

THE STARVED ACID LEACHING TECHNOLOGY (SALT) FOR NI/CO RECOVERY FROM LEAN ORES AND RESIDUES

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ABSTRACT

New sources of nickel production are required to meet the growing demand of the recovering world economy. The discovery rate on new nickel sulfide deposits is low. The promise of the HPAL treatment of limonite ores has largely been characterized by high capital and operating costs and poor ramp up rates. The conventional processing of high grade nickel saprolite deposits is typically by ferro-nickel or matte smelting. Low grade nickel saprolites are not treated. The Caron process for nickel laterite treatment produces residues that still contain significant quantities of un-leached nickel and cobalt.

The Starved Acid Leach Technology (SALT) was developed to recover nickel and cobalt from below cut-off grade saprolites and Caron plant residues. Relatively small amounts of acid are applied to ground saprolites or Caron residues in order to selectively leach nickel and cobalt. The leaching is performed atmospherically under low free acid conditions. Nickel and cobalt may be recovered from the leachate as a mixed hydroxide. The mixed hydroxide may be added directly to a saprolite smelting operation or refined separately. The simplicity of the SALT technology provides the benefits of low capital and operating cost and an expected rapid ramp up to full production.

A conceptual engineering study has been completed by SNC Lavalin for the SALT processing plant and associated residue storage facility. The study predicted a capital cost of \$282 MUS\$ for a plant to treat 2,000,000 tpa (dry) of below cut – off grade saprolite (at 1.33% Ni) for an overall recovery of 14,800 tpa of Ni in mixed hydroxide product. This equates to a cost of \$19,000 US/annual t of Ni production or \$8.66 US/annual lb of Ni production. The operating cost from ore to mixed hydroxide product was estimated to be \$3.60 US/lb of Ni.

INTRODUCTION

Commercial processes for laterite ore treatment vary according to laterite type and local conditions. The two general categories of laterite ore are saprolites and limonites. Saprolites are generally treated by pyrometallurgical processes for reduction of nickel to matte or ferro nickel alloys. The high cost of drying, chemical reduction and electric furnace smelting requires that the saprolite grade for economic processing is of the order of 1.8% nickel or higher. There are of course large amounts of saprolite that fall below this “cut-off” grade that are not currently economic. Limonites can be treated by the Caron Process which involves drying of the ore, a reduction roast of the whole ore to reduce nickel and cobalt (and some iron) to the metallic state followed by processing in ammonium – ammonium carbonate solution to leach nickel and cobalt. Nickel and cobalt recovery are low (70-85 % for Ni and 20-50% for Co). The Caron plant tailings therefore contain high levels of unrecovered nickel and cobalt. Relatively few plants have used the Caron Process and new installations are not anticipated due to the high energy cost of drying and reduction roasting. Nevertheless, plants in Brazil (Niquelandia), Australia (Yabulu) and Cuba continue to operate.

In contrast to the success of the smelting and Caron plant processes, hydrometallurgical extraction of nickel and cobalt has generally been marked by difficult plant startups and financial failure. The high pressure acid leach (HPAL) process was pioneered at Moa Bay in Cuba in the 1960's in the treatment of limonites. In the 1990's the HPAL process was touted to revolutionize the nickel industry with low capital and operating costs for new installations. Murrin Murrin, Cawse and Bulong in Australia were started and in each case the technical difficulties of operating at extreme temperature and pressure conditions and the challenges in “ramping up” a complex hydrometallurgical facility conspired to cause these plants to fail economically. At present, Cawse and Bulong are closed and Murrin Murrin continues to operate after undergoing a financial restructuring.

Since the beginning of this century, a number of additional HPAL plants have been built and are either operating (Coral Bay Nickel), or restarting (Ravensthorpe), or “ramping up” (Goro, Ambatovy and others). The success of these projects is varied. The Coral Bay facility in the Philippines is touted as a technical and financial success. The plant was constructed at a brownfields site with careful engineering and scaleup of largely known technology. The plant produces a mixed sulfide product of nickel and cobalt for refining by the Sumitomo refinery in Japan. The Ravensthorpe nickel laterite plant was constructed by BHP Billiton Stainless Steel Materials as a showcase for the Enhanced Pressure Acid Leach (EPAL) process incorporating upgrading of laterite, HPAL for limonite and atmospheric leaching of saprolites. The plant encountered technical difficulties during startup during a period of low nickel prices and BHP Billiton elected to close the facility. The plant has been sold to First Quantum Minerals who report (www.first-quantum.com);

First Quantum Minerals acquired Ravensthorpe as a decommissioned mine in February 2010 for US\$ 340 million and proceeded to make significant modifications and improvements to the process and facilities in the following 18 months. Nickel was first produced in October 2011 during commissioning and first exports were shipped on 25 November 2011.

This plant is expected to be technically and financially successful under First Quantum but following the Australian laterite plant pattern, lost substantially all of the investment for the original owner (BHP Billiton).

On April 25, 2001, INCO announced “INCO to Proceed with a US\$1.4 Billion nickel-cobalt project at Goro, New Caledonia”. This announcement was based on many years of metallurgical testwork, a \$50 Million USD pilot plant in New Caledonia and detailed engineering and cost studies. As the plant construction started, costs began to increase excessively and the project stopped. The project has been re-engineered and restarted and is currently in startup. The total capital cost of the facility has now risen to at least \$6 Billion USD for a facility to produce 60,000 tonnes of nickel per annum (when it reaches operating capacity). The specific investment is of the order of \$100,000 USD per annual tonne of nickel capacity. The ramp up is going slowly. The production of nickel was reported as 4,000 t in 2012 and 16,000 t in 2013 (7% and 27% of design respectively) (Vale production report dated February 26, 2014).

In a similar vein, the Ambatovy nickel project in Madagascar uses HPAL to process a limonite ore. The Ambatovy project has achieved commercial production (defined as 70% of ore throughput of nameplate capacity in the pressure acid leach circuit on average over a thirty-day period). Sherritt

reports on the capital costs for Ambatovy in the release of their 4th quarter results (dated April 30, 2014, www.sherritt.com).

“Ambatovy ceased capitalizing project costs on January 31, 2014. Cumulative spending on capital at Ambatovy was US\$5.3 billion (100% basis), excluding financing charges, working capital and foreign exchange, below the US\$5.5 billion (100% basis) estimate established in June 2011. Cumulative total project costs at January 31, 2014 (including operating costs, financing charges, working capital and foreign exchange, and net of sales revenue) were US\$7.2 billion (100% basis), with US\$49.9 million (100% basis) spent in January 2014.”

This report illustrates clearly how the costs associated with any delay in achieving commercial production can exacerbate an already excessively high capital cost. Significantly, Ambatovy is expected to have a period of continued “ramping” up before achieving nameplate capacity.

Based on this review, there are a number of critical technical and economic issues that need to be summarized.

1. Caron plant tailings represent a potential source of value. At a grade of 0.4-0.5% Ni and ~0.1% Co, there is significant in-situ value in previously mined, ground, surface tailings deposits at Caron facilities.
2. Economic processing of “below cut-off” grade saprolites represent an economic target for consideration. At 1.5% Ni grade, the in-situ value of below cut-off grade saprolites is \$330 USD/t using a nickel price of \$20,000/t.
3. The economic model for HPAL treatment of greenfield nickel laterite deposits seems fatally flawed. At greater than \$100,000 investment per annual tonne of nickel production, a combination of high nickel price and low operating cost are required for economic attractiveness. Long term nickel prices are not expected to be “high” and the operating costs for HPAL plants, while often projected to be low in engineering feasibility studies, seldom achieve this result.

Against this techno-economic backdrop there has been attempt to find a process that could be technically and economically successful. I.e. the process must be simple to engineer, construct, commission and “ramp up” and must do so at a profit.

Unfortunately, there are few examples in the nickel industry of any existing process with these characteristics. There is however a number of reference points for such a concept in other hydrometallurgical industries. The gold industry economically processes materials that are often even less than 1 g/t of Au content. This equates to an in-situ ore value of around \$40/t at current gold prices of \$1200-1400/ounce of gold (to be compared with \$330 USD of Ni/t of below cut-off grade saprolite ore). Perhaps closer to the mark is the case of the Sepon Copper Project in Laos, which finished construction and started in March 2005. Oxiana Limited were the developer of the project (successor companies include Oz Minerals and now MMG Resources). The Sepon project treats a weathered copper ore by atmospheric acid leaching [1,2] followed by SX-EW for cathode production. The project cost \$227 Million US for a production rate of 60,000 tpa of Cu. This equates to \$3,780/annual tonne of copper production, less than 4% of the current going rate for a nickel HPAL plant development. The plant was designed to treat 170 dry tonnes per hour of ore. This is equivalent to an investment of \$152/annual tonne of ore treated. The ramp up for Sepon was 2/3 of nameplate production for period April – December 2005 and then at or above nameplate from 2006 to present. The analysis presented here is simplified and building a nickel recovery plant is not the same as either a copper or a gold plant. However, the lessons learned from these examples in the gold and copper industries ought to provide some guide to development of a future process for treatment of nickel ores.

The Starved Acid Leach Technology (SALT) has been developed in order to (1) meet the opportunities related to Caron Plant tailings and (2) develop a method to treat “below cut off” grade saprolite ores [3, 4]. The technology has progressed from the bench to the pilot plant. InCoR Technologies Limited (a subsidiary of InCoR Holdings Plc. (www.incorholdings.com)) acquired the SALT technology from Search Minerals in November of 2013. InCoR Technologies is actively pursuing the application of SALT at a number of nickel processing sites worldwide.

The results of the initial bench testing of the SALT leaching process applied to Caron Plant tailings and below cut-off grade saprolites are described in this paper. The results of an economic study for application of SALT to below cutoff grade saprolites originating from the Pomalaa deposit in Indonesia are summarized.

STARVED ACID LEACHING TECHNOLOGY DEVELOPMENT

SALT development has proceeded on the bench scale in two steps. Leaching tests have been conducted on (1) Caron Plant tailings from the Votorantim plant at Niquelandia in Brazil and (2) below cut-off grade saprolite samples from PT ANTAM's Pomalaa deposit in Indonesia. These samples were obtained under a "Heads of Agreement" between Search Minerals and PT ANTAM to cooperate on the development of SALT, prior to the agreement by InCoR Technologies to purchase SALT.

