Title: NICKEL AND COBALT REFINING BY QNI PTY LTD, YABULU, QLD

Author:　　　　　**John E. Fittock**

Publication:　　　　**"Mawby" - AUSIMM Monograph 19 Volume**

Contact Person:　　**John Fittock**

Postal address:　　**QNI Pty Ltd**
　　　　　　　　　　Private Mail Bag 5
　　　　　　　　　　Mail Centre
　　　　　　　　　　Townsville Queensland 4818

Telephone:　　　　**07 47206114**

Fax:　　　　　　　**07 47206111**

e-mail:　　　　　**jefittock@qni.com.au**

Title: NICKEL AND COBALT REFINING BY QNI PTY LTD, YABULU, QLD

Author: **John E. Fittock**

Affiliation: **Principal Development Chemist**
QNI Pty Ltd

Postal address: **QNI Pty Ltd**
Private Mail Bag 5
Mail Centre
Townsville Queensland 4818

ABSTRACT

QNI Pty Ltd (QNI), a wholly owned subsidiary of BHP Billiton Plc, owns the Yabulu Nickel Refinery, which is located on the North Queensland coast approximately 25 km north-northwest of the city of Townsville. The refinery is operated on a 24 hour per day basis and has a permanent work force of 800 people. The Yabulu Refinery processes 3.75 million wet t/a of nickel laterite ore imported from New Caledonia, Indonesia and the Philippines, and 190 000 wet t/a of mixed nickel-cobalt hydroxide intermediate from Ravensthorpe Nickel in Western Australia. Combined production is 76 kt/a of nickel and 3.5 kt/a of cobalt, contained in a range of products.

The Yabulu Refinery was constructed in the early 1970's. The original design was based on the Caron process of reduction roasting and ammonia ammonium carbonate leaching to dissolve nickel and cobalt as ammine complexes. Cobalt was separated by semi-selective sulfiding, and the mixed nickel-cobalt sulfide intermediate with Ni:Co >2 was sold to overseas refineries. The remaining nickel was precipitated, calcined, partially reduced, and sold as intermediate grade Ni/NiO rondels containing 85 to 90% Ni.

Processing of nickel laterite ore from the Greenvale Mine, located 220 km west-northwest of Yabulu, commenced in 1974. Ore was transported from Greenvale to Yabulu by rail. Processing of imported nickel laterite ore from a number of mines in New Caledonia, Indonesia and the Philippines commenced in 1986. The imported ore is shipped to Townsville Port and then railed 25 km to the plant. Mining at Greenvale ceased in 1992 and the last Greenvale ore was processed at Yabulu in 1993. Small annual tonnages of beneficiated nickel laterite ore were obtained by rail from the Brolga Mine near Rockhampton between 1993 and 1995. In 1996 the number of ore reduction roasters was increased from ten to twelve, increasing the ore treatment capacity to 2.4 million dry t/a. Current production levels from nickel laterite ore are 31 kt/a of nickel and 2.0 kt/a of cobalt.

A novel solvent extraction process for separation of nickel and cobalt was developed and installed during the transition to imported ore in 1989, producing step change improvements in the quality of both products. This lead to the development of a new generation cobalt refinery, with commercial production of a chemical grade cobalt product commencing in 1997.

A major brownfield expansion of the nickel and cobalt refining sections of the Yabulu Refinery was undertaken from 2004 to 2007, increasing the production capacity of Yabulu Refinery to process a mixed nickel-cobalt hydroxide precipitate (MHP) from the Ravensthorpe Nickel project in southwest Western Australia. MHP is containerised at Ravensthorpe, shipped from the port at Esperance to Townsville Port and then railed to the plant. Contained metal values in the MHP are 45 kt/a nickel and 1.5 kt/a cobalt.

