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 teaching 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 2 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_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 3 – 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

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.

3

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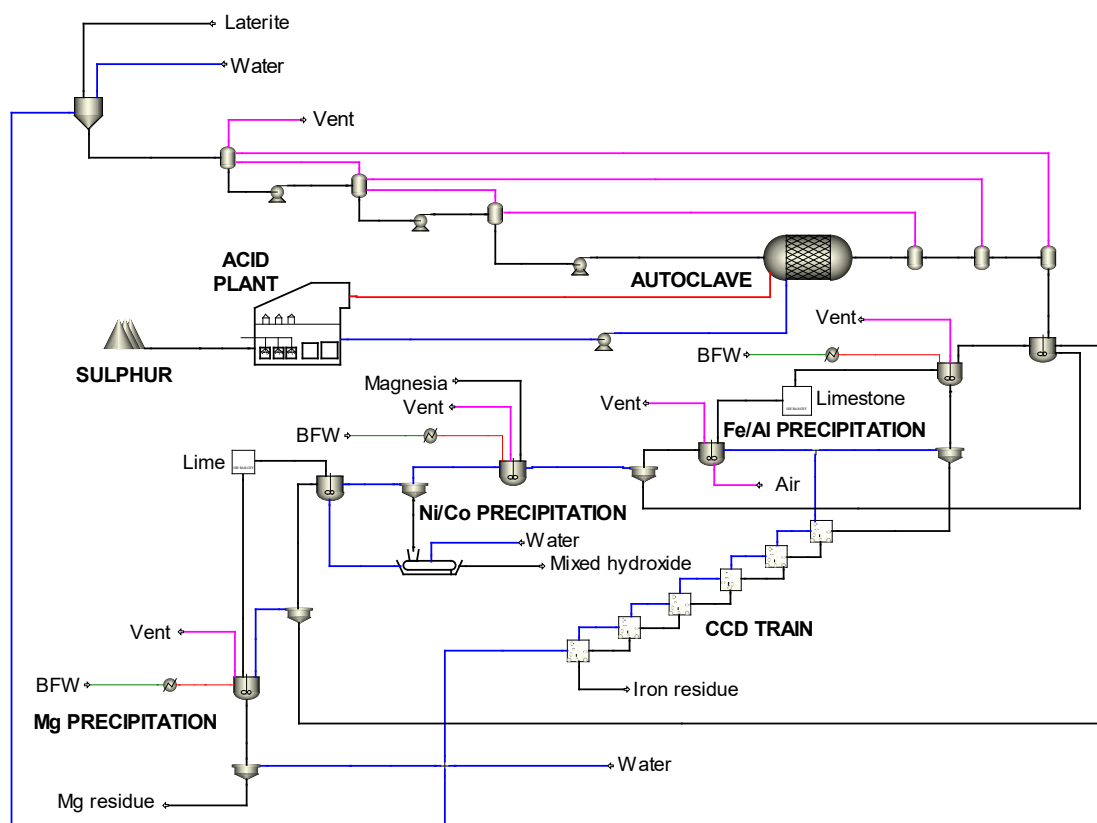
The sulphate based technologies all include a sulphur burning acid plant generating sulphuric acid and steam. This is a change from what was presented in part 2, in which purchased acid and steam were assumed.

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.

Figure 1 – PAL model

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 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 as hydroxide and gypsum 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.

The feed mixing tank was assumed to be an agitated vessel of brick lined steel, with the default residence time of five minutes. The pre-heating stages ahead of the autoclaves and the depressurisation vessels after the autoclaves were assumed to be brick-lined steel, with the default five minute liquid residence time for each vessel.

A limitation in the equipment database used in this exercise is that it does not cover the specialised high pressure slurry pumps used in the autoclave train. The workaround used was to assume conventional pumps but specify the material of construction as either Inconel or Monel, whichever resulted in the higher cost, and to assume two pumps per pumping stage.