Caron Plant Tailing Leaching

A sample of Caron plant tailings was obtained from the Votorantim Niquelandia site in Brazil. The analysis is shown below. The sample was tested as received without milling.

Table 1: Caron Plant Tailings Sample Analysis

SiO ₂	Cu	Zn	Co	Ni	Cr	Mn	Al	Ca	Fe	Mg
%	ppm	ppm	ppm	ppm	ppm	ppm	%	%	%	%
29.76	1144	436	538	5072	14364	7452	3.39	0.44	30.86	3.40

A 100 g sample of tailings was slurried with 1 L of solution containing 10, 25, 50 or 100 g/L of sulphuric acid. This corresponds to acid additions of 100, 250, 500 or 1000 kg/t. The slurry was heated to the target temperature and leached for 48 h under stirring conditions in a 2 L leach reactor. The temperature was maintained constant at 50 °C for the first 4 experiments (at 4 acid levels) and then increased from 50 to 80 °C while maintaining 100 kg/t acid addition and addition of oxygen (to oxidize ferrous to ferric). Samples of slurry were taken at 1, 2, 4, 8, 24, 32 and 48 h. Each of the intermediate samples was filtered to produce a solution for assay. At the end of the 48 h the remaining slurry was filtered and washed. The filtrate and wash solution were collected separately. The washed solid residue was dried and weighed. Note that for the first 4 tests with variable acid addition, the solids removed from the intermediate samples were NOT returned to the leach. This procedure was modified for the final 4 tests to ensure that all solids were leached for the full 48 h.

All solution and solid samples were sent to International Plasma Laboratory (IPL) in Vancouver for analysis. The solids were analyzed using a multi-acid digestion/whole rock assay suite and the solutions were analyzed by the ICP 30 water package analysis.

An example of the type of data obtained from leaching of Caron plant tailings is summarized in Table 2 below. The nickel and cobalt leaching results of the 8 acid leach experiments are summarized in Table 3 below.

Experiments 1-4 show increasing nickel and cobalt extraction with increasing acid addition over the range of 100-1000 kg/t of tailings. Clearly addition of acid to ever higher levels produces only an incremental benefit in overall metal extraction. At 10 X the acid addition (1000 kg/t versus 100 kg/t) the nickel extraction doubles and the cobalt extraction is only about 50% higher. The iron extraction rises from 5% to 89% over the same range of acid addition. The terminal pH and calculated acid consumptions (for dissolved species) indicates that most of the acid was consumed in each case. As the acid addition was increased to 1000 kg/t, the main consequence was the dissolution of iron.

For the temperature series with oxygen addition the calculated acid consumption is around 100 kg/t (same as addition) but does decrease slightly with increasing temperature. This is consistent with greater re-precipitation of iron at higher temperature, likely as hydronium jarosite, which would account for additional acid consumption (but in the solids rather than in solution). The major soluble-species acid consumption is due to iron, aluminum, magnesium and manganese. The magnesium component of acid consumption is relatively minor in the case of leaching of Caron plant tailings due to the relatively low levels of magnesium in the solid feed.

Table 2: Results of the Caron Tails Leaching Test with 100 kg/t Acid Addition. 80 °C and 48 h leach time.

			Analysis (% or mg/L)									
			Al	Ca	Co	Cu	Fe	Mg	Mn	Ni	Zn	
Input			Mass (g)/Vol (mL)									
Feed Sample			100	3.39	0.44	538	1144	30.86	3.40	7452	5072	436
Units (g)			Total	3.390	0.440	0.054	0.114	30.860	3.400	0.745	0.507	0.044
Output												
Residue			88	3.42	0.43	213	365	35.28	3.12	3453	3425	338
Solutions	pH (80°C)	Eh (80°C)										
1	0.82	576	10.0	332	25	26	54.5	1398	453	334	127.8	2.8
2	2.01	541	10.0	442	31	35	71.8	1724	589	423	170.3	3.7
3	2.05	522	10.0	467	29	35	73.8	1774	642	450	190.4	4.2
4	2.10	506	10.0	481	29	37	75.0	1723	696	444	200.8	5.1
5	2.27	468	10.0	472	31	37	75.1	1442	838	466	227.9	4.7
6	2.39	450	10.0	481	32	38	77.1	1379	983	516	264.8	4.9
7	2.39	449	800.0	446	32	37	72.9	823	1100	560	290.0	4.9
8	2.41	414	1150.0	14	6	1	3.0	34	37	17	8.9	0.3
Units (g)												
Residue				3.010	0.378	0.019	0.032	31.046	2.746	0.304	0.301	0.030
Solutions				0.003	0.000	0.000	0.001	0.014	0.005	0.003	0.001	0.000
1				0.004	0.000	0.000	0.001	0.017	0.006	0.004	0.002	0.000
2				0.005	0.000	0.000	0.001	0.018	0.006	0.004	0.002	0.000
3				0.005	0.000	0.000	0.001	0.017	0.007	0.004	0.002	0.000
4				0.005	0.000	0.000	0.001	0.014	0.008	0.005	0.002	0.000
5				0.005	0.000	0.000	0.001	0.014	0.010	0.005	0.003	0.000
6				0.357	0.026	0.030	0.058	0.659	0.880	0.448	0.232	0.004
7				0.017	0.006	0.002	0.003	0.039	0.042	0.019	0.010	0.000
8												
			Total	3.410	0.412	0.052	0.098	31.838	3.710	0.797	0.555	0.034
			Balance	-0.6%	6.3%	2.9%	14.2%	-3.2%	-9.1%	-7.0%	-9.5%	21.5%
			Extraction (%)	11.74	8.18	64.11	67.27	2.49	25.99	61.89	45.74	13.10
			Acid Cons. (g)	2.18	0.08	0.06	0.10	2.09	3.89	0.88	0.42	0.01
				22.45%	0.82%	0.62%	1.03%	21.52%	40.06%	9.06%	4.33%	0.10%
			Total Acid Cons.	9.7	g							
				97.1	kg/t							

Table 3: Starved Acid Leach Experiments Treating Caron Plant Tailings

Test	T (°C)	Acid (kg/t)	O ₂ Addn	Ni Ext (%)	Co Ext (%)	Fe Ext (%)	Final pH
1	50	100	No	36.51	65.34	5.00	2.93
2	50	250	No	52.93	76.47	23.75	2.28
3	50	500	No	61.94	82.10	52.72	1.68
4	50	1000	No	72.19	91.32	88.92	0.94
5	50	100	Yes	41.90	64.56	4.49	2.58
6	60	100	Yes	45.74	64.11	2.49	2.41
7	70	100	Yes	51.13	69.73	1.67	2.32
8	80	100	Yes	53.29	70.66	0.93	2.41

The Ni, Co, Zn, Cu, Mn levels increased during each leach experiment. The Fe levels in solution peak and stay high for most of the first series of four experiments with variable acid addition. For the second series of four higher temperature experiments (with oxygen sparging) there was a marked decrease in the iron in solution with time due to oxidation and precipitation as hydronium jarosite. The jarosite precipitation is a favourable reaction as the acid released upon Fe precipitation continues to leach Ni and other metals. The nickel extraction rose with increased temperature of leaching.



Saprolite Leaching

A series of 9 saprolite samples were obtained from PT ANTAM's Pomalaa deposit in Indonesia and tested. Table IV shows the composition of the major elements. The nickel content ranged from 0.97 to 2.05%. The magnesium content ranges from 12 to 21%, consistent with saprolite mineralogy. The iron levels are relatively low compared to the Caron plant tailings.

Table 4: Chemical Analysis (%) of Saprolite Samples Tested

Sample	Ni	Co	CO ₃	SiO ₂	Al	Ca	Cr	Fe	Mg	Mn
1	2.05	0.025	0.03	38.31	0.85	0.68	0.43	10.08	12.75	0.15
2	1.68	0.011	0.06	43.17	0.44	0.33	0.19	5.32	20.60	0.07
3	1.25	0.015	<0.01	43.18	0.70	0.68	0.33	7.62	17.27	0.13
4	1.04	0.015	0.10	42.12	0.47	0.57	0.24	7.60	21.01	0.13
5	1.12	0.017	<0.01	47.81	0.63	0.81	0.31	8.37	12.35	0.14
6	1.02	0.014	<0.01	42.73	0.48	0.67	0.21	7.05	19.51	0.12
7	1.35	0.023	<0.01	44.52	0.79	0.64	0.37	10.35	13.94	0.19
8	1.45	0.018	<0.01	43.27	0.51	0.44	0.27	8.81	17.59	0.15
9	0.97	0.016	0.23	42.13	0.46	0.67	0.20	7.50	19.68	0.13

The saprolite samples were leached in the same way as the Caron plant tailings above. A volume of 1 L of solution containing sulfuric acid and 250 g of dry saprolite sample were mixed in an aerated reactor for 8 hours at 85 °C. Samples were taken periodically and analyzed for dissolved metals. The acid addition values ranged from 150-450 kg H₂SO₄/t. The initial particle size was approximately 50 µm (P80).

Saprolite Leach Results. The results of the testing are shown in Figures below. Each of the nine samples has a particular "fingerprint" with respect to nickel and acid addition. In general, the nickel extraction is equal to or greater than the magnesium extraction. This is consistent with some portion of the ore consisting of "available" nickel and the balance in the magnesium silicate ore matrix. Sample 1 has the biggest difference between the nickel and magnesium curves (nickel extraction is approximately 20% higher than magnesium extraction). The cobalt extraction is variable but generally trends with nickel extraction. The iron extraction is low and in many cases negligible, indicating that the host ore matrix is highly basic, leading to iron hydrolysis and precipitation.

These test results confirm the principle of SALT leaching – i.e. that small amounts of acid can leach a significant fraction of the available nickel and cobalt in Caron plant tailings and below cut-off grade saprolite ores.