The original flowsheet used heavy fuel oil and naphtha for ore drying and roasting, power and steam generation, ammonia production and nickel calcining. Partial conversion to coal firing occurred following the 1970's oil crisis. Further conversion to coal seam methane firing occurred in 2006 following delivery of gas by pipeline to the Townsville region.

PROCESS FLOWSHEET OVERVIEW

The Yabulu Refinery carries out three major functions – laterite ore processing, mixed nickel-cobalt hydroxide processing, and nickel and cobalt refining. Figure 1 is an aerial view of the Yabulu Nickel Refinery prior to the major brownfield expansion, while Figure 2 shows the Yabulu Nickel Refinery flowsheet in 2007 after the major brownfield expansion.

Figure 1. Aerial view of the Yabulu Refinery prior to the major brownfield expansion

Figure 2. Yabulu Refinery flowsheet in 2007 after the major brownfield expansion

The ore processing section comprises receival of shipments of imported nickel laterite ore, blending, drying, milling and reduction roasting at about 750°C using Herreshoff multiple hearth furnaces. The reduced ore is cooled and then leached in ammonia ammonium carbonate liquor at atmospheric pressure to selectively dissolve nickel and cobalt. Product liquor (PL, typically 10 to 12 g/L Ni and 0.6 g/L Co) is separated from the tailings solids in an eight thickener washing circuit. Tailings slurry is steam stripped to recover ammonia prior to disposal in storage ponds.

The PL from the ore processing section contains excess ammonia, part of which is utilised to dissolve further nickel and cobalt in the mixed nickel-cobalt hydroxide processing section. The

resultant concentrated product liquor (CPL, typically 23 g/L Ni and 1.1 g/L Co) is separated from the residual solids (mainly manganese and magnesium carbonates), which are sent to the washing circuit in the ore processing section.

In the metal refining section of the refinery, the remaining excess ammonia in the CPL is removed by controlled steam stripping, enabling nickel to be separated from cobalt using QNI's patented ammoniacal solvent extraction (ASX) technology. Nickel is selectively extracted, and then stripped from the nickel-loaded solvent using concentrated ammonia ammonium carbonate solution. The resultant nickel loaded strip liquor (Ni LSL, typically 75 to 80 g/L Ni, 0.05 g/L Cu and 0.02 g/L Co) is further purified by selective copper sulfide precipitation to 0.001 g/L Cu, and then steam stripped to precipitate basic nickel carbonate solids (BNC, typically 52% dry wt Ni). The slurry is dewatered using thickeners and pressure filters. The BNC solids are calcined in rotary kilns, either in an oxidising atmosphere to produce nickel oxide products (NiO, typically 78% Ni, granular and powder forms), or in a reducing atmosphere to produce a nickel metal/oxide calcine (approximately 60% nickel metal and 40% nickel oxide, typically 90 to 92% Ni). The nickel metal/oxide calcine is further reduced to typically 97.5% Ni in travelling-belt furnaces using sawdust and hydrogen reductants. The resultant nickel powder is pressed into compacts, which are sintered and further reduced to typically 99% Ni in travelling-belt furnaces using hydrogen reductant, prior to packaging for sale.

The raffinate from the ammoniacal solvent extraction circuit (typically 1.0 g/L Co and 0.02 g/L Ni) is sulfided using ammonium hydrosulfide solution. Cobalt sulfide solids are separated using a thickener, allowing the barren liquor to be recycled to the washing circuit in the ore processing section. Cobalt sulfide solids are oxidatively dissolved using air and oxygen to produce a crude cobalt sulfate solution, from which the zinc, iron, nickel, sulfur, magnesium and calcium impurities are selectively removed using three solvent extraction stages and one ion exchange stage. The resultant high purity cobalt ammine solution (typically 65 g/L Co and <0.001 g/L Ni) is steam stripped to precipitate chemical grade cobalt oxide/hydroxide solids. The slurry is dewatered using a thickener, pressure filter and fluid bed drier, and the cobalt oxide/hydroxide solids are packaged for sale.

