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TECHNICAL & COST COMPARISON OF LATERITE TREATMENT PROCESSES
PART 2

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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_2O_3	5.61	0.86
Cr_2O_3	3.36	0.67
Fe_2O_3	64.31	11.99
CaO	0.06	0.15
MgO	1.54	27.33
MnO	1.51	0.18
Na_2O	0.17	0.05
K_2O	0.01	0.03
SiO_2	7.62	40.00

Table 2 – Hypothetical ore mineralogy

Mineral	Limonite	Saprolite
NiO	1.55	0.00
Ni_2SiO_4	0.00	2.86
CoO	0.25	0.03
$Al_2O_3 \bullet H_2O$	6.60	1.01
$FeCr_2O_4$	4.95	0.99
$Fe_2O_3 \bullet H_2O$	45.28	0.00
$Fe(OH)_3$	29.25	15.58
$Ca_3Si_2O_7 \bullet 3H_2O$	0.12	0.31
$Mg_3Si_2O_5(OH)_4$	0.00	41.05
$Mg_4Si_6O_{15}(OH)_2 \bullet 6H_2O$	6.17	37.71
$MnSiO_3$	2.79	0.33
Na_2SiO_3	0.33	0.10
K_2SiO_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

$$NiO + 2H_2SO_4 \rightarrow Ni^{2+} + 2HSO_4^- + H_2O$$
$$Ni_2SiO_4 + 4H_2SO_4 \rightarrow 2Ni^{2+} + 4HSO_4^- + 2H_2O + SiO_2$$
$$CoO + 2H_2SO_4 \rightarrow Co^{2+} + 2HSO_4^- + H_2O$$
$$Al_2O_3 \bullet H_2O \rightarrow Al_2O_3 + H_2O$$
$$FeCr_2O_4 + 2H_2SO_4 \rightarrow Fe^{2+} + Cr_2O_3 + 2HSO_4^- + H_2O$$
$$2FeOOH + 6H_2SO_4 \rightarrow 2Fe^{3+} + 6HSO_4^- + 4H_2O \rightarrow Fe_2O_3 + 6H_2SO_4 + H_2O$$
$$2Fe(OH)_3 + 6H_2SO_4 \rightarrow 2Fe^{3+} + 6HSO_4^- + 6H_2O \rightarrow Fe_2O_3 + 6H_2SO_4 + 3H_2O$$
$$Ca_3Si_2O_7 \bullet 3H_2O + 6H_2SO_4 \rightarrow 3Ca^{2+} + 6HSO_4^- + 6H_2O + 2SiO_2$$
$$Mg_3Si_2O_5(OH)_4 + 6H_2SO_4 \rightarrow 3Mg^{2+} + 6HSO_4^- + 5H_2O + 2SiO_2$$
$$Mg_4Si_6O_{15}(OH)_2 \bullet 6H_2O + 8H_2SO_4 \rightarrow 4Mg^{2+} + 8HSO_4^- + 11H_2O + 6SiO_2$$
$$MnSiO_3 + 2H_2SO_4 \rightarrow Mn^{2+} + 2HSO_4^- + H_2O + SiO_2$$
$$Na_2SiO_3 + 2H_2SO_4 \rightarrow 2Na^+ + 2HSO_4^- + H_2O + SiO_2$$
$$K_2SiO_3 + 2H_2SO_4 \rightarrow 2K^+ + 2HSO_4^- + H_2O + SiO_2$$

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

$$CaCO_3 + 2H_2SO_4 + H_2O \rightarrow CaSO_4 \bullet 2H_2O\downarrow + CO_2$$
$$Fe_2(SO_4)_3 + 3CaCO_3 + 7H_2O \rightarrow 2FeOOH\downarrow + 3CaSO_4 \bullet 2H_2O\downarrow + 3CO_2$$
$$Al_2(SO_4)_3 + 3CaCO_3 + 7H_2O \rightarrow 2AlOOH\downarrow + 3CaSO_4 \bullet 2H_2O\downarrow + 3CO_2$$
$$Cr_2(SO_4)_3 + 3CaCO_3 + 7H_2O \rightarrow 2CrOOH\downarrow + 3CaSO_4 \bullet 2H_2O\downarrow + 3CO_2$$