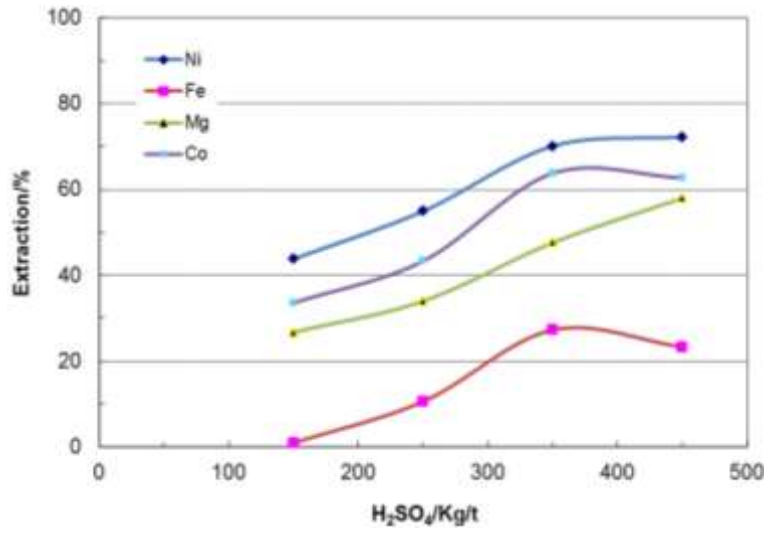


Figure 1: Metal Extraction Versus Acid Addition for Sample 1.

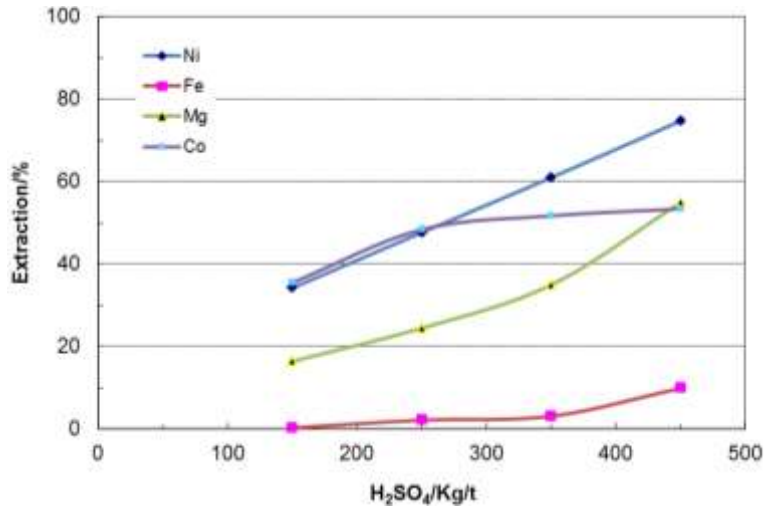


Figure 2: Metal Extraction Versus Acid Addition for Sample 2.

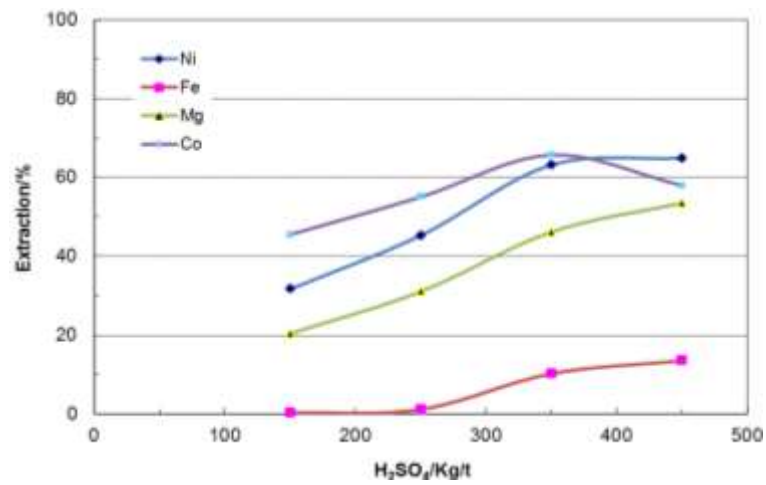


Figure 3: Metal Extraction Versus Acid Addition for Sample 3.

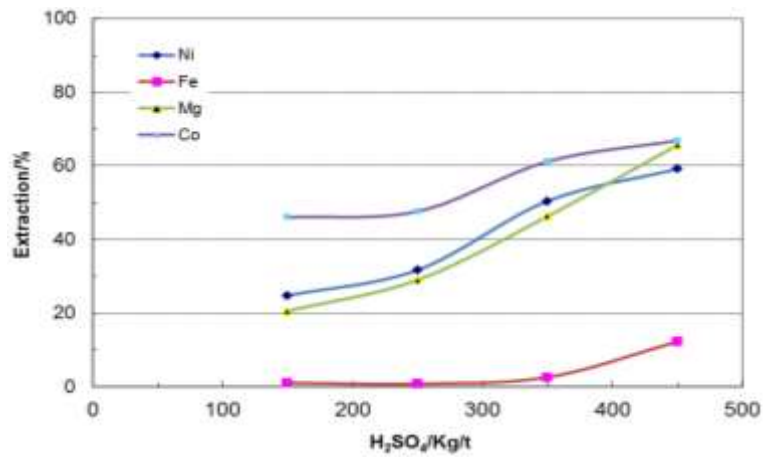


Figure 4: Metal Extraction Versus Acid Addition for Sample 4.

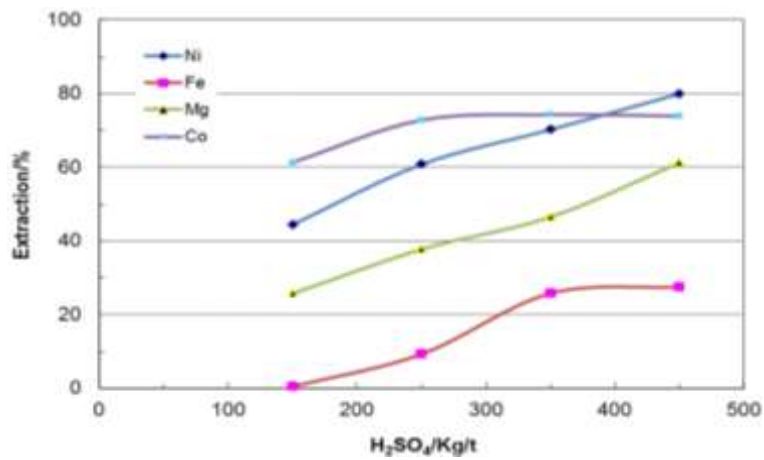


Figure 5: Metal Extraction Versus Acid Addition for Sample 5.

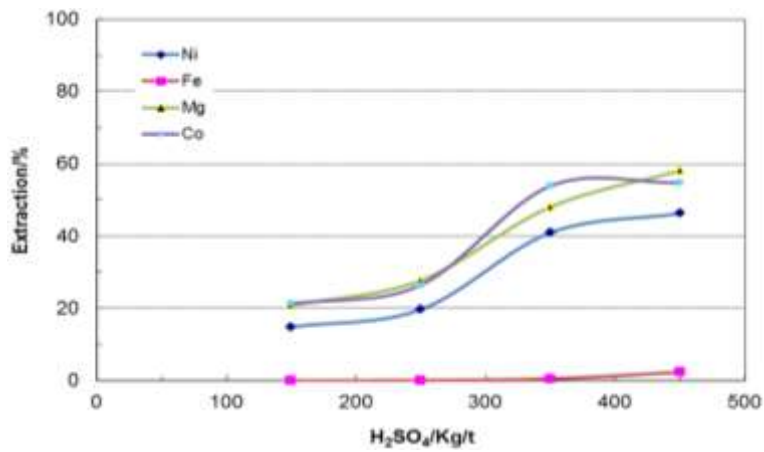


Figure 6: Metal Extraction Versus Acid Addition for Sample 6.

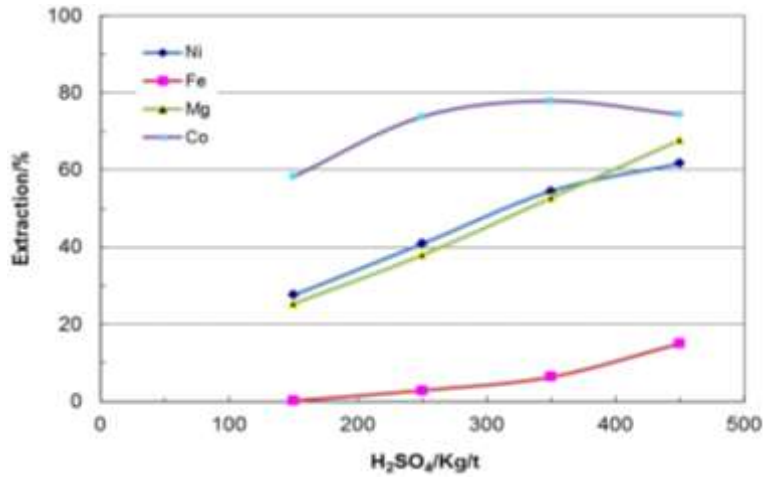


Figure 7: Metal Extraction Versus Acid Addition for Sample 7.

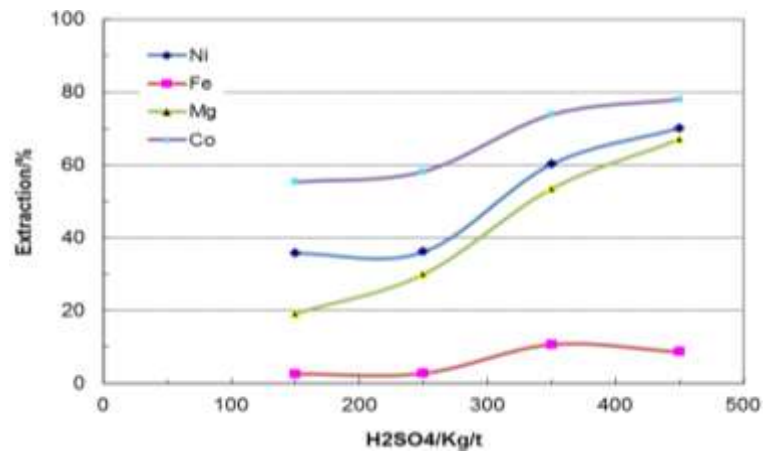


Figure 8: Metal Extraction Versus Acid Addition for Sample 8.