NICKEL LATERITE ORE TREATMENT

Characteristics of imported ore

The Yabulu Refinery processes 3.75 million wet t/a of imported ore from over 15 mines in New Caledonia, Indonesia and the Philippines. The ores vary in mineralogy, moisture content and handling ability. Imported ore is a combination of high iron limonitic ore (primarily goethite), manganese-iron oxides (asbolan) and high magnesium, high silica saprolitic ore (primarily serpentine). Nickel and cobalt association with each ore type is variable. Moisture content averages 35% and varies between 30 and 40% depending on ore types and weather circumstances at the mine, load port, Townsville and Yabulu. The moisture content affects the ore handling ability, with low moisture content leading to excessive dustiness, and high moisture content causing stickiness. The laterite ore feed to the Yabulu Refinery has the typical composition shown in Table 1.

Table 1. Typical composition of imported nickel laterite ore

Bulk materials handling

Approximately 75 ore shipments of between 40 kt and 60 kt are received each year. QNI's ore handling infrastructure at Townsville Port includes an automated ore unloader capable of handling 1200 t/h, wharf-side hoppers and conveyor system, grizzly and crushing station, a dedicated rail loop and automatic train load-out station. Rocks larger than 15 cm are screened using the triple deck grizzly, and then crushed. The ore is automatically loaded onto trains at a rate of 1920 t/h. The ore is railed 25 km to Yabulu Refinery; each train has 33 wagons of 60 t capacity each.

Ore receival

The wagons are unloaded by means of a Noyes rotary tippler. An apron feeder beneath the unloader feeds the ore onto a conveyor system, which transfers the ore to a radial stacker and onto the 30 kt capacity live stockpile. A 16 kg sample is extracted from each ore carriage using a platform-mounted backhoe. A representative composite of all wagon samples for each ore shipment is prepared and assayed for payment.

Solar drying and ore blending

Yabulu Refinery's ore dryers were designed to process Greenvale ore, which typically contained 25% moisture. Solar drying is undertaken to reduce the moisture content of the imported ores from typically 35% to 28%, where it can be readily handled in the feed system to the plant's three dryers. Gordon (2003) described the history, development and benefits of solar drying of ore at Yabulu.

Excavators load the wet ore from the rail car receival stockpile into haul trucks, which transport the ore to designated areas. Trucks dump the ore in rows on the lay-down pads, which are then leveled with a bulldozer and/or excavator into a 30 cm thick layer. The ore is then ploughed and turned over on a daily basis for four to six days, until it is <30% moisture. It is then wheel rolled by empty trucks and another layer is constructed on top of it. Once the required number of layers is in place and the pad is full and dried to target, the reclaiming begins; this is done with excavators scraping the face of the stockpile, taking a vertical cut through each layer, loading trucks that run to the reclaim feeder. The horizontal multi-layer lay-down and vertical cut reclaim supports ore blending requirements.

In wet weather the drying pads are wheel rolled by empty trucks driving back and forth over the surface. A layer of approximately 30 kt, laid out in an area of 130 m x 270 m, can be sealed in one hour by the use of four haul trucks offset by a wheel width. The resulting seal is effective even in heavy rain.

The benefits of the solar drying process are improved ore handling with less blockages and downtime, reduced fuel consumption in drying, and blending of substantially different ores from a variety of mines. This allows for long-term treatment of ores without large grade variations and delivers downstream stabilisation of plant and processes to maximise throughput and recoveries.

Ore drying

The ore drying plant consists of one coal seam gas-fired and two coal-fired rotary dryers, each with a nominal water evaporation capacity of 50 t/h. The 34 m long dryers are co-currently fired. The ore is dried to nominally 6% moisture content. The discharge gas from the dryers is cleaned

to recover the entrained fines in two stages - multiclones and electrostatic precipitators. The dust, which constitutes up to 10% of total throughput, is pneumatically pumped directly to the roaster feed bins, bypassing the grinding plant.