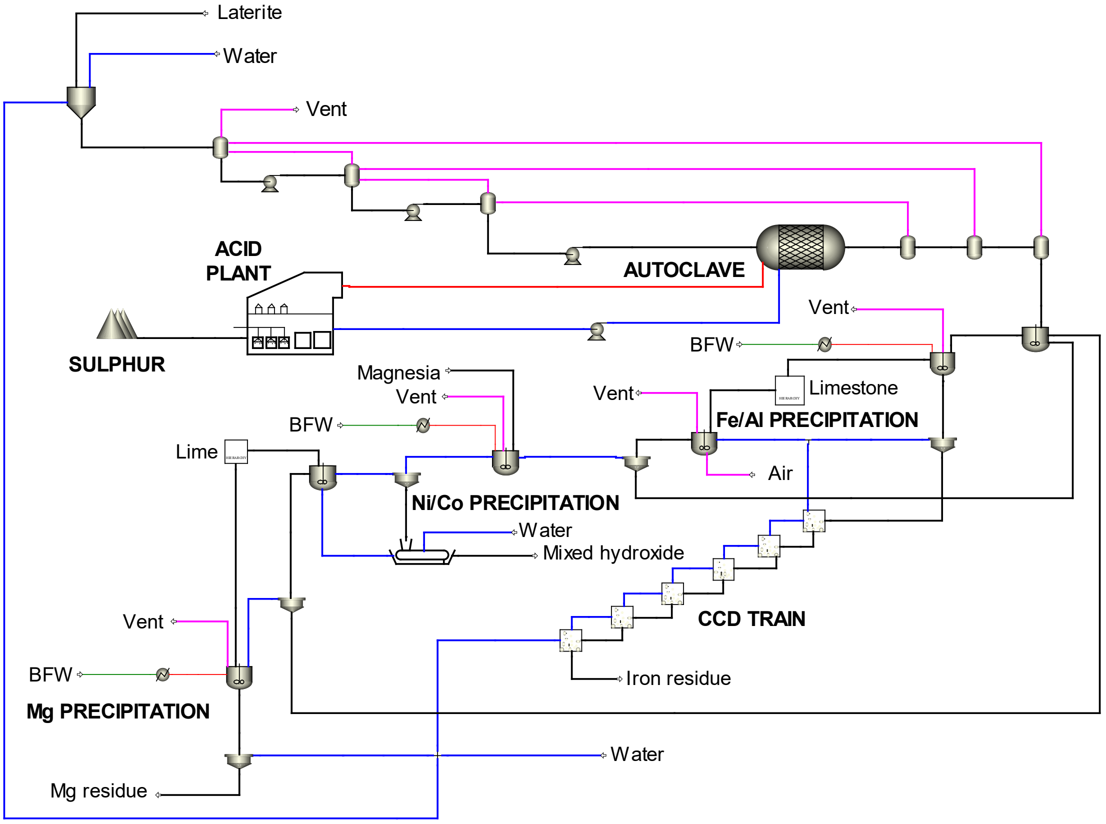
$$NiSO_4 + MgO + H_2O \rightarrow Ni(OH)_2\downarrow + MgSO_4$$
$$CoSO_4 + MgO + H_2O \rightarrow Co(OH)_2\downarrow + MgSO_4$$

$$MgSO_4 + Ca(OH)_2 + 2H_2O \rightarrow Mg(OH)_2\downarrow + CaSO_4 \bullet 2H_2O\downarrow$$
$$MnSO_4 + Ca(OH)_2 + 2H_2O \rightarrow Mn(OH)_2\downarrow + CaSO_4 \bullet 2H_2O\downarrow$$

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

$$Ni_2SiO_4 + 2H_2SO_4 \rightarrow 2Ni^{2+} + 2SO_4^{2-} + 2H_2O + SiO_2$$

$$CoO + H_2SO_4 \rightarrow Co^{2+} + SO_4^{2-} + H_2O$$

$$Ca_3Si_2O_7 \bullet 3H_2O + 3H_2SO_4 \rightarrow 3Ca^{2+} + 3SO_4^{2-} + 6H_2O + 2SiO_2$$

$$Mg_3Si_2O_5(OH)_4 + 3H_2SO_4 \rightarrow 3Mg^{2+} + 3SO_4^{2-} + 5H_2O + 2SiO_2$$

$$Mg_4Si_6O_{15}(OH)_2 \bullet 6H_2O + 4H_2SO_4 \rightarrow 4Mg^{2+} + 4SO_4^{2-} + 11H_2O + 6SiO_2$$

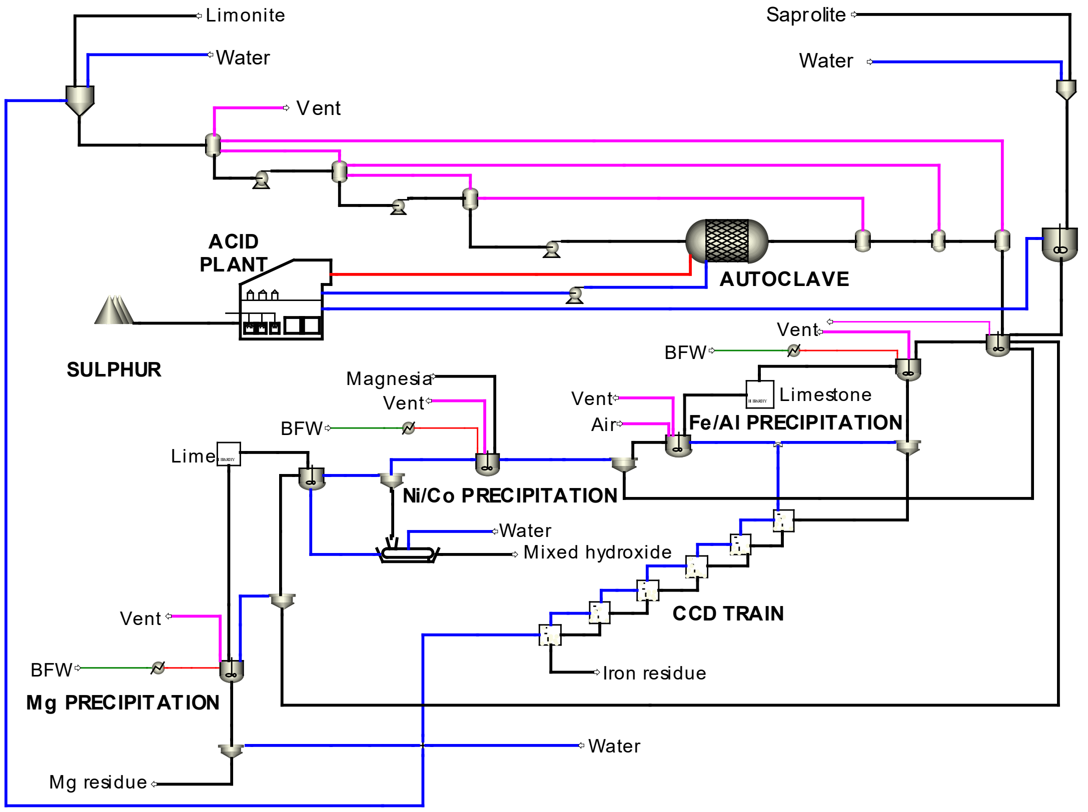
$$MnSiO_3 + H_2SO_4 \rightarrow Mn^{2+} + SO_4^{2-} + H_2O + SiO_2$$