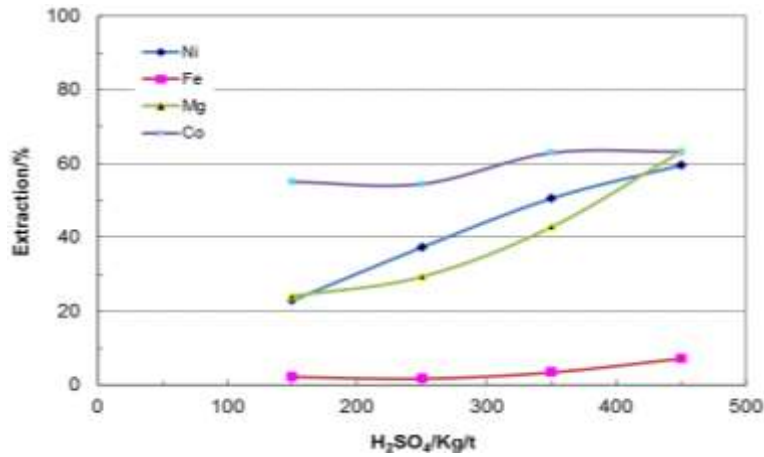


Figure 9: Metal Extraction Versus Acid Addition for Sample 9.

SALT FLOWSHEET DESIGN

SNC Lavalin has prepared a scoping level report on the application of SALT technology at the PT ANTAM site at Pomalaa. This site was selected in light of the test data available on leaching of a range of Pomalaa samples under the historical cooperation between PT ANTAM and Search Minerals to collaborate on the development of the SALT process. The project location is shown in Figure 10.

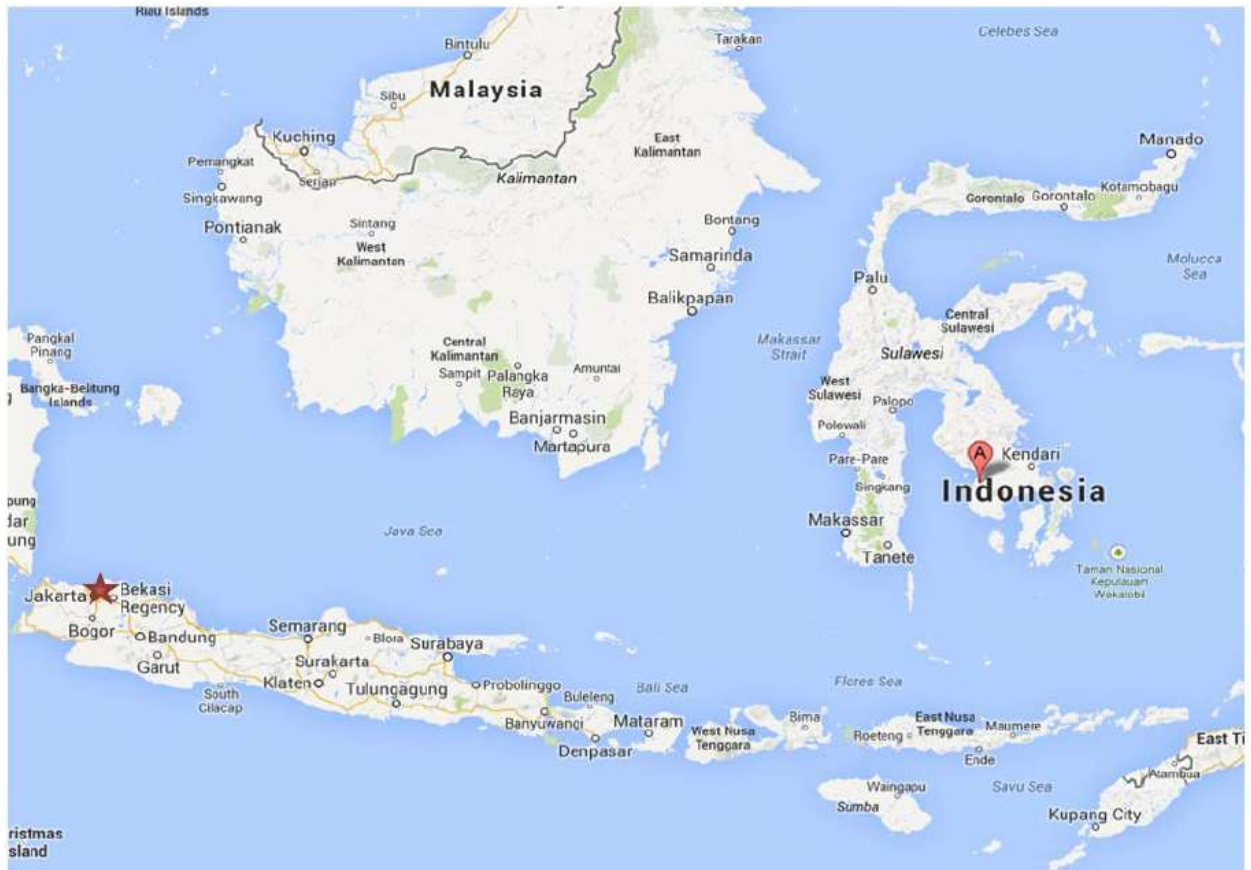


Figure 10: Pomalaa Site Location for SNC Lavalin Study

The basis for engineering design was the treatment of 2 MTPA (dry) of below cut of grade saprolite ore. The ore contains 1.33% Ni, 0.02% Co, 8.1% Fe and 17.2% Mg as an average of the samples received. Based on the testwork reported above, the extraction levels for SALT were set at 57.9% (Ni), 65% (Co), 10% (Fe) and 46.5% (Mg) at an acid addition rate of 350 kg/t of dry ore. The overall recovery to a mixed hydroxide product was predicted to be greater than 99% for Ni and Co. Based on these parameters, the plant would produce 116,600 tpa of wet mixed hydroxide product containing 14,800 tpa of Ni and 214 tpa of Cu. The overall nickel recovery from ore to MHP was 55%.

The flowsheet for the SALT plant is shown in Figure 11. The key steps are feed preparation, SALT leaching, pre-neutralization, CCD washing, MHP precipitation stages 1 and 2 (with MgO and CaO respectively), MHP thickening, washing and filtration, manganese removal and final environmental treatment of the tailings. These steps are described below.

Ore Preparation

The ore preparation circuit prepares the slurry for leaching. The composition of the saprolite feed is shown in Table 4. The ore is assumed to carry 20% moisture. The process includes crushing of the whole ore and grinding to increase the surface area available for leaching. The ground ore is then classified and thickened before feeding to the SALT circuit.

Table 4: Saprolite Feed Composition (dry)

Ni	Co	Al	Ca	Cr	Fe	Mg	Mn	SiO ₂
1.33	0.02	0.59	0.61	0.29	8.10	17.20	0.14	42.7

The overall ore preparation flow is shown in Figure 12. The ore is ground to a P80 of 105 μm and thickened to 40% solids in a deep bed paste thickener. A series of two surge tanks provides 12 h of capacity between the ore preparation and SALT leaching circuit.