Ore grinding

The dried ore is conveyed to a 2 kt surge bin, which feeds two parallel Hardinge 3 MW dry-grinding ball mills. The ore milling rate is typically 150 to 200 dry tonnes per mill operating hour, and the power requirement is typically 8 to 12 kWh per dry tonne of ore milled. Each 4.7 m diameter by 4.6 m long mill contains a ball load of approximately 150 t and is charged with 7.5 cm diameter forged steel balls. The mills are air swept with a classification system returning the oversize material to the mill feed. The typical particle size distribution of the product is:

<div align="center">

99.5%   < 850 microns

95%   < 355 microns

80%   < 150 microns

70%   < 90 microns

55%   < 45 microns

</div>

The dried, ground ore is transported pneumatically to four 3 kt live storage silos in the ore reduction section.

Ore reduction

Ore reduction is conducted in twelve 8 m diameter by 23 m high Nichols Herreshoff multiple hearth reduction furnaces, each containing 17 bricked hearths with interconnecting 'drop holes'. The finely ground, dry ore is delivered from the ore storage silos at, typically, 25 t/h per roaster by a pneumatic feed system to a paddle-mixer where heavy fuel oil is added and mixed at a rate of 2.5% w/w as primary reductant. This ore/oil blend is screw conveyed into the top hearth of each roaster. Hollow central shafts fitted with four rabbling arms per hearth rotate slowly to move the ore across the alternating inwards and outwards hearths as it progresses downwards through the roasters.

Figure 3. Herreshoff roasters used for reduction of laterite ore

Eleven external combustion chambers installed at hearths 7, 9, 11, 13, 15 and 17 (numbered from the top) on each roaster are fired with heavy fuel oil at nominally 60% combustion stoichiometry to provide the necessary energy for roasting. Secondary air is added in hearths 5 and 7 to combust the remaining hydrogen and carbon monoxide.

As the ore moves downwards through the roasters it is heated by counter-current contact with combustion gases and radiant energy from the brickwork to initiate dehydroxylation and then reduction reactions. Dehydroxylation occurs over a range of temperatures from around 350°C (goethite) up to 650°C (serpentine). Once the metal oxides have been liberated, reduction with carbon monoxide and hydrogen (from oil decomposition and sub-stoichiometric operation of combustion chambers) occurs at temperatures up to 750°C. The nickel, cobalt and iron oxides in the ore are reduced to reactive metals to varying extents. Metal sulfides are also formed by reaction of sulfur (from the 3% sulfur fuel oil) with the ore. Typically less than 20% of the iron oxide in the ore feed is reduced to metal, with the majority of iron being present as magnetite in the reduced ore.

Reduced ore discharges from hearth 17 of each roaster at >700°C and is cooled to <250°C inside 3.0 m diameter by 26 m long rotary coolers partly immersed in water baths.

The off-gas from the top of each roaster passes through a set of cyclones and the recovered dust is returned to the furnace. An induced draught fan on each roaster off-gas flue delivers the gases to electrostatic precipitators for final cleaning.

Willenborg (1985) and Hobson and Piper (2002) described some of the many improvements in ore reduction and roaster operations that have been achieved at Yabulu.

Reduced ore leaching and washing

Archimedes screws at the discharge end of the reduced ore coolers transfer the solids into sealed chutes connected to launders, where the reduced ore is mixed with the ammonia ammonium carbonate plant leach solution. Approximately 1800 m^3/h of leach liquor from thickener No 1 overflow is cooled to about 25°C in two banks of plate heat exchangers, using water-cooling

followed by refrigeration. The slurry of reduced ore and leach liquor is collected in the agitated quench slurry tanks, and then pumped to distribution boxes that feed the four parallel leaching trains, each comprised of a series of four 150 m^3 agitated aeration tanks.