$$Na_2SiO_3 + 2H_2SO_4 \rightarrow 2Na^{+} + SO_4^{2-} + H_2O + SiO_2$$

$$K_2SiO_3 + 2H_2SO_4 \rightarrow 2K^{+} + SO_4^{2-} + H_2O + SiO_2$$

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



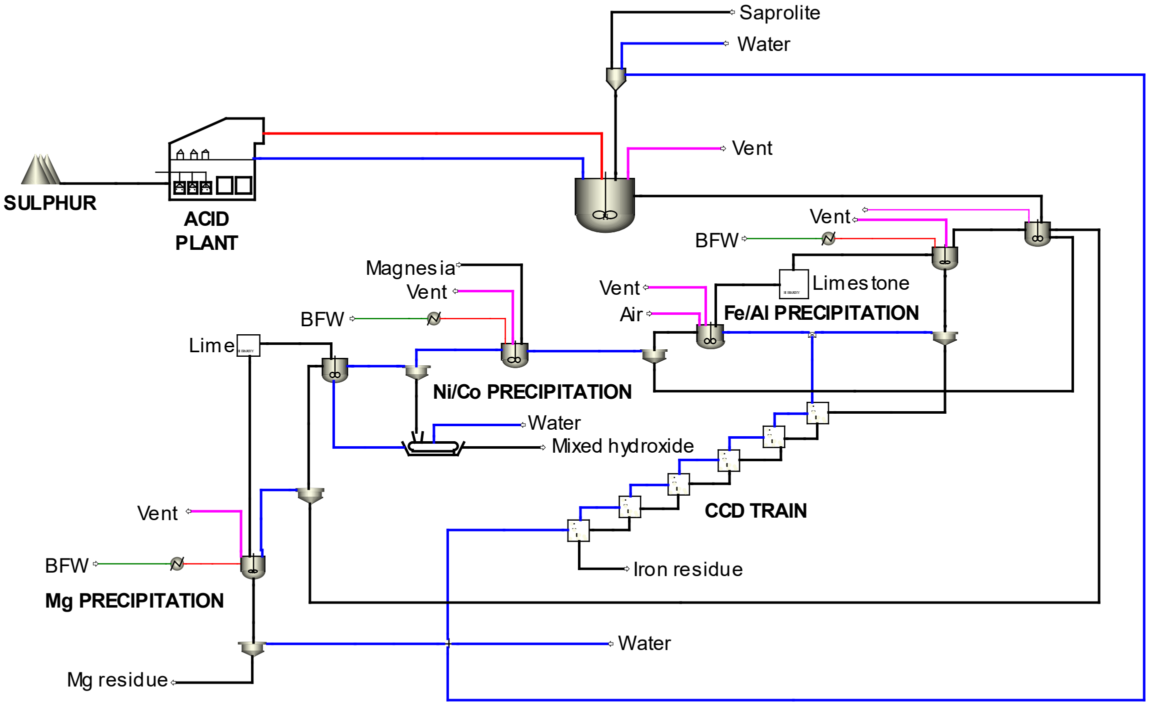
The limonite is treated as in the PAL model. As outlined in the part 1 companion paper, the saprolite is mixed with water and pre-leached at atmospheric pressure with acid from the acid plant to destroy minerals such as carbonates (although the assumed feed does not contain these minerals). The pre-leached 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