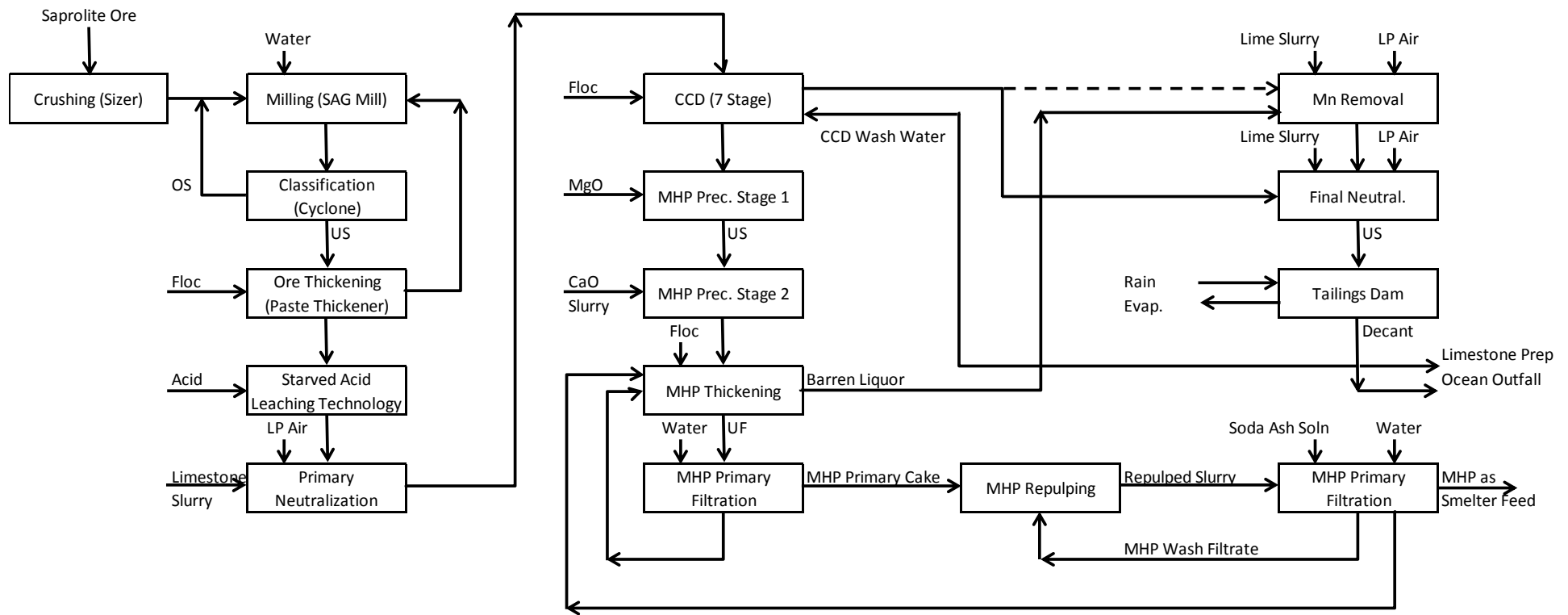


Figure 11: SALT Process Plant Block Flow Diagram

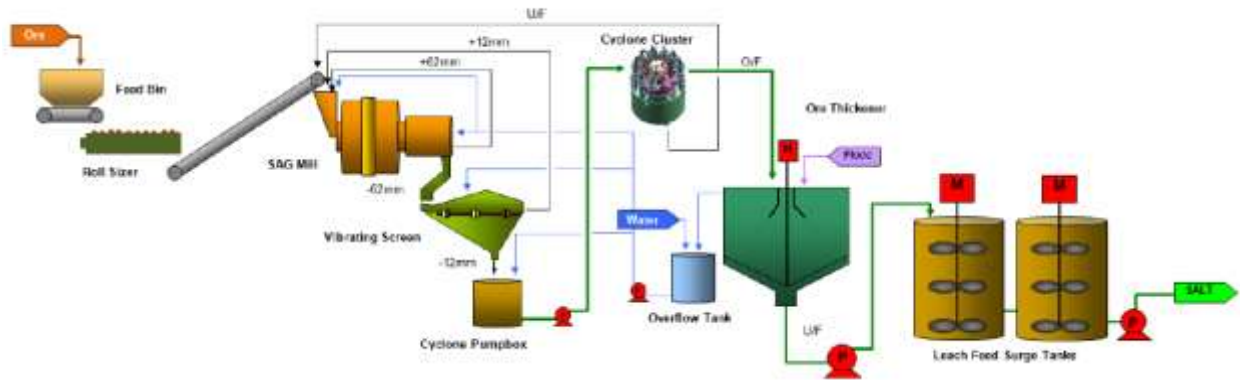


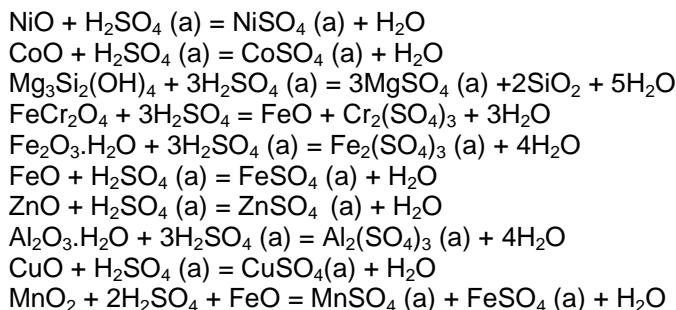
Figure 12: Ore Preparation Circuit

SALT Leaching Circuit

The starved acid leaching technology extracts the nickel and cobalt values from the saprolite ore by leaching with restricted amounts of sulphuric acid at elevated temperature under atmospheric conditions. By minimizing the acid addition, nickel extraction is reduced but with the enormous benefit of reduced extraction of iron, magnesium, aluminum, etc.

The thickened slurry from the Leach Feed Surge Tanks is pumped to SALT circuit for leaching with concentrated sulphuric acid (98.5% w/w H₂SO₄). The leaching takes place in two brick-lined reactors followed by two SAF 2205 alloy reactors in series. A total of 8 hours retention time is allowed. All acid is added to the first reactor. Tanks are designed to maximize effective leaching and minimize short circuiting using upcomers and baffling. The simplified circuit drawing is shown in Figure 13.

The exothermic reaction generated by the acid addition during mixing, increases the temperature of the process to about 100-105°C. Hence, no additional steam is required, except during start-up. The key reactions for SALT leaching are shown below.



Brick-lined carbon steel materials were chosen for the first two SALT tanks due to the potential for incomplete mixing resulting in pockets of high acidity. Although all acid is added in the first tank, the second tank is treated in a similar manner for cases when the first reactor is offline for maintenance. The brick-lined tanks are more than double the costs of SAF 2205 tanks, but considerably more resistant should there be any localised high acid concentrations. The SALT leach slurry is directed to primary neutralization.

Primary Neutralization of the Leach Slurry

The primary neutralization circuit neutralizes the free acid in the SALT leach slurry and precipitates iron, aluminium and chromium with limestone addition. Air is injected to oxidize ferrous to ferric and enhance the removal of iron.

The circuit is comprised of four SAF 2205 agitated tanks, with a total reaction time of 160 minutes to ensure effective neutralisation. Limestone slurry is added to achieve pH 3.8 while the slurry temperature remains close to 100°C

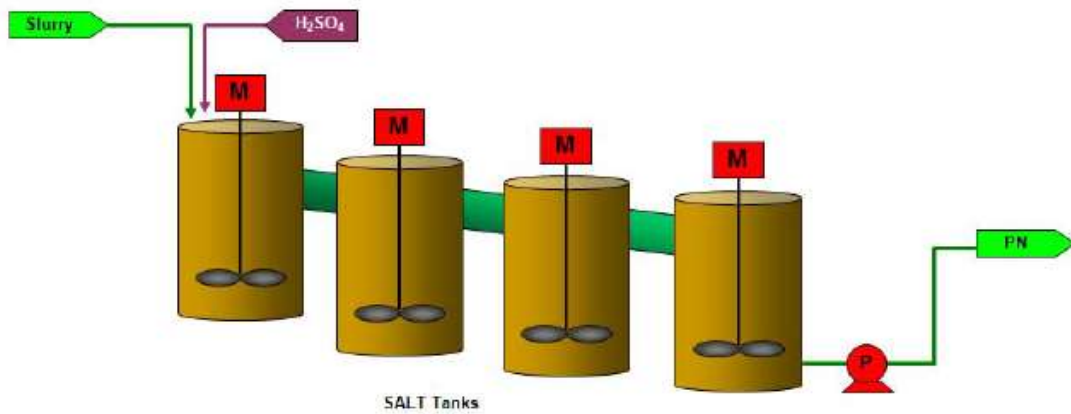


Figure 13: SALT Leaching Circuit

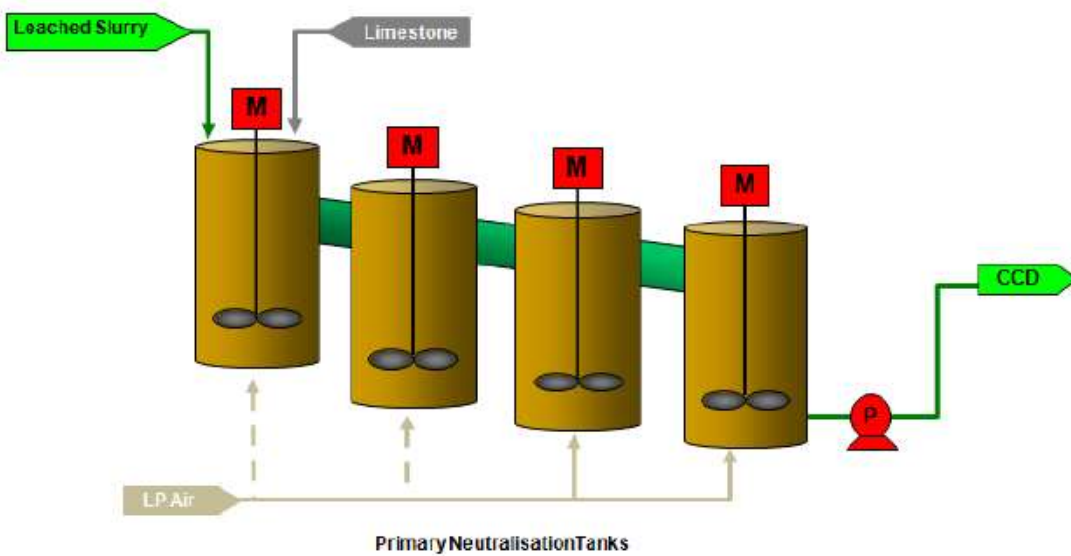


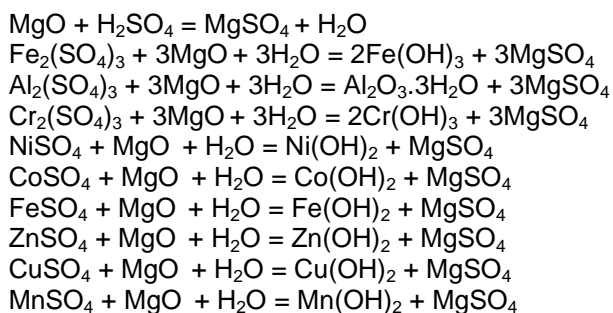
Figure 14: Primary Neutralization

Counter Current Decantation (CCD)

The counter current decantation (CCD) circuit washes the leached solids and recovers a pregnant leach solution (PLS) containing nickel and cobalt values (Figure 15). A 7-stage CCD circuit has been specified with 40% solids in each thickener underflow. The CCD 7 underflow is directed to neutralization and tailings disposal. The CCD 1 overflow is collected as the final PLS solution for mixed hydroxide recovery.

Mixed Hydroxide Precipitation

The PLS solution from the PLS pond is reclaimed to a series of reactors for mixed hydroxide precipitation. After pre-mixing the PLS with a recycle of “seed” from the MHP thickener, the first stage precipitation is performed with magnesia and the second stage with lime. The first stage reactions are shown below.



