

Technical & Cost Comparison of Laterite Treatment Processes

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Introduction

This booklet is a compilation of a series of four papers presented at ALTA conferences during 2014, 2015 and 2016. The first paper was authored by Alan Taylor, ALTA Metallurgical Services, Australia, and the final three by Mike Dry, Arithmetek Inc., Canada.

The aim of the series is to provide a technical and economic evaluation and comparison of commercially applied hydrometallurgical and pyrometallurgical processes and a range of developing technologies for treating nickel-cobalt laterite. A hypothetical laterite orebody containing potentially economic zones of limonite and saprolite is used as a basis.

For any specific project, of course, the exercise contained in these four papers would need to be repeated using the composition of the laterite in question.

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**TECHNICAL & COST COMPARISON OF LATERITE TREATMENT PROCESSES
PART 1**

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ABSTRACT

This paper presents a technical overview of the commercially applied processes for nickel-cobalt laterites together with a range of processes under development. For each process, an outline flowsheet and brief process description are presented and the applicability to a hypothetical laterite orebody is assessed and compared from the technical viewpoint. An economic comparison is presented in Part 2. Processes covered include PAL, EPAL, Caron, RKEF (Ferronickel), Sintering/Blast Furnace (NPI), Sintering/SAF (NPI), HL, AL, Neomet and Direct Nickel.

INTRODUCTION

Future nickel supplies will have to be produced largely from lateritic deposits, including lower grade limonite and low grade saprolite not suitable for RKEF ferronickel smelting, the workhorse of the industry. This has resulted in the construction of large scale plants based on the pressure acid leaching (PAL) process first applied at Moa Bay Cuba in the late 1950s. However, these plants have a high capital cost and a number have suffered from commissioning problems which has led would be producers and technology developers to focus on potentially less expensive processes operating at atmospheric pressure, including heap leaching (HL) and agitated tank leaching (AL) with sulphuric acid, hydrochloric acid and nitric acid. In addition, the rising cost of nickel and rapidly increasing demand, especially in China, has led to the rise of numerous nickel pig iron (NPI) smelting operations, initially of small capacity, but becoming progressively larger in size.

All of these processes, as well as likely future trends, should be assessed in developing the optimum strategy for exploiting a laterite resource. An additional complication is that laterite deposits typically contain ore zones with significantly different mineralogy, so that a particular process may be applicable to only a portion of the resource. This paper presents a technical review of the proven and various developing processes and their potential application to a hypothetical laterite orebody producing 30,000 t/a nickel. Comparative economics are presented in the separate Part 2 companion paper authored by Dr Mike Dry owner of Arithmetek Inc. in Canada. Other than cobalt, the comparison excludes possible by-products unless they are an integral part the process.

SELECTED PROCESSES

The processes included in this review are shown in Table 1. The commercial category includes the processes which are currently applied to laterites on a stand-alone basis. Heap leaching and atmospheric tank leaching with sulphuric acid are categorized as being commercially applied as satellite operations to PAL. The selected developing processes are Neomet, one of a number of chloride processes, and the Direct Nickel nitric acid leaching process. Some of the processes have a number of possible downstream processing and product options. For the comparative purposes of this paper, the processes are limited to the production of intermediates and onsite refining is excluded. Hydroxide or oxide products have been selected for the hydrometallurgical processes, and ferronickel or nickel pig iron for the smelting processes (production of matte via FeNi smelting is excluded). Both fresh water and sea water are assumed to be available close to the site. The process and product abbreviations in Table 1 are used throughout this paper.

Processes	Product
Commercially Applied	
Pressure Acid Leaching (PAL)	Mixed Nickel-Cobalt Hydroxide (MHP)
Enhanced Pressure Acid Leaching (EPAL)	Mixed Nickel-Cobalt Hydroxide (MHP)
Reduction Roast-Ammonia Leach (Caron)	Nickel Oxide Sinter (NiO) plus Mixed Nickel-Cobalt Sulphide (MSP)
Rotary Kiln Electric Furnace Smelting (RKEF)	Ferronickel (FeNi)
Sintering/Blast Furnace Smelting (BF)	Nickel Pig Iron (NPI)
Sintering/Submerged Arc Smelting (SAF)*	Nickel Pig Iron (NPI)
Commercially Applied as Satellite Operations	
Heap Leaching (HL) with Sulphuric Acid	Mixed Nickel-Cobalt Hydroxide (MHP)
Atmospheric Tank Leaching (AL) with Sulphuric Acid	Mixed Nickel-Cobalt Hydroxide (MHP)
Developed to Pilot Plant Stage	
Neomet Chloride leach (Neomet)	Nickel Oxide (NiO) plus Cobalt Oxide (CoO)
Direct Nickel Nitric Acid Leach (DNi)	Mixed Nickel-Cobalt Hydroxide (MHP)

Table1: Selected Processes & Products

*Note: There is a recent trend towards using the RKEF process instead of Sintering/SAF for NPI production, which is not considered in this paper.

PROCESSES COMMERCIALY APPLIED

Pressure Acid Leaching

PAL is the most common hydrometallurgical process for low grade laterite ores. It is applicable to limonite and mixed limonite/saprolite ores with a magnesium content of up to about 5% Mg. Above this level there is a risk of increasingly heavy scaling in the PAL autoclaves and excessive acid consumption. Within this constraint, the maximum economic magnesium content will be determined by the cost for producing or importing sulphuric acid and the nickel price. The ore grade in existing commercial operations is in the range of around 1.1-1.5%. Nickel extraction is high, typically 90-95%. Cobalt recovery is also high and is a potentially valuable by-product, though the price tends to be volatile and may well be depressed by numerous copper-cobalt producers coming on stream in Central Africa. Acid consumption is typically in the range of 300-400 kg/t. To minimize initial capex, downstream processing can be limited to production of an intermediate product for sale, ie mixed sulphide (MSP) or mixed hydroxide (MHP). On-site refining to high purity nickel and cobalt can be added later if appropriate. Direct solvent extraction (DSX) is a potentially lower cost option for on-site refining from the onset of the project. For the purposes of this paper, the production of MHP is selected.

The basic flowsheet is shown in Fig. 1. It is assumed that the ore is not amenable to upgrading, with only a minor amount of coarse barren material rejected in ore preparation which consists of pulping with fresh water, screening, grinding and thickening. PAL with sulphuric acid is carried out at 245-250°C, 4,000 kpa and 40% solids in steam heated 6-compartment, mechanically agitated, horizontal autoclaves for 60-90 minutes to extract the nickel and cobalt. A reductant, typically sulphur, is added as required to promote leaching of manganese minerals and release cobalt and to suppress the formation of chromium (VI) which would cause problems in downstream processing. Three stages of direct contact preheating and flash heat recovery are typically included. The residue reports to a recycle leach step, then limestone is added to neutralize the residual acid and raise the pH to 2-3 to precipitate iron, aluminium and chromium for rejection with the subsequent CCD tailings. The CCD circuit comprises 6 stages and produces washed tailings for disposal. Additional neutralization of the tailings may be included if required. The pH of the CCD O/F is raised to pH 5.0-5.5 with limestone with aeration for final purification and the precipitate is sent to the recycle leach stage to recover co-precipitated nickel and cobalt. Magnesia is added to the clarified solution to precipitate mixed nickel-cobalt hydroxide at pH 6-7 and 50°C, which is washed then dewatered using a high pressure filter to form the final product. Lime is used to precipitate residual nickel and cobalt at pH 8.0-8.5 from the residual solution which is sent to recycle leach, then additional lime is used to raise the pH to 8.5-9.0 to remove manganese before recycling the barren solution for CCD washing. The manganese precipitate reports to tailings via the final CCD thickener. The manganese precipitate reports to tailings via the final CCD thickener.

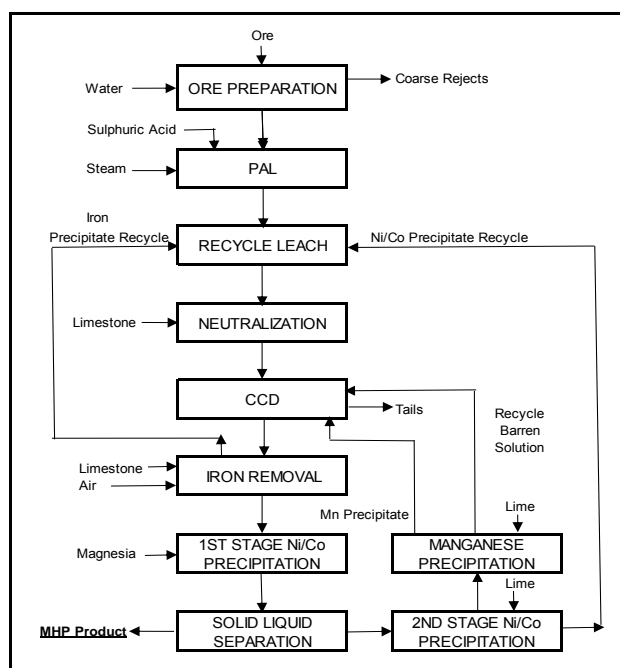


Figure 1: PAL Flowsheet for Limonite

Enhanced Pressure Acid Leaching

EPAL is currently limited to the Ravensthorpe operation in Western Australia which was designed and constructed by BHP Billiton and is now owned and operated by First Quantum. It produces MHP which was originally intended to be treated at the Yabulu Caron operation but is now exported to overseas customers. EPAL is a hybrid process aimed at processing both limonite and saprolite ore zones. PAL is used to treat limonite ore, while saprolite is pre-leached in an AL facility then mixed with PAL discharge slurry in a second AL step to utilize the residual acid to complete the leaching of the saprolite. The pre-leaching step enables more saprolite to be processed than that would be possible using the PAL residual acid alone. Nickel and cobalt extractions are in the range of 90-95% and acid consumption 400-600 kg/t ore. For the purposes of this paper, the production of MHP is selected as at Ravensthorpe.

The flowsheet in Fig. 2 does not include ore upgrading, though upgrading is practiced at Ravensthorpe. As at Ravensthorpe sea water is used for process purposes. The limonite ore preparation and PAL flowsheet is similar to the previous PAL only option. For saprolite, ore preparation includes crushing, screening, grinding and thickening followed by pre-mixing with sulphuric acid to initiate leaching and decompose carbonates. The partially leached saprolite is then mixed with PAL discharge in acid-brick lined mechanically agitated atmospheric pressure leaching vessels. AL discharge is treated with limestone to reduce the acid content to about 5 g/L and induce iron precipitation as jarosite with minimum losses of nickel and cobalt and improved settling in the downstream CCD circuit. Limestone is then added in primary neutralization to achieve pH 2.5 to precipitate the remaining ferric iron, followed by 6 Stages of CCD. The remainder of the flowsheet is similar to the previous PAL only case.

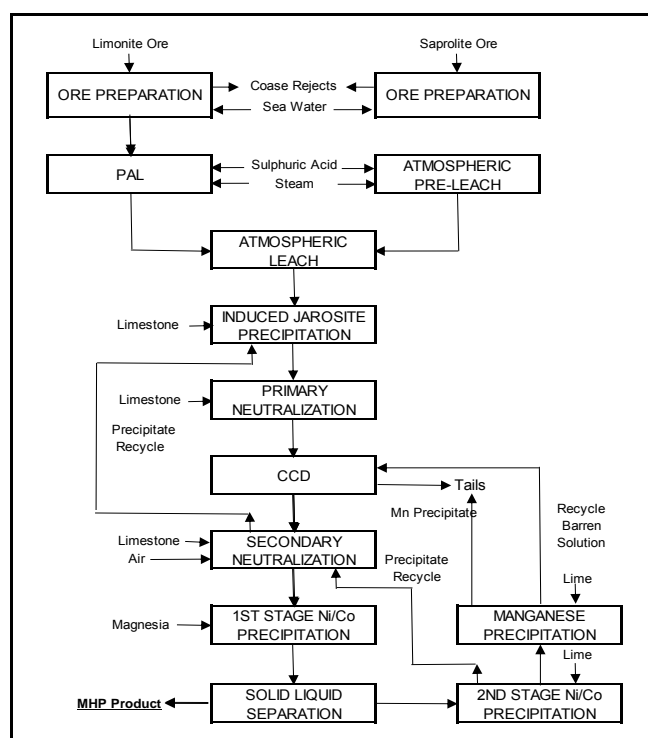


Figure 2: EPAL Flowsheet for Limonite & Saprolite

Caron Process

There are currently 4 Caron facilities (Nicaro in Cuba was reported closed in late 2012, though not officially confirmed). Compared with PAL, nickel (80-85%) and cobalt (35-45%) recoveries are significantly lower and energy consumption is higher due to the initial drying step. The Caron process is applicable to limonite and mixed limonite/saprolite with a minimum blended iron level of about 35% Fe. (Nickel recovery falls away as the proportion of low iron saprolite is increased). Various downstream processing variations and product options are available. For the purposes of this paper, cobalt is precipitated from the leach solution and recovered as a mixed Ni-Co sulphide (MSP) and nickel is produced as a partially reduced nickel oxide sinter containing 85% Ni.

In Figure 3, ore preparation consists of crushing and screening with rejection of coarse low grade material. This is followed by drying in a coal fire rotary kiln, dry grinding, then reduction roasting in multi-hearth furnaces at 750°C using heavy fuel oil as reductant which forms hydrogen and carbon monoxide. The reduced ore is cooled then leached with ammoniacal ammonium carbonate solution in aerated mechanically agitated tanks. The resulting leach pulp is passed through a 7 stage CCD washing circuit and the tailings are steam stripped to recover ammonia prior to disposal. Cobalt is precipitated as MSP from the CCD O/F solution with ammonium hydrosulfide which is thickened and dried for shipping to a refinery. The solution is then steam stripped to remove ammonia and carbon dioxide and precipitate nickel carbonate. The ammonia and carbon dioxide are recovered and re-used in leach. The nickel carbonate is calcined to form oxide in a rotary kiln then sintered and partially reduced with syn gas to form the NiO product.

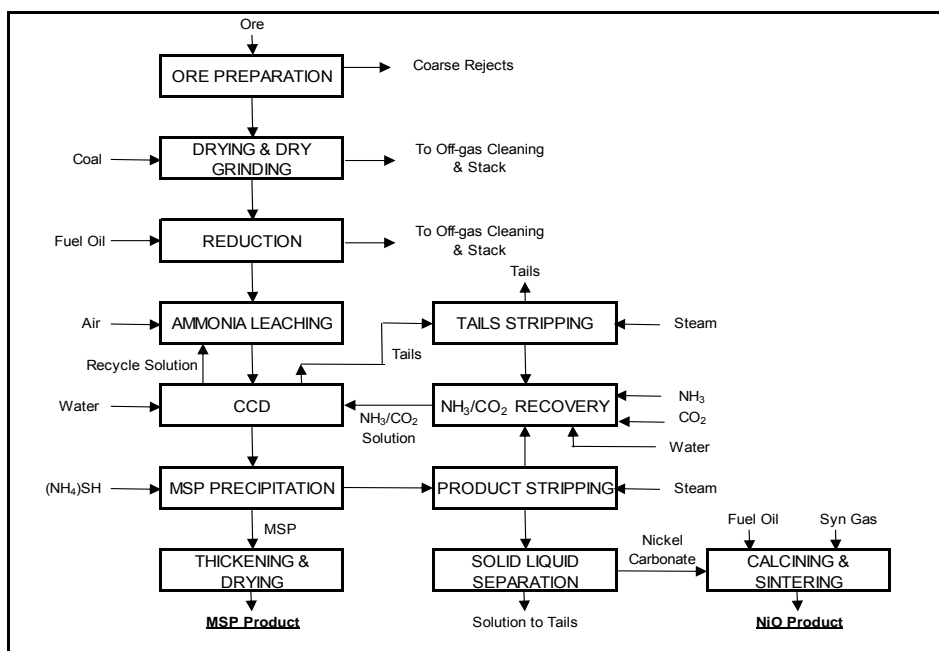


Figure 3: Caron Flowsheet for Limonite

Rotary Kiln Electric Furnace Smelting

The rotary kiln electric furnace (RKEF) smelting process is the industry standard process for producing ferronickel from saprolite ores containing >1.5% nickel. The basic process produces an iron-nickel product with a minimum of 15% nickel used for making stainless steel. Nickel extraction is high, typically 90-95%. There is no cobalt credit for ferronickel. Several operations add a sulphur source to form a Ni/Fe sulphide matte in the electric furnace, which is sent to a refinery for production of nickel and cobalt products. For this paper, a ferronickel process yielding a 25% nickel product is selected.

In Figure 4 the ore is prepared by coarse crushing and screening for the rejection of coarse barren material, followed by rotary drying at about 250°C to achieve a residual moisture of about 15-20% which is needed to avoid excessive dusting. This is followed by further crushing and screening to minus 10 mm with the rejection of additional coarse barren material. Calcination and pre-reduction is then carried out in a rotary kiln at about 800°C, which optimizes the use of energy from the reductant (coal) and fuel (oil or natural gas) thus reducing the energy input to the electric furnace. Smelting at 1550-1650°C completes the reduction of nickel and rejects gangue minerals and impurities as slag which is dumped. A final refining step using oxygen and additives is included to eliminate undesirable impurities such as sulphur, silica, carbon, and phosphorous. Methods used include ladle furnaces, electric furnaces, and oxygen blown converters. Sulphur can originate from ore as well as from the fuels used.

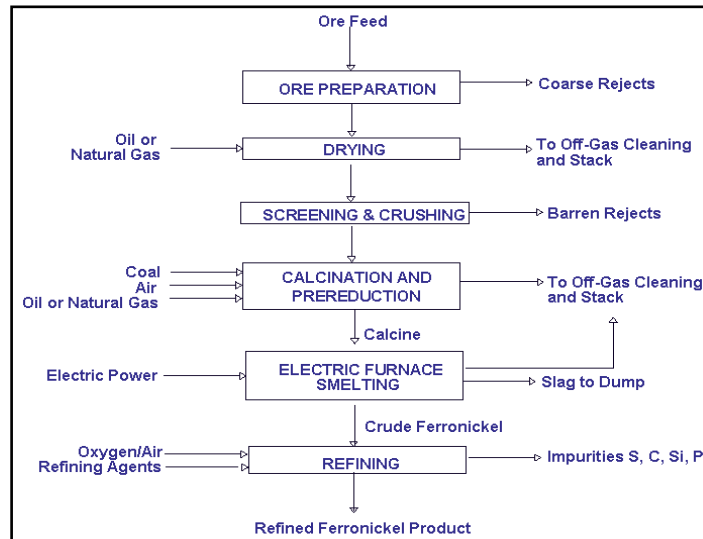


Figure 4: RKEF Smelting Flowsheet for Saprolite

Sintering/Blast Furnace Smelting

The sintering/blast furnace (BF) smelting process is commonly used to produce NPI in China, especially by smaller producers. It is tending to be phased out due to environmental issues, lower energy efficiency, lower recovery (80%+) and lower product grade (2-8% Ni) than other processes, though new plants are being established in Indonesia. It is particularly suited to limonite ores. For this paper, an 8% nickel NPI product is assumed. There is no cobalt credit for NPI.

In Figure 5, ore preparation consists of crushing and screening with coarse rejects. The ore is dried, mixed with coke and fluxes, then sintered to agglomerate prior to smelting in a blast furnace at 1400-1500°C to reduce the metal oxides. The NPI product is cast in molds.

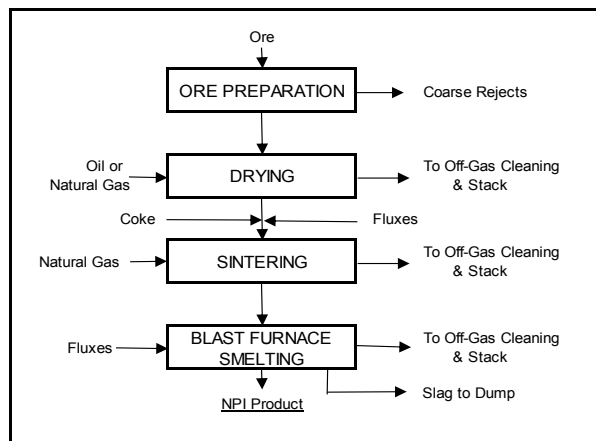


Figure 5: Sintering/Blast Furnace Smelting Flowsheet for Limonite

Sintering/Submerged Arc Smelting

The sintering/submerged arc furnace (SAF) smelting process is also commonly used to produce NPI in China, particularly for larger operations treating higher grade saprolite ores. It has lower environmental impact, higher efficiency, higher recovery (90%+) and higher product grade (10-15% Ni) than the BF process. For this paper a 15% nickel NPI product is assumed. There is no cobalt credit.

The process flowsheet is similar to Figure 5, with SAF smelting replacing BF smelting.

PROCESS COMMERCIALY APPLIED AS SATELLITE OPERATIONS

Heap Leaching

Significant effort has been invested by numerous organizations into the development of sulphuric acid heap leaching as a potentially lower cost alternative to PAL. However, only one commercial project has been developed to date, namely a satellite operation to Minara's Murrin Murrin PAL plant in Western Australia, processing grinding circuit scats. Murrin Murrin also successfully tested HL of fresh ore but are not practicing it commercially, preferring to treat all fresh ore by PAL. The closest to a stand-alone HL operation to date was at Çaldağ, in Turkey where European Nickel ran a demonstration plant. The project was purchased by Caldag Nickel, VTG Nickel, who are now testing the alternative AL process. No stand-alone HL commercial operation has been established as yet. HL is typically more suited to saprolite ores which are generally easier to leach than limonite. Problems with limonite include high fines content which affects percolation, and high iron content which results in high mass loss and heap instability. Compared with PAL and AL, HL yields lower nickel and cobalt extractions, typically 70-80%, and higher acid consumption of 500-700 kg/t ore.

The process flowsheet in Figure 6 is similar to the heap leaching of copper and uranium oxide ores with sulphuric acid. The ore is crushed, screened and agglomerated in a drum with concentrated sulphuric acid and a binder (commonly needed), and conveyed to a radial stacker for careful placement on a prepared leach pad. After allowing an initial curing period, the heap is irrigated with sulphuric acid solution in fresh water in a two-stage countercurrent leaching system for a combined cycle time of about 18 months. The primary stage is used to maximize the concentration of nickel and cobalt in the PLS while the secondary stage is a scavenging operation to maximize metal extraction and produce ILS as feed to the primary stage. Finally, water is applied to wash residual nickel and cobalt for collection in the ILS pond. The PLS is treated through downstream processing similar to the that applied to CCD overflow solution in the above PAL and EPAL flowsheets to produce MHP. The flowsheet shows an on/off pad arrangement in which the spent ore is reclaimed and dumped together with overburden.

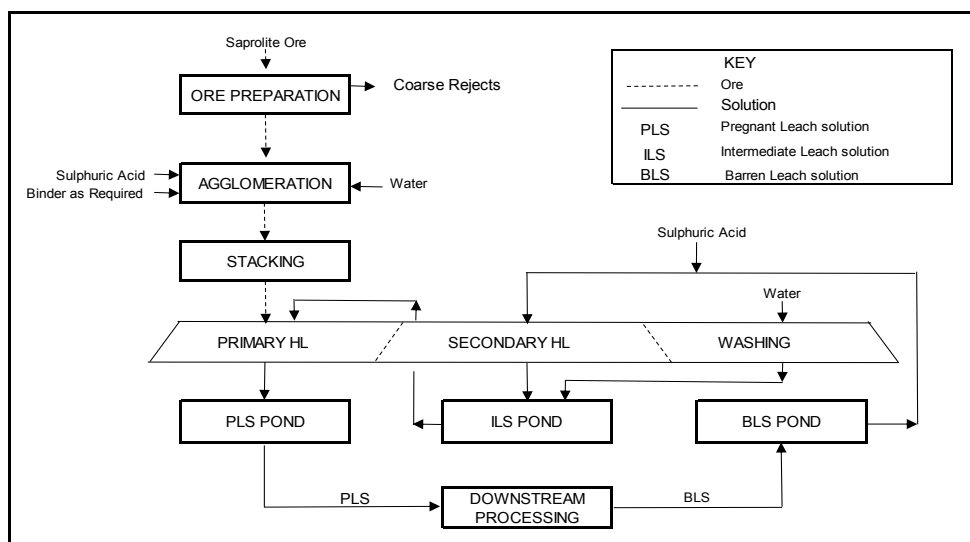


Figure 6: HL Flowsheet for Saprolite

Atmospheric Tank Leaching

There is a recent trend away from HL towards sulphuric acid atmospheric tank leaching despite the higher capex and higher acid consumption, because of higher extraction, smaller footprint, much shorter testwork program, shorter ramp up time, and lower environmental impact. Nickel and cobalt extractions are typically 85-95% and acid consumption 700-900 kg/t ore, depending on mineralogy, ore grade and the particular process concept adopted. Various AL concepts have been tested including single stage leaching for treating saprolite, two-stage leaching for treating limonite and saprolite sequentially and single stage leaching for treating a saprolite/limonite blend. Counter-current leaching systems have also been tested. To date, only one single stage AL facility treating saprolite has been commercialized as a satellite to PAL in the Ravensthorpe EPAL operation. (It has been reported that a stand-alone AL plant is in operation at Jiangxi Jiangli Sci-Tech Co.,Ltd, in Jiangli City, China, but there are no published details.) The Ravensthorpe AL facility uses sea

water to precipitate iron as jarosite with liberation of a portion of the acid (patented by BHP Billiton). For the purposes of this paper a stand-alone AL for saprolite using fresh water is selected.

In the Figure 6 flowsheet, ore preparation includes crushing, screening, grinding and thickening followed by leaching at atmospheric pressure in a series of mechanically agitated acid-brick lined vessels with steam injection as required to supplement the heat input from acid dilution. AL discharge is treated with limestone to achieve pH 2.5 to precipitate the remaining ferric iron with minimal nickel and cobalt losses followed by 6 Stages of CCD. The remainder of the flowsheet is similar to the PAL and EPAL cases.

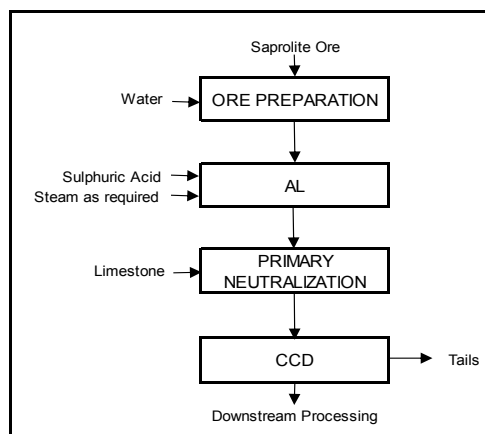


Figure 7: AL Flowsheet for Saprolite

PROCESSES DEVELOPED TO PILOT PLANT STAGE

Neomet Process

This Neomet Process is being developed by Neomet Technologies, Canada, as a potentially low capex process able to process both limonite and saprolite ores. It is an atmospheric hydrochloric acid tank leaching process coupled with a patented “atmospheric autoclave” system to regenerate the acid. Nickel and cobalt extractions are high (> 90% in reported testwork data). No secondary neutralization is needed to remove residual iron. Nickel and cobalt are recovered as basic chlorides which can be further processed to metals or calcined to form oxides. For the purposes of this paper mixed a nickel/cobalt oxide intermediate product for sale is selected. The flowsheet is a closed loop in that there are no liquid effluents and solids are environmentally benign. The process has been developed to a continuous integrated pilot plant level.