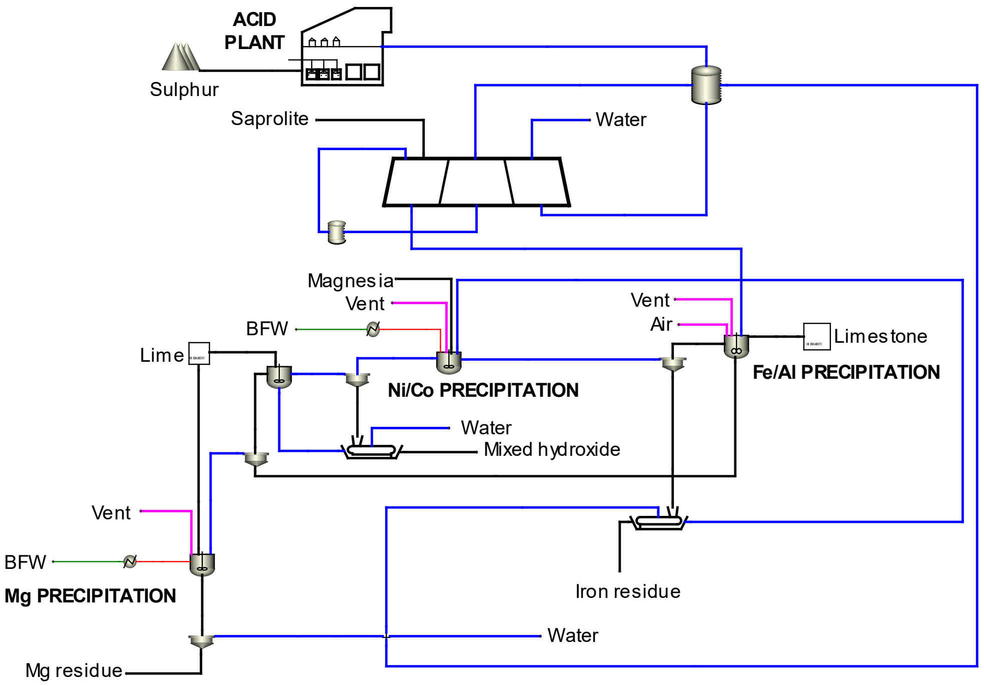


Heap leaching

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

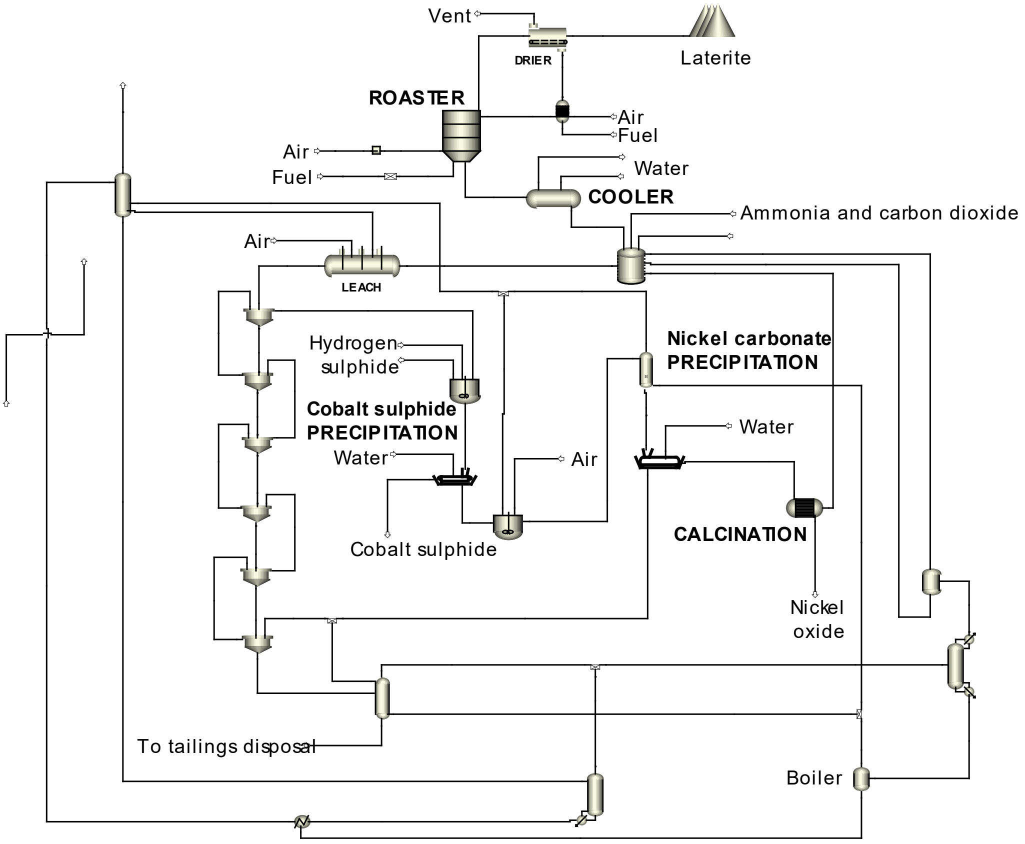
The Caron process is applicable to limonite. It uses a reducing roast to convert the oxidized iron minerals and the nickel and cobalt oxides in those minerals to a metallic state. An air plus aqueous ammonia-ammonium carbonate leach is used to selectively dissolve the metallic nickel and cobalt from 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