The residence time in the autoclaves was assumed to be 75 minutes. Assuming the active volume to be 60 percent of the total volume and autoclaves 5 meters in diameter and 35 meters in length led to three autoclave trains in parallel, each train consisting of the three pre-heating stages and 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 4. 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.
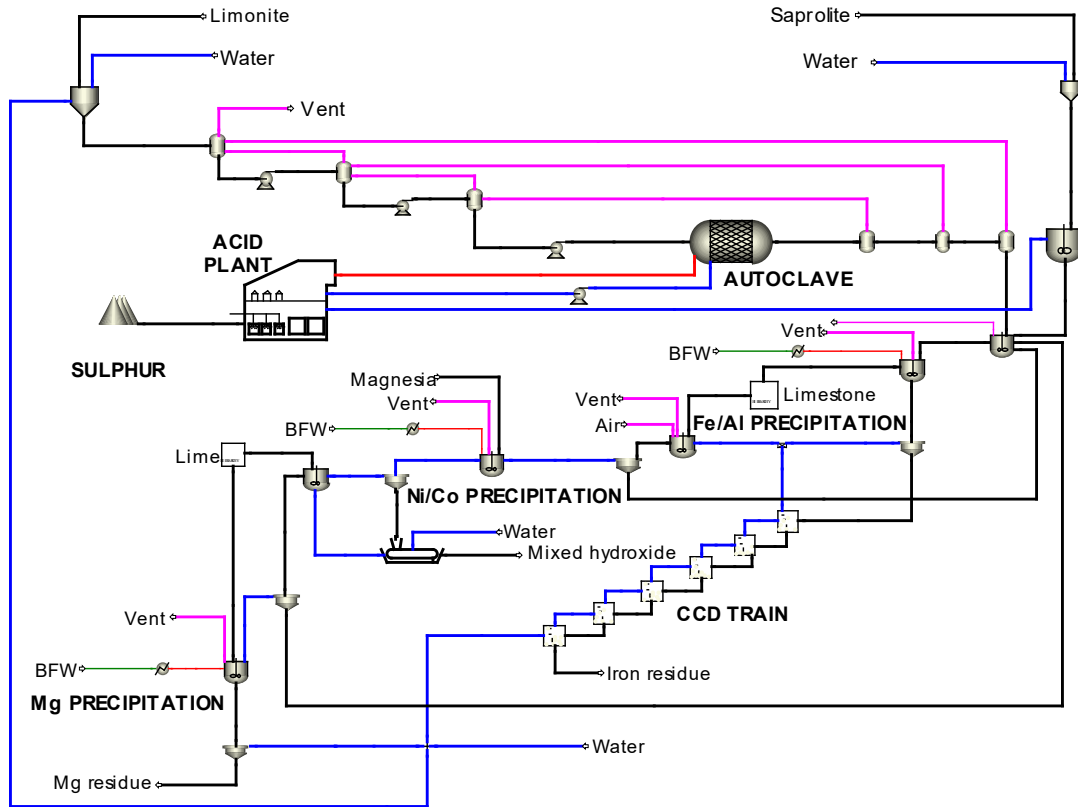
Figure 2 – EPAL model