The flowsheet shown in Figure 8 is drawn from the Neomet paper at ALTA 2011⁽¹⁾ and the Arithmetek paper at ALTA 2012⁽³⁾. Ore is leached at 100-110°C under atmospheric conditions with recycled HCl. The leach is aggressive in order to affect maximum dissolution of pay-metals. The solid/liquid separation is effected by thickening and followed by a vacuum belt filter. The resulting solution is concentrated then iron and aluminium are subsequently removed by hydrolysis as oxides at 180-190°C during the acid regeneration process, and HCl is released and recycled to the leaching stage. The unique “atmospheric autoclave” system makes use of an inert solvent matrix which remains fluid at temperatures up to 200-250°C (i.e. autoclave temperatures), but which also remains open to the atmosphere, thereby allowing the formed HCl to be removed from the system. A bleed of the solvent matrix is taken and steam injected to recover nickel and cobalt as tri-basic chlorides. (If copper is present, an effective separation of copper from nickel and cobalt can be achieved, since copper hydrolyzes preferentially, and a two-stage process can be employed.) The solids are filtered on a vacuum filter, then calcined at 200-400°C to produce a mixed Ni/Co oxide product, with the released HCl being recycled to the leaching stage. Magnesium is then removed as MgO in a similar manner.

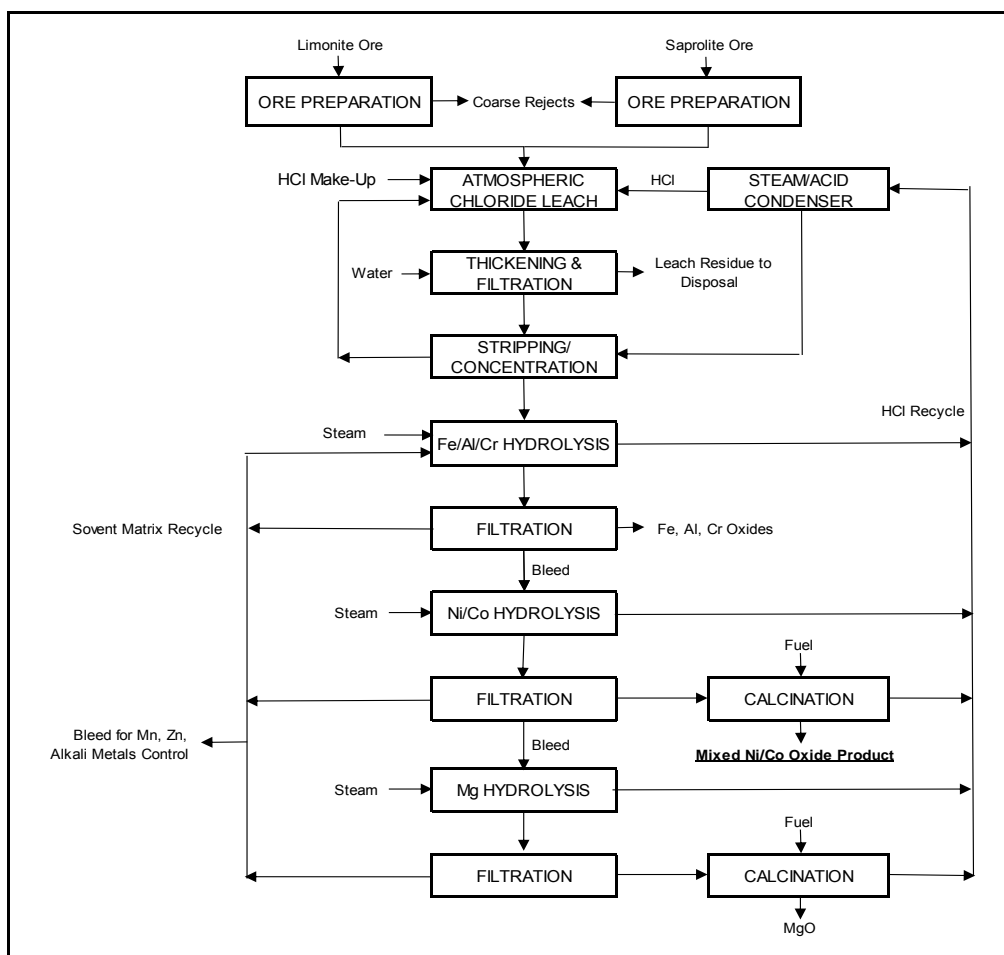


Figure 8: Neomet Flowsheet for Limonite & Saprolite

Direct Nickel Process

The DNi process under development by Direct Nickel Limited, Australia, able to process both limonite and saprolite ores. It is an atmospheric nitric acid tank leaching process coupled with a patented acid regeneration system. Nickel and cobalt extractions are high (reported to be > 90%). No secondary neutralization is needed to remove residual iron. Nickel and cobalt are recovered as MHP or further processed to mixed oxide. For the purposes of this paper mixed a nickel/cobalt oxide intermediate product for sale is selected. The flowsheet is a closed loop in that there are no liquid effluents and solids are environmentally benign. The process has been demonstrated at a continuous integrated pilot plant level. A feasibility Study has commenced for a DNi Process Plant at ANTAM's Buli operation in Halmahera, Indonesia, adjacent to Antam's new ferronickel smelter which is under construction.

The flowsheet shown in Figure 9 is drawn from the Direct Nickel paper at ALTA 2011⁽²⁾ and the Arithmetek paper at ALTA 2012⁽³⁾. Ore is leached in less than 5 hours at just over 100°C under atmospheric conditions with recycled HNO₃, then passed through a CCD circuit. The CCD overflow solution is concentrated by evaporation then iron and chromium and aluminium are removed as oxides by thermal hydrolysis. The pH of the iron-free solution is raised with recycled MgO to precipitate an intermediate product which contains the Ni, Co, Mn, Al and other base metals leaving a barren solution of mainly magnesium nitrate. The intermediate precipitate is separated from the barren solution by filtering and thickening, and is then re-leached. The re-leach solution is heated to remove aluminium by thermal hydrolysis and the pH of the resulting solution is then raised with MgO to produce a high quality Ni/Co hydroxide as a final product. The barren solution is evaporated to form a nitrate melt then fed to a thermal decomposition in which NO_x gases are liberated and MgO powder formed. 99% of the NO_x gases are recovered through a series of absorbers and scrubbers to form a 55% strength nitric acid. The acid and MgO are recycled to the process.

Applicability

The applicability of the selected processes to the orebody for the purposes of the economic comparison in the Part 2 paper is as follows:

- Limonite ore alone: PAL, Caron, Sintering/BF.
- Saprolite ore alone: RKEF, Sintering/SAF, AL, HL.
- Limonite and Saprolite: EPAL (separate feed of Limonite and Saprolite), Neomet (blended feed), DN_i (blended feed).

It is acknowledged that for a real project additional cases would likely be evaluated such as:

- Partial blends: for example for PAL and Caron it may be feasible to blend a portion of the saprolite with the limonite without moving outside the limits based on industry experience namely about 5% Mg maximum for PAL and about 35% Fe minimum for Caron.
- Combination of two processes with compatible features such as:
 - PAL for limonite coupled with HL for saprolite, with interlocking downstream processing.
 - PAL for limonite coupled with AL for saprolite, with interlocking downstream processing (EPAL is a proprietary example of this processing strategy.)
 - Sintering/BF for limonite coupled with Sintering/SAF or RKEF for saprolite.

It should be noted that for orebodies with different ratios of limonite and saprolite than the 50:50 assumed for this paper, the applicability of the processes can change. For example:

- If the saprolite zone is small, PAL, Caron or Sintering/BF may be applicable to the entire orebody by blending.
- If the limonite zone is small, AL, HL or Sintering/SAF may be applicable to the entire orebody by blending.
- The inclusion of a nontronite zone such as in the Australian dry laterite deposits could also affect process applicability.

PERFORMANCE

The optimum metal recovery for the application of each process for a real project is affected by number of variables other than the inherent characteristics of the process, some of which are site specific. These include:

- Ore grade
- Ore mineralogy
- Unit cost of power
- Cost of key reagents
- Reagent transport cost
- Type of downstream process.

Table 2 presents indicative metal recovery figures for the selected processes and tropical laterites drawn from industry experience commercially applied processes and projections from testwork for processes under development. Sulphuric acid consumptions are presented where applicable.

Processes	%Recovery	Sulphuric Acid Kg/T Ore
Pressure Acid Leaching (PAL)	Ni & Co 90-95	300-400
Enhanced Pressure Acid Leaching (EPAL)	Ni & Co 90-95	400-600
Caron Reduction Roast-Ammonia Leach (Caron)	Ni 75-85, Co 35-45	-
Rotary Kiln Electric Furnace Smelting (RKEF)	Ni 90-95	-
Sintering/Blast Furnace Smelting (BF)	Ni 80+	-
Sintering/Submerged Arc Smelting (SAF)	Ni 90+	-
Heap Leaching (HL) with Sulphuric Acid	Ni & Co 90-95	500-700
Atmospheric Tank Leaching (AL) with Sulphuric Acid	Ni & Co 90-95	700-900
Neomet Chloride leach (Neomet)	Ni & Co 90+	-
Direct Nickel Nitric Acid Leach (DN _i)	Ni & Co 90+	-

More detailed performance data, including energy consumption, are presented in the Part 2 paper.

SUMMARY AND CONCLUSIONS

For the treatment of nickel-cobalt laterites, commercially applied processes include PAL, Caron, RKEF, Sintering/BF and Sintering/SAF. Processes commercially applied as satellite operations to PAL include HL and AL. There are also a number of developing processes which have reached the pilot plant stage, including Neomet and Direct Nickel.

For the purposes of the economic comparison in the Part 2 paper, based on a hypothetical orebody with a 50:50 ratio of limonite and saprolite, Caron, and Sintering/BF are applicable to the limonite ore alone; RKEF and Sintering/SAF, AL and HL are applicable to the saprolite ore alone; EPAL Neomet and DNi are applicable to the entire orebody.

It is acknowledged that for a real project, additional cases would likely be evaluated such as partial blends and combinations of compatible processes, but these options are outside of the scope of this comparative review.

It should also be noted that for orebodies with different ratios of limonite and saprolite, the applicability of the processes can change.

The optimum metal recoveries for the application of each process for a real project are affected by number of variables other than the inherent characteristics of the process; some of which are site specific. Indicative metal recovery figures and sulphuric acid consumptions (where applicable) are presented for the selected processes. More detailed performance data, including energy consumption, are presented in the Part 2 paper

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ABSTRACT

This paper presents an analysis of the reagent and utility costs of commercially applied processes for nickel-cobalt laterites and also for two processes under development. For each process, a numerically rigorous mass-energy balance (process model) was used to quantify reagent and utility requirements. These numbers were used to calculate the variable portion of the operating cost for processing a hypothetical laterite deposit. Part 1 of this paper outlines the processes examined and their applicability to limonite and saprolite.

The established hydrometallurgical processes examined in this exercise are:

- Pressure acid leaching;
- Enhanced pressure acid leaching;
- Agitated tank leaching at atmospheric pressure;
- Heap leaching;
- Caron reduction roast, ammonia leach.

The pyrometallurgical processes examined are:

- Rotary kiln calcination with electric furnace smelting;
- Sintering with blast furnace smelting;
- Sintering with submerged arc smelting.

The two developing technologies examined are:

- Neomet;
- Direct Nickel.

The hydrometallurgical processes and the developing processes were assumed to produce nickel and cobalt as intermediate products (mixed nickel-cobalt hydroxide or oxide, cobalt sulphide in the Caron process). The pyrometallurgical processes were assumed to produce ferronickel or nickel pig iron.

Pressure acid leaching appears to be the option with the lowest reagent and utility cost for limonite and the option with the highest reagent and utility cost for saprolite. Of the established sulphate based technologies, atmospheric tank and heap leaching gave the lowest reagent and utility costs for saprolite. At the unit costs assumed, the cost of energy makes the pyrometallurgical options appear to be less attractive than the established hydrometallurgical options, at least in terms of the variable operating cost. The two developing technologies appear to be potentially competitive with the established options for processing saprolite. Assuming revenue for the nickel and cobalt in the intermediate products from the hydrometallurgical processes and only for the nickel in the alloys from the pyrometallurgical processes, net revenue gives the same ranking as the reagent and utility costs. If by-product credit for hematite and for magnesium oxide can be realized as well as the revenue for nickel and cobalt, the developing technologies could well turn out to be distinctly superior to all of the established technologies, because they also produce hematite and magnesium oxide.

INTRODUCTION

Increasingly, future nickel supplies will have to be produced from lateritic deposits, including the lower grade limonite zones which are not suitable for ferronickel smelting by the workhorse of the industry, rotary kiln/electric-furnace technology. This has led to the rise of other technologies, for example pressure acid leaching. However, pressure acid leaching carries a high capital cost and a number of plants have suffered from commissioning problems. That has led to less capital intensive processes operating at atmospheric pressure, including heap leaching and agitated tank leaching with sulphuric acid. Novel processes using hydrochloric acid and nitric acid are also under development. In addition, the rising cost of nickel and rapidly increasing demand, especially in China, have led to nickel pig iron smelting operations, initially of small capacity, but becoming larger.

The established and the newer technologies offer a range of options for exploiting new laterite resources. A complication is that laterite deposits typically contain ore zones with significantly different mineralogy, so that a particular process may be applicable to only a portion of the new resource. This paper presents a review of the reagent and utility costs associated with the proven and the developing processes, for processing a hypothetical laterite deposit. The technical status of the processes examined is presented in the separate Part 1 companion to this paper, authored by Alan Taylor of ALTA Metallurgical Services, Australia. The Part 1 paper contains most of the literature references relevant to the processes covered in this Part 2 paper, and those references are not duplicated here.

SELECTED PROCESSES

The processes included in this review are listed in Table 1. The commercial category is 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. The selected developing processes are Neomet, a chloride process, and the Direct Nickel nitric acid leaching process. Some of the processes have a number of possible downstream processing and product options. For the comparative purposes of this paper, the processes are limited to the production of intermediates and onsite refining is excluded. Hydroxide or oxide products have been selected for the hydrometallurgical processes and ferronickel or nickel pig iron for the smelting processes.

Table1 - Selected Processes & Products

Processes	Product
Commercially Applied	
Pressure Acid Leaching	Nickel-cobalt hydroxide
Enhanced Pressure Acid Leaching	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	
Heap leaching with sulphuric acid	Nickel-cobalt hydroxide
Atmospheric tank teaching with sulphuric acid	Nickel-cobalt hydroxide
Developed to Pilot Plant Stage	
Neomet chloride leach	Nickel-cobalt oxide
Direct Nickel nitric acid leach	Nickel-cobalt hydroxide

FEED

The hypothetical laterite deposit assumed for this exercise has the following characteristics:

- Tropical type laterite of the sort found in South East Asia.
- Simplistically, it consists of a limonite zone underlaid with a saprolite zone.
- The ratio of limonite to saprolite is 50:50 (dry tonnage basis).
- The moisture content is 30% for both limonite and saprolite.

Table 1 lists the assays assumed for the limonite and saprolite in this exercise. The hypothetical mineralogy listed in Table 2 back-calculates to the assays in Table 1.

Table 1 - Hypothetical ore analysis

Assay	Limonite	Saprolite
Ni	1.22	1.60
Co	0.20	0.02
Al ₂ O ₃	5.61	0.86
Cr ₂ O ₃	3.36	0.67
Fe ₂ O ₃	64.31	11.99
CaO	0.06	0.15
MgO	1.54	27.33
MnO	1.51	0.18
Na ₂ O	0.17	0.05
K ₂ O	0.01	0.03
SiO ₂	7.62	40.00

Table 2 – Hypothetical ore mineralogy

Mineral	Limonite	Saprolite
NiO	1.55	0.00
Ni ₂ SiO ₄	0.00	2.86
CoO	0.25	0.03
Al ₂ O ₃ •H ₂ O	6.60	1.01
FeCr ₂ O ₄	4.95	0.99
Fe ₂ O ₃ •H ₂ O	45.28	0.00
Fe(OH) ₃	29.25	15.58
Ca ₃ Si ₂ O ₇ •3H ₂ O	0.12	0.31
Mg ₃ Si ₂ O ₅ (OH) ₄	0.00	41.05
Mg ₄ Si ₆ O ₁₅ (OH) ₂ •6H ₂ O	6.17	37.71
MnSiO ₃	2.79	0.33
Na ₂ SiO ₃	0.33	0.10
K ₂ SiO ₃	0.02	0.05
SiO ₂	2.68	0.00
Total	100.00	100.00

In this exercise minor elements were excluded for simplicity, though it is recognized that they may have important implications in specific situations and some may offer by-product opportunities. Some of the processes examined in this exercise are suited to only limonite or only saprolite. For those processes the feed was assumed to be only limonite or only saprolite. For processes suitable for processing either limonite or saprolite, and for pressure acid leaching, the models were run for three cases each, assuming the feed to be only limonite, only saprolite, or half limonite and half saprolite.

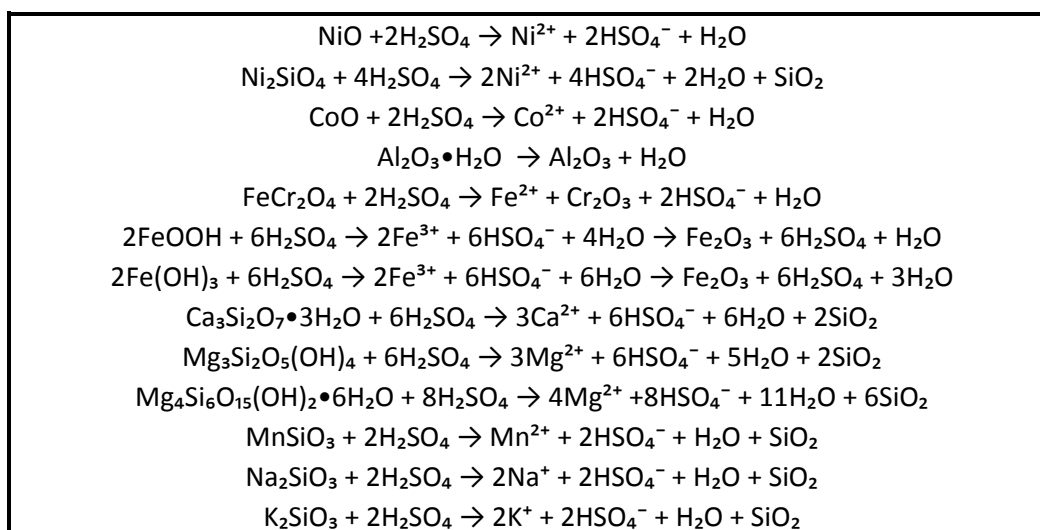
PROCESS MODELS

Commercially available process simulation software known as Aspen Plus was used to build a process model (numerically rigorous mass/energy balance) for each process considered in this exercise, and the appropriate laterite feed (limonite, saprolite, or 50:50 blend). Ore preparation was excluded from this exercise because that would be essentially the same for all the processes examined.

Pressure Acid Leaching

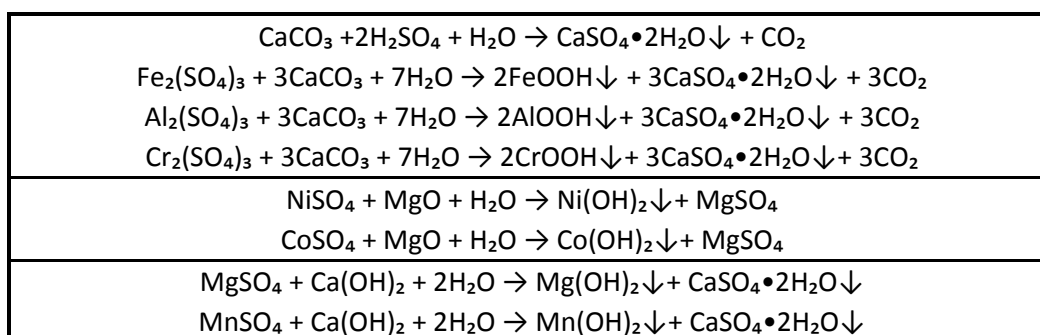
Pressure acid leaching (PAL) technology exploits the chemistry of ferric iron, in that at elevated temperature (achieved by using autoclaves operating at elevated pressure) the minerals containing ferric iron are dissolved, consuming acid, but the ferric iron re-precipitates as ferric oxide, releasing the corresponding acid back into solution and thereby substantially reducing the overall acid requirement. This makes the PAL technology better suited to limonite than to saprolite, as limonite contains more oxidized iron minerals than saprolite, which contains more magnesium silicate minerals which are high acid consumers, with no way of recovering the acid. A further factor in the acid requirement in PAL technology is that, at the temperatures used in the autoclave, the second dissociation of sulphuric acid does not occur, and each molecule of H_2SO_4 releases only one proton for acid leaching. Table 3 shows stoichiometry representing the autoclave chemistry for the PAL technology. Chromium is treated simplistically in this exercise, the assumption being that it does not form any hexavalent chromium species. In reality, a little reducing agent such as sulphur is added to prevent the formation of hexavalent chromium.

Table 3 – PAL leach stoichiometry



Once the laterite is leached, the slurry from the autoclave is progressively neutralized, first with limestone to neutralise residual free acid and precipitate ferric iron and aluminium from solution, then with magnesium oxide to precipitate nickel and cobalt (and copper, if present), and finally with lime to precipitate magnesium and manganese. Table 4 shows the stoichiometry used to represent the downstream chemistry.

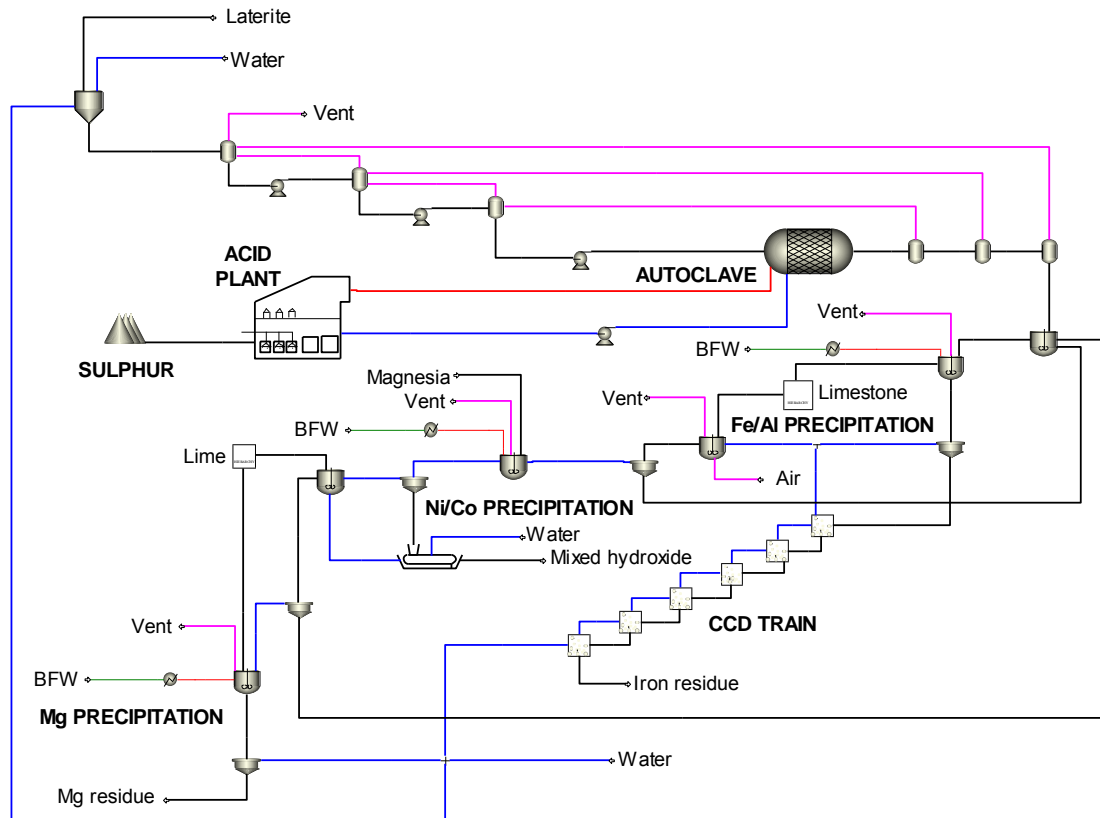
Table 4 – PAL precipitation stoichiometry



Not shown in the above chemistry, monovalent ions (Na, K) are rejected in jarosite that forms in the autoclave or as the residual ferric iron is precipitated.

Figure 1 illustrates the process model used to represent pressure acid leaching. It includes a standard sulphur-burning acid plant that produces the required amount of sulphuric acid and more than the required amount of steam for heating the autoclave. 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 to 250°C. 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.

Figure 1 – PAL model



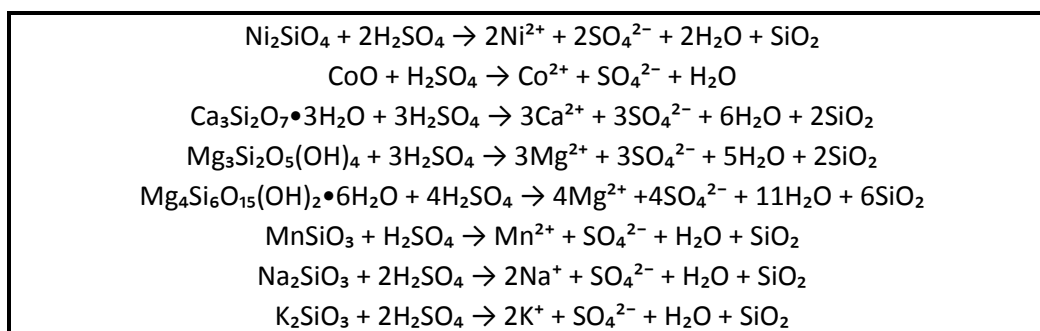
The leached slurry is first contacted with recycled hydroxide precipitates from the second stage of the precipitation of iron and aluminium and the first stage of the precipitation of manganese and magnesium, to re-dissolve co-precipitated nickel and cobalt and to partially consume the free acid left after the leach. The resulting 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 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 of the 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 fresh water. The washed filter cake leaves the circuit as the required hydroxide intermediate product. The supernatant and filtrate are combined and neutralized again, this time using lime, to precipitate the remaining nickel and cobalt into hydroxide-gypsum slurry that is 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.

