounds (H-H), the long term average metal prices and reagent costs (A-A), metal prices and reagent costs at their respective lower bounds (L-L) and metal prices at their lower bounds with reagent prices at their upper bounds (L-H).

Over any twenty year period beginning between 1941 and 1992, the economic performance of the HPAL circuit assumed for this series of papers would have been substantially better than predicted by the most two conservative scenarios, low metal prices and high or low reagent prices. Similarly, the two most optimistic scenarios, high metal prices and low or high reagent prices, appreciably over-estimated the economics of the project. From Figure 9, it would appear that using the long term average prices for metals and reagents is a plausible approach.

The exercise presented in this paper is necessarily generalised. For a real situation there would be other aspects to consider as well, such as the ability of the secondary refinery to handle the various impurities in the intermediate hydroxide.

SUMMARY AND CONCLUSIONS

The exercise presented in this paper was done to examine the merits of processing a hypothetical laterite ore to cathode nickel and cobalt instead of to an intermediate hydroxide product that is sold for onward processing by others. Three processing options were examined for producing nickel and cobalt metal instead of the intermediate hydroxide:

1. Leaching the intermediate product in sulphate electrolyte from nickel electrowinning, solvent extraction of cobalt, electrowinning cobalt from the loaded strip solution and electrowinning nickel from the raffinate.
2. Leaching the intermediate product in an ammonia-ammonium sulphate system, with solvent extraction and electrowinning of the nickel, sulphide precipitation of cobalt from the resulting raffinate, oxidative leaching of the cobalt sulphide and electrowinning cobalt from the resulting solution.
3. Eliminating production of the intermediate hydroxide and using synergistic solvent extraction to recover nickel directly from the solution from which the intermediate product would otherwise be produced, stripping the organic phase with spent electrolyte and electrowinning nickel from the loaded strip liquor. Then using the same synergistic solvent extraction system to extract cobalt from the nickel raffinate, with stripping and electrowinning of cobalt in the same way as nickel.

For the conditions assumed for this exercise, if the intermediate hydroxide product can be sold for about 80 percent or more of the contained value of the nickel and cobalt, the overall economics of the operation would probably not be significantly improved by making nickel and cobalt metal. If the intermediate hydroxide can only be sold for less that about 80 percent of its metal content, further processing to metal products may well be worth consideration.

Of the three options examined, route 1 (leaching the hydroxide in spent nickel anolyte, etc.) appears to be the strongest, economically. The weakest option, economically, appears to be replacing the hydroxide precipitation step with synergistic solvent extraction. It should be noted, however, that a synergistic solvent extraction route that extracts nickel and cobalt directly from the leach solution, before removal of the iron and aluminium, has not yet been evaluated.

Using the long term averages for metal and reagent prices would appear to be a realistic way of evaluating the economics of processes such as the processing of laterite.

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