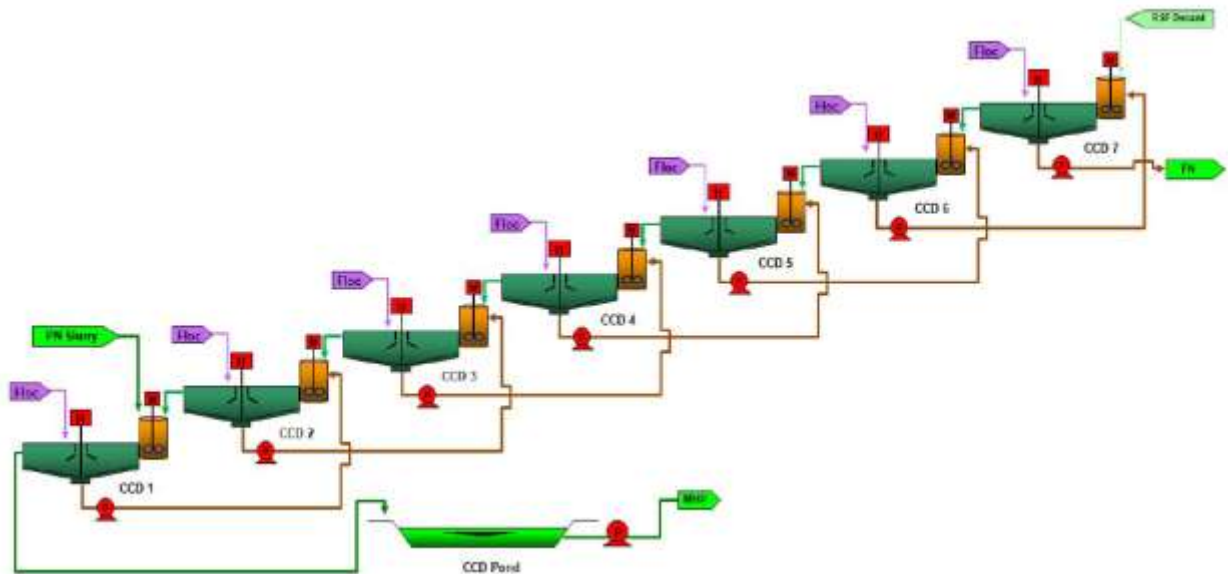


Figure 15: The Counter Current Decantation Circuit

MHP Stage 1 is comprised of three reactors in series with a total of 60 minutes retention time. Approximately 91% of the nickel and 95% of cobalt in the Pre-mix discharge solution are precipitated in Stage 1 along with ferrous, manganese, aluminium, copper, chromium and zinc.

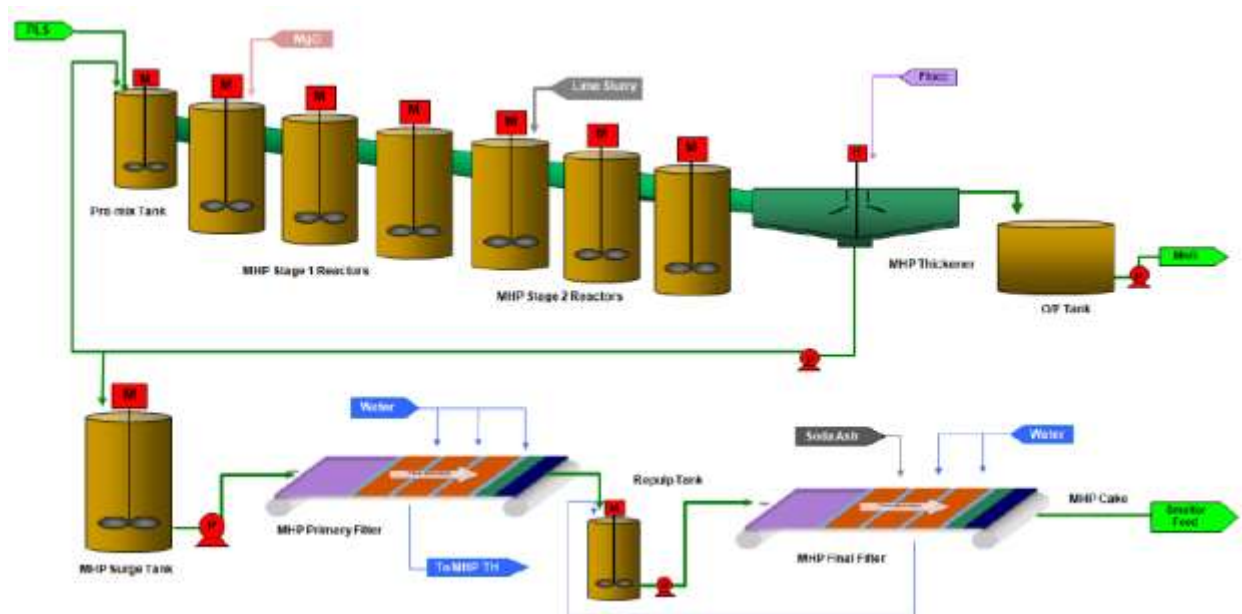
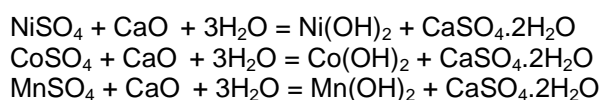


Figure 16: The Mixed Hydroxide Precipitation Circuit

Subsequently, the discharge slurry from reactor 3 overflows by gravity to MHP Stage 2 reactors. Lime slurry is added as precipitant to scavenge the remaining aqueous nickel and cobalt. MHP Stage 2 is composed of three reactors in series with a total of 60 minutes retention time. More than 99% of the residual nickel and cobalt in the Stage 1 barren solution are recovered along with the majority of manganese. The stage 2 reactions are shown below.



The MHP Stage 2 discharge slurry is thickened to 20% solids. The overflow is directed to a storage tank and then on to the manganese removal circuit. The underflow is split. The first part is returned as a seed to the MHP circuit and the second part is filtered, repulped and filtered again. The second filtration includes provision for washing with soda ash to remove sulfate from the solids. The wash

liquor from the second filtration stage is recycled to the repulp tank to maximize the efficient use of soda ash for sulfate removal. The final filtered MHP product at 45% solids is conveyed to the ferro nickel smelter complex for smelting to ferro nickel product.

The commercial product generated from the SALT plant is a Mixed Hydroxide Precipitate in the form of a filter cake. This was expected to be fed into the neighbouring PT Antam Ferronickel Smelter for refining (under the historical PT ANTAM – Search Minerals Agreement). The MHP filter cake product has the following composition (Table 5). This composition can be varied by decreasing the amount of magnesia added and compensating by an increased addition of lime in the MHP area. This will result in a lower nickel and higher sulphur in the product.

Table 5: MHP Composition

Component	Composition (Wt. %)
Ni	28
Co	0.4
Mn	3.3
Mg	3.4
Fe	3.9
Al	1.7
Cr	0.2
Cu	0.1
Si	1.9
Na	0.1
Zn	0.1
S, as gypsum	3.9
S, as basic nickel sulfate	1.0
Moisture (wet % w/w)	55

Manganese Removal and Final Neutralization

The barren liquor from MHP Thickener Overflow Tank and other minor waste solution streams are pumped to the Manganese Removal circuit, where lime slurry is added to precipitate the manganese at pH 9.5-10. At this pH range, other heavy metals such as nickel, cobalt, chromium, copper and zinc are also removed as stable hydroxide precipitates. Air is aerated into the tanks to precipitate a portion of the manganese as manganese dioxide, potentially reducing the lime consumption. The MnR discharge slurry is expected to contain a residual concentration of <2 mg/L Mn in the solution. A portion of the CCD 7 U/F is used as seed during the precipitation process. Manganese removal is conducted in a series of three tanks with a total retention time of 90 minutes. The discharge slurry exiting the last tank flows by gravity to the Final Neutralisation circuit for mixing with the CCD 7 U/F.

The CCD 7 U/F is pumped to a series of three aerated final neutralisation reactors, where it is combined with the MnR discharge slurry. A portion of the CCD 7 U/F is sent to the MnR removal as seed. The initial contact with basic MnR solution enables the neutralisation of free acid and bulk precipitation of heavy metals such as nickel, cobalt, ferric, aluminium, copper and zinc in the CCD U/F slurry. Lime slurry is added to the first two tanks to increase the pH to 8.5-9.0. Manganese and other heavy metals are precipitated as hydroxides after 90 minutes of reaction time. Air addition aids in the mixing as well as oxidize the ferrous iron to ferric and manganese to manganese dioxide. The final neutralisation slurry is expected to contain a residual aqueous manganese concentration of <1.4 mg/L. This aqueous manganese concentration is reduced to 1 mg/L by rainfall dilution in the Residue Storage Facility and this is considered suitable for disposal of the excess liquor to ocean outfall. The neutralised slurry is forwarded to the Residue Storage Facility by a series of two centrifugal pumps.

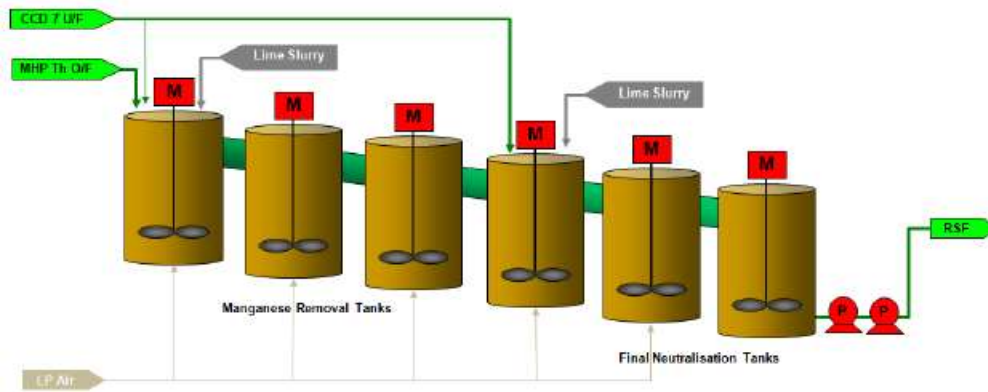


Figure 17: Manganese Removal and Final Neutralization

Process Plant Services and Utilities

Sulfuric Acid

Sulphuric acid is the major reagent of the plant used for leaching the saprolite ore. The imported sulphuric acid is received and unloaded in the port via a pipeline to a storage tank close to the tank. A stainless steel acid tanker trailer and prime mover delivers the acid to the process plant. Two additional acid storage tanks are allowed at the process plant site with 30 days acid storage in total (including tank at the port).