$CH_4 + 1\frac{1}{2}O_2 \rightarrow CO + 2H_2O$
$Al_2O_3 \bullet H_2O \rightarrow Al_2O_3 + H_2O$
$2FeOOH \rightarrow Fe_2O_3 + H_2O$
$2Fe(OH)_3 \rightarrow Fe_2O_3 + 3H_2O$
$Ca_3Si_2O_7 \bullet 3H_2O \rightarrow 3CaO + 2SiO_2 + 3H_2O$
$Mg_3Si_2O_5(OH)_4 \rightarrow 3MgO + 2SiO_2 + 2H_2O$
$Mg_4Si_6O_{15}(OH)_2 \bullet 6H_2O \rightarrow 4MgO + 6SiO_2 + 7H_2O$
$NiO + CO \rightarrow Ni + CO_2$
$CoO + CO \rightarrow Co + CO_2$
$FeCr_2O_4 + CO \rightarrow Fe + CO_2 + Cr_2O_3$
$3Fe_2O_3 + CO \rightarrow 2Fe_3O_4 + CO_2$
$Fe_3O_4 + 4CO \rightarrow 3Fe + 4CO_2$
$Ni + 6NH_3 + \frac{1}{2}O_2 + H_2O \rightarrow Ni(NH_3)_6^{2+} + 2OH^-$
$Co + 6NH_3 + \frac{3}{4}O_2 + \frac{3}{4}H_2O \rightarrow Co(NH_3)_6^{3+} + 3OH^-$
$Fe + 6NH_3 + O_2 + H_2O \rightarrow Fe(NH_3)_6^{2+} + 2OH^-$
$2Fe(NH_3)_6^{2+} + \frac{1}{2}O_2 + 2H_2O \rightarrow Fe_2O_3 + 8NH_3 + 4NH_4^+$

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

$$NiO + 2HCl \rightarrow Ni^{2+} + 2Cl^- + H_2O$$
$$Ni_2SiO_4 + 4HCl \rightarrow 2Ni^{2+} + 4Cl^- + 2H_2O + SiO_2$$
$$CoO + 2HCl \rightarrow Co^{2+} + 2Cl^- + H_2O$$
$$Al_2O_3 \bullet H_2O + 6HCl \rightarrow 2Al^{3+} + 6Cl^- + 4H_2O$$
$$FeCr_2O_4 + 8HCl \rightarrow Fe^{2+} + 2Cr^{3+} + 8Cl^- + 4H_2O$$
$$2FeOOH + 6HCl \rightarrow 2Fe^{3+} + 6Cl^- + 4H_2O$$
$$Fe(OH)_3 + 3HCl \rightarrow Fe^{3+} + 3Cl^- + 3H_2O$$
$$Ca_3Si_2O_7 \bullet 3H_2O + 6HCl \rightarrow 3Ca^{2+} + 6Cl^- + 6H_2O + 2SiO_2$$
$$Mg_3Si_2O_5(OH)_4 + 6HCl \rightarrow 3Mg^{2+} + 6Cl^- + 5H_2O + 2SiO_2$$
$$Mg_4Si_6O_{15}(OH)_2 \bullet 6H_2O + 8HCl \rightarrow 4Mg^{2+} + 8Cl^- + 11H_2O + 6SiO_2$$
$$MnSiO_3 + 2HCl \rightarrow Mn^{2+} + 2Cl^- + H_2O + SiO_2$$
$$Na_2SiO_3 + 2HCl \rightarrow 2Na^+ + 2Cl^- + H_2O + SiO_2$$
$$K_2SiO_3 + 2HCl \rightarrow 2K^+ + 2Cl^- + H_2O + SiO_2$$

$$2Fe^{3+} + 6Cl^- + 3H_2O \rightarrow Fe_2O_3 + 6HCl\uparrow$$
$$2Cr^{3+} + 6Cl^- + 3H_2O \rightarrow Cr_2O_3 + 6HCl\uparrow$$
$$2Al^{3+} + 6Cl^- + 3H_2O \rightarrow Al_2O_3 + 6HCl\uparrow$$

$$3NiCl_2 + H_2O \rightarrow Ni_3(OH)_5Cl + 5HCl\uparrow$$
$$3CoCl_2 + H_2O \rightarrow Co_3(OH)_5Cl + 5HCl\uparrow$$

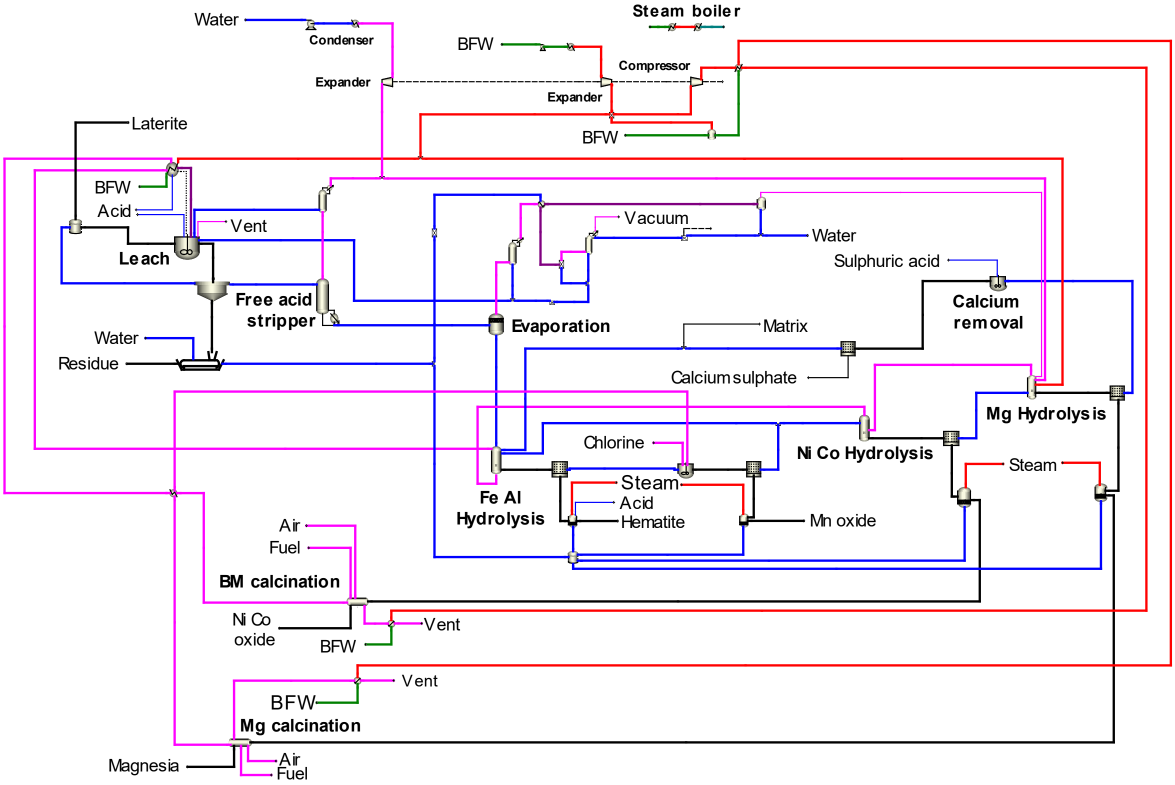
$$MgCl_2 + H_2O \rightarrow MgOHCl + HCl\uparrow$$