Figure 2 – EPAL model

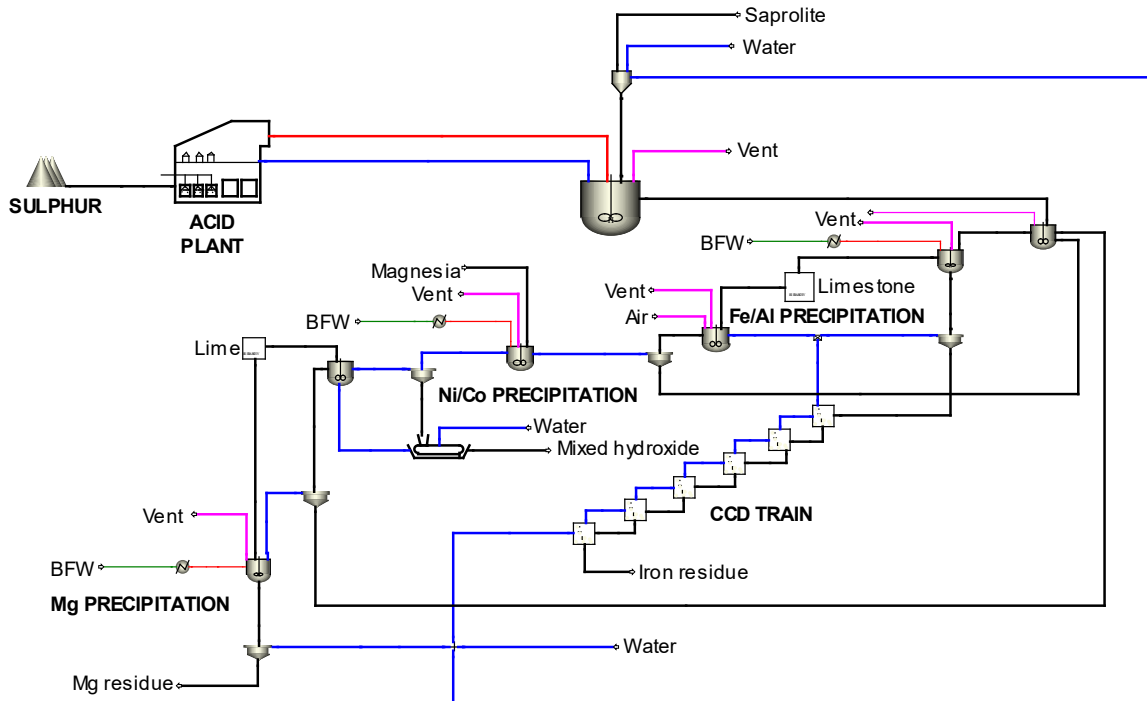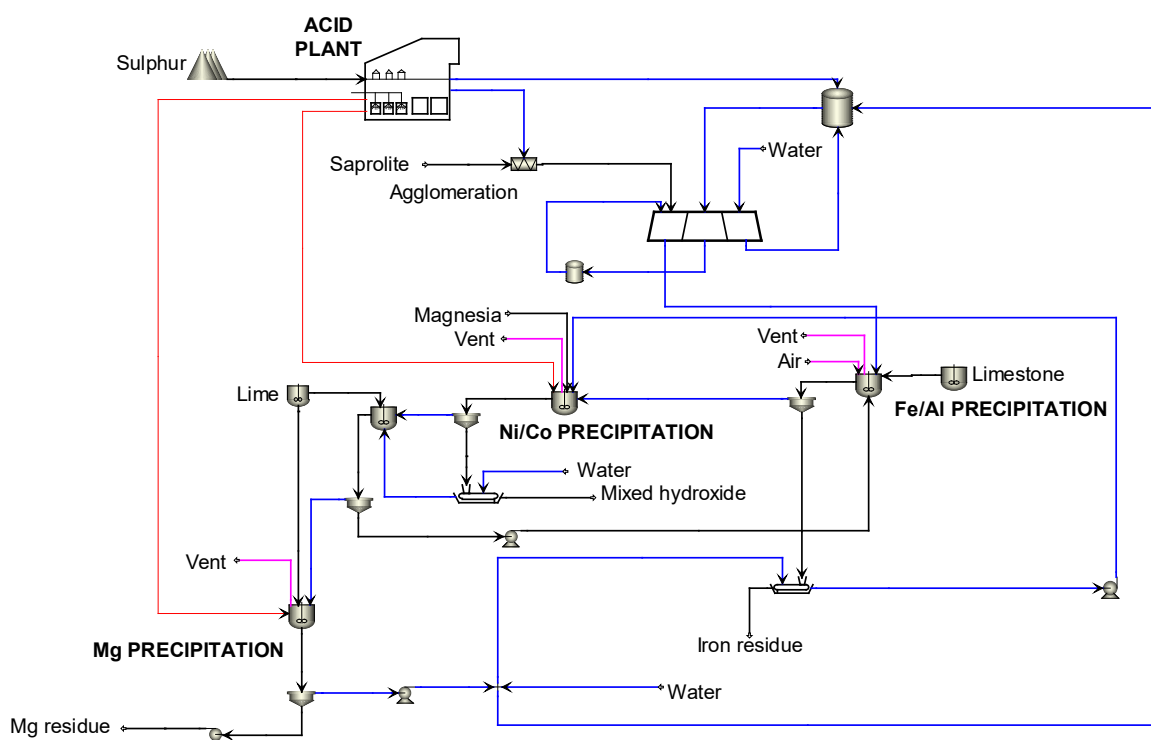## Figure 3 – ATL model