Limestone

The Limestone Plant provides limestone slurry for neutralisation of free acid and impurity removal of the SALT leached slurry at Primary Neutralisation circuit. Locally sourced crushed limestone is delivered by truck to the process plant and stockpiled on the limestone storage pad near the Limestone plant. Crushed limestone is retrieved from the stockpile by a front-end loader and loaded to a conveyor for slurring in the ball mill. Limestone slurry (30% w/w solids) is stored in a storage tank and distributed to the PN tanks via a ring main.

Lime Kiln and Slaking Plant

The consumption of lime in the SALT process is significant with respect to scavenging the Ni/Co in the 2nd stage MHP process, neutralization of barren solution to remove manganese and neutralization of the final washed solids prior too residue disposal. The decision was taken to install a lime kiln and slaking plant to minimize the ongoing operating cost of lime use.

The plant consists of a coal fired limestone calciner and lime slaking facilities. Crushed limestone is fed to a large rotary kiln operating at temperatures up to 1100°C. The Lime Slaking Plant converts the quicklime into milk-of-lime slurry using a slaker. The quicklime produced from the Lime Plant is cooled and stored in a silo. It is then conveyed and fed into the Vertimill™, a vertical ball mill. Filtered water is added to slake the quicklime and control the slurry concentration to 20% w/w solids.

Magnesia

Magnesia is used to precipitate the nickel and cobalt as mixed hydroxides in the MHP circuit. The imported MgO, which is in fine powder form, is introduced into the MHP Stage 1 reactors by pneumatic conveying.

Flocculant

Flocculant is required in Ore Preparation, CCD and MHP thickeners to aid settling. Imported flocculant which is received as dry powder in bulk bags, are loaded into a hopper. A screw feeder feeds the flocculant to the mixing tank where water is added to produce a 0.5% w/w stock solution. The flocculant solution is transferred to a storage tank for distribution.

Process Gases

Process gas requirements for the Process Plant are:

- Low pressure air
- Plant air
- Instrument air

Residue Storage Facility

The Residue Storage Facility (RSF) area includes disposal and storage facilities for the treated plant residue. The residue disposal method for the plant is wet disposal to an unlined residue storage facility. The slurry from the Final Neutralisation circuit is pumped on a 2 km pipeline to the RSF via two centrifugal pumps operating in series. At the impoundment, the residue is hydraulically deposited behind the retaining wall.

For the current study, the RSF is designed to be built in progressive cells. Each cell will have a surface area of 100 Ha (1 km L x 1 km W x 15 m H). The Phase 1 cell will be initially built in four walls. The succeeding cell will be built in adjacent to the initial cell utilising one of the Phase 1 walls, hence only requires three walls to build the total catchment. The Phase 1 cell which has an effective storage capacity of 12.5 Mm³ can accommodate the plant residue for 4 years. The plant requires 5 cells to be built progressively to store the total residue for 20 years.

Consolidation of residue over time produces a clear decant liquor which is pumped via a decant pipeline to the process plant as recycled process water or disposed to the sea as ocean outfall.

Infrastructure Requirements

The provision of infrastructure is a significant part of the overall development of the SALT Project. The proposed project location is in a brownfield site (in the immediate area of the existing PT Antam Ferronickel Plant with existing supporting infrastructures in the area. Therefore only specific project related infrastructures are required.

The infrastructure facilities to be provided for the project include:

- Temporary construction facilities
- Site works for the process plant and other facilities
- Ship unloading facility upgrade at the port for reagents such as limestone, MgO and coal
- Process plant roads, haul roads for the rejects material, and upgrade of the existing roads
- Utilities and services including water supply, power supply and telecommunications
- Site and process buildings including administration building, canteen and change building, control room building, product storage building, chemical storage building, maintenance and spare parts building, utilities building and chemical preparation building.
- Plant loading and unloading facilities like limestone, coal and ore stockyard.
- Mobile equipment

The SALT process plant will take advantage of the extensive public facilities, plant facilities and infrastructure in the vicinity of Pomalaa. Existing plant facilities and infrastructure include access roads, fuel storage and refueling station, raw water pump station, communications, power plant, ore loading port and plant mobile equipment.

The current study assumes the project power requirement is supplied by the national grid. The electric power supply is supplied in the main substation for distribution to various process and infrastructure sub-stations. The connected and used power has been estimated and summarized in Table 6.

The current port facility at Pomalaa provides for coal and ore unloading and ferronickel product shipment. This facility will require upgrading for SALT for both construction and plant operations. The cost of port upgrading has not been included in the current capital cost estimate. Table 7 summarizes the quantities of reagents to be received at the port facility on an annual basis.

Table 6: Power Requirements for the SALT Facility

Area	Connected Power (kW)	Used Power (kW)
Ore Preparation	6900	5100
SALT + Primary Neutralization	1000	640
CCD	1800	810
MHP Precipitation	900	470
Residue Neutralization	2100	1000
Reagent Preparation and Storage	290	140
Limestone Slurry Preparation	220	160
Lime Calcining and Slaking Plant	2300	1800
Utilities and Services	1500	900
Residue Disposal and RSF	800	350
Total	18000	11000

Table 7: Annual Reagent Requirements for SALT Facility

Reagent	Annual amount (tpa)
Acid	730,000
Diesel	700
Limestone, crushed	460,000
Coal	65,000
MgO	13,000
Flocculant	820
Soda Ash	1,000

Plant site buildings will be required for product storage, production building, utility building, canteen and change building, administration and first aid building, chemical storage building, maintenance workshop and spare parts building, flocculant preparation building, magnesium oxide preparation building, guard houses and toilet blocks. Contractor site offices, kitchen and mess hall, plant workshop and all temporary construction support buildings are included in the estimate. These facilities would remain on the site after completion of construction for future use.

Plant mobile equipment has been estimated to support the various unit operations and equipment. It has been assumed that some of the equipment from the ferronickel plant could be shared by the SALT plant, as such chiefly dedicated mobile equipment have been included in current estimate.

ECONOMIC STUDY

The capital and operating costs for a SALT facility at Pomalaa have been developed. The scope of work for the estimate prepared by SNC Lavalin was to cost the facilities inside the process plant fence/footprint and the Residue Storage Facility and associated overland pipelines.

Capital Cost

The direct and indirect costs are outlined as follows;

Direct costs

- Process plant from ore preparation through to final neutralization
- Reagents and process support packages such as sulphuric acid storage and supply, limestone plant, lime plant and reagents preparation
- Utilities and services such as water, power and air supply and distribution, mobile equipment and plant control system
- Process plant infrastructure including site development, process plant, buildings, unloading and storage facilities at the plant site
- Off-site infrastructure including Phase 1 Residue Storage Facility

Indirect Costs

- EPCM
- Capital and commissioning spares
- First fill

- Vendor representatives
- Contingency

The estimated capital costs are summarized in Table 8. The total cost of \$282 MUS\$ equates to a unit capital investment of \$19,000 US\$/annual tonne of Ni production or \$8.66 US\$/annual lb of Ni production.

Table 8: Capital Cost Estimate

Area Description	Cost (US\$)
Ore Preparation	20,400,000
Starved Acid Leach	5,740,000
Primary Neutralization	1,810,000
Counter Current Decantation	20,000,000
Mixed Hydroxide Precipitation	9,470,000
Manganese Removal	1,260,000
Final Neutralization	14,600,000
Residue Disposal	7,260,000
Sulphuric Acid Storage and Supply	4,890,000
MgO Distribution	302,000
Flocculant Preparation	1,890,000
Limestone Slurry Preparation	2,230,000
Lime Calcining Plant	39,470,000
Kiln Firing System	4,350,000
Lime Storage and Slaking Plant	5,890,000
Water Supply and Distribution	9,810,000
Air Supply and Distribution	4,940,000
Piperacks	1,750,000
Residue Storage Facility (Phase 1)	28,580,000
Mobile Equipment	7,510,000
Total Direct Cost	192,100,000
EPCM	48,250,000
Capital Spares	1,700,000
Commissioning Spares	280,000
Vendor Representatives	1,000,000
Contingency	38,600,000
Total Indirect Cost	89,830,000
Total Project Cost	282,000,000

Cost Basis

The costs were developed using the factored estimate technique and an in house database of historical equipment of similar projects and location. The costs were developed for February 2014 US\$ basis. The important exchange rates used in the estimate were 1 US\$ = 1.10 AUD = 11,737 IDR = 0.73 EUR.

Process Plant Equipment

Process plant equipment estimates have largely been derived from similar processing projects for which vendor pricing was received for the bulk of the equipment items. Costs were escalated using the CE Plant Cost Index, the M+S economic indices or escalation data from the Australian Bureau of Statistics. The six tenths rule was used to adjust cost to match design equipment size, when not available directly from the database.

Bulk Materials

Bulk materials costs were estimated based on their unit of measure. This includes:

- Earthworks

- Civil Works
- Plate Work
- Structural Steel
- Pipework
- Electrical
- Instrumentation

Bulk material costs have been factored based on the ex-works cost of mechanical equipment by facility. Costs were then split into materials and labour to reflect construction man-hours in the estimate. It is assumed that bulk materials will be sourced within Indonesia or other suppliers in the region. Where possible major tank work has been costed based on material take-offs (platework) derived from calculations to determine wall thickness. These quantities were then used to cost the tanks based on unit material costs and unit labour costs derived from previous projects in the area. The unit costs for bulk materials include costs for Quality Assurance/Quality Control to be present in the workshops. This includes expatriate personnel providing advice and assistance to local contractors.

Freight

Freight costs have been estimated as 7.5% of the ex-works material cost for each piece of equipment or bulk material shipment. This percentage is not applied to earthworks, civil works and buildings as these activities are generally based on sub-contract packages and freight is included in the package.

Residue Storage Facility (RSF)

The RSF sizes are based on the amount of residue generated by plant within 20 years of full operation (as derived from the mass balance) and assumed earthworks data from a similar project. The installed costs of these facilities are developed from a nickel project in Northern Philippines, where similar RSF model is costed. The project requires five cells that will be constructed in phases. The costs for phase 1 have been incorporated into the capital costs, while the costs for phases 2 to 5 (including the decommissioning of Phase 1) are treated as sustaining capital cost in the operating cost estimate.