Enhanced pressure acid leaching

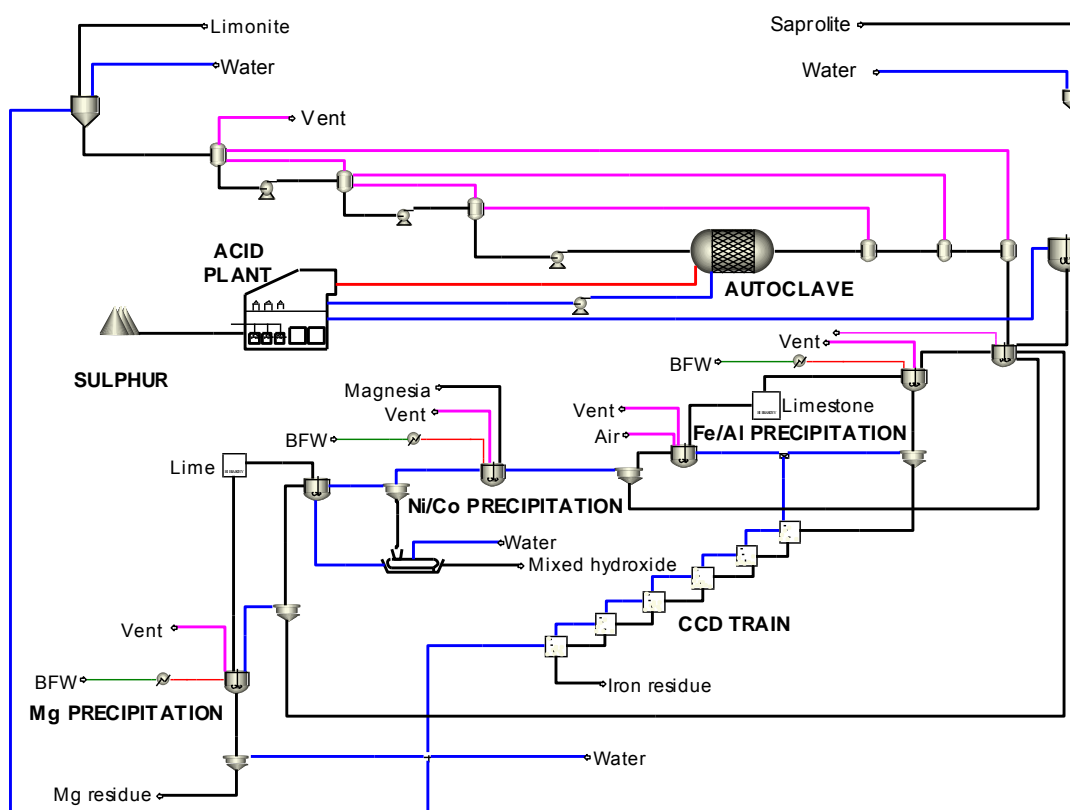
The main benefit of PAL technology, the reduction in acid requirements for leaching oxidized iron minerals, is diminished when the feed contains more magnesium silicate minerals, as in the case of saprolite. The magnesium minerals dissolve more easily than the oxidized iron minerals, but the magnesium does not re-precipitate and release acid. This means that, in the autoclave, each molar unit of magnesium requires two molar units of sulphuric acid. Enhanced pressure acid leaching technology (EPAL) seeks to minimise the overall acid requirement by feeding limonite to the autoclave and saprolite to a sulphuric acid leach at atmospheric pressure and lower temperature to consume the acid released as the slurry from the autoclave is cooled and the second dissociation of sulphuric acid occurs, and to extract nickel from the saprolite. Stoichiometry representing the atmospheric-pressure leach in EPAL technology is shown in Table 5. The autoclave chemistry and the downstream chemistry in EPAL are the same as in PAL.

Table 5 – EPAL leach stoichiometry for saprolite



The process model used to represent EPAL is illustrated in Figure 2. For this model the feed laterite was assumed to be half limonite and half saprolite, mined selectively, the limonite going to the autoclave and the saprolite to the atmospheric pressure leach. This, of course, is a somewhat artificial situation, set up for the exercise presented in this paper. In reality, it may be that the only saprolite needed is that consumed by the residual acid in the slurry from the autoclave, or that the extent of selective mining assumed is not achievable.

Figure 2 – EPAL model



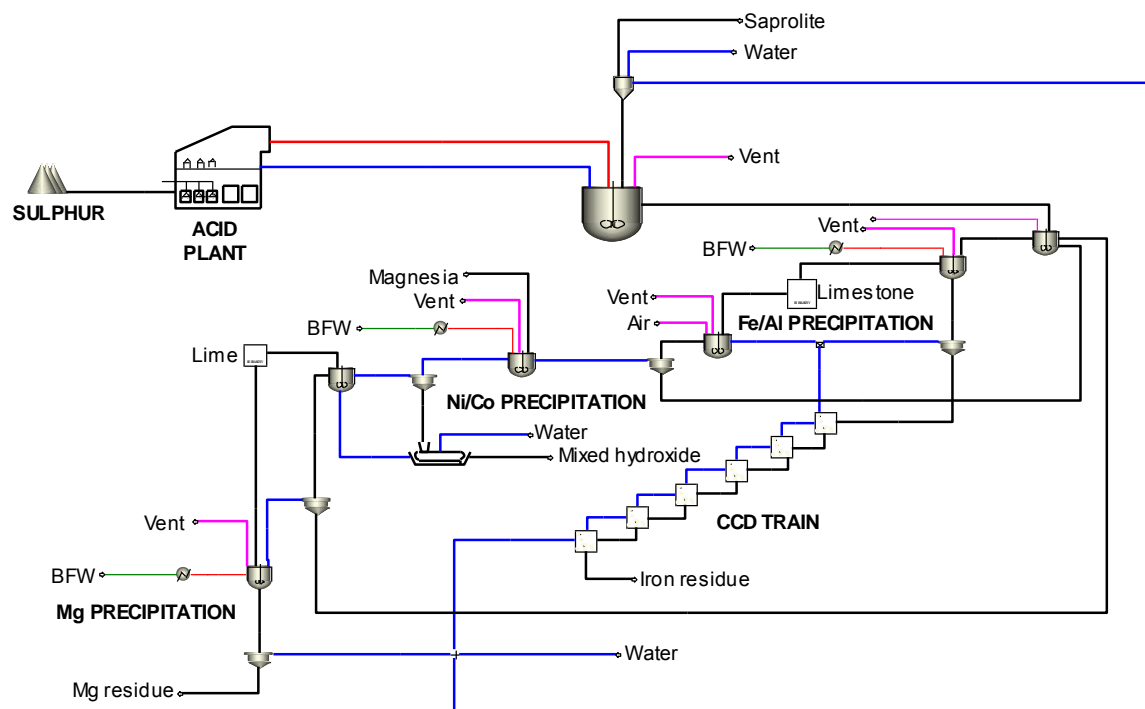
The limonite is treated as in the PAL model. As outlined in the part 1 companion paper, the saprolite is mixed with water and pre-leached at atmospheric pressure with acid from the acid plant to destroy minerals such as carbonates (although the assumed feed does not contain these minerals). The pre-leached saprolite slurry and the pressure leached limonite slurry are combined at the step in which recycled hydroxides are used to partially neutralize the residual acid from leaching, and the nickel-bearing magnesium silicate minerals are dissolved. From this stage on, the circuit is the same as that of the PAL model.

Atmospheric tank leaching

Atmospheric tank leaching of laterite (AL) seeks to dispense with the autoclave altogether, and is applicable where the laterite is amenable to atmospheric leaching. Although there are exceptions, that usually means saprolite. The leach chemistry is as illustrated in Table 5 and the downstream chemistry is the same as for PAL and EPAL technology.

Figure 3 illustrates the process model used to represent atmospheric-pressure tank leaching. In this model the feed laterite is assumed to be all saprolite. The incoming saprolite is mixed with water and leached in agitated tanks with sulphuric acid from the acid plant. The leach is heated by addition of steam from the acid plant. The leached saprolite is contacted with recycled hydroxides to partially neutralize the residual acid from the leach. From there on, the circuit is the same as in the preceding two models.

Figure 3 – AL model



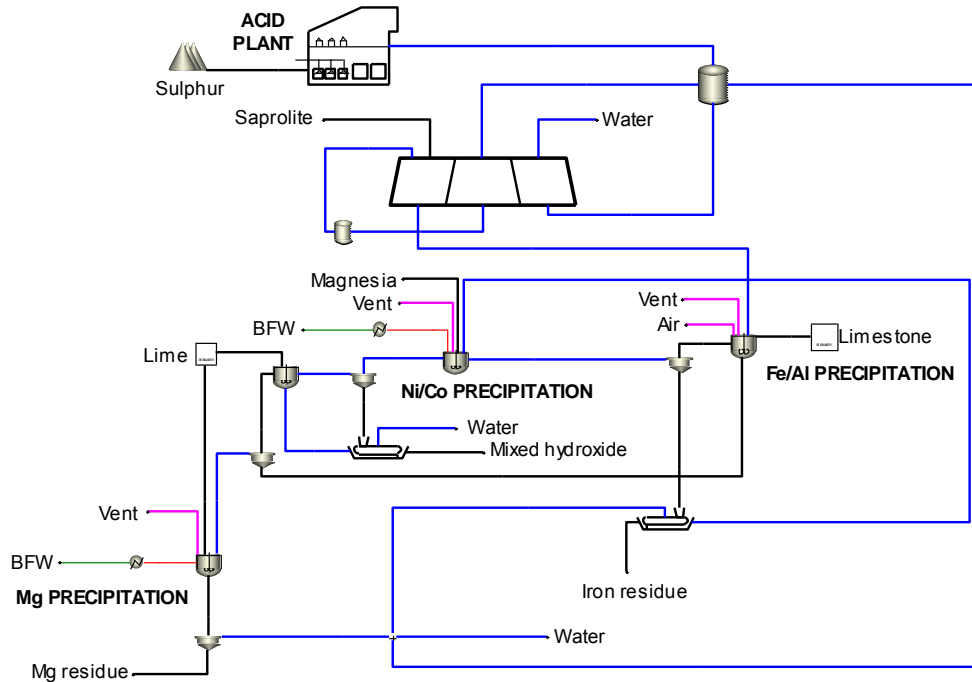
Heap leaching

Heap leach (HL) technology seeks to do away with the agitated tanks of AL technology, as well as the bulk of the solid-liquid separation following the leach. HL technology is applicable to saprolite laterite, where there is not enough value in the oxidized iron minerals to justify either more acid in an atmospheric-pressure leach, or PAL technology. The process chemistry is essentially the same as for AL technology. HL works only where the nature of the ore allows it to undergo agglomeration.

The process model used to represent heap leach (HL) technology is illustrated in Figure 4. As for the AL model, the HL model assumes that the feed is all saprolite. The heap leach is done in three stages, the leached part of the heap being washed with water, the solution from this wash step being supplemented with sulphuric acid from the acid plant and used to fully leach the laterite. The solution from this part of the heap is then passed over fresh laterite in the newest part of the heap to

maximise the concentration of nickel and minimise that of free sulphuric acid. The solution from this step is contacted with recycled hydroxides to partially neutralize the remaining free acid. From this point onwards the circuit is the same as for the preceding models.

Figure 4 – HL model



Caron

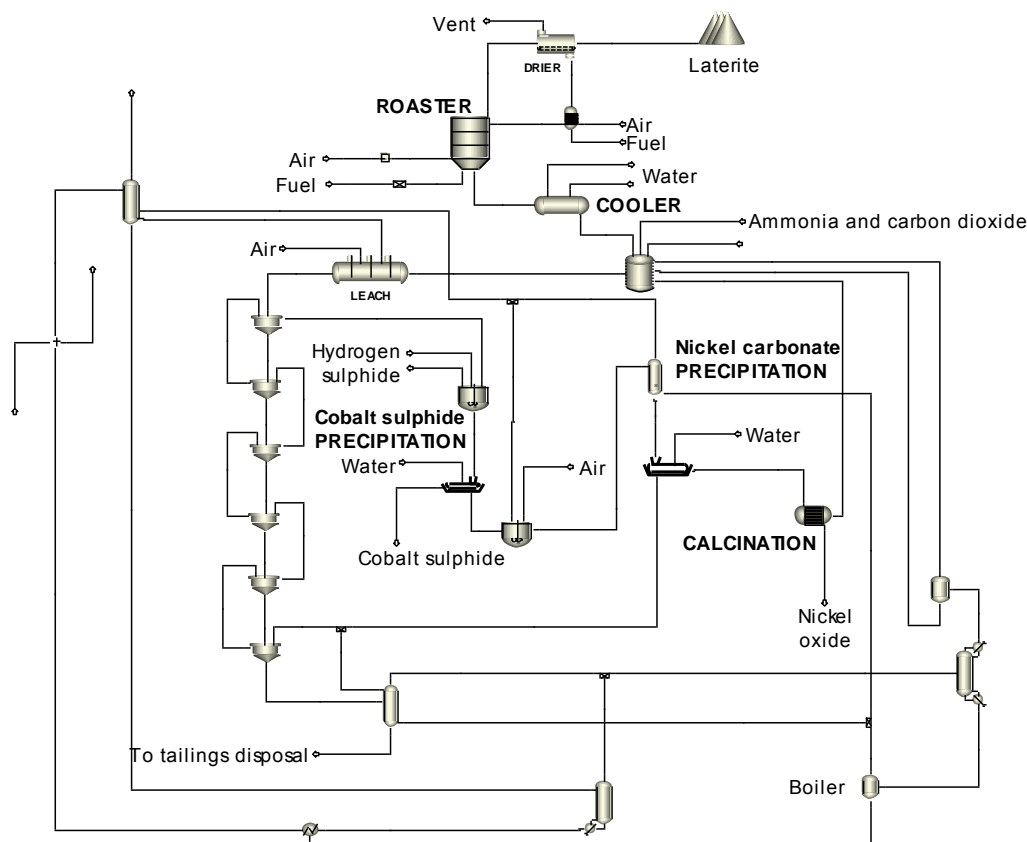
The Caron process is applicable to limonite. It uses a reducing roast to convert the oxidized iron minerals and the nickel and cobalt oxides in those minerals to a metallic state. An air plus aqueous ammonia-ammonium carbonate leach is used to selectively dissolve the metallic nickel and cobalt from the calcine, forming dissolved ammine complexes. The metallic iron is oxidized to solid ferric oxide. Table 6 lists stoichiometry representing the roast and leach stages of the Caron process. The reactions listed directly after the first reaction (representing the partial combustion of methane to carbon monoxide and water, giving the heat and the reducing atmosphere required for the roast) represent dehydration of the minerals, consuming energy. The middle five represent the reduction of iron, nickel and cobalt. The lower four represent the leach.

Table 6 – Caron stoichiometry, roast and leach

$\text{CH}_4 + 1\frac{1}{2}\text{O}_2 \rightarrow \text{CO} + 2\text{H}_2\text{O}$
$\text{Al}_2\text{O}_3 \cdot \text{H}_2\text{O} \rightarrow \text{Al}_2\text{O}_3 + \text{H}_2\text{O}$
$2\text{FeOOH} \rightarrow \text{Fe}_2\text{O}_3 + \text{H}_2\text{O}$
$2\text{Fe}(\text{OH})_3 \rightarrow \text{Fe}_2\text{O}_3 + 3\text{H}_2\text{O}$
$\text{Ca}_3\text{Si}_2\text{O}_7 \cdot 3\text{H}_2\text{O} \rightarrow 3\text{CaO} + 2\text{SiO}_2 + 3\text{H}_2\text{O}$
$\text{Mg}_3\text{Si}_2\text{O}_5(\text{OH})_4 \rightarrow 3\text{MgO} + 2\text{SiO}_2 + 2\text{H}_2\text{O}$
$\text{Mg}_4\text{Si}_6\text{O}_{15}(\text{OH})_2 \cdot 6\text{H}_2\text{O} \rightarrow 4\text{MgO} + 6\text{SiO}_2 + 7\text{H}_2\text{O}$
$\text{NiO} + \text{CO} \rightarrow \text{Ni} + \text{CO}_2$
$\text{CoO} + \text{CO} \rightarrow \text{Co} + \text{CO}_2$
$\text{FeCr}_2\text{O}_4 + \text{CO} \rightarrow \text{Fe} + \text{CO}_2 + \text{Cr}_2\text{O}_3$
$3\text{Fe}_2\text{O}_3 + \text{CO} \rightarrow 2\text{Fe}_3\text{O}_4 + \text{CO}_2$
$\text{Fe}_3\text{O}_4 + 4\text{CO} \rightarrow 3\text{Fe} + 4\text{CO}_2$
$\text{Ni} + 6\text{NH}_3 + \frac{1}{2}\text{O}_2 + \text{H}_2\text{O} \rightarrow \text{Ni}(\text{NH}_3)_6^{2+} + 2\text{OH}^-$
$\text{Co} + 6\text{NH}_3 + \frac{3}{4}\text{O}_2 + \frac{3}{4}\text{H}_2\text{O} \rightarrow \text{Co}(\text{NH}_3)_6^{3+} + 3\text{OH}^-$
$\text{Fe} + 6\text{NH}_3 + \text{O}_2 + \text{H}_2\text{O} \rightarrow \text{Fe}(\text{NH}_3)_6^{2+} + 2\text{OH}^-$
$2\text{Fe}(\text{NH}_3)_6^{2+} + \frac{1}{2}\text{O}_2 + 2\text{H}_2\text{O} \rightarrow \text{Fe}_2\text{O}_3 + 8\text{NH}_3 + 4\text{NH}_4^+$

The process model representing the Caron process is illustrated in Figure 5. This model assumes that the feed is all limonite.

Figure 5 – Caron model



The incoming limonite is dried using the sensible heat in the gas from the next step, in which the dried limonite is heated to about 800°C under a reducing atmosphere. This reduces the nickel and cobalt to a metallic state, along with part of the iron. The hot gas leaving the roaster contains carbon monoxide. Air is added to burn that to carbon dioxide, then the hot gas is used as the heat source for drying the incoming limonite. The hot reduced calcine is cooled in a rotating tube that is externally cooled with water. The cooled calcine is mixed with a recycled solution of aqueous ammonia and ammonium carbonate and leached with that and air, causing the metallic nickel and cobalt to dissolve as ammine complexes. The metallic iron is converted into insoluble ferric oxide. The leached slurry is thickened and washed with recycled barren solution, in a six-stage counter-current decantation train.

The supernatant from the counter-current decantation train is contacted with hydrogen sulphide, precipitating the cobalt and some of the nickel as a mixed sulphide that, after recovery by filtration and washing with water, leaves the circuit as an intermediate cobalt product. The filtrate is contacted with air to oxidize any residual sulphide in the solution, and the oxidized solution is steam-stripped to convert the aqueous ammonia and ammonium carbonate to gaseous ammonia and carbon dioxide that are removed with the stripping steam, causing the nickel to be precipitated as a basic nickel carbonate that is calcined to nickel oxide that leaves the circuit as an intermediate nickel product. The steam, ammonia and carbon dioxide mixture is condensed and recycled to the leach.

The washed underflow from the counter-current decantation train is stripped with steam to convert the dissolved ammonia and ammonium carbonate to gaseous ammonia and carbon dioxide. The stripped underflow leaves the circuit as leach residue.

The air leaving the leach and the steam, ammonia and carbon dioxide from the stripped tailings are scrubbed with water to capture the ammonia and carbon dioxide before the remaining water-saturated air is discharged to the atmosphere. The dilute solution of aqueous ammonia and ammonium carbonate from the scrubber is distilled to recover the ammonia and carbon dioxide in a more concentrated gaseous form and to regenerate water that is recycled to the scrubber.

The steam, ammonia and carbon dioxide joins the stream distilled from the scrubber bottoms and is separated by distillation into a concentrated ammonia/ammonium carbonate solution and water. The strong solution returns to the leach. Part of the water from this step is boiled to raise steam for stripping the tailings and the rest joins the water going to the scrubber, along with make-up fresh water.

Neomet

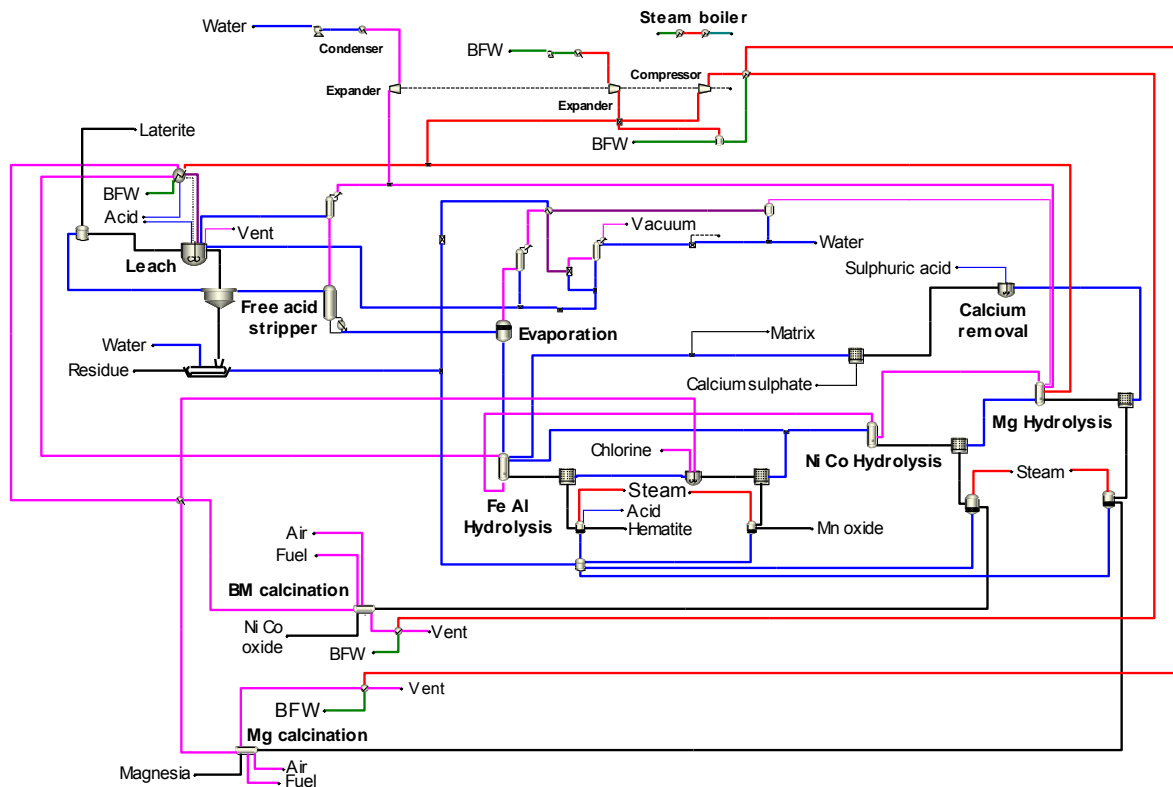
Leaching laterite with sulphuric acid entails adding the required amount of acid, then neutralising the solution after leaching to recover the valuable elements. The sulphuric acid is used only once. The sulphuric acid processes, therefore, necessarily entail the consumption of a fixed quantity of sulphur per unit of valuable product, and generate substantial volumes of wastes which need to be contained and managed appropriately. Neomet technology seeks to avoid this by regenerating and recycling the acid used in the leach. To do so, Neomet technology uses hydrochloric acid to leach the laterite, dissolving essentially all of the iron, aluminium and magnesium with the valuable metals. Since the iron in laterite comes from oxidized iron minerals, the resulting leach solution contains high levels of ferric chloride. If necessary, the solution can be oxidized to convert any ferrous chloride to ferric chloride. The ferric chloride solution is evaporated to remove excess water and the concentrated solution is heated to about 185°C, at atmospheric pressure, in a circulating "matrix" of molten salt hydrate. Steam is added, hydrolysing the ferric chloride to solid hematite and gaseous hydrochloric acid. The hydrochloric acid is recycled and the remaining molten salt is heated further with more steam addition to cause the nickel and cobalt to hydrolyse to solid hydroxychlorides and again to hydrolyse the magnesium chloride to solid magnesium hydroxychloride. The nickel/cobalt and the magnesium hydroxychlorides are calcined to oxides, releasing more hydrochloric acid for recycle. Stoichiometry representing the Neomet process is shown in Table 7. The first thirteen reactions represent the leach, the next three the hydrolysis of the trivalent ions, the next three the hydrolysis of nickel, cobalt and magnesium and the bottom three reactions represent the subsequent calcination steps. Apart from the laterite itself, the major input into the Neomet circuit is energy.

Table 7 – Neomet stoichiometry

$\text{NiO} + 2\text{HCl} \rightarrow \text{Ni}^{2+} + 2\text{Cl}^- + \text{H}_2\text{O}$
$\text{Ni}_2\text{SiO}_4 + 4\text{HCl} \rightarrow 2\text{Ni}^{2+} + 4\text{Cl}^- + 2\text{H}_2\text{O} + \text{SiO}_2$
$\text{CoO} + 2\text{HCl} \rightarrow \text{Co}^{2+} + 2\text{Cl}^- + \text{H}_2\text{O}$
$\text{Al}_2\text{O}_3 \cdot \text{H}_2\text{O} + 6\text{HCl} \rightarrow 2\text{Al}^{3+} + 6\text{Cl}^- + 4\text{H}_2\text{O}$
$\text{FeCr}_2\text{O}_4 + 8\text{HCl} \rightarrow \text{Fe}^{2+} + 2\text{Cr}^{3+} + 8\text{Cl}^- + 4\text{H}_2\text{O}$
$2\text{FeOOH} + 6\text{HCl} \rightarrow 2\text{Fe}^{3+} + 6\text{Cl}^- + 4\text{H}_2\text{O}$
$\text{Fe}(\text{OH})_3 + 3\text{HCl} \rightarrow \text{Fe}^{3+} + 3\text{Cl}^- + 3\text{H}_2\text{O}$
$\text{Ca}_3\text{Si}_2\text{O}_7 \cdot 3\text{H}_2\text{O} + 6\text{HCl} \rightarrow 3\text{Ca}^{2+} + 6\text{Cl}^- + 6\text{H}_2\text{O} + 2\text{SiO}_2$
$\text{Mg}_3\text{Si}_2\text{O}_5(\text{OH})_4 + 6\text{HCl} \rightarrow 3\text{Mg}^{2+} + 6\text{Cl}^- + 5\text{H}_2\text{O} + 2\text{SiO}_2$
$\text{Mg}_4\text{Si}_6\text{O}_{15}(\text{OH})_2 \cdot 6\text{H}_2\text{O} + 8\text{HCl} \rightarrow 4\text{Mg}^{2+} + 8\text{Cl}^- + 11\text{H}_2\text{O} + 6\text{SiO}_2$
$\text{MnSiO}_3 + 2\text{HCl} \rightarrow \text{Mn}^{2+} + 2\text{Cl}^- + \text{H}_2\text{O} + \text{SiO}_2$
$\text{Na}_2\text{SiO}_3 + 2\text{HCl} \rightarrow 2\text{Na}^+ + 2\text{Cl}^- + \text{H}_2\text{O} + \text{SiO}_2$
$\text{K}_2\text{SiO}_3 + 2\text{HCl} \rightarrow 2\text{K}^+ + 2\text{Cl}^- + \text{H}_2\text{O} + \text{SiO}_2$
$2\text{Fe}^{3+} + 6\text{Cl}^- + 3\text{H}_2\text{O} \rightarrow \text{Fe}_2\text{O}_3 + 6\text{HCl}\uparrow$
$2\text{Cr}^{3+} + 6\text{Cl}^- + 3\text{H}_2\text{O} \rightarrow \text{Cr}_2\text{O}_3 + 6\text{HCl}\uparrow$
$2\text{Al}^{3+} + 6\text{Cl}^- + 3\text{H}_2\text{O} \rightarrow \text{Al}_2\text{O}_3 + 6\text{HCl}\uparrow$
$3\text{NiCl}_2 + \text{H}_2\text{O} \rightarrow \text{Ni}_3(\text{OH})_5\text{Cl} + 5\text{HCl}\uparrow$
$3\text{CoCl}_2 + \text{H}_2\text{O} \rightarrow \text{Co}_3(\text{OH})_5\text{Cl} + 5\text{HCl}\uparrow$
$\text{MgCl}_2 + \text{H}_2\text{O} \rightarrow \text{MgOHCl} + \text{HCl}\uparrow$
$\text{Ni}_3(\text{OH})_5\text{Cl} \rightarrow 3\text{NiO} + \text{HCl}\uparrow + 2\text{H}_2\text{O}\uparrow$
$\text{Co}_3(\text{OH})_5\text{Cl} \rightarrow 3\text{CoO} + \text{HCl}\uparrow + 2\text{H}_2\text{O}\uparrow$
$\text{MgOHCl} \rightarrow \text{MgO} + \text{HCl}\uparrow$

Figure 6 illustrates the process model of the Neomet circuit. It begins with leaching the feed in strong hydrochloric acid, thickening and filtering the leached slurry, recycling some of the thickener overflow to the leach as the method of moving the feed solids into the leach reactor train and managing the solids content in the leach, washing the residue with water and discarding the washed residue. The filtrate is concentrated by evaporation, first in a stripper to remove much of the free acid and then by boiling till the atmospheric boiling point of the concentrated solution is 140°C. The acid in the vapour from the stripper is recovered in a scrubber and returned to the leach. The remaining steam is split, part going to the hydrolysis sequence and part being passed through an expansion turbine to harvest energy and then being condensed. The vapour from the boiling step is scrubbed to capture the acid, which returns to the leach, and the remaining steam is used to concentrate the wash filtrate from the leach.