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.
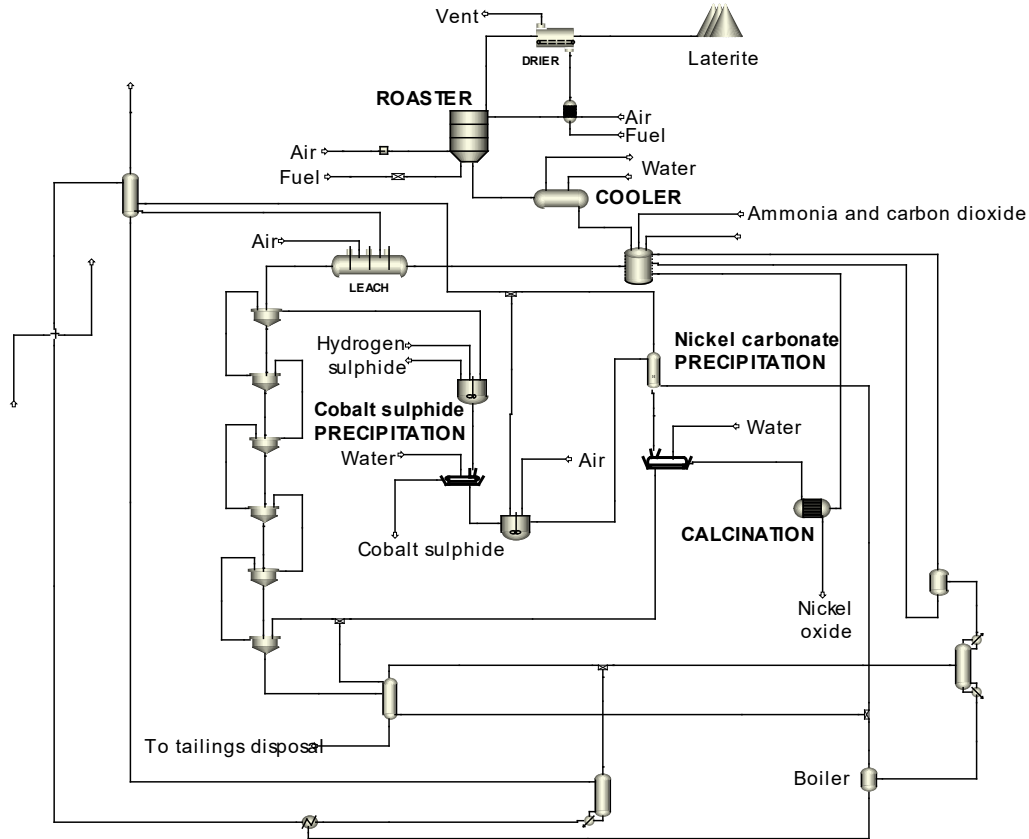
The cooled calcine is mixed with recycled solution of aqueous ammonia and ammonium carbonate and leached with that solution and air, causing the metallic nickel and cobalt to dissolve as ammine complexes. The metallic iron is converted into insoluble ferric oxide. The leach is done in a cascade of four 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 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 (using two agitated tanks in series, rubber lined steel, 30 minute total retention), 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 stripping is done in a carbon steel column, which was approximated for cost calculation as a distillation column containing ten equilibrium stages. The calcination equipment was approximated for cost calculation as a direct contact rotary drier fitted with a refractory lining and sized by the heat duty from the process model.

Figure 4 – Caron model

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 stripper was approximated for cost purposes as a stripping column made of carbon steel and containing forty theoretical stages.

The air leaving the leach, the steam, ammonia and carbon dioxide from the nickel precipitation stage and the spent air from the aeration stage after cobalt precipitation are combined and scrubbed with water to capture the ammonia and carbon dioxide before the remaining water-saturated air is discharged to the atmosphere. This is done in a scrubbing column made of carbon steel and containing fifty theoretical stages.
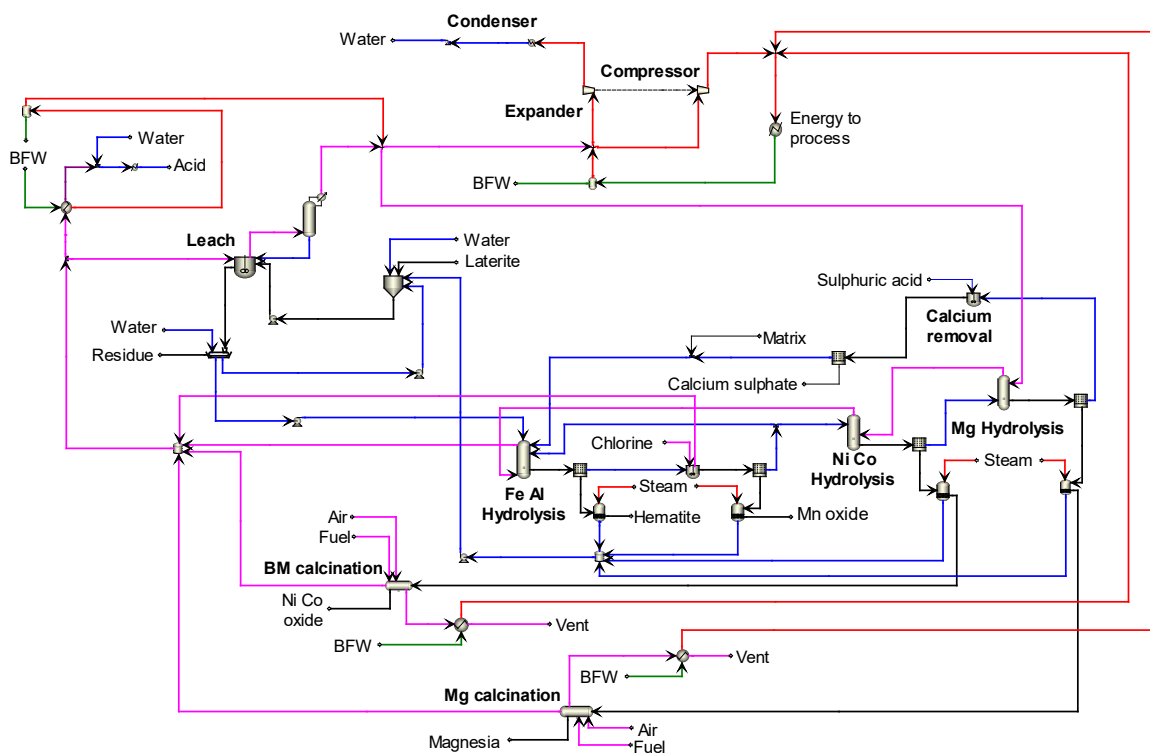
The dilute solution of aqueous ammonia and ammonium carbonate from the scrubber is stripped to recover the ammonia and carbon dioxide in a more concentrated gaseous form and to regenerate water for recycle to the scrubber, in a carbon steel stripping column containing twenty theoretical stages. The overhead vapour (steam, ammonia and carbon dioxide) joins the steam, carbon dioxide and ammonia stripped from the counter-current decantation underflow and separated by distillation into a concentrated ammonia/ammonium carbonate solution and water in a carbon steel distillation column containing twenty theoretical stages. The strong solution returns to the leach. Part of the water from this step is boiled in a carbon steel heat exchanger to raise steam for stripping the tailings and the rest joins the water going to the scrubber, along with make-up fresh water.