On-Site Buildings

The process buildings and site infrastructure cost are estimated based on a similar project within the Pomalaa area. The buildings are costed according to the factored floor area in m². The laboratory cost includes major analytical instruments such as inductively coupled plasma, X-ray fluorescence, atomic absorption spectrometer, carbon and sulphur analyser and particle size analyser.

On-Site Labour

The labour rate is based on current information on similar projects costed within Sulawesi Island and is a composite or "all in" rate which includes;

- Direct labour
- Supervision (including expat supervision)
- Small construction equipment and tools
- Fuel for construction
- Construction camp and management
- Temporary construction facilities and associated operating cost
- Contractor mobilisation and demobilisation and,
- Major craneage and scaffolding.

Productivity Factor

The labour productivity factor used in the project estimate is based on in-house experience of project construction in Southeast Asia.

On-Site Installation Man-hours

On-site installation man-hours for equipment and bulk materials installation are obtained from contractor or major supplier consultation. Where data is not available, historical experience and/or in-house database are applied.

Engineering, Procurement, Construction Management (EPCM)

The EPCM cost for project is estimated at 25% of the project's direct cost and includes detailed engineering, project management, procurement and contracts and construction supervision.

Capital and Commissioning Spares

Capital spares are also called "insurances spares", providing spares that would not normally be consumed, but are required to maintain plant operation in a timely manner if they do fail. Commissioning spares are the typical spares stockholding and consumption capitalised prior to the plant reaching a commercial production status. Capital and commissioning spares are estimated at 2.2% and 0.4% of the mechanical equipment cost respectively.

First Fill

The cost of the first fill is estimated according to the initial requirements of the plant to start up, such as sulphuric acid and other reagents, mill lubricants and balls. An allowance is provided for the first fill and initial stock of reagents for 30 days of operation.

Vendor Representatives

An allowance has been provided for pre-commissioning, commissioning assistance, third party services and vendor representatives on site.

Contingency

Contingency is provided to cover the unforeseeable ("the unknown"); the items of work which will have to be performed or elements of cost that will be incurred within the defined scope of the estimate. Based on the preliminary nature of the current study, a 20% contingency has been applied to the Direct Costs.

Estimate Exclusions

There are a number of costs that have been excluded from the capital cost estimate for the current conceptual study. These are outlined below:

- Owner's costs;
- Technology fees, permits and licenses;
- All necessary permits and licenses;
- EPCM assistance during commissioning and ramp up;
- Custom clearance and import duties;
- Local and federal taxes;
- Escalation and currency variations;
- Ore mining, delivery and stockpiling related capital costs;
- Provision of utilities e.g. water, power, ablutions;
- Off-site infrastructure (except RSF);
- Bulk earthworks other than process plant pad preparation;
- Excavation in rock beds;
- Service lay-down areas;
- Road access and power supply to RSF;
- Road access to port;
- Port upgrade.

Operating Cost

The total operating costs of the project at full nameplate capacity is estimated at US\$ 119 million per annum or US\$ 3.62 /lb of nickel produced as MHP. The main operating cost item is the

sulphuric acid, representing almost half of the production costs. The operating costs are summarized in Table 9.

Table 9: Operating Costs

Items	Cost 1000 US\$ pa			US\$/lb Ni		
	Year 1	Year 2	Year 3+	Year 1	Year 2	Year 3+
Reagents and Consumables						
H ₂ SO ₄ (98.5%)	41,200	55,800	58,800	1.80	1.79	1.80
MgO	7,200	9,770	10,300	0.31	0.31	0.31
Limestone	4,878	6,600	6,940	0.21	0.21	0.21
Sub Bit. Coal	3,400	4,610	4,850	0.15	0.15	0.15
Flocculant	1,410	1,920	2,020	0.06	0.06	0.06
Soda Ash	210	285	300	0.01	0.01	0.01
Proc. Plant Consumables	1,910	2,580	2,720	0.08	0.08	0.08
Power Consumption	7,340	8,150	8,150	0.32	0.26	0.25
Utilities and Services Reagents and Consumables	1,360	1,290	1,300	0.06	0.04	0.04
Labour	4,240	4,240	4,240	0.19	0.14	0.13
General Expenses	5,000	5,000	5,000	0.22	0.16	0.15
Maintenance Materials	3,740	3,740	3,740	0.16	0.12	0.11
Contract Services	4,040	4,030	3,770	0.18	0.13	0.12
RSF Sustaining Capital	0	0	6,530	0	0	0.20
PROJECT OPERATING COST	86,000	108,000	119,000	3.75	3.47	3.62

Scope of the Estimate

The operating cost estimate for the project is based on the process mass balance (Full METSIM Balance). The estimate was developed for the following categories:

Variable Costs:

- Reagents and consumables
- Maintenance materials
- Contract expenses

Fixed Costs:

- Labour (including administration and general expenses costs)
- Administration and General Expenses
- Sustaining capital cost

Reagent cost includes shipping and freight and therefore considered site-delivered price, but ore is considered to be mined and delivered to the ROM stockpile at no charge.

Basis of the Estimate and Ramp Up

The estimate is based on a steady ramp up to full operational performance.

- Year 1 – 70%
- Year 2 – 95%
- Year 3+ - 100%

The estimate was based on January 2014 costs and the currency is US\$.

Reagents and Consumables

The cost of reagents and consumables represents a considerable portion of the operating cost. Pricing information was sourced from sole major supplier quotation, recent commodity report and/or SNCL-Lavalin database for recent and similar projects. Reagent costs include shipping and freight and are therefore considered site-delivered price. Major consumable unit costs (delivered) are summarized below.

- H₂SO₄ - 80 US\$/t
- MgO - 788 US\$/t
- Limestone – 15 US\$/t
- Sub-bituminous coal – 75 US\$/t
- Flocculant – 2,460 US\$/t

Services and Utilities

The two major operating costs for services and utilities are for grid power (\$0.11 US/kWh) and diesel for the mobile equipment (\$1.1 US/L).

Labour

A total of 36 non-western expatriates and 332 local/national workers have been allowed in the SALT plant labour summary. It was assumed that if required, non-western expatriates may be drawn from neighbouring Asian countries such as the Phillipines and Malaysia.

- Labour has been estimated to cover the complete plant including:
- Process plant operations;
- Utilities and services personnel for acid loading/unloading, limestone slurry and lime calcining and slaking plants, water services, process gases and other systems
- Maintenance personnel to provide maintenance services to all plant areas
- Technical staff in areas such as environmental, laboratory, process plant and engineering
- Administration and general personnel providing accounting, payroll, human relations, safety and training, stores and purchasing, and other services
- Management and supervisory staff in all areas.

Administration and General Expenses

This allowance covers safety equipment and training, telecommunication costs, medical costs, environmental costs, community relations, human relations costs and business travel.

Maintenance Consumables

Costs were allowed for maintenance consumables, which are replacement parts necessary to maintain equipment. The costs were derived from either vendor recommended costs or historical in-house factors for plant areas of similar operation based on the installed equipment costs.

Sustaining Capital

The Residue Storage Facility was developed to consist of 5 cells, each with a capacity of 4 years of plant residue. Starting in year 3, the annual sustaining capital amount for constructing RSF cells was \$6,5 Million US\$ per annum.

Escalation and Contingency

There was no allowance for escalation of operating costs in the estimate. This will be addressed in the financial analysis. Similarly, there was no allowance for contingency in the development of operating costs. Potential variations in the operating costs components will be addressed in the financial sensitivity analysis for the SALT project.

Estimate Exclusions

A number of costs have been excluded including:

- Mining and ore delivery;
- Replacement cost (e.g. mobile equipment);
- Some government charges;
- Royalties
- Taxation costs;
- Marketing costs;
- Escalation cost;
- Contingency;
- Corporate consultancies and
- Duties, customs or other imposts.

CONCLUSIONS AND RECOMMENDATIONS

The conceptual engineering evaluation of SALT validates that the technology has the potential to economically process low-grade saprolite ore. The overall operation incorporates simple and proven atmospheric leaching and precipitation processes. This will enable the plant to have fast ramp-up and achieve high availabilities, two of the important technical drivers for a nickel hydrometallurgical plant. The Sepon Copper plant experience cited in the introduction is an important reference in this regard.

The order of magnitude project costs of \$282 million capital cost and \$3.60 /lb nickel operating cost indicate that the SALT process has potential to be economically attractive in comparison to other nickel laterite processes, such as PAL and AL. It is important to note that these costs are limited to the process plant and the associated residue storage facility.

The SALT process plant capital cost was calculated at \$19,000 US\$/annual t of Ni production from the saprolite ore. This compares very favourably with the greater than \$100,000 US\$/annual t of Ni production typical of an HPAL plant.

The treatment of MHP in a ferro-nickel or matte smelting operation where the MHP contains some gypsum and hence higher sulfur must be confirmed.

The environmental aspects of SALT processing of below cut-off grade nickel saprolite ore have been carefully considered in the design and costing. Leach residues are expected to be benign when placed in the residue storage facility. Excess solution from the plant is treated to ensure all quality standards for ocean outfall disposal are met.

The current study recommends the project to proceed to the next phase to optimise and confirm design parameters and to investigate further potential along with the following recommendations:

1. Further bench testing and possibly continuous pilot testwork of the integrated plant should be completed prior to the next phase of the study to establish and optimise design parameters. Bench testing should also examine the feasibility and impact of some of the flowsheet modification opportunities.
2. Trade-off study for sulphuric acid and power sources: Acid importation versus generating onsite from sulphur with subsequent power generation based on optimum power source costs. Also the feasibility of a port to plant pipeline should be investigated.
3. Locating and testing lower cost limestone sources. The testing should include hardness, competency and reactivity both as limestone and as lime.
4. Locating and testing lower cost MgO sources. The testing should include reactivity and performance in synthetic process solutions.
5. Determination of resource available to ensure testing of representative samples, better define the optimum project scale and include the notional cost of ore mining and delivery to the project.
6. Full assessment on the infrastructure required for the project is required for a complete assessment of the Project's scope and capital cost. This would include mine, port and MHP feed system to the smelter.

InCoR Technologies is actively engaged with a number of nickel laterite processing companies with the goal of building the first SALT facility to process below cut-off grade nickel saprolite ores to produce mixed hydroxide product.

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