Figure 6 – Neomet model



The concentrated solution is mixed with a circulating solvent matrix, contacted with steam from the following step and heated to 180°C, causing the ferric, aluminium and chromium chlorides to decompose to solid ferric/aluminium/chromium oxide (hematite) and gaseous wet hydrogen chloride. The resulting slurry of hematite in the molten matrix is filtered and washed using a proprietary method. The washed filter cake leaves the circuit. The wash filtrates are combined and recycled to the evaporation section. The primary filtrate is contacted with gaseous chlorine to oxidize and precipitate manganese as manganese dioxide, which is recovered and washed in the same way as the hematite, then leaves the circuit. This addition of chlorine also acts as chloride make-up in the circuit. The filtrate is split, part returning to the iron hydrolysis stage to build up the level of base metals, and a bleed proceeding to the nickel/cobalt hydrolysis step, where it is heated further and contacted again with steam to convert the nickel and cobalt to solid basic hydroxychlorides, releasing the associated chloride ions as gaseous hydrogen chloride. The nickel and cobalt hydroxychlorides are filtered out and washed in the same way as the hematite. The filtrate is heated further and contacted again with steam, causing the magnesium chloride to decompose to gaseous hydrochloric acid and solid magnesium hydroxychloride that is recovered by filtration at temperature and washed.

The base metal hydroxychlorides are calcined to a mixed nickel/cobalt oxide that leaves the circuit as the main product. The magnesium hydroxychloride is calcined to magnesium oxide that leaves the circuit. In these steps the chloride is released as gaseous hydrochloric acid that is recycled. The energy consumed in the calcination steps was assumed to come from steam, via indirect heating.

The gaseous mixture of hydrochloric acid and steam from the hydrolysis sequence is condensed in a heat exchanger, boiling water to raise steam at atmospheric pressure. The condensed acid returns to the leach. Some of the steam goes to the hydrolysis train and the balance is split, part being expanded through a turbine and then condensed, the condensate becoming boiler feed water for recycle. The mechanical power from the expansion turbine, plus the power from the turbine expanding/condensing steam from the scrubber after the free acid stripper, drives a compressor in which the other part of the atmospheric steam is compressed and used as the energy source for the acid stripping and evaporation steps of the process. The ratio of steam expanded to steam compressed is adjusted to make the amount of compressed steam produced balance the amount needed. The model assumes a gas-fired steam boiler, producing 30 Bar steam to supply the energy needed in the hydrolysis sequence.

The amount of atmospheric-pressure steam going through the hydrolysis sequence is manipulated to give 35 mass percent HCl in the steam/acid leaving the iron/aluminium hydrolysis step.

Direct Nickel

The Direct Nickel process uses nitrate chemistry to do what the Neomet process does in chloride chemistry, regenerating and recycling essentially all the nitric acid required to leach the laterite. Table 8 shows the stoichiometry used to represent the Direct Nickel technology.

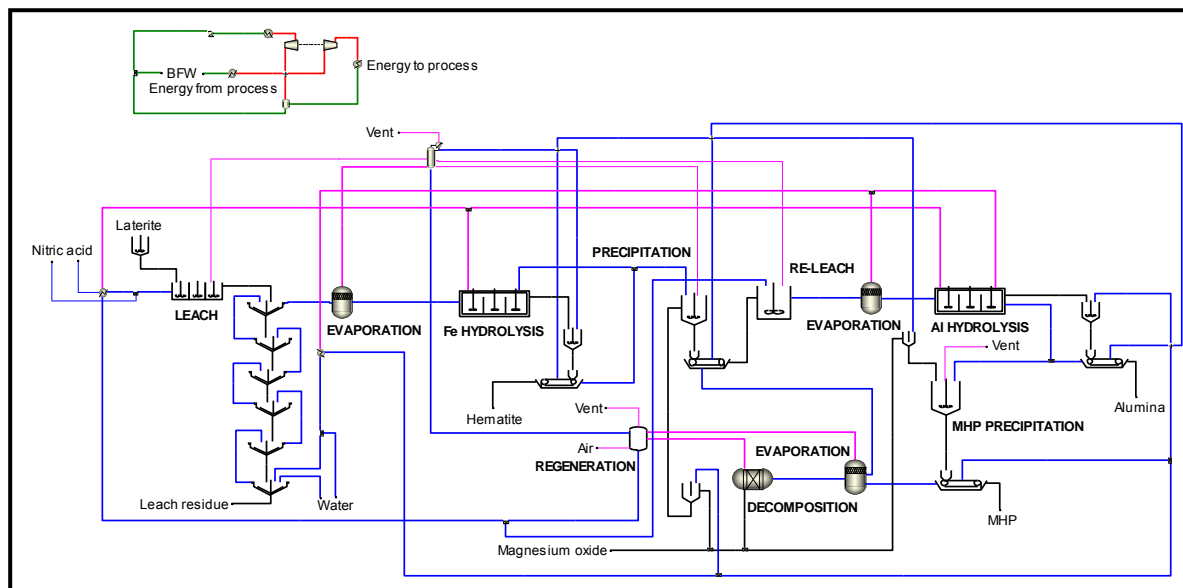
Table 8 – Direct Nickel stoichiometry

$\text{NiO} + 2\text{HNO}_3 \rightarrow \text{Ni}^{2+} + 2\text{NO}_3^- + \text{H}_2\text{O}$ $\text{Ni}_2\text{SiO}_4 + 4\text{HNO}_3 \rightarrow 2\text{Ni}^{2+} + 4\text{NO}_3^- + 2\text{H}_2\text{O} + \text{SiO}_2$ $\text{CoO} + 2\text{HNO}_3 \rightarrow \text{Co}^{2+} + 2\text{NO}_3^- + \text{H}_2\text{O}$ $\text{Al}_2\text{O}_3 \cdot \text{H}_2\text{O} + 6\text{HNO}_3 \rightarrow 2\text{Al}^{3+} + 6\text{NO}_3^- + 4\text{H}_2\text{O}$ $\text{FeCr}_2\text{O}_4 + 8\text{HNO}_3 \rightarrow \text{Fe}^{2+} + 2\text{Cr}^{3+} + 8\text{NO}_3^- + 4\text{H}_2\text{O}$ $\text{Fe}_2\text{O}_3 \cdot \text{H}_2\text{O} + 6\text{HNO}_3 \rightarrow 2\text{Fe}^{3+} + 6\text{NO}_3^- + 4\text{H}_2\text{O}$ $\text{Fe}(\text{OH})_3 + 3\text{HNO}_3 \rightarrow \text{Fe}^{3+} + 3\text{NO}_3^- + 3\text{H}_2\text{O}$ $\text{Ca}_3\text{Si}_2\text{O}_7 \cdot 3\text{H}_2\text{O} + 6\text{HNO}_3 \rightarrow 3\text{Ca}^{2+} + 6\text{NO}_3^- + 6\text{H}_2\text{O} + 2\text{SiO}_2$ $\text{Mg}_3\text{Si}_2\text{O}_5(\text{OH})_4 + 6\text{HNO}_3 \rightarrow 3\text{Mg}^{2+} + 6\text{NO}_3^- + 5\text{H}_2\text{O} + 2\text{SiO}_2$ $\text{Mg}_4\text{Si}_6\text{O}_{15}(\text{OH})_2 \cdot 6\text{H}_2\text{O} + 8\text{HNO}_3 \rightarrow 4\text{Mg}^{2+} + 8\text{NO}_3^- + 11\text{H}_2\text{O} + 6\text{SiO}_2$ $\text{MnSiO}_3 + 2\text{HNO}_3 \rightarrow \text{Mn}^{2+} + 2\text{NO}_3^- + \text{H}_2\text{O} + \text{SiO}_2$ $\text{Na}_2\text{SiO}_3 + 2\text{HNO}_3 \rightarrow 2\text{Na}^+ + 2\text{NO}_3^- + \text{H}_2\text{O} + \text{SiO}_2$ $\text{K}_2\text{SiO}_3 + 2\text{HNO}_3 \rightarrow 2\text{K}^+ + 2\text{NO}_3^- + \text{H}_2\text{O} + \text{SiO}_2$
$2\text{Fe}(\text{NO}_3)_3 + 3\text{H}_2\text{O} \rightarrow \text{Fe}_2\text{O}_3 + 6\text{HNO}_3$ $2\text{Al}(\text{NO}_3)_3 + 3\text{H}_2\text{O} \rightarrow \text{Al}_2\text{O}_3 + 6\text{HNO}_3$ $2\text{Cr}(\text{NO}_3)_3 + 3\text{H}_2\text{O} \rightarrow \text{Cr}_2\text{O}_3 + 6\text{HNO}_3$
$2\text{Al}(\text{NO}_3)_3 + 3\text{Mg}(\text{OH})_2 \rightarrow 3\text{Mg}(\text{NO}_3)_2 + 2\text{Al}(\text{OH})_3$ $\text{Co}(\text{NO}_3)_2 + \text{Mg}(\text{OH})_2 \rightarrow \text{Mg}(\text{NO}_3)_2 + \text{Co}(\text{OH})_2$ $\text{Ni}(\text{NO}_3)_2 + \text{Mg}(\text{OH})_2 \rightarrow \text{Mg}(\text{NO}_3)_2 + \text{Ni}(\text{OH})_2$
$\text{Mg}(\text{NO}_3)_2 \cdot 2\text{H}_2\text{O} \rightarrow \text{MgO} + \text{NO}_2 + \text{NO} + \text{O}_2 + 2\text{H}_2\text{O}$
$\text{NO} + \text{O}_2 \rightarrow 2\text{NO}_2$ $4\text{NO}_2 + \text{H}_2\text{O} + \text{O}_2 \rightarrow 4\text{HNO}_3$

Figure 7 illustrates the process model developed to examine the Direct Nickel process. The incoming laterite is leached in strong nitric acid, the leached residue is separated and washed with water in a counter-current decantation train and the washed residue leaves the circuit. The supernatant from the counter-current decantation train is concentrated by evaporation to an atmospheric boiling point of 140°C, the steam and free nitric acid evaporated going to a scrubber to capture the acid for recycle. The concentrated solution is heated to 180°C and excess steam is added, causing the ferric nitrate to hydrolyse to solid hematite and gaseous nitric acid. The residual steam and the gaseous nitric acid are recycled to the leach. The slurry from the hydrolysis reactor is quenched into water and the hematite is recovered by filtration and washed. The washed

hematite leaves the circuit. Part of the combined filtrate returns to the iron hydrolysis reactor to control the solids content of the slurry in that reactor and the balance is neutralized with recycled magnesium oxide to precipitate aluminium and the base metals. The precipitated slurry is filtered and the filter cake is re-dissolved in recycled nitric acid. The resulting solution is re-concentrated by evaporation. The steam evolved is split, part going to the counter-current decantation train where it is condensed and added to the wash water and the balance going to the aluminium hydrolysis step, in which the concentrated solution from the evaporation step is heated to 190°C to hydrolyse the aluminium nitrate to solid aluminium oxide and gaseous nitric acid. The excess steam and the gaseous nitric acid from this step are recycled.

Figure 7 – Direct Nickel model



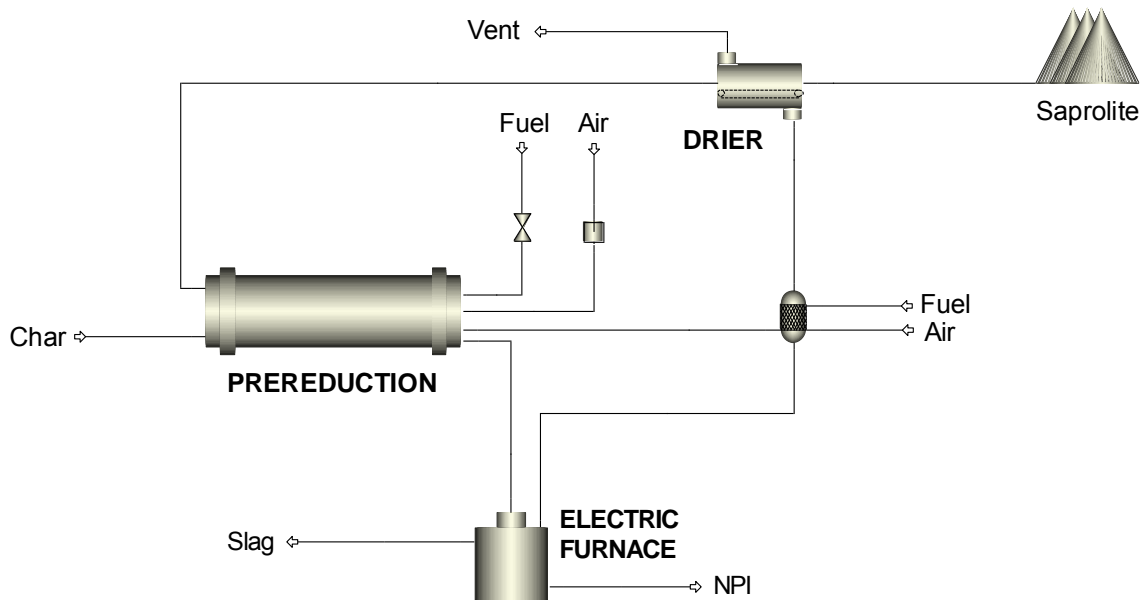
The hot slurry from the aluminium hydrolysis reactor is quenched into water, the aluminium oxide is separated from the diluted slurry by filtration and the filter cake is washed with water. The washed filter cake leaves the circuit and the combined filtrate is split, part recycling to the aluminium hydrolysis reactor to control the solids level in the reactor and the balance being contacted with recycled magnesium oxide to precipitate the base metals as a mixed nickel/cobalt hydroxide that is filtered out, washed and leaves the circuit as the product. The filtrate from the base metal recovery stage, essentially a solution of magnesium nitrate, is again concentrated by evaporation, such that it becomes a molten salt hydrate, essentially magnesium nitrate di-hydrate. This molten salt hydrate is heated to 500°C, causing it to decompose to solid magnesium oxide, steam and gaseous O₂, NO and NO₂. The magnesium oxide is partly recycled to the two precipitation steps and the excess leaves the circuit. The NO/NO₂ is converted back to nitric acid by a patented method in which NO reacts with HNO₃ in the presence of dissolved trivalent nitrogen, then the solution is oxidized to nitric acid with air. The gas from the thermal decomposition step and the steam from the preceding evaporation step are contacted with the aqueous nitric acid scrubbed from the steam from the evaporation step after the leach, and air is added. The NO/NO₂ is completely oxidized, giving aqueous nitric acid (60% HNO₃) that returns to the leach and the re-dissolution step after the first precipitation step.

The steam and gaseous nitric acid leaving the iron and aluminium hydrolysis stages, plus the steam, acid and nitrogen oxides leaving the evaporation and decomposition stages, carry latent heat that can be recycled. The heat exchanger shown above the label "Energy from process" in the separate little circuit at the top left of Figure 7 represents waste heat from the process being used to raise steam at atmospheric pressure from boiler feed water. That steam is split, part being expanded through a turbine and condensed. The power from the expansion turbine drives a compressor that raises the pressure of the balance of the steam such that its saturation temperature is high enough for it to be used as a heat source in the evaporation section. The heat exchanger labelled "Energy to process" represents that. The high temperature condensate is flashed to atmospheric pressure and the steam released joins the atmospheric steam raised from the process waste heat. The two condensate streams become boiler feed water for recycle. The energy input to the circuit not covered by the energy recycled as condensing compressed steam is supplied from an external source. The process model assumes natural gas, heating value 55.2 MJ/kg, as the heating utility.

Rotary kiln electric furnace

The rotary kiln, electric furnace (RKEF) smelting route is established technology for producing ferronickel from saprolite. The furnace power predicted by the model produced to examine this technology was calibrated against data published by Hatch¹. The ore is partially dried, then dehydrated and reduced before being smelted to produce ferronickel and slag. Figure 8 illustrates the process model used to examine this process.

Figure 8 – RKEF model



The incoming saprolite is dried using hot gas from the pre-reduction kiln, then mixed with coal char and heated to 800°C, and the hot pre-reduced mixture is then smelted in an electric furnace. The nickel, cobalt and much of the iron form a ferronickel alloy that is tapped and recovered as the desired product. The other constituents are rejected to a slag phase. Table 9 lists the stoichiometry used to represent this process. The first six reactions represent dehydration of the relevant minerals, the next seven represent reduction and the last two represent the combustion needed to generate the required heat in the kiln. The reduction reactions are completed in the furnace. The carbon monoxide in the hot gas leaving the furnace is burned to carbon dioxide with air and the resulting hot gas is used in the drying step. The model allows for fuel to also be burned in the hot furnace gas, but this was found to be unnecessary.

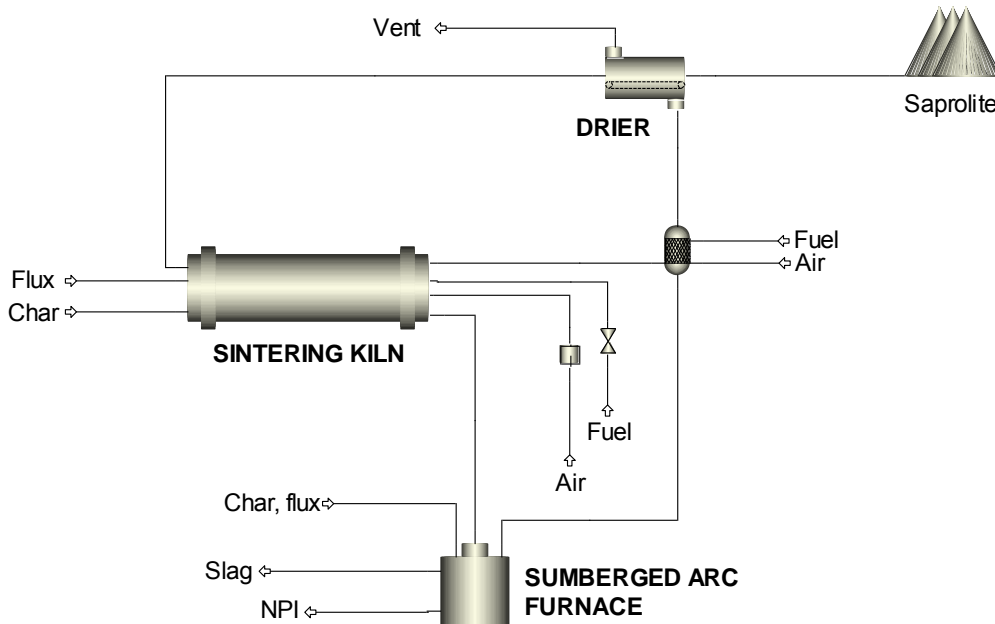
Table 9 – Pre-reduction stoichiometry

$\text{Al}_2\text{O}_3 \cdot \text{H}_2\text{O} \rightarrow \text{Al}_2\text{O}_3 + \text{H}_2\text{O}$ $\text{Fe}_2\text{O}_3 \cdot \text{H}_2\text{O} \rightarrow \text{Fe}_2\text{O}_3 + \text{H}_2\text{O}$ $2\text{Fe}(\text{OH})_3 \rightarrow \text{Fe}_2\text{O}_3 + 3\text{H}_2\text{O}$ $\text{Ca}_3\text{Si}_2\text{O}_7 \cdot 3\text{H}_2\text{O} \rightarrow 3\text{CaO} + 2\text{SiO}_2 + 3\text{H}_2\text{O}$ $\text{Mg}_3\text{Si}_2\text{O}_5(\text{OH})_4 \rightarrow 3\text{MgO} + 2\text{SiO}_2 + 2\text{H}_2\text{O}$ $\text{Mg}_4\text{Si}_6\text{O}_{15}(\text{OH})_2 \cdot 6\text{H}_2\text{O} \rightarrow 4\text{MgO} + 6\text{SiO}_2 + 7\text{H}_2\text{O}$
$\text{MgO} + \text{Fe}_2\text{O}_3 \rightarrow \text{MgFe}_2\text{O}_4$ $\text{CaO} + \text{Fe}_2\text{O}_3 \rightarrow \text{CaFe}_2\text{O}_4$ $\text{FeCr}_2\text{O}_4 + 4\text{C} \rightarrow \text{Fe} + 2\text{Cr} + 4\text{CO}$ $\text{FeCr}_2\text{O}_4 + 4\text{C} \rightarrow \text{Fe} + 2\text{Cr} + 4\text{CO}$ $\text{Fe}_2\text{O}_3 + 3\text{C} \rightarrow 2\text{Fe} + 3\text{CO}$ $\text{Ni}_2\text{SiO}_4 + 2\text{C} \rightarrow 2\text{Ni} + 2\text{CO} + \text{SiO}_2$ $\text{CoO} + \text{C} \rightarrow \text{Co} + \text{CO}$
$2\text{C} + \text{O}_2 \rightarrow 2\text{CO}$ $\text{CH}_4 + 1\frac{1}{2}\text{O}_2 \rightarrow \text{CO} + 2\text{H}_2\text{O}$

Sintering and submerged arc smelting

Sintering and submerged-arc smelting (SAF) is another established technology. In the context of process modelling, it is similar to the RKEF technology. The incoming laterite is dried and then mixed with coal char and flux (CaO in this exercise) and sintered at 1200°C, the heat coming from the combustion of natural gas. The sintered mixture is then smelted in an electric arc furnace. The stoichiometry used to model this process is similar to that shown in Table 9. Figure 9 illustrates the process model used to examine the SAF route.

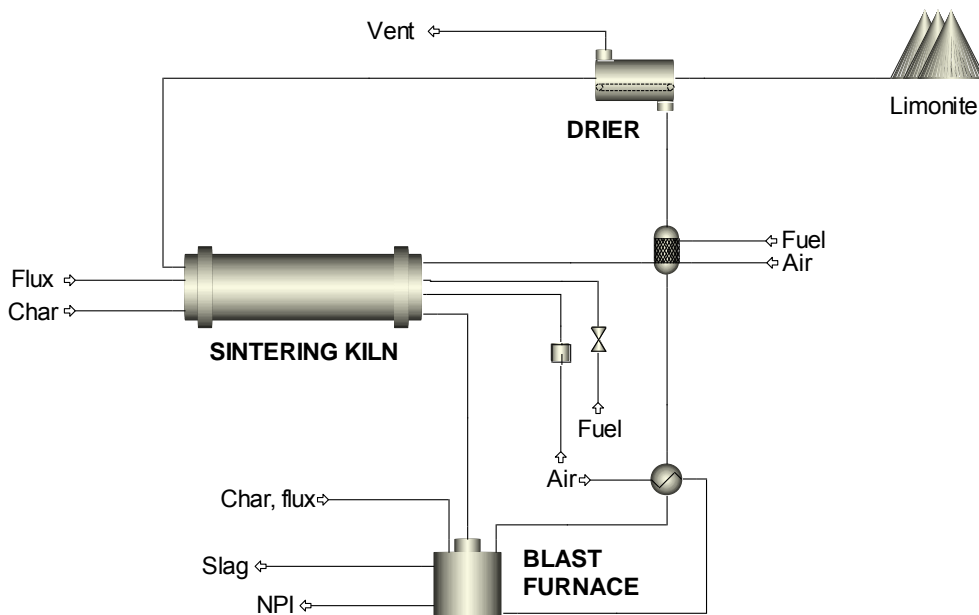
Figure 9 – SAF model



Sintering and blast furnace smelting

Sintering and smelting in a blast furnace (BF) is similar to sintering and smelting in a submerged arc furnace, except that the energy in the blast furnace comes from the combustion of excess coal char in the blast furnace. The excess char required by the blast furnace also results in the reduction of more of the iron, making for a product that contains a lower percentage of nickel. Figure 10 illustrates the process model used to examine the BF option.

Figure 10 – BF model



RESULTS AND DISCUSSION

The overall costs associated with all of the process options examined consist of capital costs, fixed operating costs and variable operating costs, which are those associated with reagents and utilities. The process models were set up assuming 30 thousand tonnes per year of nickel in the feed laterite. This exercise examines the variable operating costs, normalised to the cost per unit of nickel in the relevant intermediate product, so the exact throughput is not important in this exercise. Because the different processes have different nickel recoveries and use different parts of the orebody, i.e. limonite and saprolite, limonite alone or saprolite alone, the amounts of laterite consumed per unit of nickel produced vary, therefore the comparison includes an allowance for the variable portion of the mining cost of the ore. This was previously estimated² as \$5 per tonne of ore mined. Table 10 lists the unit costs used in this exercise for reagents and utilities. The quantities consumed by the various processes examined are listed in Table 11,

Table 12 and Table 13. Table 14 lists the calculated costs for the various processes examined, at the unit costs listed in Table 10.