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 6 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.
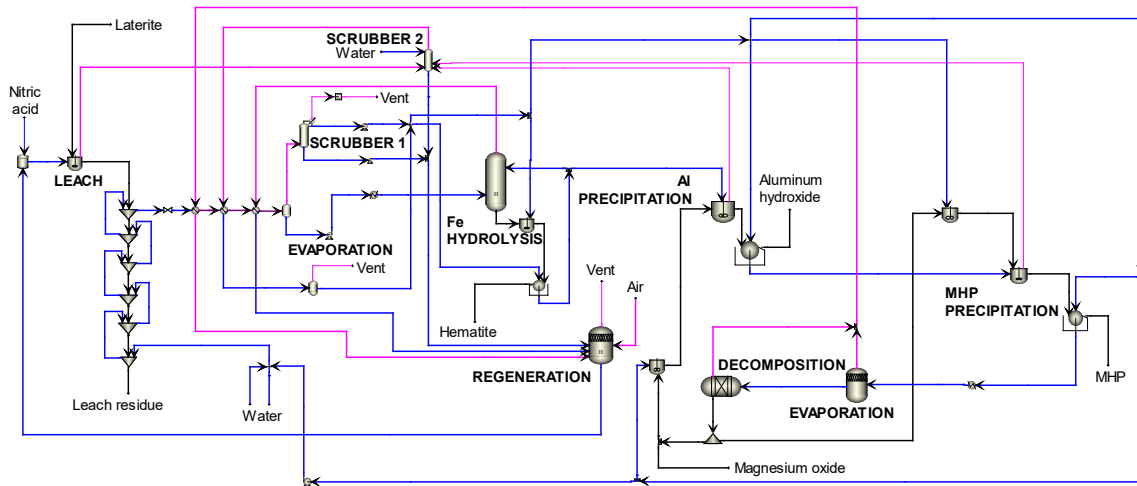
Direct Nickel (DNi)
The DNi process has also evolved since Part 2 of this series. Figure 7 illustrates the process model used to examine the DNi process, as that model presently stands. Except where stated otherwise, the material of construction throughout the DNi process is stainless steel (type 304).