$$Ni_3(OH)_5Cl \rightarrow 3NiO + HCl\uparrow + 2H_2O\uparrow$$
$$Co_3(OH)_5Cl \rightarrow 3CoO + HCl\uparrow + 2H_2O\uparrow$$

$$MgOHCl \rightarrow MgO + HCl\uparrow$$

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

$NiO + 2HNO_3 \rightarrow Ni^{2+} + 2NO_3^- + H_2O$
$Ni_2SiO_4 + 4HNO_3 \rightarrow 2Ni^{2+} + 4NO_3^- + 2H_2O + SiO_2$
$CoO + 2HNO_3 \rightarrow Co^{2+} + 2NO_3^- + H_2O$
$Al_2O_3 \bullet H_2O + 6HNO_3 \rightarrow 2Al^{3+} + 6NO_3^- + 4H_2O$
$FeCr_2O_4 + 8HNO_3 \rightarrow Fe^{2+} + 2Cr^{3+} + 8NO_3^- + 4H_2O$
$Fe_2O_3 \bullet H_2O + 6HNO_3 \rightarrow 2Fe^{3+} + 6NO_3^- + 4H_2O$
$Fe(OH)_3 + 3HNO_3 \rightarrow Fe^{3+} + 3NO_3^- + 3H_2O$
$Ca_3Si_2O_7 \bullet 3H_2O + 6HNO_3 \rightarrow 3Ca^{2+} + 6NO_3^- + 6H_2O + 2SiO_2$
$Mg_3Si_2O_5(OH)_4 + 6HNO_3 \rightarrow 3Mg^{2+} + 6NO_3^- + 5H_2O + 2SiO_2$
$Mg_4Si_6O_{15}(OH)_2 \bullet 6H_2O + 8HNO_3 \rightarrow 4Mg^{2+} + 8NO_3^- + 11H_2O + 6SiO_2$
$MnSiO_3 + 2HNO_3 \rightarrow Mn^{2+} + 2NO_3^- + H_2O + SiO_2$
$Na_2SiO_3 + 2HNO_3 \rightarrow 2Na^+ + 2NO_3^- + H_2O + SiO_2$
$K_2SiO_3 + 2HNO_3 \rightarrow 2K^+ + 2NO_3^- + H_2O + SiO_2$

$2Fe(NO_3)_3 + 3H_2O \rightarrow Fe_2O_3 + 6HNO_3$
$2Al(NO_3)_3 + 3H_2O \rightarrow Al_2O_3 + 6HNO_3$
$2Cr(NO_3)_3 + 3H_2O \rightarrow Cr_2O_3 + 6HNO_3$