Table 10 – Unit costs

Reagent/Utility	Unit cost,\$
Sulphur (S)	300/tonne
Sulphuric acid (H ₂ SO ₄)	100/tonne
Hydrogen sulphide (H ₂ S)	500/tonne
Limestone (CaCO ₃)	50/tonne
Lime (CaO)	100/tonne
Reactive magnesia (MgO)	300/tonne
Nitric acid (HNO ₃)	750/tonne
Chlorine (Cl ₂)	300/tonne
Ammonia (NH ₃)	500/tonne
Ammonium bicarbonate (NH ₄ HCO ₃)	260/tonne
Fresh water	1/tonne
Coal char	140/tonne
Silica sand	20/tonne
Natural gas (52 GJ/t, \$10/GJ)	520/tonne
Electrical power	\$30/GJ

Table 11 – Reagents and utilities (sulphate routes)

Reagent Utility	Consumption, kg/kg Ni in product					
	PAL			EPAL	AL	HL
	Limonite	Blend	Saprolite	Blend	Saprolite	Saprolite
Ore	124	107	94	78	66	82
S	8.4	20.9	35.6	10.8	15.6	14.8
CaCO ₃	13.7	32.8	60.4	3.1	1.3	2.2
MgO	0.6	0.6	0.6	0.6	0.6	0.6
CaO	3.7	15.5	25.7	13.4	24.7	24.8
Water	2.5	7.7	13.4	2.9	4.9	2.1

Table 12 – Reagents and utilities (smelting processes)

Reagent Utility	Consumption, per kg Ni in product		
	RKEF	SAF	BF
	Saprolite	Saprolite	Limonite
Ore	62	94	147
Gas, kg	13.1	8.2	18.5
Char, kg	1.4	2.1	32.5
CaO, kg	0	2.4	12.6
SiO ₂ , kg	0	0	26.3
Power, MJ	97	258	8

Table 13 – Reagents and utilities (developing processes and Caron)

Reagent Utility	Consumption, per kg Ni in product						
	Caron	Neomet			Direct Nickel		
	Limonite	Limonite	Blend	Saprolite	Limonite	Blend	Saprolite
Ore	103	123	109	94	119	104	90
NH ₃	0.04	-	-	-	-	-	-
NH ₄ HCO ₃	0.05	-	-	-	-	-	-
H ₂ S	0.07	-	-	-	-	-	-
HNO ₃	-	-	-	-	0.07	0.03	0.03
H ₂ SO ₄	-	0.1	0.1	0.2	-	-	-
Cl ₂	-	1.9	1.0	0.2	-	-	-
Water	126	34	61	45	0	0	10
Gas, kg	10	7	7	7	17	16	15

Table 14 – Calculated variable cost, \$/lb Ni

Process	Feed		
	Limonite	Blend	Saprolite
Pressure acid leach	2.33	5.10	8.07
Enhanced pressure acid leach	-	2.68	-
Atmospheric tank leach	-	-	3.63
Heap leach	-	-	3.54
Rotary kiln, electric furnace	-	-	4.64
Sinter, arc furnace	-	-	5.91
Sinter, blast furnace	7.44	-	-
Caron process	2.67	-	-
Neomet process	2.37	2.17	1.99
Direct Nickel process	4.43	3.91	3.71

Based on the numbers in Table 14, pressure acid leaching (PAL) would appear to be the best option for the limonite feed and the worst option for the saprolite feed. Both of the developing processes (Neomet and Direct Nickel) would appear to offer low variable operating costs relative to the established technologies, especially for the saprolite feed.

A reason for the variable cost of the Direct Nickel process being appreciably higher than that of the Neomet process is that the process model for the Direct Nickel process assumes conventional washing of the hematite filter cake with water, while the Neomet process used a different technique that substantially reduces the use of wash water, and thus the evaporation of water in the circuit. According to an earlier paper³ comparing these two processes to sulphate technology, if the Direct Nickel process could use the washing method of the Neomet process the variable cost of the Direct Nickel process would be reduced appreciably.

The above analysis was done purely on the variable costs associated with the reagents and utilities consumed per pound of nickel in the products. In addition to nickel, the processes examined generate by-products, the value of which would influence their economics. The sulphate based processes and the Direct Nickel process produce cobalt hydroxide in the mixed hydroxide intermediate product. The Caron process produces a mixed cobalt-nickel sulphide. The Neomet process produces cobalt oxide in its mixed oxide product. The Neomet and Direct Nickel processes also produce magnesium oxide and hematite. The pyrometallurgical processes produce ferronickel or nickel pig iron containing iron and chromium.

While there is no by-product credit to be had for cobalt in ferronickel or nickel pig iron, the metallic iron would displace iron from iron ore in the manufacture of stainless steel. For the next part of this exercise, the assumption used was that the nickel and cobalt in the mixed hydroxide, mixed oxide and sulphide products can be sold for 75 percent of the LME metal value. The iron in the ferronickel and the nickel pig iron would directly displace iron from other iron ore in the manufacture of stainless steel, giving it the value of pig iron. A brief search of the internet (May 2014) found the prices listed in Table 15 for the metals relevant to this exercise. Table 16 lists the revenue minus the variable costs, as calculated from these values and the various consumption numbers.

Table 15 – Metal values

Nickel	\$18 000/tonne
Cobalt	\$30 000/tonne
Pig iron	\$400/tonne

Table 16 – Revenue less variable cost (\$/lb Ni)

Process	Feed		
	Limonite	Blend	Saprolite
Pressure acid leach	5.46	1.84	-1.79
Enhanced pressure acid leach	-	4.39	-
Atmospheric tank leach	-	-	2.66
Heap leach	-	-	2.75
Rotary kiln, electric furnace	-	-	2.28
Sinter, arc furnace	-	-	1.52
Sinter, blast furnace	2.69	-	-
Caron process	4.29	-	-
Neomet process (Ni and Co only)	5.44	4.87	4.27
Neomet process (all by-products)	7.94	7.45	6.93
Direct Nickel process (Ni and Co only)	3.36	3.13	2.58
Direct Nickel process (all by-products)	5.84	4.91	3.69

If the pyrometallurgical processes are to be credited with by-product values, then it would seem reasonable to also credit the Neomet and Direct Nickel processes with some value for their by-products, which would be hematite (also containing the chromium as Cr_2O_3) and magnesium oxide. Assuming \$300/tonne for magnesium oxide and \$100/tonne for hematite significantly increases the net revenue for these two options. The chromium and any aluminium in the hematite would be rejected to the slag during smelting, which might require appropriate flux addition and diminish the value of the hematite accordingly. The Neomet developers believe that removing aluminium from the hematite would be fairly simple, though, should that be warranted.

For the limonite, pressure acid leaching still comes out as the most profitable of the established options. The next best of the established technologies, for limonite, would appear to be the Caron process. For the saprolite, pressure acid leaching would appear to be a non-starter. Of the established technologies, heap leaching and atmospheric tank leaching would seem to be options of choice for the saprolite. Of the pyrometallurgical options, accepting the argument for value in the iron, the sinter and blast furnace smelting option would appear to have both higher variable costs and higher net revenue. That would be because this option makes the alloy with the lowest nickel content, thus more iron per unit of nickel. The exercise presented here may be too simplistic to place particular reliance on the ranking found for the pyrometallurgical options.

Capital costs would influence the technology ranking, but that aspect was omitted from this exercise.

The two developing technologies examined, Neomet and Direct Nickel, would appear to be potentially competitive with the established options on the basis of revenue for nickel and cobalt. If their by-products can actually be sold for the prices assumed in this exercise, these two processes would appear to be distinctly superior to the established technology.

SUMMARY AND CONCLUSIONS

The exercise presented in this paper was done to examine the reagent and utility portions of the operating costs of the processes examined, for a hypothetical laterite ore. The conclusions of this exercise are as follows, for the unit costs used:

- Pressure acid leaching offers the lowest reagent and utility costs of the established options for processing limonite.
- Atmospheric tank leaching and heap leaching appear to be the best sulphate based options for processing saprolite.
- The pyrometallurgical processes appear to be less attractive, for both laterite and saprolite, than the established hydrometallurgical options.
- The two developing processes, Neomet and Direct Nickel, both appear to be potentially competitive with the established processes, and distinctly superior if their by-products prove marketable.

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TECHNICAL & COST COMPARISON OF LATERITE TREATMENT PROCESSES PART 3

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ABSTRACT

This is the third paper in a series examining the economics of processing a hypothetical nickel-cobalt laterite using commercially applied processes and two processes that are still under development. Part 1 of this series 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 extends the comparison to the fixed operating and capital costs and uses simple financial modelling to compare the different processes.

The established hydrometallurgical processes examined in this exercise are:

- Pressure acid leaching;
- Enhanced pressure acid leaching;
- Agitated tank leaching at atmospheric pressure;
- Heap leaching;
- Caron reduction roast, ammonia leach.

The pyrometallurgical processes examined are:

- Rotary kiln calcination with electric furnace smelting;
- Sintering with blast furnace smelting;
- Sintering with submerged arc smelting.

The two developing technologies examined are:

- Neomet;
- Direct Nickel (DNi).

The hydrometallurgical processes and the developing processes were assumed to produce nickel and cobalt as intermediate products (mixed nickel-cobalt hydroxide or oxide, and cobalt sulphide in the Caron process). The pyrometallurgical processes were assumed to produce nickel in metallic iron, i.e. ferronickel or nickel pig iron.

Pressure acid leaching appears to offer the best economics for treating the limonite if the Neomet and DNi processes do not produce saleable by-products. If the by-product hematite and magnesia turn out to be saleable, particularly if the hematite can be sold for more than the price of iron ore, the Neomet and DNi processes look distinctly superior.

Of the established processes for saprolite, heap leaching would appear to offer the best economics, but if the by-products are saleable Neomet offers better economics for processing saprolite.

The pyrometallurgical processes appear to be economically inferior to the hydrometallurgical processes.

INTRODUCTION

Increasingly, future nickel supplies will have to be produced from lateritic deposits, including the lower grade limonite zones which are not suitable for ferronickel smelting by the workhorse of the industry, rotary kiln/electric-furnace technology. This has led to the rise of other technologies, for example pressure acid leaching and processes operating at atmospheric pressure, including heap leaching and agitated tank leaching with sulphuric acid. Novel processes using hydrochloric acid and nitric acid are also under development. In the past few years, rising cost of nickel and rapidly increasing demand, especially in China, led to nickel pig iron smelting operations that now appear to have been adversely affected by the current lower nickel price.

Laterite deposits typically contain ore zones with significantly different mineralogy, so any particular process may be applicable to only a portion of a new resource. This paper is the third in a series examining the processing of a hypothetical limonite ore and a hypothetical saprolite ore using established and developing technologies. The technical status of the processes examined was presented in the Part 1 of this series⁽¹⁾. Part 2 of the series⁽²⁾ presented a review of the reagent and utility costs associated with the proven and the developing processes. This paper extends the comparison to include capital and fixed operating costs.

SELECTED PROCESSES

The processes included for this exercise 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 selected developing processes are Neomet, a chloride process, and the Direct Nickel nitric acid leaching process (DNi). Some of the processes have a number of possible downstream processing and product options. For the comparative purposes of this series, the processes are limited to the production of intermediates and onsite refining is excluded. Hydroxide or oxide (and cobalt sulphide, in the Caron process) products have been selected for the hydrometallurgical processes and ferronickel or nickel pig iron for the smelting processes.

Table1: Selected 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 leaching with sulphuric acid	Nickel-cobalt hydroxide
Developing Technologies	
Neomet chloride leach	Nickel-cobalt oxide
Direct Nickel nitric acid leach	Nickel-cobalt hydroxide

FEED

The hypothetical laterite deposit assumed for this exercise has the following characteristics:

- Tropical type laterite of the sort found in South East Asia.
- Simplistically, it consists of a limonite zone and a saprolite zone.
- The ratio of limonite to saprolite is 50:50 (dry tonnage basis).
- The moisture content is 30% for both limonite and saprolite.

Table lists the assays assumed for the limonite and saprolite in this exercise. The hypothetical mineralogy listed in Table 3 back-calculates to the assays in Table 2.

Table 2: Hypothetical ore analysis

Assay	Limonite	Saprolite
Ni	1.22	1.60
Co	0.20	0.02
Al ₂ O ₃	5.61	0.86
Cr ₂ O ₃	3.36	0.67
Fe ₂ O ₃	64.31	11.99
CaO	0.06	0.15
MgO	1.54	27.33
MnO	1.51	0.18
Na ₂ O	0.17	0.05
K ₂ O	0.01	0.03
SiO ₂	7.62	40.00

Table 3: Hypothetical ore mineralogy

Mineral	Limonite	Saprolite
NiO	1.55	0.00
Ni ₂ SiO ₄	0.00	2.86
CoO	0.25	0.03
Al ₂ O ₃ •H ₂ O	6.60	1.01
FeCr ₂ O ₄	4.95	0.99
Fe ₂ O ₃ •H ₂ O	45.28	0.00
Fe(OH) ₃	29.25	15.58
Ca ₃ Si ₂ O ₇ •3H ₂ O	0.12	0.31
Mg ₃ Si ₂ O ₅ (OH) ₄	0.00	41.05
Mg ₄ Si ₆ O ₁₅ (OH) ₂ •6H ₂ O	6.17	37.71
MnSiO ₃	2.79	0.33
Na ₂ SiO ₃	0.33	0.10
K ₂ SiO ₃	0.02	0.05
SiO ₂	2.68	0.00
Total	100.00	100.00

Some of the processes examined in this exercise are suited to only limonite or saprolite. For those processes, the feed was assumed to be only limonite or only saprolite. For processes suitable for processing either limonite or saprolite, and for pressure acid leaching, the models were run for three cases each, assuming the feed to be only limonite, only saprolite, or 50 percent limonite and 50 percent saprolite.

MODELLING

Commercially available process simulation software known as Aspen Plus[®] (AP) was used to build a process model (numerically rigorous mass/energy balance) for each process and the appropriate laterite feed (limonite, saprolite, or 50:50 blend) considered in this exercise.

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 and the desired process equipment was specified for the various unit operations. The desired material of construction and the required residence time was either specified or left as a default for each item of process equipment. The results obtained were checked against published information in a separate validation exercise.

The APEA software uses volumetric flow information from the imported balance and the specified or default residence time to size each item of process equipment. The required thickness of the material specified for pressure vessels is calculated from the pressure and temperature, using pressure vessel codes. If a material is not suitable the software does not calculate a cost for that item. Where materials are suitable, the software extracts the quantities of material and manpower needed to fabricate and install the item concerned from a database of process equipment. Using those numbers, it calculates the fabrication and installation cost from a database of material and manpower costs. These databases were developed from a large number of real projects and are updated regularly. The APEA software uses default allowances for process piping and valves, instrumentation, wiring, etc. It calculates the weight of each item of process equipment, empty and filled with water, and uses a set of civil engineering rules to calculate quantities and costs for structural steel and for concrete foundations. The software also assigns operating and supervision manpower to the circuit and calculates the manpower cost using labour rates from its database.

Solid-liquid separation is important in hydrometallurgical circuits and the relevant equipment cannot be accurately sized without experimental solid-liquid separation numbers, which are not available for this exercise because the feed is hypothetical. The assumption used to work around this is that all the hydrometallurgical processes have similar solid-liquid separation characteristics. This assumption might be somewhat unfair to the Neomet and DNi processes. Neomet does not use thickeners at all because it produces solids with excellent filtration characteristics. DNi does have a thickener train but since it also leaches in very strong acid it might well be that the leach residue, which is washed in a thickener train, settles better than the residue produced in the sulphate based processes. The assumption that all the settling and filtration characteristics are the same was used because the relevant numbers were not available for this exercise.

The settling rate used for all the thickeners in the various processes is 100 kg/h per square meter of thickener cross-section. The filtration steps were sized from the flow of dry solids to filtration, assuming the default filtration rate (200 kg/h/m²) in the equipment database. For cost estimation the filters were all assumed to be rotary drum filters.

The chemistry associated with the processes examined is given in Part 2 of this series⁽²⁾, and only the descriptions are repeated here, expanded to include materials of construction and the equipment sizing parameters not taken automatically from the models.

The sulphate based technologies all include a sulphur burning acid plant generating sulphuric acid and steam. This is a change from what was presented in Part 2, in which purchased acid and steam were assumed.

Pressure Acid Leaching (PAL)

Figure 1 illustrates the process model used to represent the PAL 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.

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 to 250°C. 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.

feed pumps, the autoclave itself and the three depressurisation stages (Figure 1 shows only a single train for simplicity). The autoclave shells were assumed to be titanium-lined steel. Each autoclave was assumed to have six agitated compartments and each agitator was assumed to be made from Monel alloy (the equipment database used does not have titanium for agitators, hence Monel as a cost proxy).

The counter-current decantation train was assumed to be six thickeners, with rubber lined steel as the material of construction. The re-dissolution and precipitation stages were each assumed to require an overall residence time of one hour, in a train of three agitated tanks in series made of rubber-lined steel for the steps between the autoclave and the first precipitation of nickel-cobalt hydroxide and of carbon steel for the second precipitation of nickel-cobalt hydroxide (recycled to the re-leach step) and the final precipitation of magnesium hydroxide, the last two precipitation stages being at pH values at which carbon steel is a suitable material of construction. The thickeners following each precipitation stage were assumed to require the same material of construction as the precipitation tanks. The filter on which the mixed nickel-cobalt hydroxide is filtered and washed was assumed to be a rotary drum filter, made of epoxy lined carbon steel.

The limestone and lime used in the precipitation steps were assumed to be slurried (and slaked, in the case of lime) in two agitated tanks in series, each having a residence time of 15 minutes. The slurries of limestone and lime were assumed to be moved to the relevant precipitation stages using centrifugal pumps made of carbon steel.

The re-dissolution stage and the precipitation stages were assumed to require pumping to transfer the slurry to the following stage. Two centrifugal pumps, one in service and one on standby, were specified for each stage. Where the material of construction for the agitated tanks was set to rubber-lined steel, the material for the pumps was set to stainless steel. Where the agitated tanks were specified as carbon steel, the pumps were specified as carbon steel.

Enhanced Pressure Acid Leaching (EPAL)

The process model used to represent EPAL is illustrated in Figure 2. For this model the feed laterite was assumed to be half limonite and half saprolite, mined selectively, the limonite going to the autoclaves and the saprolite to the atmospheric pressure leach.

The limonite is treated as in the PAL model. The saprolite is mixed with water and pre-leached at atmospheric pressure with acid from the acid plant. The pre-leached saprolite slurry and the pressure leached limonite slurry are combined at the step in which recycled hydroxides are used to partially neutralize the residual acid from leaching, and the nickel-bearing magnesium silicate minerals are dissolved. From this stage on, the circuit is the same as that of the PAL model. The materials of construction and the residence times are also as in the PAL model. The residence time in the saprolite pre-leach is four hours and in the re-dissolution stage is also four hours. These two stages each have four, brick lined steel, agitated tanks in series. Because it processes less limonite, this circuit has two autoclave trains instead of the three in the PAL circuit.

Atmospheric Tank Leaching (ATL)

Figure 3 illustrates the process model used to represent the ATL circuit. In this model, the feed laterite is assumed to be all saprolite. The incoming saprolite is mixed with water and leached in agitated tanks with sulphuric acid from the acid plant. The leach is heated by addition of steam from the acid plant. The leached saprolite is contacted with recycled hydroxides to partially neutralize the residual acid from the leach. From there on, the circuit is the same as in the preceding two models.

The atmospheric leach was assumed to require a residence time of four hours, in a train of four agitated tanks in series, made of brick lined steel. The rest of the equipment is the same as for the PAL circuit.

Heap leaching (HL)

The process model used to represent heap leach (HL) technology is illustrated in Figure . As for the ATL model, the HL model assumes that the feed is all saprolite. The heap leach is done in three stages, the leached (third stage) part of the heap being washed with water. The solution from the washed part of the heap is supplemented with sulphuric acid from the acid plant and used to fully leach the laterite in the second stage of the heap. The solution from this part of the heap is then

passed over fresh laterite in the newest part of the heap to maximise the concentration of nickel and minimise that of free sulphuric acid. The solution from this step is contacted with recycled hydroxides to partially neutralize the remaining free acid. From this point onwards the circuit and the process equipment are the same as for the preceding models.

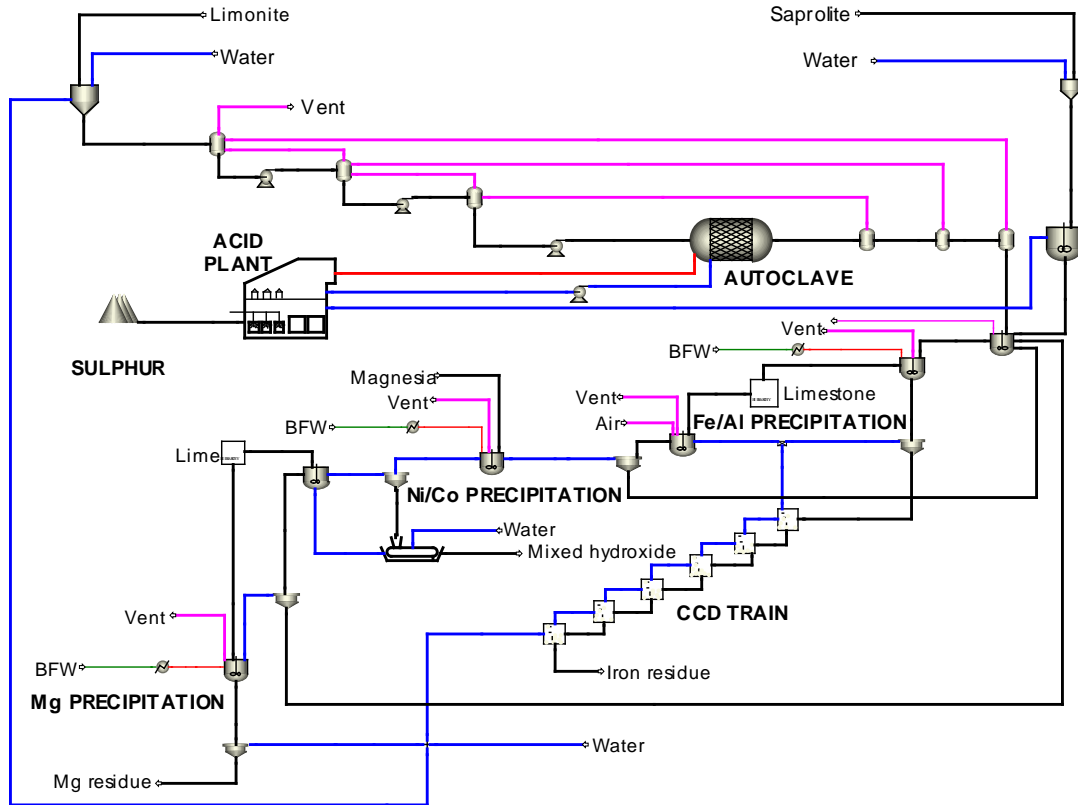


Figure 2: EPAL model

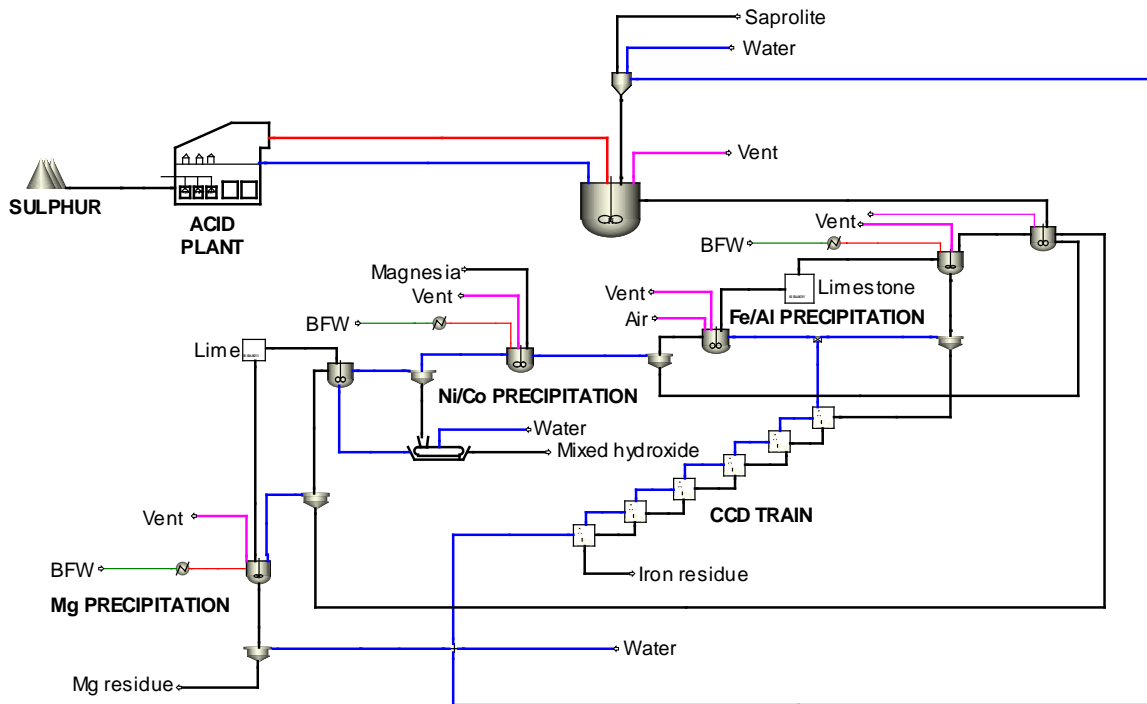


Figure 3: ATL model

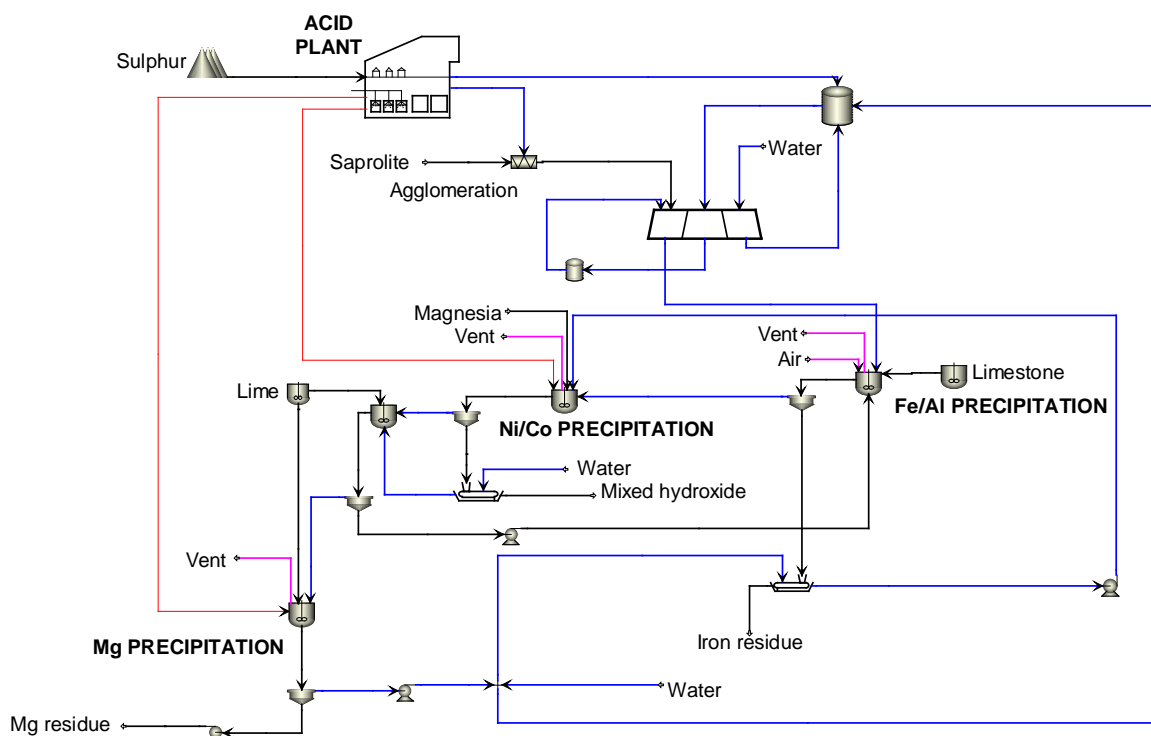


Figure 4: HL model

The total leaching time for a single heap was assumed to be 540 days and the height of the heap was assumed to be six meters. Those assumptions gave a heap 660 meters square.