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 NOx from the decomposition of magnesium nitrate, the second condenses steam from the above mentioned scrubber and the third exchanger condenses steam/HNO$_3$ 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 O_2, NO and NO_2. 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/NO_2 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 NOx 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. For cost calculations an agitated tank (15 minute retention) was assumed for contacting the NO/NO_2, air and aqueous acid, coupled with a vertical process vessel (10 minute retention) to represent the absorber capturing the gaseous NO/NO_2 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.
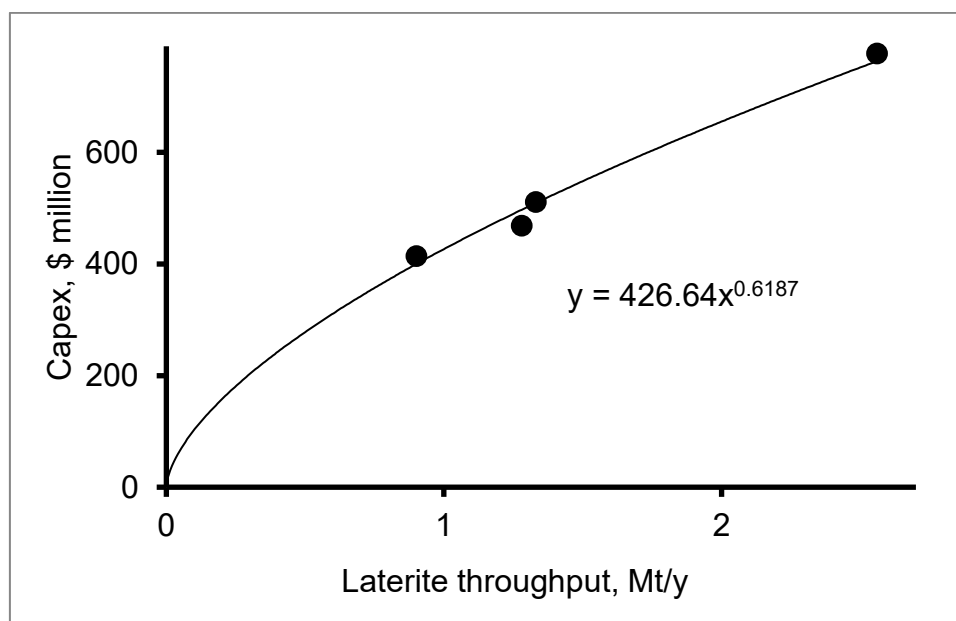
Sintering and submerged-arc smelting (SAF) is an established iron making technology that has been applied to making nickel pig iron. In the context of process modelling, it is similar to the RKEF technology. The incoming laterite is dried and then mixed with coal char and flux (CaO in this exercise) 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

$y = 426.64x^{0.6187}$

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 eleswhere[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.

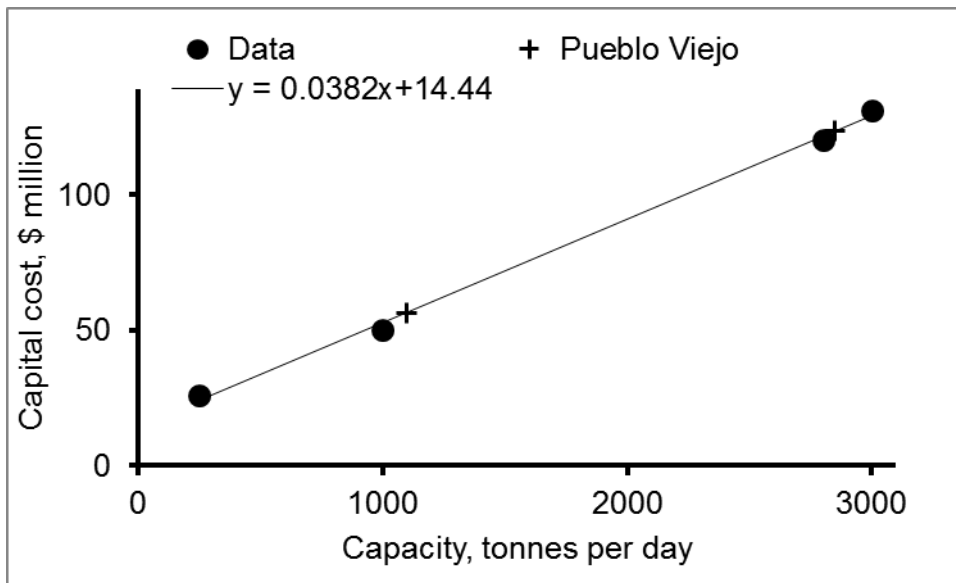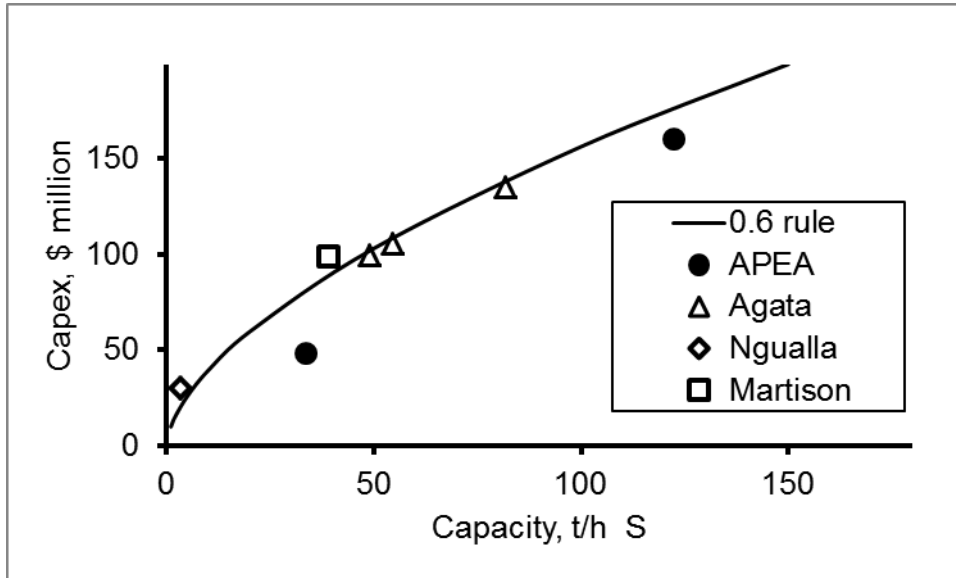**Figure 9 – Capital costs of conventional air separation plants**

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.