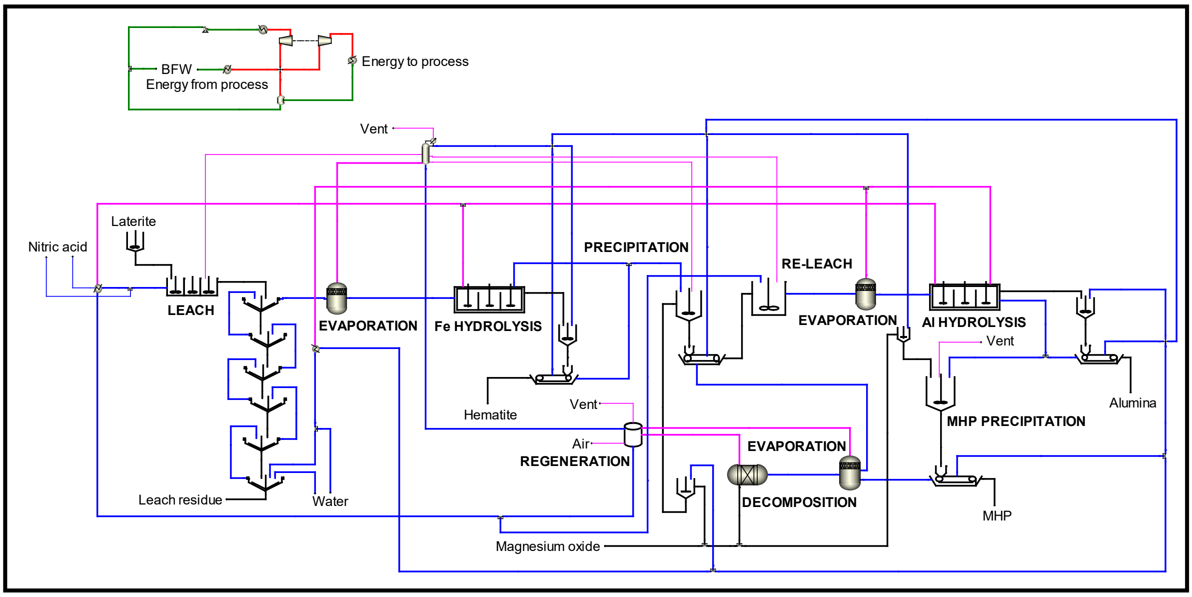
$2Al(NO_3)_3 + 3Mg(OH)_2 \rightarrow 3Mg(NO_3)_2 + 2Al(OH)_3$
$Co(NO_3)_2 + Mg(OH)_2 \rightarrow Mg(NO_3)_2 + Co(OH)_2$
$Ni(NO_3)_2 + Mg(OH)_2 \rightarrow Mg(NO_3)_2 + Ni(OH)_2$

$Mg(NO_3)_2 \bullet 2H_2O \rightarrow MgO + NO_2 + NO + O_2 + 2H_2O$

$NO + O_2 \rightarrow 2NO_2$
$4NO_2 + H_2O + O_2 \rightarrow 4HNO_3$

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



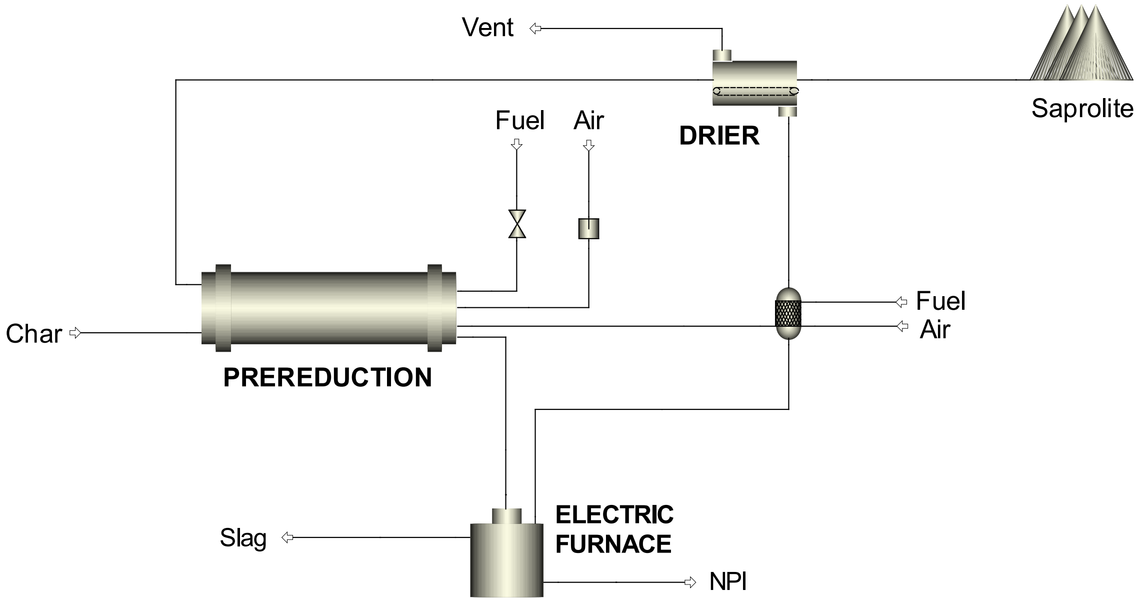
The hot slurry from the aluminium hydrolysis reactor is quenched into water, the aluminium oxide is separated from the diluted slurry by filtration and the filter cake is washed with water. The washed filter cake leaves the circuit and the combined filtrate is split, part recycling to the aluminium hydrolysis 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 O_2, NO and NO_2. The magnesium oxide is partly recycled to the two precipitation steps and the excess leaves the circuit. The NO/NO_2 is converted back to nitric acid by a patented method in which NO reacts with HNO_3 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/NO_2 is completely oxidized, giving aqueous nitric acid (60% HNO_3) 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 Hatch[1]. 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.