The assumption used for calculating an estimate of the capital cost for the HL option is that the heaps are built, leached, washed and then left in place as the tailings. The heaps are built using conveyors to move and stack the ore onto a layer of clay sandwiched between two geo-polymer linings, with HDPE drainage piping spaced at 3 meter intervals for collecting the solution from the bottom of the heap. Heap irrigation is provided by a grid of HDPE piping on top of the heap, at two meter spacing.

Caron

The process model representing the Caron process is illustrated in Figure 5. The feed is limonite. The incoming limonite is dried using the sensible heat in the gas from the next step, in which the dried limonite is heated to about 800°C under a reducing atmosphere. This reduces most of the nickel and much of the cobalt to a metallic state, along with part of the iron. The hot gas leaving the roaster contains carbon monoxide. Air is added to burn that to carbon dioxide, then the hot gas is used as the heat source for drying the incoming limonite.

Drying of the incoming laterite is done using direct contact rotary driers made of carbon steel. Five of the largest such driers in the equipment database are required for the total drying duty. The reductive roasting stage is done in rotary kilns, which are approximated by ten rotary driers (the largest in the equipment database) each with a refractory lining. The driers and the rotary kilns use natural gas as fuel (and as reductant in the roasting stage). Cooling of the hot reduced calcine is done in five units similar to the rotary driers, cooled externally with water.

The cooled calcine is mixed with recycled solution of aqueous ammonia and ammonium carbonate and leached with that solution and air, causing the metallic nickel and cobalt to dissolve as ammine complexes. The metallic iron is converted into insoluble ferric oxide. The leach is done in a cascade of four agitated tanks in series, the material of construction being carbon steel and the total residence time eighty minutes⁽³⁾.

The leached slurry is thickened and washed with recycled barren solution in a six-stage counter-current decantation train. The thickeners are made of carbon steel.

The supernatant from the counter-current decantation train is contacted with hydrogen sulphide in a cascade of three agitated tanks (rubber lined steel, one hour total residence time), precipitating the

The heat exchangers and columns in the Caron circuit were sized by the process modelling software, assuming 70 percent efficiency per stage in the columns and default heat transfer correlations for the heat exchangers. The heating utility is natural gas and the cooling utility is water.

Neomet

Figure 5 illustrates the process model of the Neomet circuit. This is developing technology that has evolved since it was described in Part 2 of this series. It begins with mixing the incoming laterite with recycled wash filtrates and water (agitated tank, brick lined steel, 10 minute residence time) and pumping the slurry to a leaching train of four agitated tanks in series (brick lined steel, two hours total residence time) where it is leached with strong hydrochloric acid that is sparged into the laterite slurry as a vapour containing steam and hydrogen chloride. The latent heat in the incoming acid drives the leach temperature up to about 135°C and the incoming steam/HCl does not fully condense but a greater proportion of the HCl does, because it is consumed in the leach reactions. The uncondensed steam/HCl passes to a scrubber column in which the acid is captured by partial condensation of the steam and returned to the leach. The remaining steam is split, part going to the hydrolysis sequence and the balance to an energy recovery step which is described further on. The scrubber has ten theoretical stages. The column and internals are made from Teflon lined steel. The condenser has stainless steel tubes and a carbon steel shell (cooling water on the shell side) and the reflux drum and pump are of stainless steel.

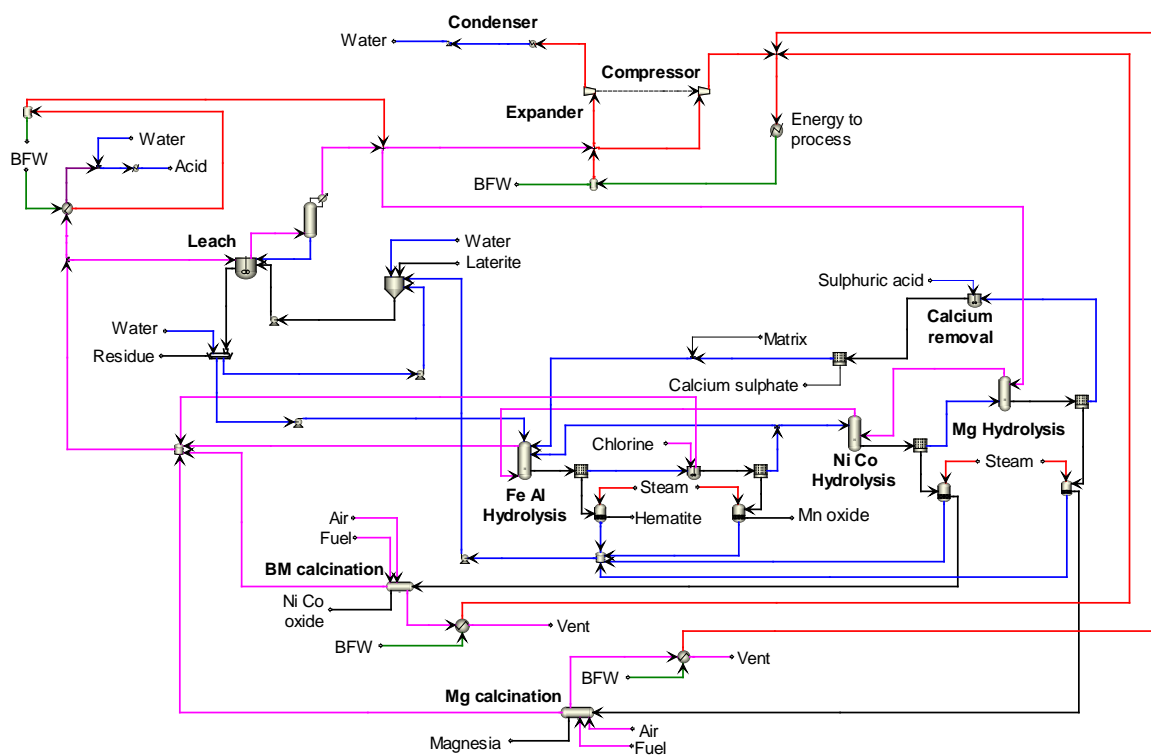


Figure 5: Neomet model

Adding the acid to the leach as a vapour is a change since the last time this process was presented. Doing it this way eliminates the need for an evaporation stage between the leach and the following iron hydrolysis section. It also eliminates the need for cooling in the leach, which greatly simplifies the leach reactor train. This method of acid addition was successfully demonstrated in the Magnola plant in Canada.

The leached slurry is filtered and washed with water (rotary drum filter, epoxy covered steel for costing purposes) and the washed filter cake leaves the circuit. The wash filtrate is recycled to the feed mixing tank and the primary filtrate is sent to the iron hydrolysis section, where it is mixed with a circulating solvent matrix, contacted with steam from the following step and heated to 180°C by indirect heat exchange with condensing high pressure steam. The ferric, aluminium and chromium chlorides react with the steam, forming solid ferric, aluminium and chromium oxides (referred to collectively as hematite because that is the major oxide) and gaseous hydrogen chloride. The resulting slurry of hematite in the molten matrix is filtered and washed using a proprietary filter that

uses steam to wash the filter cake. The washed filter cake leaves the circuit. The wash filtrates are combined and recycled to the leach section. A “true” capital cost for the proprietary filter not being available, a rotary drum filter (epoxy lined steel) was used as a capital cost proxy.

The amount of atmospheric-pressure steam going through the hydrolysis sequence is manipulated to give 35 mass percent HCl in the steam/acid leaving the iron/aluminium hydrolysis step.

The primary filtrate from the iron hydrolysis step is contacted with gaseous chlorine to oxidize and precipitate manganese as manganese dioxide, which is recovered and washed in the same way as the hematite (cost proxy also the same), then leaves the circuit. This addition of chlorine also acts as chloride make-up in the circuit. The filtrate is split, part returning to the iron hydrolysis stage to build up the level of base metals, and a bleed proceeding to the nickel/cobalt hydrolysis step, where it is contacted again with steam from the subsequent (hotter) magnesium hydrolysis step, which heats it enough to convert the nickel and cobalt to solid basic hydroxychlorides, releasing the associated chloride ions as gaseous hydrogen chloride. The nickel and cobalt hydroxychlorides are filtered out and washed in the same way as the hematite (same cost proxy again). The filtrate is heated further by indirect heat exchange with condensing high pressure steam and contacted with steam from the scrubber in the leach section, causing the magnesium chloride to decompose to gaseous hydrochloric acid and solid magnesium hydroxychloride that is recovered by filtration and washed in the same way (and with the same cost proxy) as the hematite.

The Neomet process is under development and the design of the full-scale hydrolysis reactors has not yet been published. As the chemistry is currently understood, the rate of the hydrolysis reactions is controlled by the rate at which the energy required to drive the endothermic hydrolysis reactions can be transferred into the reacting mixture. Although this is unlikely to be the actual design, one way in which the hydrolysis reactors could be configured would be an agitated tank in which steam is mixed with the incoming liquid (molten salt containing the inert matrix, the reactant chlorides and some water), with a substantial pump-around loop through a heat exchanger condensing high pressure steam on the other side. This configuration was assumed for the iron and the magnesium hydrolysis reactors. The nickel/cobalt hydrolysis, according to the energy balance, does not need the pump-around loop and heat exchanger, just the agitated tank. The agitated tanks were assumed to be brick lined steel. The heat exchangers were assumed to need Hastelloy⁽⁴⁾ tubes and carbon steel shells (high pressure steam on the shell side). The pump was taken for costing purposes as a centrifugal pump made from Inconel which is similar in cost to Hastelloy (the equipment database has Inconel but not Hastelloy pumps). A more cost effective material might be silicon carbide, but equipment costs for that material were not available for this exercise.

The base metal hydroxychlorides are calcined to a mixed nickel/cobalt oxide that leaves the circuit as the main product. The magnesium hydroxychloride is calcined to magnesia that leaves the circuit as a by-product. In these steps the chloride is released as gaseous hydrochloric acid that is recycled to the leach along with the gaseous steam and hydrogen chloride from the hydrolysis section. Calcination of the base metal and magnesium hydroxychlorides was assumed to require externally heated kilns, with the kiln tubes made from Hastelloy and the energy coming from the combustion of natural gas. The hot gas leaving the heating side of each kiln is passed through a carbon steel heat exchanger in which high pressure steam (40 bar, 250°C saturation temperature) is raised from boiler feed water, for use as part of the heat source in the hydrolysis section.

Some of the steam leaving the scrubber in the leach section goes to the hydrolysis train and the balance is split, part being expanded through a condensing turbine that drives a compressor in which the other part of the steam is compressed to 40 bar and combined with the steam from the base metal and magnesium calcination section for use as the energy source in the hydrolysis section. The ratio of steam expanded to steam compressed is adjusted to give the required pressure in the compressed steam. The model assumes a gas-fired steam boiler, producing 40 bar steam, to supply the balance of the energy needed in the hydrolysis sequence.

Direct Nickel (DNi)

The DNi process has also evolved since Part 2 of this series. Figure 6 illustrates the process model used to examine the DNi process, as that model presently stands. Except where stated otherwise, the material of construction throughout the DNi process is stainless steel (type 304).

The incoming laterite is leached adiabatically with strong nitric acid, in four agitated tanks in series (two hours total residence time). The heat of reaction causes vapour to be released that is scrubbed with water in a scrubber operating at atmospheric pressure (Scrubber 2 in Figure 7). This

scrubber contains twenty theoretical stages. The leached residue is separated and washed with water in a counter-current decantation train of six thickeners. The washed residue leaves the circuit.

The supernatant from the counter-current decantation train is passed through a pressure reduction step (to 0.05 bar absolute, using a valve) and concentrated by evaporation in three heat exchangers. The first heat exchanger cools and partially condenses steam and NO_x from the decomposition of magnesium nitrate, the second condenses steam from the above mentioned scrubber and the third exchanger condenses steam/HNO₃ from the iron hydrolysis stage. The resulting gas-liquid mixture is separated in a flash drum and the concentrated liquid is pumped back to atmospheric pressure and heated to 140°C prior to entering the iron hydrolysis stage. The vapour goes to a scrubber (Scrubber 1 in Figure 7) where it is condensed with cooling water and the acid in the vapour is captured in the scrubber bottoms, pumped back to atmospheric pressure and sent to the acid regeneration section. The water not reporting to the scrubber bottoms leaves the scrubber as a condensed overhead product. It is pumped back to atmospheric pressure and used elsewhere in the circuit (MgO slaking, filter wash, CCD wash).

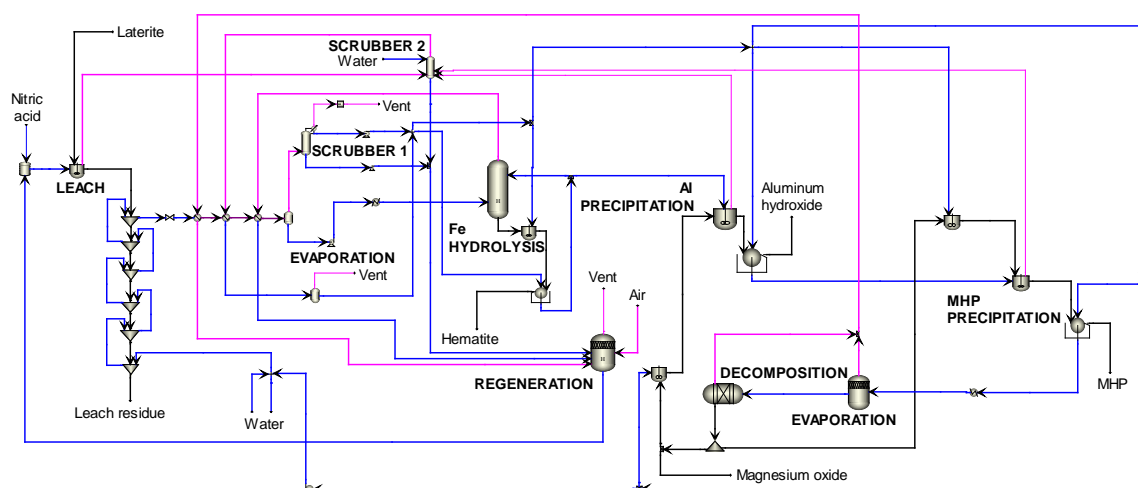


Figure 6: Direct Nickel model

The hot concentrated solution is heated to 165°C, causing the ferric nitrate to hydrolyse to solid hematite and gaseous nitric acid, also evolving some steam. The equipment assumed for this step is similar to that assumed for the corresponding step in the Neomet circuit, i.e. an agitated tank and a pump-around through a heat exchanger, the material of construction being stainless steel (304). As for the Neomet circuit, this configuration is probably not the final design, it is merely a proxy used for estimating a capital cost. The underlying assumption is that heat transfer controls the rate of the hydrolysis reaction and thus also the size and cost of the associated process equipment.

The steam and gaseous nitric acid are sent, via the heat exchanger in the evaporation section, to the acid regeneration stage. The slurry from the hydrolysis reactor is quenched into water in an agitated tank (ten minute retention) and the hematite is recovered by filtration and washed. A rotary drum filter (stainless steel) was assumed as a costing proxy. The washed hematite leaves the circuit.

Part of the combined filtrate returns to the iron hydrolysis reactor to control the solids content of the slurry in that reactor and the balance is neutralized with recycled magnesium oxide to precipitate aluminium hydroxide. The equipment assumed is three agitated tanks in series (one hour total retention). The magnesium oxide is slaked with water in two agitated tanks in series (carbon steel, 30 minute total retention) before being used to precipitate the aluminium. The precipitated slurry is filtered and washed with water, assuming a rotary drum filter as a proxy for calculating a capital cost.

The filtrate from the aluminium precipitation step is contacted with more slaked magnesium oxide to precipitate the base metals as a mixed nickel/cobalt hydroxide that is filtered out, washed and leaves the circuit as the product. The process equipment is three agitated tanks in series (one hour total retention) for the precipitation and two carbon steel agitated tanks in series (30 minute total retention) for slaking the MgO. The base metal precipitate is filtered and washed, and the washed filter cake leaves the circuit as the mixed hydroxide product. As before, a stainless steel rotary drum filter was assumed for capital cost calculation.

The filtrate from the base metal recovery stage, essentially a solution of magnesium nitrate, is again concentrated by evaporation, such that it becomes a molten salt hydrate, effectively magnesium nitrate di-hydrate. The equipment selected for this step is a process furnace fired with natural gas, as commonly used in the petrochemical industry, with a flash vessel to separate the vapour generated from the liquid phase. The molten salt hydrate is heated to 500°C, causing it to decompose to solid magnesium oxide, steam and gaseous O₂, NO and NO₂. That process equipment was assumed to be a screw feeder combined with a process furnace fired with natural gas. Part of the magnesium oxide is slaked with water and recycled to the two precipitation steps and the excess leaves the circuit as a by-product.

The NO/NO₂ from the decomposition step is combined with the steam from the preceding evaporation step and partially condensed in the evaporation section between the leach and the iron hydrolysis stage. The partially condensed mixture is routed to the acid regeneration section where the NO_x is converted back to nitric acid by a patented method in which NO reacts with HNO₃ in the presence of dissolved trivalent nitrogen, then the solution is oxidized to nitric acid with air. For cost calculations an agitated tank (15 minute retention) was assumed for contacting the NO/NO₂, air and aqueous acid, coupled with a vertical process vessel (10 minute retention) to represent the absorber capturing the gaseous NO/NO₂ into the acid, and heat removal using a cooling coil in the agitated tank. The regenerated acid is recycled to the leach.

The energy recycled by heat exchange does not fully close the overall energy balance. A high pressure steam boiler, fired with natural gas, was assumed for generating the steam needed to close the overall energy balance.

Pyrometallurgy

The rotary kiln, electric furnace (RKEF) smelting route is established technology for producing ferronickel from saprolite. The ore is partially dried, then dehydrated and reduced before being smelted to produce ferronickel and slag. The incoming saprolite is dried using hot gas from the pre-reduction kiln, then mixed with coal char and heated to 800°C, and the hot pre-reduced mixture is smelted in an electric furnace. The nickel and much of the iron form a ferronickel alloy that is tapped and recovered as the desired product. The other constituents are rejected to a slag phase.

Sintering and submerged-arc smelting (SAF) is an established iron making technology that has been applied to making nickel pig iron. In the context of process modelling, it is similar to the RKEF technology. The incoming laterite is dried and then mixed with coal char and flux (CaO in this exercise) and sintered at 1200°C, the heat coming from the combustion of natural gas. The sintered mixture is then smelted in an electric arc furnace.

Sintering and smelting in a blast furnace (BF) is another iron making technology that has been used to make nickel pig iron. It is similar to sintering and smelting in a submerged arc furnace, except that the smelting energy comes from the combustion of excess coal char in the blast furnace. The excess char required by the blast furnace also results in the reduction of more of the iron, making for a product that contains a lower percentage of nickel.

The database used for equipment cost estimation contains process equipment for the oil and gas and the chemical process industries (which is the type of equipment used in hydrometallurgy), but not for pyrometallurgical equipment such as electric or blast furnaces. An attempt was made to approximate these items as composites of steel process vessels, refractory linings, electrical transformers, etc., but the resulting capital cost estimate was about half what would be expected from published capital costs for other RKEF projects, and consequently the approach used for the hydrometallurgical circuits was not used for estimating capital costs for the pyrometallurgical circuits.

Fortunately for this exercise, some studies have been published in sufficient detail for a correlation of the capital cost versus laterite throughput to be possible for the RKEF process equipment. Figure 8 shows capital costs for process equipment extracted from published studies on the Araguaia⁽⁵⁾, Mayaniquel⁽⁶⁾ and Onça Puma⁽⁷⁾ nickel projects. These are not the full capital costs published for these projects; those costs include mining and infrastructure that were deliberately excluded from the capital costs calculated for the various processes examined in this work. The values in Figure 8 have been converted from the values extracted directly from the various published studies to 2014 values using the Chemical Engineering Plant Cost Index (CEPCI).

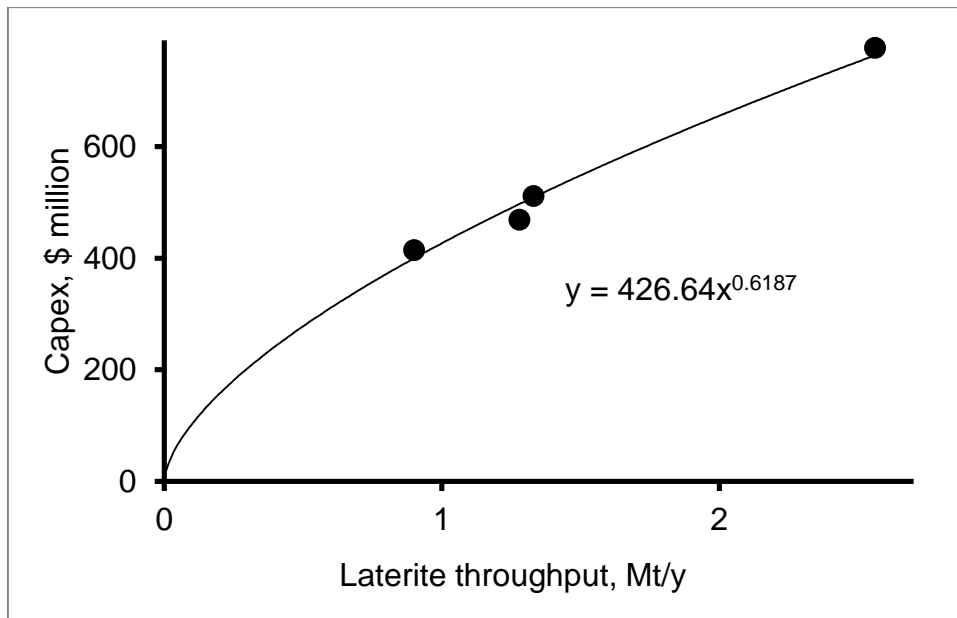


Figure 7: RKEF capital costs

According to a report published in May 2010 by ETSAP (Energy Technology Systems Analysis Programme)⁽⁹⁾ the capital cost of blast furnaces for the manufacture of pig iron is \$148-275 per annual ton of hot metal and the variable cost for operation and maintenance is around \$90 per ton of hot metal. For electric arc furnaces, the investment cost is about \$80 per annual ton of steel capacity. According to a news item in Asia Miner⁽¹⁰⁾, ANTAM recently agreed to build a nickel pig iron plant to produce 300 thousand tonnes of nickel pig iron per year, the estimated capital cost being US\$280 million, which equates to US\$933 per ton of hot metal. According to another presentation⁽¹¹⁾, the capital cost of blast furnace technology for making nickel pig iron is US\$10-15 per annual pound, presumably of contained nickel. (If the NPI contains 2% nickel, \$148-275 per annual ton of hot metal translates to roughly \$4-7 per annual pound of contained nickel.) The same presentation gives the capital cost for HPAL technology as US\$30-40 per annual pound, which is appreciably higher than the capital cost calculated for the PAL process in this work.

A study by Tata Steel Consulting⁽¹²⁾ on an iron project contains data on the capital costs for the various parts of a blast furnace operation (coke production, sinter production, blast furnace). The correlations from this study are listed in Table 4. The capacity, x , in these correlations refers to the mass throughput for the relevant plant.

Table 4: Capital costs for blast furnace processing of iron ore⁽¹²⁾

Plant	Capital cost, M\$ (2011 basis)
Coke production	$473.6x^{0.938}$
Sinter production	$45.936x^{0.811}$
Blast furnace	$196.128x^{0.922}$

x = capacity, Mt/year

CALIBRATION

The method used for estimating the capital costs of hydrometallurgical processes is explained in more detail elsewhere⁽¹³⁾, but further validation is thought to be desirable. The Pueblo Viejo Project, exploiting of a refractory gold deposit in the Dominican Republic, was evaluated by Roscoe Postle Associates Inc., for Barrick Gold Corporation. The results, as documented in a published NI 43-101 technical report⁽¹⁴⁾, were used for a verification exercise. The ore is refractory, with gold and silver occurring as submicron particles and solid solution in pyrite. Processing it entails comminution, pressure oxidation (POX - 230°C, 75 minutes), a hot cure to dissolve basic ferric sulphate, a three-stage counter-current decantation sequence to separate the acidic solution from the oxidised solids which then undergo a two stage lime boil treatment to destroy silver jarosite. The resulting slurry is cooled and the gold is recovered by cyanidation and carbon in leach (CIL). The process is described in detail in the NI 43-101 report⁽¹⁴⁾.

The AP software was used to build a process model representing the circuit as reported, the resulting balance was exported into the APEA software and a capital cost estimate was generated in the same way as was done for the laterite circuits examined in this work. The Pueblo Viejo project includes two oxygen plants (2850 and 1100 tonnes/day), as well as a separate lime plant in which limestone is crushed, some is milled and used directly and the rest is calcined, milled and slaked. The lime plant was modelled separately and a capital cost estimate was calculated. The capital cost of the oxygen plants was interpolated from published capital costs of conventional oxygen plants⁽¹⁵⁾. Figure 9 shows these costs in 2008 dollars. The Chemical Engineering Plant Cost Index (CEPCI) for 2008 is 575.4 and for 2014 it is 576.1, so the 2008 costs shown are essentially also the 2014 costs.