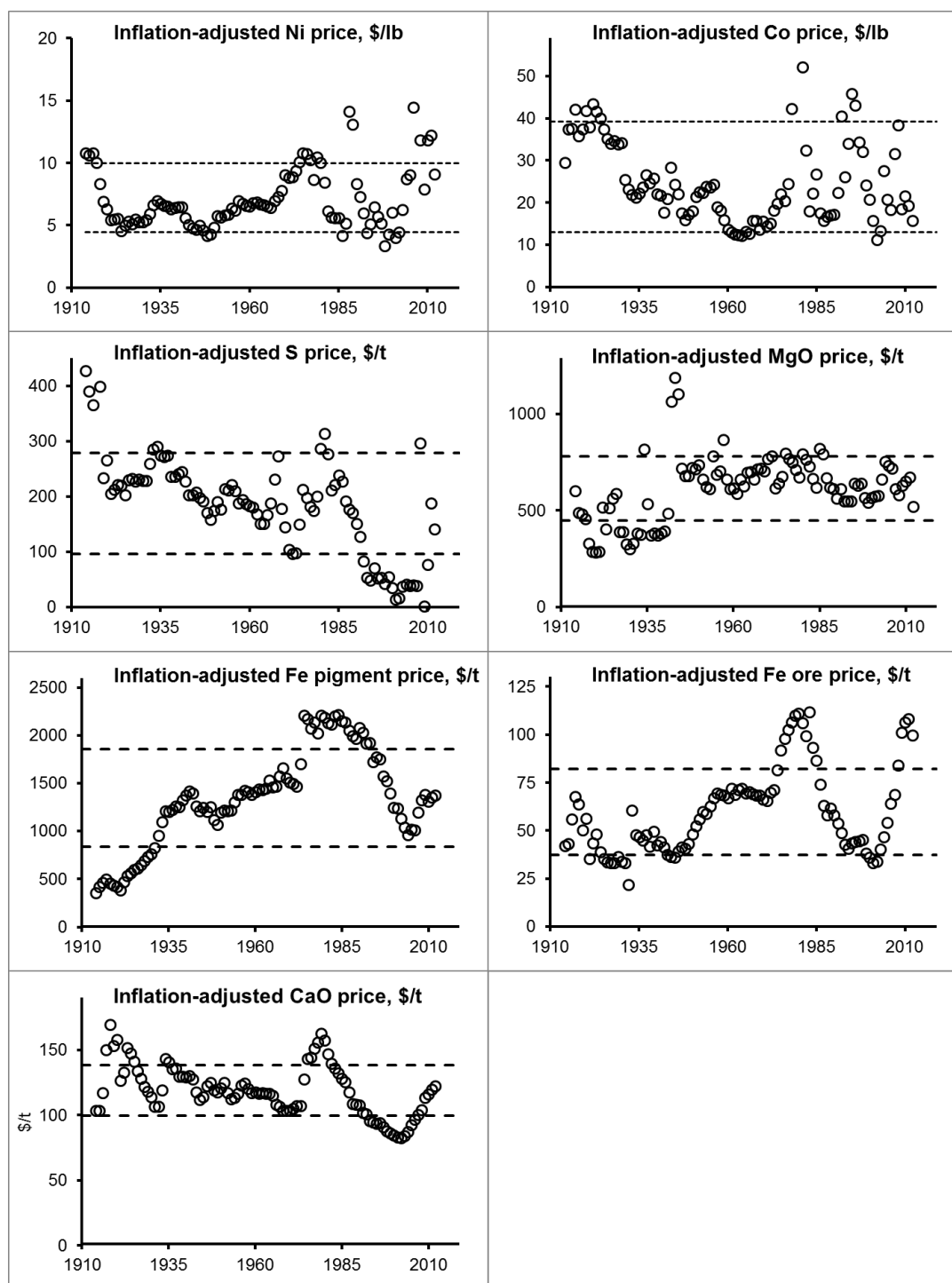**Figure 10 – Capital costs for sulphuric acid plants**

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.