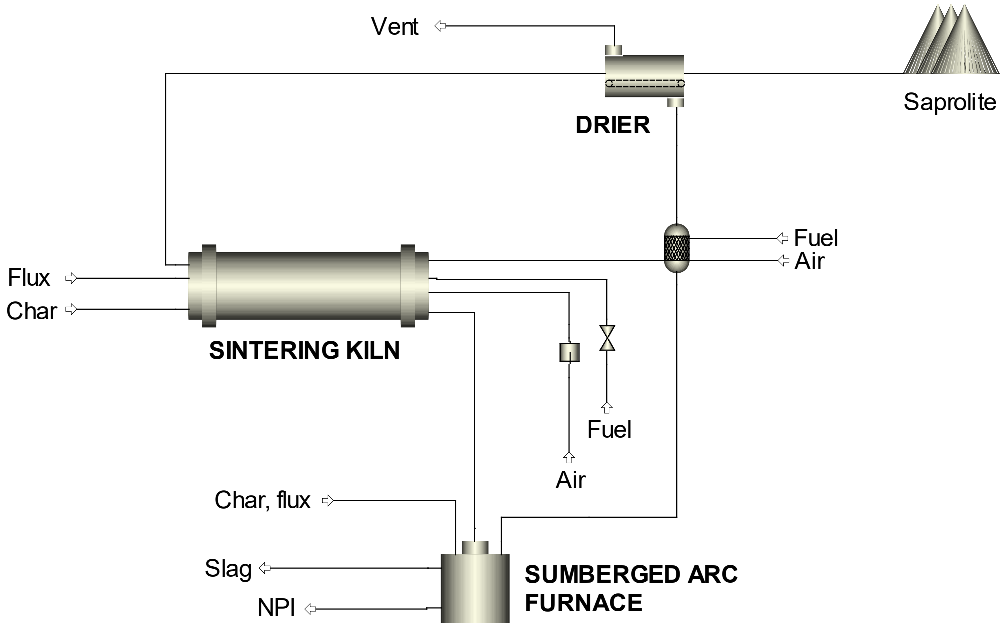
Table 9 – Pre-reduction stoichiometry

$Al_2O_3 \bullet H_2O \rightarrow Al_2O_3 + H_2O$
$Fe_2O_3 \bullet H_2O \rightarrow Fe_2O_3 + H_2O$
$2Fe(OH)_3 \rightarrow Fe_2O_3 + 3H_2O$
$Ca_3Si_2O_7 \bullet 3H_2O \rightarrow 3CaO + 2SiO_2 + 3H_2O$
$Mg_3Si_2O_5(OH)_4 \rightarrow 3MgO + 2SiO_2 + 2H_2O$
$Mg_4Si_6O_{15}(OH)_2 \bullet 6H_2O \rightarrow 4MgO + 6SiO_2 + 7H_2O$
$MgO + Fe_2O_3 \rightarrow MgFe_2O_4$
$CaO + Fe_2O_3 \rightarrow CaFe_2O_4$
$FeCr_2O_4 + 4C \rightarrow Fe + 2Cr + 4CO$
$FeCr_2O_4 + 4C \rightarrow Fe + 2Cr + 4CO$
$Fe_2O_3 + 3C \rightarrow 2Fe + 3CO$
$Ni_2SiO_4 + 2C \rightarrow 2Ni + 2CO + SiO_2$
$CoO + C \rightarrow Co + CO$
$2C + O_2 \rightarrow 2CO$
$CH_4 + 1\frac{1}{2}O_2 \rightarrow CO + 2H_2O$

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.

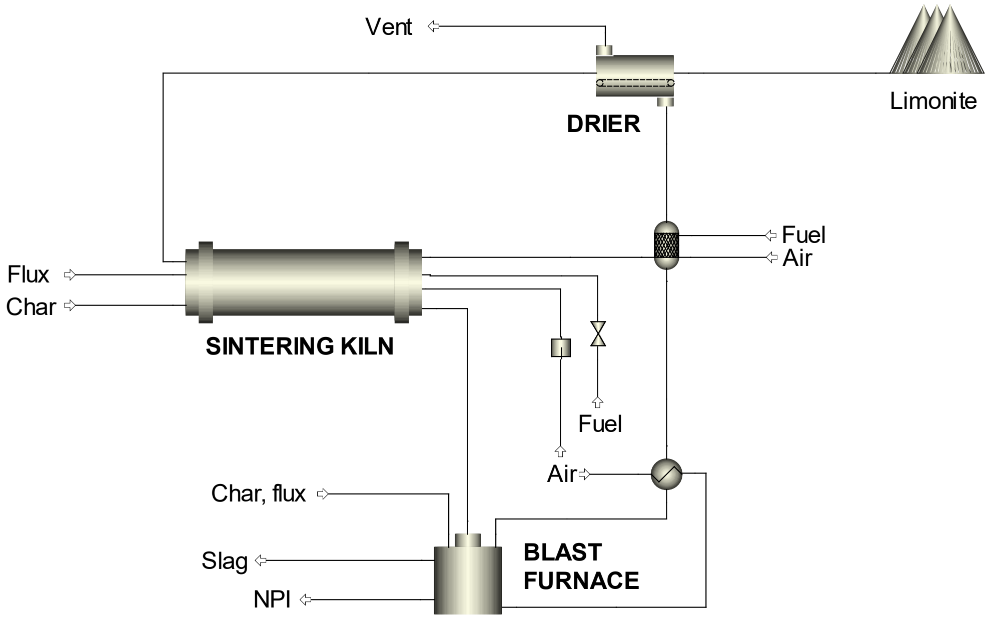
Figure 9 – SAF model



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 estimated[2] 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_2SO_4)	100/tonne
Hydrogen sulphide (H_2S)	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_4HCO_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_4HCO_3	0.05	-	-	-	-	-	-
H_2S	0.07	-	-	-	-	-	-
HNO_3		-	-	-	0.07	0.03	0.03
H_2SO_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 paper[3] 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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