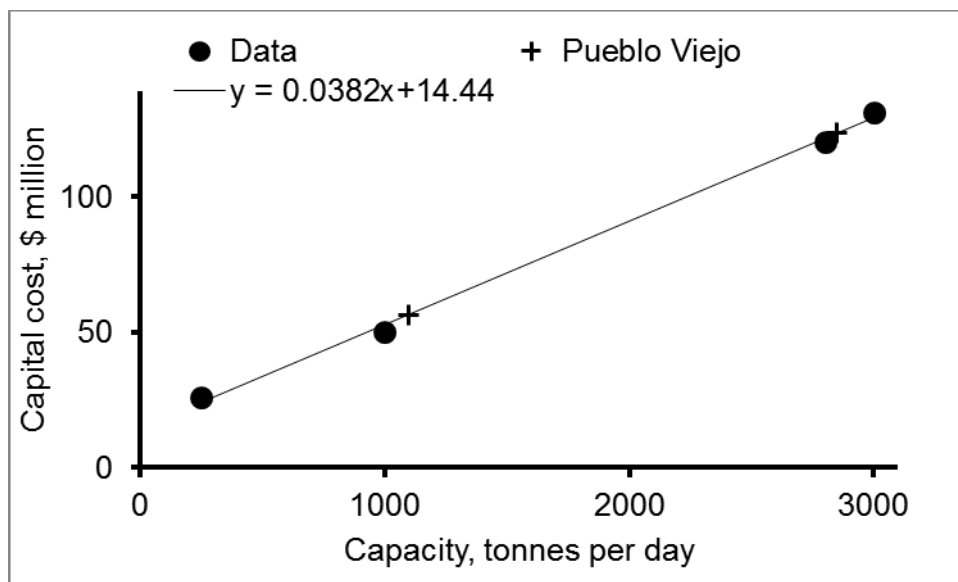


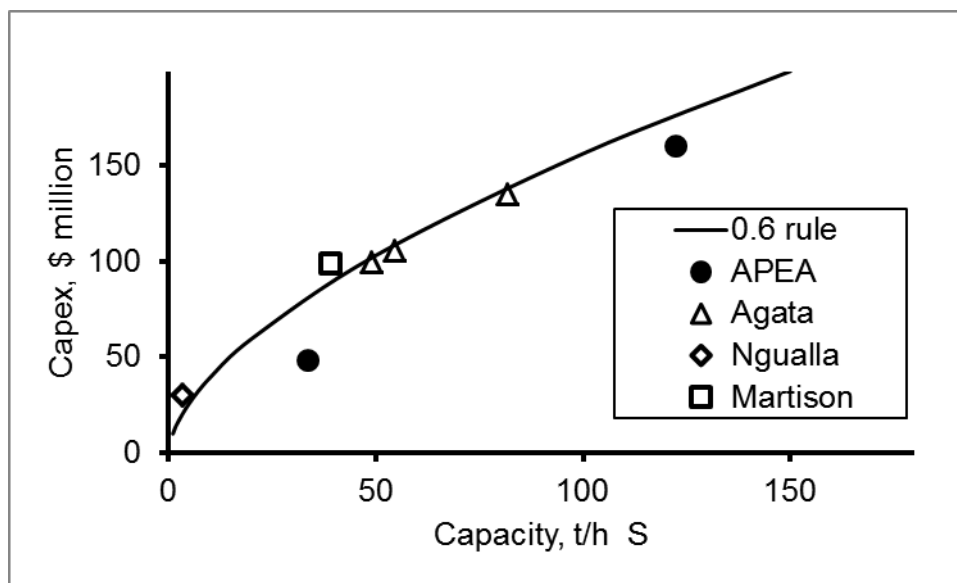
Figure 9: Capital costs of conventional air separation plants

Table 5 shows the results of this validation exercise – the total direct field calculated by the APEA software and the installed equipment cost from the NI 43-101 report differ by about five percent. The NI 43-101 report does not give a breakdown of the capital costs for the process equipment, and it lists, for example, owner's indirect costs and other indirect costs that do not appear to be part of the \$926 million given for the process capital cost. The APEA software normally calculates direct field costs, indirect field costs and non-field costs and includes these components in its tally for the cost of the process equipment, but in this case the indirect and non-field costs were not included in the APEA number called the total installed equipment cost. While this does make for a little uncertainty in the comparison, the numbers indicate that the method used in this work is satisfactory for preliminary estimates.

Table 5: Calibration results

Component	Cost, \$ million
Main process circuit, direct field cost for installed equipment	502
Main process circuit, indirect field costs and non-field costs	158
Lime plant, direct field cost for installed equipment	145
Lime plant, indirect field costs and non-field costs	38
Oxygen plants, capital cost interpolated from published costs	180
Total direct field cost calculated using APEA	972
Total installed equipment cost calculated using APEA	1 059
Installed equipment cost from the NI 43-101 report	926

As further calibration, the costs of sulphuric acid plants burning elemental sulphur were extracted from published information on other projects^{16,17,18}. These costs are shown in Figure 10, along with the capital costs calculated by the APEA software for the acid plants for the PAL circuits processing limonite and saprolite. In this case the calculated capital cost for the larger of the two acid plants in the PAL model (processing the saprolite) lies quite close to the “0.6 rule” shown by the line passing through the data points from the literature. The capital cost calculated for the smaller acid plant (in the PAL model processing limonite) is lower than predicted by the “0.6 rule” but the difference is less than the contingency allowance in the capital cost estimate for that PAL circuit.

**Figure 10: Capital costs for sulphuric acid plants**

RESULTS AND DISCUSSION

There are, of necessity because otherwise this exercise could not be done, several rather gross assumptions on the equipment, sizing and materials of construction in the capital cost estimates. These assumptions are thought to be realistic (or at least plausible in the context of cost estimation), but even so the results presented here should not be taken as anything approaching bankable. Small differences in the numbers generated are most probably not significant – the idea was to find and examine the larger differences. Costs for items common to all the processes, such as mining and infrastructure, which would normally be included in a capital cost estimate and economic evaluation for a real project, were not calculated for each process because this exercise is about comparing the various processes and not about the absolute capital costs for any specific project.

The process models were set up assuming thirty thousand tonnes per year of nickel in the feed laterite. The different processes have different nickel recoveries and use different parts of the orebody, i.e. limonite and/or saprolite. This makes the amounts of laterite consumed per unit of nickel produced vary slightly. To allow for that the comparison includes the variable portion of the mining cost of the ore, which was previously estimated⁽¹⁹⁾ as \$5 per tonne of ore mined.

Comparing the variable cost alone, as presented in Part 2 of this series, may be a reasonable way of ranking the different processes, but that does not give the full picture. The capital costs and the differences between revenue and operating costs are needed for a more complete analysis. This is a challenge because the various prices involved are not constant. Picking single numbers for the prices of nickel, cobalt and the major inputs to the various processes would be somewhat less than realistic when the venture does not yet exist and would be intended to operate for two decades or longer. One way of tackling this challenge is to see whether or not the various commodity prices can be reduced to sensible ranges. Inflation data⁽²⁰⁾ and historical price data⁽²¹⁾ were used to calculate the various commodity prices shown in Figure 11, over the period 1914 through 2012, in 2014 dollars. The upper dashed line in each graph shows the average price over the entire period plus one standard deviation, and lower dashed line shows the average price minus one standard deviation.

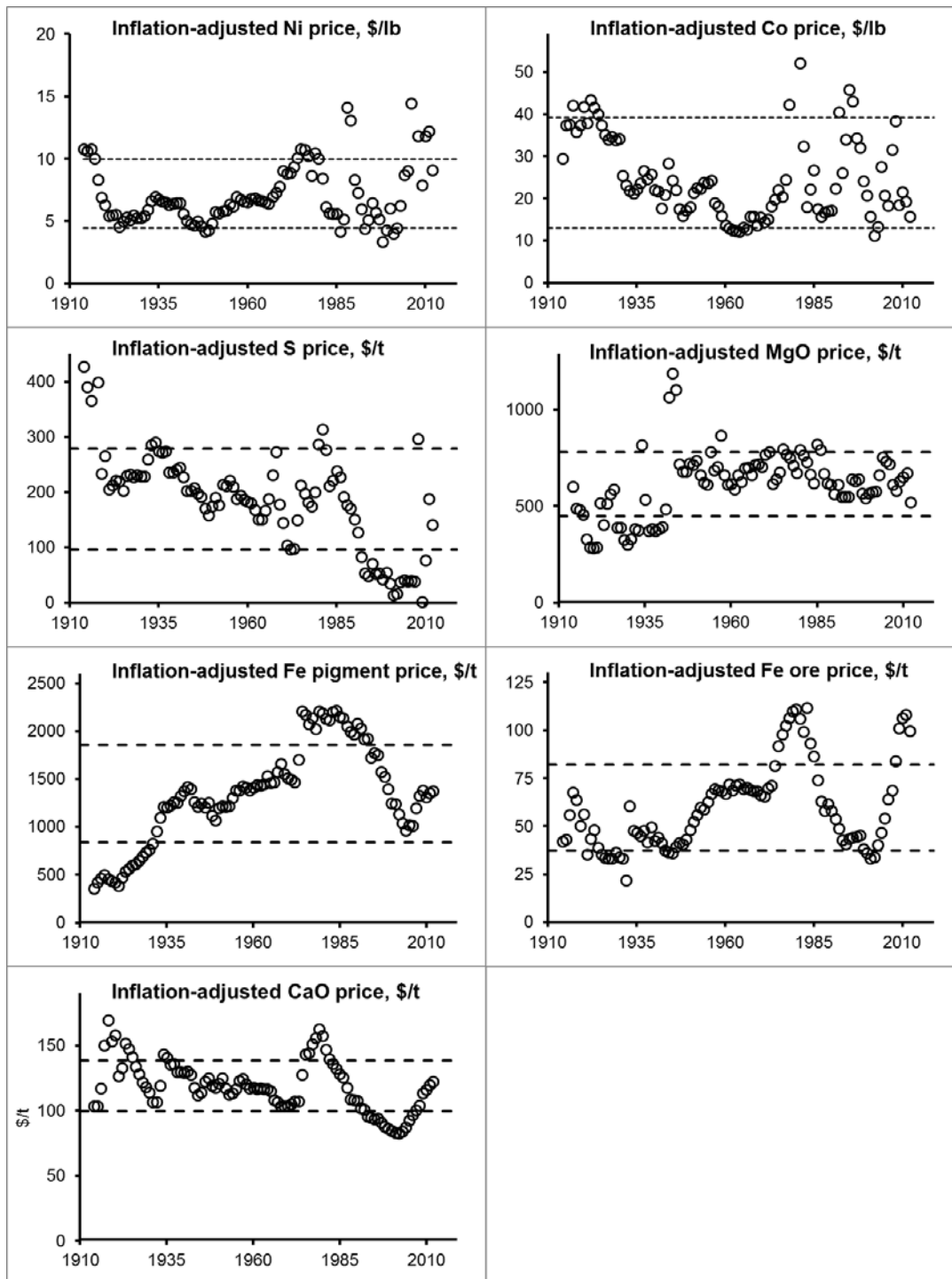


Figure 11: Inflation-adjusted commodity prices (2014 dollars)

For most of that period, the price of nickel, adjusted for inflation, was between \$4.4/lb and \$10/lb, with excursions to the high side between 1974 and 2011. The price of cobalt ranged between \$12.9/lb and \$39.2/lb, with excursions above that range before 1924 and again between 1978 and 1996. The price of sulphur ranged between \$96/t and \$278/t, with an excursion to the low side between 1992 and 2010. Magnesium oxide ranged between \$449/t and \$782/t since 1945. Iron ore ranged between \$37/t and \$82/t for most of the period covered, with two significant excursions to the high side. Iron oxide pigment ranged between \$842/t and \$1861/t from 1931 to 1973, above that from 1974 to 1992 and inside that range since. (These prices are in US\$ per US ton, one ton being 2000 lb.)

From the data graphed in Figure 11, it would not seem unreasonable to say that each time a price went above the upper dashed line, or below the lower one, it ultimately reverted to inside the range defined as the long term average plus or minus one standard deviation. It would thus seem reasonable to use these two limits as a plausible price range for examining the economics of the

various technologies selected for this series of papers. A more conservative yet still reasonable approach would be to use the long term averages and the upper limits for calculating input costs and the long term averages and the lower limits for calculating revenues.

The other commodities of interest are fuel, reductant and electricity. A search of the internet yielded historical price data⁽²²⁾ for coking coal from 1987 to 2014, liquefied natural gas (LNG) from 1984 to 2015 and electricity⁽²⁴⁾ from 1990 to 2013. A typical coke yield from coking coal is 64 percent⁽²³⁾. Figure 12 shows the data on coke, LNG and electricity prices, adjusted to 2014 dollars. The dashed lines show the average prices plus or minus one standard deviation. While the time period spanned by the data for coke, LNG and electricity is considerably smaller than the time period in Figure 11, the general behaviour looks similar.

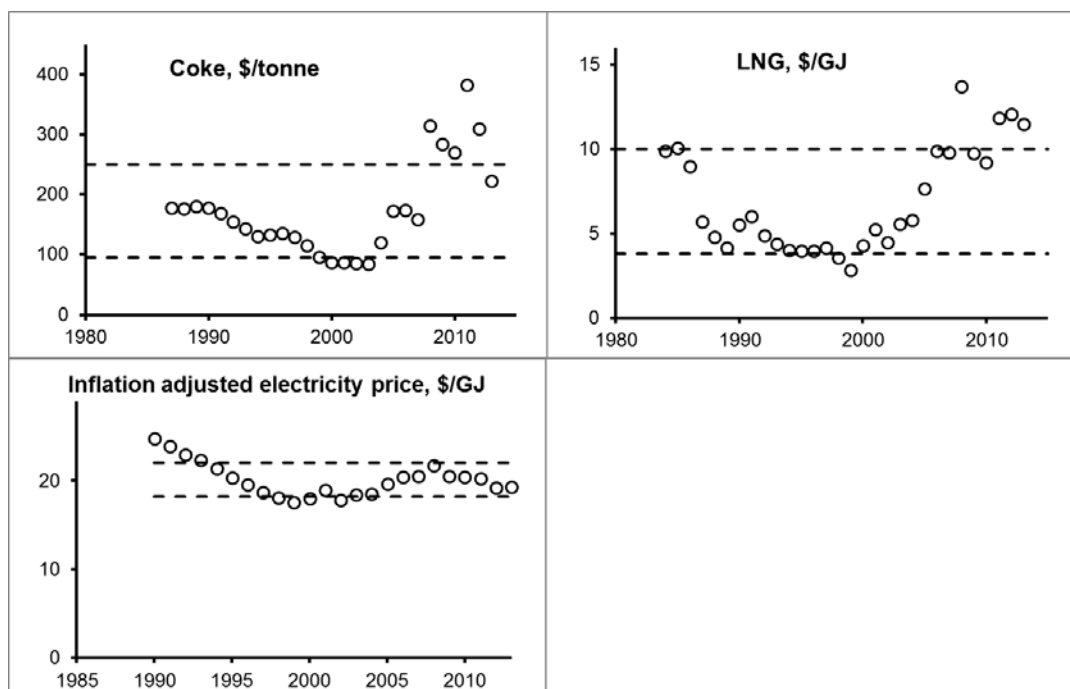


Figure 12: Coke and LNG prices

For each of the processes examined, capital costs were calculated for the installed process equipment and the tailings disposal area. Capital costs for common items like the mine and other infrastructure can be set to the same number for all the processes. While the exact number for non-process capital costs might not matter particularly because it would cancel out in comparisons between the different processes, a plausible number is better. The published studies^(5,6,7) from which the process capital cost for the RKEF technology was interpolated also gave capital costs for the total project, covering the mine and other infrastructure. Subtracting the process capital from the total capital gives the non-process capital cost data shown by the open circles in Figure 13, plotted against the laterite mining rate. The other two symbols mark the tonnages of limonite and saprolite containing 30 thousand tonnes of nickel per year, at the compositions used in this work. Interpolating the published data gives \$666 million for limonite and \$606 million for saprolite, as the capital cost of the mine and other non-process infrastructure. The number for equal amounts of limonite and saprolite would be \$638 million.

Table 6, Table 7 and Table 8 show the fixed and total variable operating costs and the capital costs calculated for the various processes and feeds, along with the nickel price required to give ten year internal rates of return (IRR) of 30, 20 and 10 percent, arranged in ascending order of nickel price. The numbers in bold map out the region that requires a nickel price that is above the long term price range. The lower the number found for the nickel price required for any given internal rate of return, the more favourable the economics for that case. For these calculations, the cobalt price was assumed to be three times the nickel price (the long term average is 3.6 and the range is 2.9 to 3.9). The various input costs were calculated using the long term averages of the commodity prices from Figure 11 and Figure 12 as well as the upper limits (average plus one standard deviation). The nickel prices in the columns headed L were calculated assuming the long term averages of the various input commodities. The prices in the columns headed H were calculated using the upper limits (average plus one standard deviation) for the input commodities. For the Neomet and DNi processes, the calculations were done assuming the long term averages and the lower limits

(average minus one standard deviation) for the by-product selling prices of magnesia and ore-grade hematite. Minor variable cost items such as make-up acid or chlorine were calculated using the reagent and utility prices given in Part 2 of this series.

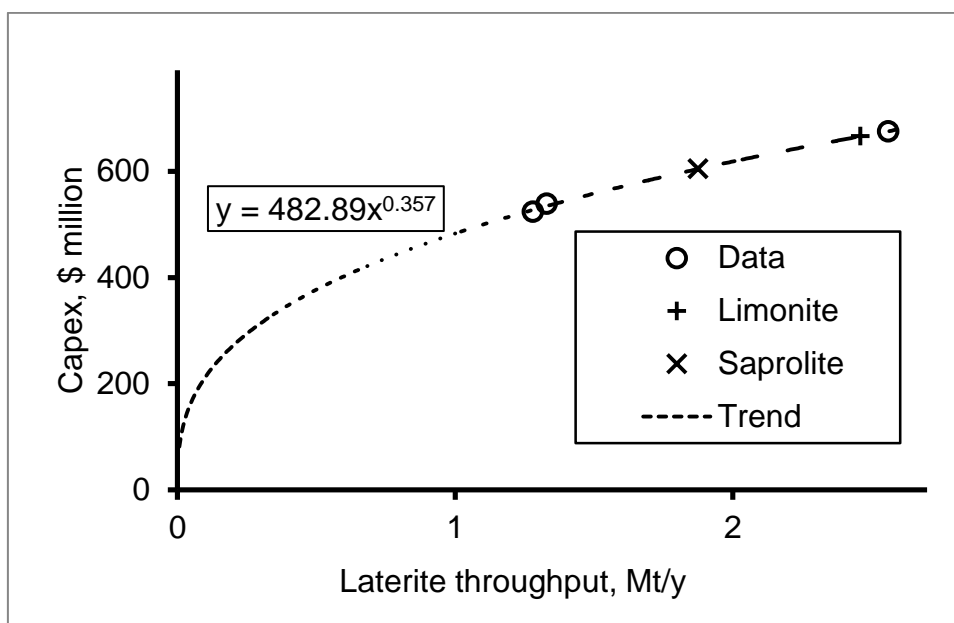


Figure 13: Non-process capital costs

The calculations for the Neomet and DNi processes were done three ways. In the variations marked (a) the by-products were assumed to be magnesia and pigment grade hematite, in the variations marked (b) the hematite was assumed to be saleable as iron ore and in the variations marked (c) no by-product revenues were assumed. The pigment-grade hematite was assumed to be saleable for half the lower limit shown for iron oxide pigment in Figure 11, i.e. \$421/t.

Table 6: Costs and required Ni price, limonite

Process	Operating costs		Capital cost \$ million	Nickel price needed, \$/lb					
	Fixed \$ million/y	Variable \$/lb Ni&Co		30% IRR		20% IRR		10% IRR	
				L	H	L	H	L	H
Neomet (a)	8.3	2.24	399+666	2.68	3.23	<0	<0	<0	<0
DNi (a)	7.8	2.49	463+666	3.60	4.26	0.28	0.96	<0	<0
PAL	6.2	2.05	479+666	9.95	10.32	7.15	7.54	4.93	5.35
Neomet (b)	8.3	2.24	399+666	10.27	11.31	7.02	8.08	4.51	5.54
DNi (b)	7.8	2.49	463+666	10.48	11.61	7.15	8.31	4.58	5.71
Neomet (c)	8.3	2.24	399+666	11.53	12.08	8.33	8.88	5.80	6.34
DNi (c)	7.8	2.49	463+666	11.84	12.44	8.54	9.17	5.95	6.58
Caron, CS	5.5	3.79	256+666	12.22	13.47	9.04	10.28	6.60	7.84
Caron, SS	6.2	3.79	323+666	15.00	16.42	10.93	12.39	7.84	9.30

(a) With by-product magnesia and pigment grade hematite

(b) With by-product magnesia and iron ore grade hematite

(c) No by-products

In the case of the processes treating limonite, the Neomet and DNi processes appear to offer similar overall economics to the established PAL process, if the by-product hematite is sold as iron ore. If the hematite is such that it can be sold for significantly more than the price of iron ore, the Neomet and the DNi process economics become distinctly superior to PAL. Excluding revenue from by-product hematite and magnesia makes the PAL process economically superior to Neomet and DNi for processing limonite. This makes it important for Neomet and DNi to demonstrate that

the hematite and magnesia are indeed by-products and not residues, and it will be important to determine whether or not the hematite can be sold for significantly more than the price of iron ore.

The Caron process was found to have the lowest capital cost of all the processes treating limonite, especially if the process equipment can be made of carbon steel instead of stainless steel. Two variations of the capital cost estimate were done for the Caron process, one assuming carbon steel and the other assuming stainless steel as the material of construction (except for the section in which cobalt sulphide is precipitated, which was assumed to require rubber lined steel). The cost calculations predict that using stainless steel would make the process equipment about 26 percent more expensive than carbon steel. The variable cost is what makes the Caron process less favourable than the PAL process, if the IRR is used as the criterion.

While not yet proven because they are both still fledgling processes, it would seem from the results of this exercise that the Neomet and DNi processes may have lower capital costs than the PAL process, for treating limonite.

Table 7: Costs and required Ni price, saprolite

Process	Operating costs		Capital cost \$ million	Nickel price needed, \$/lb					
	Fixed \$ million/y	Variable \$/lb Ni&Co		30% IRR		20% IRR		10% IRR	
				L	H	L	H	L	H
Neomet (a)	4.9	1.93	249+606	11.24	11.83	7.49	8.08	4.61	5.18
Neomet (b)	4.9	1.93	249+606	12.79	13.47	9.06	9.74	6.13	6.82
Neomet (c)	4.9	1.93	249+606	13.12	13.64	9.40	9.95	6.48	7.04
HL	12.4	3.65	221+606	13.65	14.70	10.40	11.45	7.87	8.92
DNi (a)	7.8	2.48	423+606	13.71	15.16	9.42	10.89	6.09	7.52
DNi (b)	7.8	2.48	423+606	14.44	16.25	9.76	11.57	6.13	7.89
DNi (c)	7.8	2.48	423+606	15.59	16.49	11.36	12.26	8.02	8.91
ATL	4.4	3.48	397+606	15.85	16.87	11.78	12.80	8.60	9.63
RKEF	12.6	4.8	629+606	16.47	17.19	12.26	12.95	8.97	9.65
Sinter-BF	17.4	7.47	145+606	16.72	20.41	13.98	17.67	11.86	15.54
Sinter-SAF	12.3	5.95	614+606	17.85	18.97	13.45	14.56	10.02	11.29
PAL	8.7	7.95	991+606	27.60	29.93	21.10	23.41	16.03	18.34

- (a) With by-product magnesia and pigment grade hematite
- (b) With by-product magnesia and iron ore grade hematite
- (c) No by-products

The numbers calculated for the PAL process contain a major difference in capital cost for the process equipment, between the circuit treating limonite (\$479 million) and the one treating saprolite (\$991 million). The higher magnesium content of the saprolite leads to a significantly higher consumption of sulphuric acid and thus a larger and more expensive acid plant, which accounts for 30% of the overall difference. Because the second dissociation of sulphuric acid does not occur appreciably at the temperature in the PAL autoclaves and the magnesium in the saprolite makes it consume more acid than the limonite, the solution in the slurry from the autoclaves has a much higher level of free acid in the case of saprolite than limonite. That, in turn, requires substantially more limestone to neutralise the free acid, which leads to the formation of significantly more gypsum. The increased gypsum means more tailings to disposal, making the increased tailings dam size account for another 33 percent of the difference in capital cost between the two PAL circuits. The increased quantity of gypsum in the saprolite case also requires larger thickeners in the counter-current decantation train, accounting for a further 23 percent of the difference in the capital costs.

The atmospheric tank leaching (ATL) and heap leaching (HL) circuits have lower variable costs than the PAL circuit because they do not suffer from sulphuric acid delivering only one proton per molecule in the leaching stage, therefore they require less acid and limestone for leaching saprolite than the HPAL process does. The fixed operating cost calculated for the HL process is higher than the fixed cost calculated for the ATL process and the capital cost is lower for HL than for ATL.

because the assumption used for the HL process is that the heaps are built, leached, washed and then left as tailings. That assumption means that the major portion of the cost for the tailings dump, i.e. the base (a layer of clay between two geo-membranes, plus drainage piping) is switched from the capital cost to the operating cost. Instead of a single large base being laid down as part of the construction, the first leach pad (similar to the base of a tailings dam – an impervious layer between two geo-membranes and a drainage system) is made during the construction phase and the first heap is built on that pad. Subsequent pads and heaps are built as part of the ongoing operation of the circuit. The assumption used in the fixed cost calculation is that a new heap would be built each year for leaching during the subsequent year.

The Sinter-SAF and the Sinter-BF processes require nickel prices higher than the long term upper limit, even at the 10 percent IRR level, which eliminates these two pyrometallurgy routes from further consideration. For the RKEF process, the required nickel price is below the upper long term limit for 10 percent IRR, but above the limit at higher IRR numbers. That probably eliminates the RKEF process as well.

Table 8: Costs and required Ni price, blend

Process	Operating costs		Capital cost \$ million	Nickel price needed, \$/lb					
	Fixed \$ million/y	Variable \$/lb Ni&Co		30% IRR		20% IRR		10% IRR	
				L	H	L	H	L	H
Neomet (a)	6.4	2.13	360+636	6.59	7.15	3.02	3.57	0.24	0.77
DNi (a)	7.5	2.55	441+636	9.67	10.64	5.96	6.92	3.08	4.04
EPAL	5.8	2.41	415+636	11.14	11.64	8.07	8.59	5.66	6.18
Neomet (b)	6.4	2.13	360+636	11.68	12.58	8.12	9.02	5.35	6.23
DNi (b)	7.5	2.55	441+636	11.97	13.24	8.26	9.55	5.39	6.65
Neomet (c)	6.4	2.13	360+636	12.57	13.10	9.03	9.58	6.23	6.80
DNi (c)	7.5	2.55	441+636	13.46	14.21	9.79	10.55	6.90	7.65

- (a) With by-product magnesia and pigment grade hematite
- (b) With by-product magnesia and iron ore grade hematite
- (c) No by-products

In the case of the mixed feed (50% limonite, 50% saprolite), if the sale of by-product hematite for significantly more than the price of iron ore is a realistic assumption, Neomet and DNI would appear to be superior to the EPAL technology. If the hematite is sold as iron ore, the economics of the Neomet, DNI and EPAL processes appear to be similar. In the absence of by-product revenue the EPAL process takes the lead. This again emphasises the need for Neomet and DNI to demonstrate that the hematite and magnesia are saleable by-products, and to determine whether or not the hematite can realistically be expected to sell for a premium over the price of iron ore.

The chromium and aluminium in the limonite and saprolite all hydrolyse and become trivalent oxides in the iron hydrolysis step in the Neomet process. In DNI processes, aluminium is not hydrolysed - it is removed by precipitation after the iron hydrolysis step. Chromium was assumed to follow iron in the DNI model. Manganese, calcium, sodium and potassium do not hydrolyse in either process. In the Neomet circuit magnesium is hydrolysed and rejected selectively, with separate steps for purging manganese, calcium, sodium and potassium. In the DNI circuit the entire magnesium nitrate solution from the precipitation of mixed hydroxide is concentrated by evaporation and then decomposed thermally. The magnesium nitrate becomes magnesium oxide, but the calcium, sodium and potassium remain as nitrates and end up in the magnesia. The manganese would also remain in the magnesia, presumably as manganese dioxide. Depending on the amounts of magnesium, manganese, calcium, sodium and potassium in the feed (and on how much of each is leached), this may or may not affect the quality of the magnesia produced by the DNI process. Similarly, the aluminium and chromium could affect the quality of the hematite produced by one or both of these two processes. Table 9 and Table 10 list the compositions of the hematite and magnesia calculated by the process models for each feed, for the Neomet and the DNI processes. Whether or not by-products of these compositions can be successfully marketed, and at what prices, is something that both Neomet and DNI will need to establish in due course.