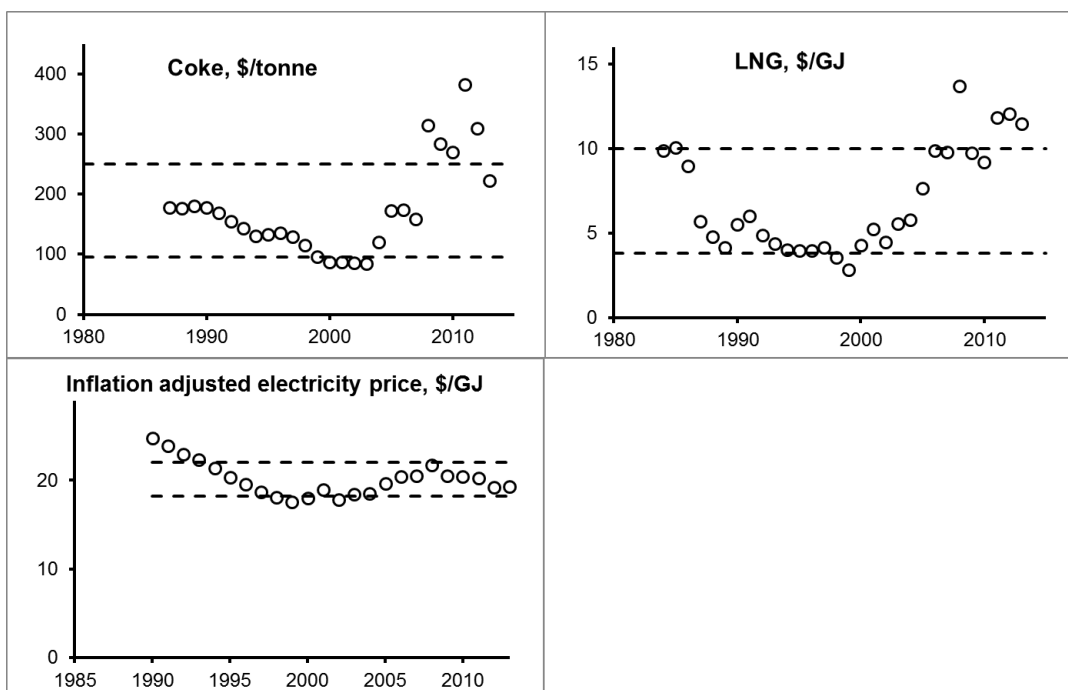## Figure 11 – Inflation-adjusted commodity prices (2014 dollars)

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.

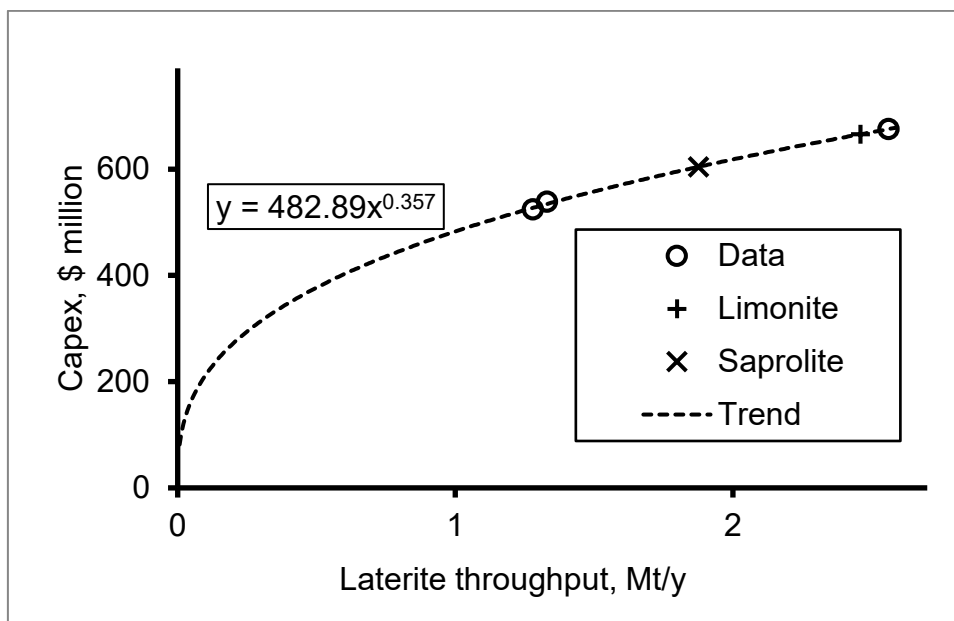**Figure 12 – Coke and LNG prices**

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_2O_3	87.2	87.3	87.9	93.7	93.7	94.5
Cr_2O_3	4.7	4.7	5.1	4.9	4.9	5.2
Al_2O_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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