Table 9: Hematite composition, mass %

Component	Neomet			DNi		
	Limonite	Blend	Saprolite	Limonite	Blend	Saprolite
Fe ₂ O ₃	87.2	87.3	87.9	93.7	93.7	94.5
Cr ₂ O ₃	4.7	4.7	5.1	4.9	4.9	5.2
Al ₂ O ₃	7.8	7.7	6.5	0	0	0
MnO ₂	0	0	0	1.4	1.4	0.9
Salts	0.3	0.3	0.5	0	0	0

Table 10: Magnesia composition, mass %

Component	Neomet			DNi		
	Limonite	Blend	Saprolite	Limonite	Blend	Saprolite
MgO	99.9	99.8	99.8	79.4	90.6	94.5
MnO ₂	0	0	0	2.7	1.6	0.2
Salts	0.1	0.2	0.2	17.9	7.8	5.3

SUMMARY AND CONCLUSIONS

The exercise presented in this paper was done to examine the capital, fixed and variable operating costs of the processes examined, for hypothetical limonite and/or saprolite ore. The conclusions of this exercise are as follows:

- In the absence of by-product revenue, the PAL process appears to offer the strongest overall economics for processing limonite.
- If the hematite produced in the Neomet and DNi processes can be sold for appreciably more than the price of iron ore, the economics of these two processes could well be distinctly superior to the economics of PAL for processing limonite.
- The Neomet and DNi processes appear to be the most attractive for processing saprolite if the by-products generate the revenue assumed and especially if the hematite can be sold for appreciably more than the price of iron ore. This is a change from the results presented in Part 2 of this series. Both Neomet and DNi have evolved since Part 2 was presented.
- Of the established processes, heap leaching would appear to offer the best economics for processing saprolite.
- The pyrometallurgical processes appear to be economically inferior to all the hydrometallurgical processes for saprolite.

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Dr. Bryn Harris of Neomet and Dr. Fiona McCarthy of DirectNickel reviewed the mass-energy balances and the capital cost calculations produced for the Neomet and DNi processes, respectively, and gave very useful feedback on the evaluations. Their contributions significantly enhanced this paper, and are gratefully acknowledged.

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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.

Keywords: Laterites, Nickel, Cobalt, Intermediate Products, Mixed Hydroxide, Downstream Processing to Metal, Cost Comparison

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 ₂ O ₃	Cr ₂ O ₃	Fe ₂ O ₃	CaO	MgO	MnO	Na ₂ O	K ₂ O	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.

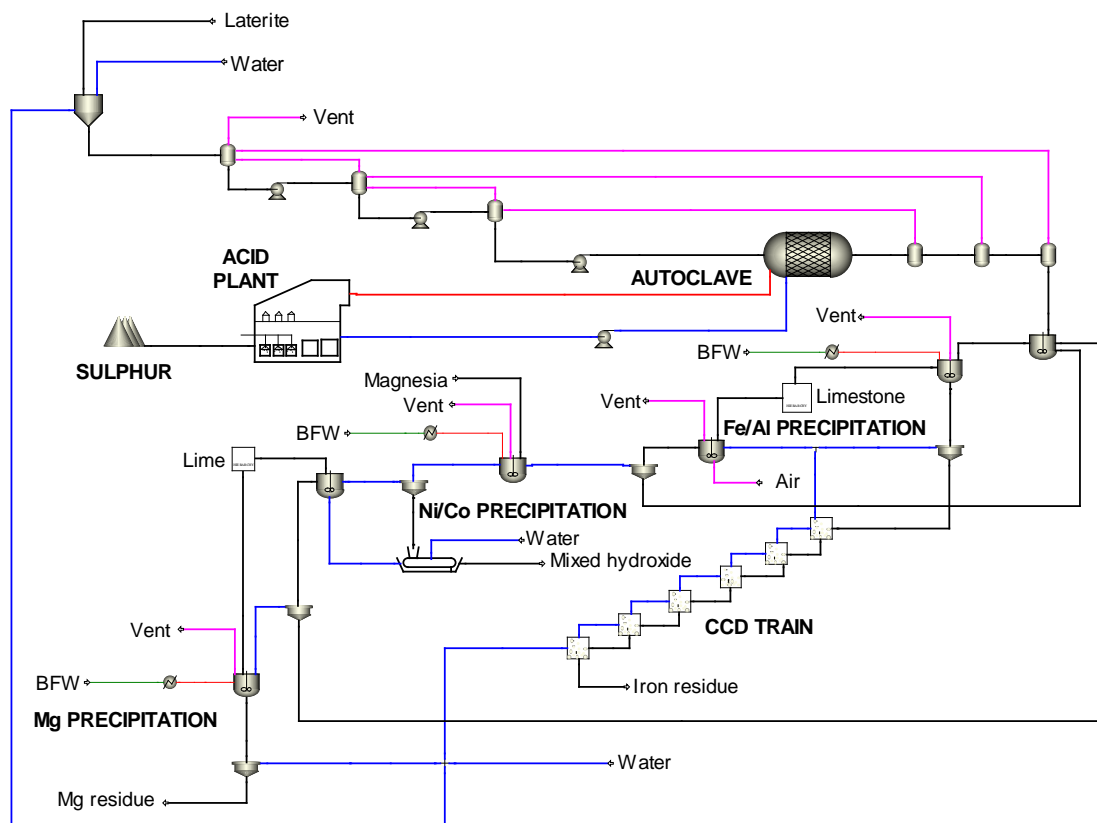


Figure 1: HPAL circuit

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.

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

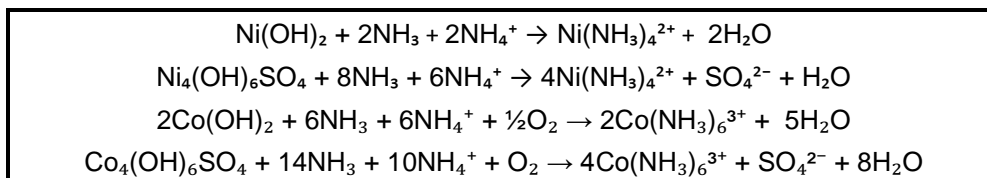


Table 4: Nickel SX and EW chemistry

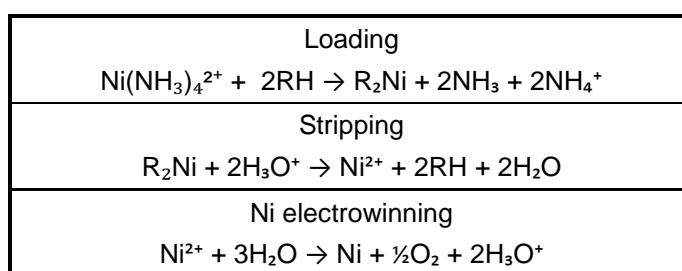
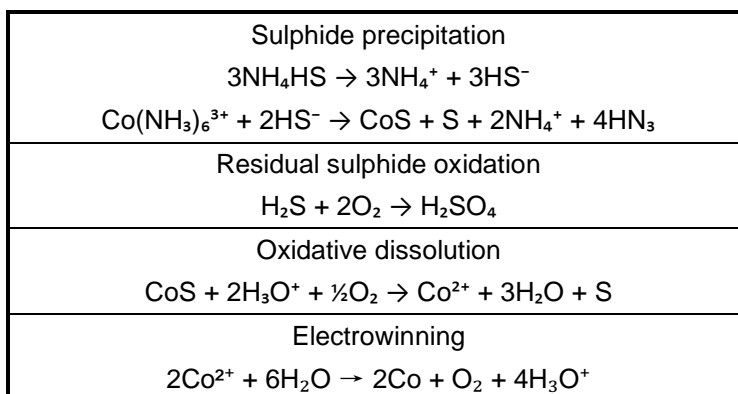


Table 5: Cobalt precipitation, oxidation and EW chemistry



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 Paulo 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.

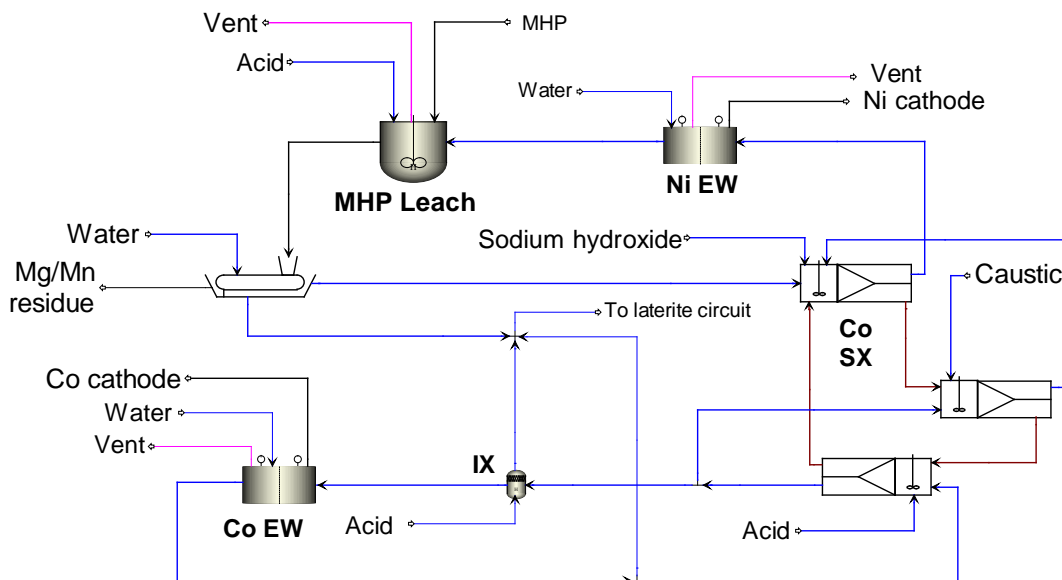


Figure 3: Sulphuric acid leach circuit

Table 6: Sulphuric acid leach stoichiometry

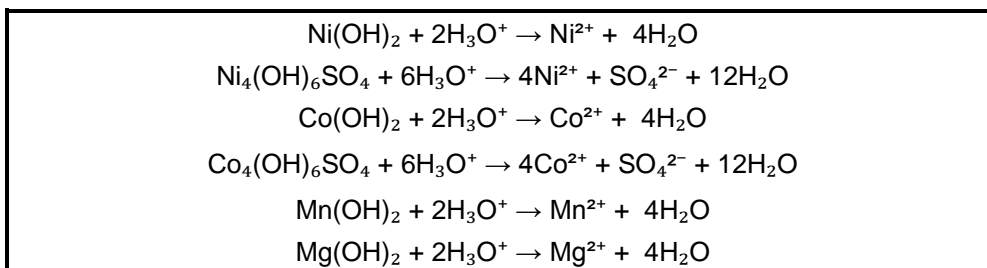
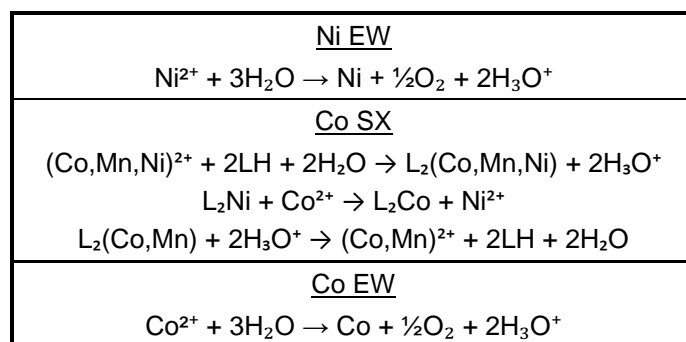


Table 7: Ni EW, Co SX, IX and EW stoichiometry



The MHP is dissolved using fresh sulphuric acid and spent electrolyte from the nickel electro-winning 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 electro-winning section. The loaded strip liquor, less the part used as scrub solution, goes to the cobalt electro-winning 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 electro-winning. 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.

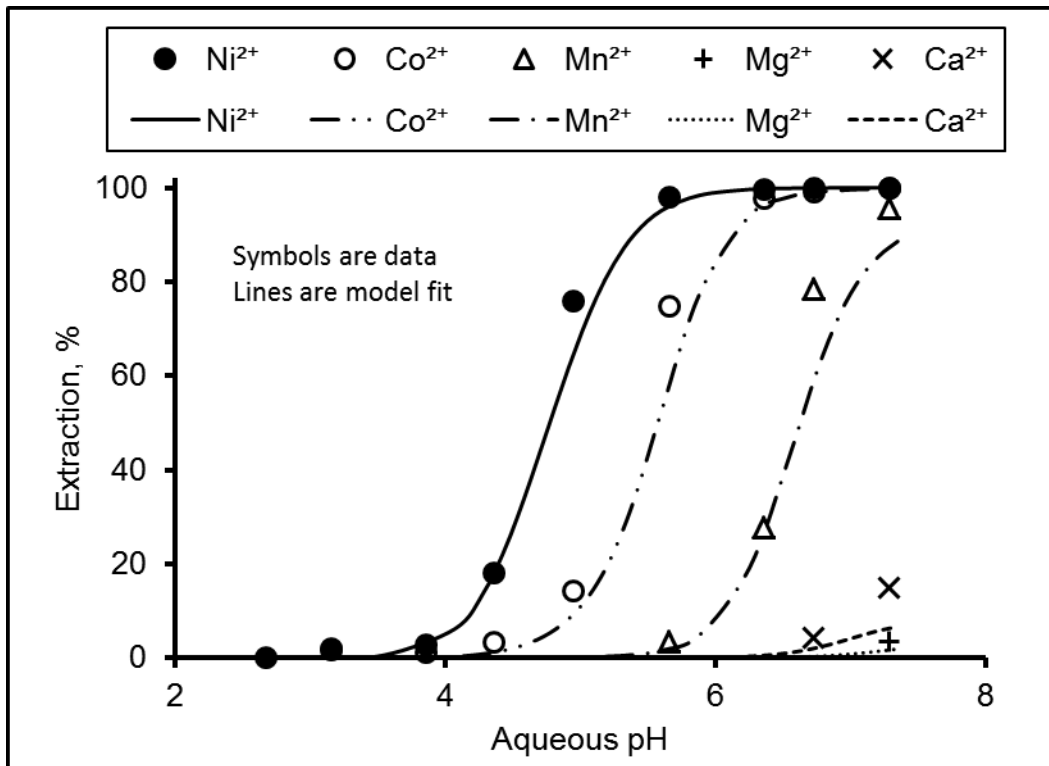
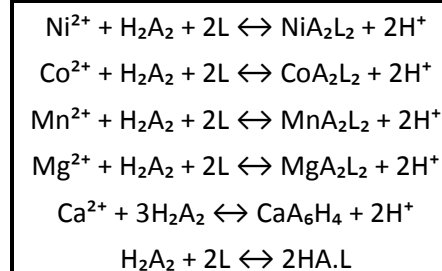


Figure 4: SSX data and model fit

Table 8: SSX reaction stoichiometry



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.

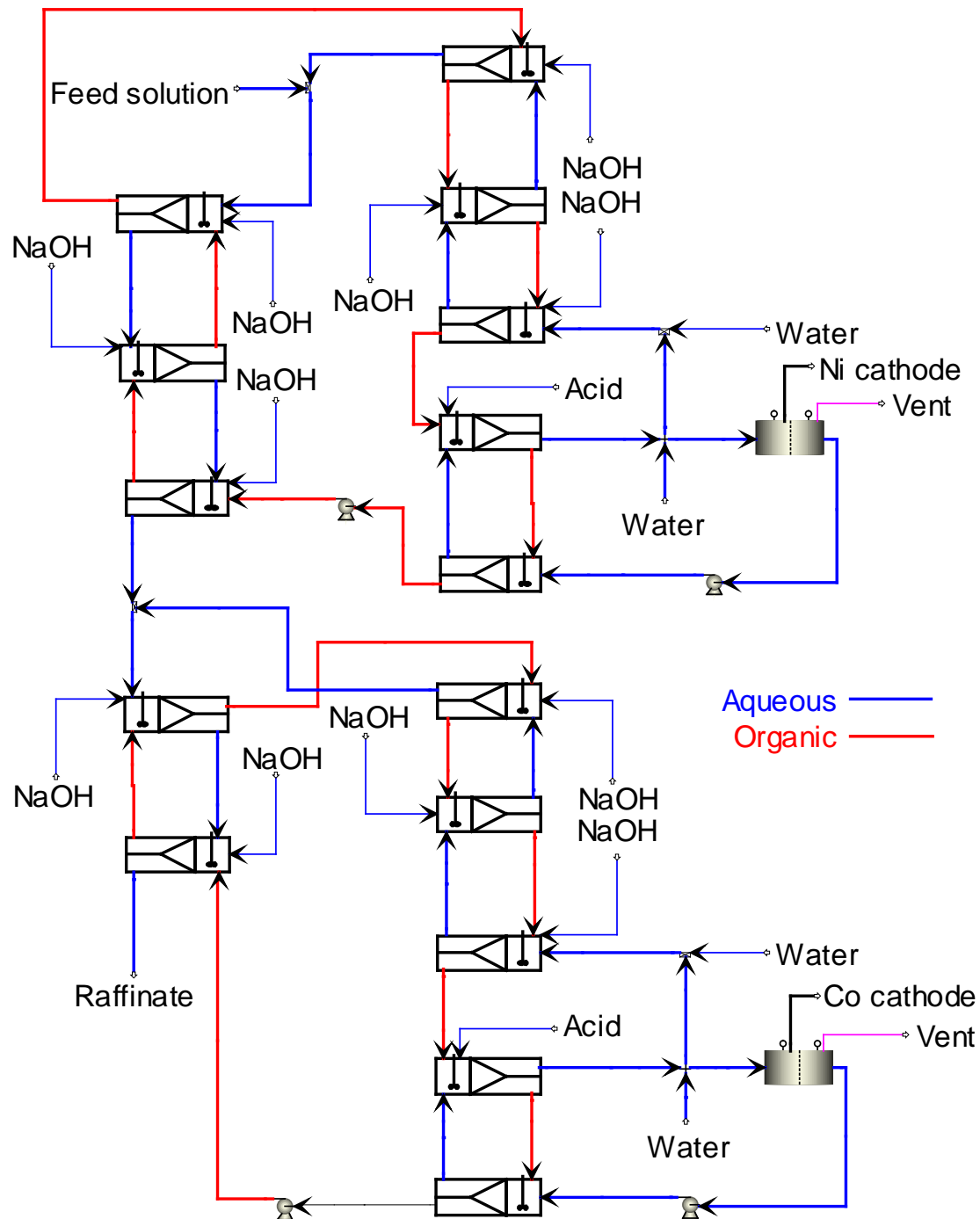


Figure 5: SSX circuit

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 electro-winning 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 electro-winning 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 literature³. 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 ₂ S	\$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	6
Variable costs			32	65	130	130	130	130	130	130	130	130
Revenue			151	303	605	605	605	605	605	605	605	605
Gross margin	-572	-572	113	232	470	470	470	470	470	470	470	470
Tax	0	0	0	0	0	42	141	141	141	141	141	141
Net margin	-572	-572	113	232	470	428	329	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	3
Variable costs			4	8	16	16	16	16	16	16	16	16
Revenue			27	53	107	107	107	107	107	107	107	107
Gross margin	-156	-156	20	42	88	88	88	88	88	88	88	88
Tax	0	0	0	0	0	0	4	26	26	26	26	26
Net margin	-156	-156	20	42	88	88	84	62	62	62	62	62
IRR, 10 year	10%											
IRR, 20 year	17%											

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

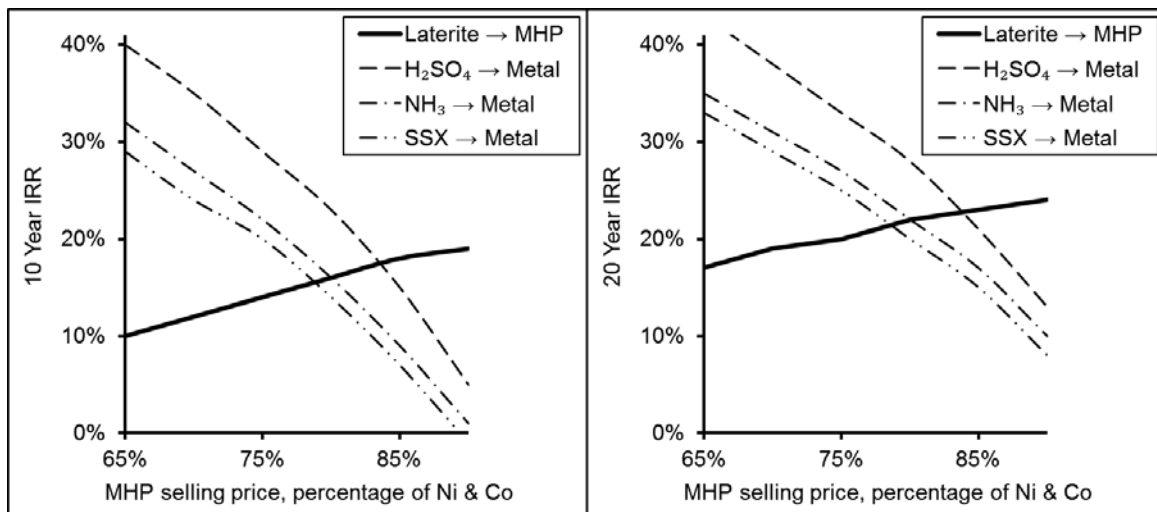
Cash flow \$ million	Year											
	1	2	3	4	5	6	7	8	9	10	11→	
Capital cost	118	118										
Fixed costs			2	2	2	2	2	2	2	2	2	2
Variable costs			5	9	19	19	19	19	19	19	19	19
Revenue			27	53	107	107	107	107	107	107	107	107
Gross margin	-118	-118	20	42	86	86	86	86	86	86	86	86
Tax	0	0	0	0	0	0	25	26	26	26	26	26
Net margin	-118	-118	20	42	86	86	61	60	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	3
Variable costs			4	9	18	18	18	18	18	18	18	18
Revenue			27	53	107	107	107	107	107	107	107	107
Gross margin	-174	-174	19	41	86	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.

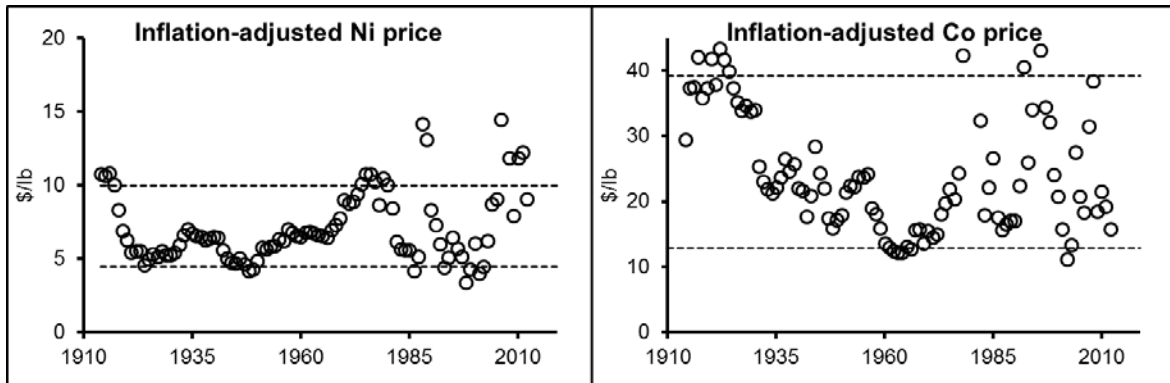


Figure 7: Historical nickel and cobalt prices

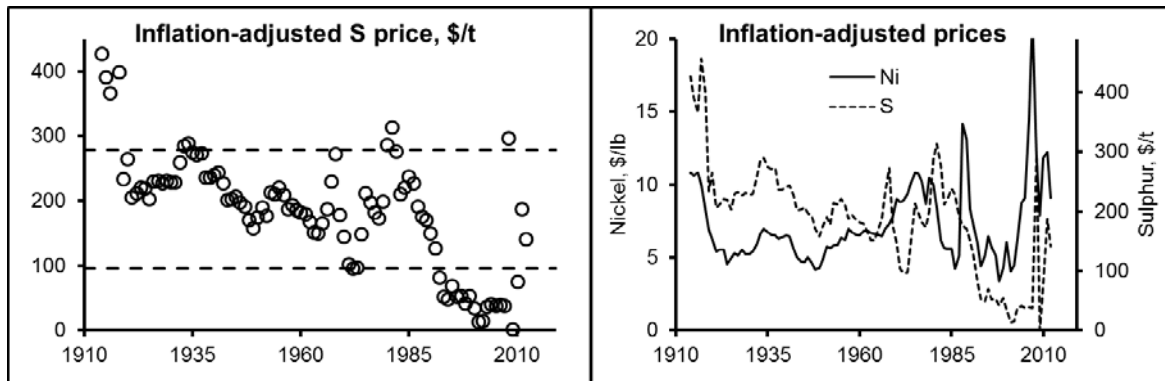


Figure 8: Sulphur and nickel prices

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.

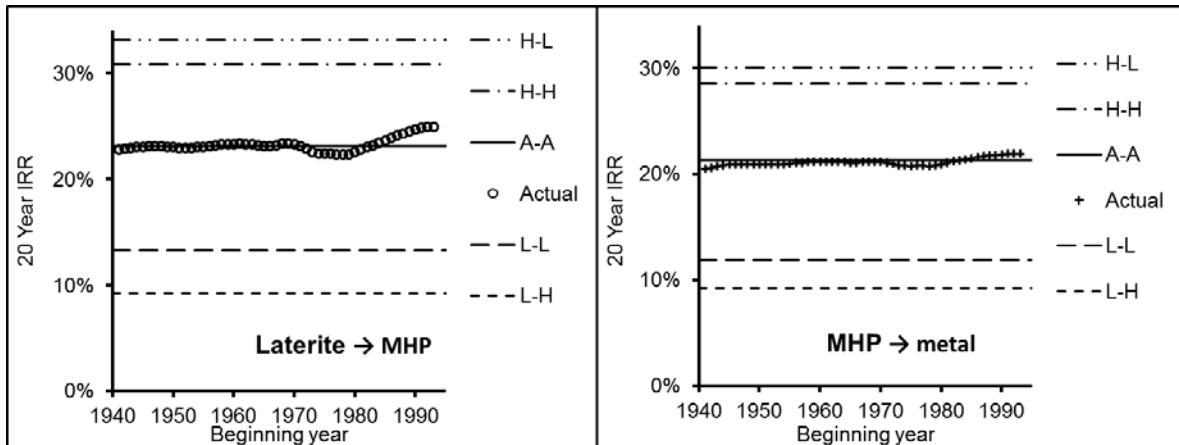


Figure 9: IRR calculations using historical price data

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 than 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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