Up to 600 m^3/h of leach liquor from thickener No 2 overflow is cooled to about 35°C in a bank of plate heat exchangers using water-cooling, and then partially sulfided to precipitate cobalt as a mixed cobalt-nickel sulfide, which is separated in a clarifier for use as a reductant in the MHP treatment section. The cobalt-depleted leach liquor is pumped to the distribution boxes that feed the reduced ore leaching trains.

Iron, nickel and cobalt present in the reduction roasted laterite ore as alloys and sulfides react with oxygen from the air and dissolve in the ammonia ammonium carbonate leach solution. Iron dissolves initially as a ferrous (Fe^{2+}) ammine complex, and is rapidly oxidised to ferric (Fe^{3+}) and precipitates. Nickel forms stable ammine complexes such as $Ni(NH_3)_6^{2+}$. Cobalt initially forms cobaltous (Co^{2+}) ammine complexes, which are further oxidised to very stable cobaltic (Co^{3+}) ammine complexes. Sulfide is oxidised and dissolves initially as thiosulfate ($S_2O_3^{2-}$), which is slowly further oxidised to sulfite (SO_3^{2-}) and then sulfate (SO_4^{2-}).

Mackenzie and Lumsdaine (2002) showed that some soluble cobalt is lost by co-precipitation and also adsorption with the fresh iron precipitate. Lowering the soluble cobalt concentration in the leach liquor from thickener No 2 overflow has the benefit of decreasing these losses. The leaching reactions are exothermic, and the temperature of the leach slurry increases to 40 - 45°C. Cooling the leach liquor prior to its mixing with the reduced ore is critical as both the soluble cobalt loss and evaporation of ammonia increase with temperature during leaching. The oxygen-depleted air from the aeration process carries considerable ammonia and carbon dioxide vapours, which are recovered as weak leach liquor (WLL) by scrubbing with water.

Product liquor (PL, typically 10 to 12 g/L Ni, 0.6 g/L Co, 95 g/L NH_3, 60 g/L CO_2 and pH 10.5) is separated from the waste solids by counter-current decantation (CCD) washing with barren ammonia ammonium carbonate solution in a series of eight thickeners. The 45 m diameter thickeners are fitted with cable torque arms and are fully covered to prevent loss of ammonia and

carbon dioxide. Final metal dissolutions from reduced ore after washing are typically 80 to 82% Ni and 50 to 60% Co.

Disposal of washed tailings

The underflow slurry from thickener No 8 in the CCD circuit (typically 60% w/w solids) contains the reduced ore tailings, MHP leach tailings and a substantial amount of ammonia and carbon dioxide. The tailings pulp is completely steam stripped of ammonia and carbon dioxide in four parallel tailings stills, which include a direct preheater tower, an agitated magnesium crystallisation tank, a 2.4 m diameter by 20 m high 12-tray stripping still and a bottoms flash vessel. The internals of each tailings still are cleaned using high pressure water blasting every 4 weeks to remove magnesium carbonate scale. The stripped tailings discharge to a sump and are pumped to the tailings ponds for storage.

MIXED NICKEL-COBALT HYDROXIDE TREATMENT

Characteristics of MHP

Ravensthorpe MHP is an amorphous hydroxide, containing nickel and cobalt in conjunction with lesser amounts of manganese, magnesium, sulfate, and other trace elements. The MHP is "as received" filter cake product and has not been dried or extruded. MHP has the following typical composition.

Table 2.   Typical composition of Ravensthorpe MHP

MHP receival

MHP from Ravensthorpe is transported by approximately 17 shipments per annum of 430 x 26 t high-top containers by ship to Townsville Port, and then delivered by rail to a storage area at the Yabulu Refinery at a peak rate of approximately 50 containers per day.

MHP leaching

Three stages of leaching of MHP in ammoniacal liquors are used for the extraction of nickel and cobalt.

Repulp

A container tipper rotates the full MHP containers and dumps the MHP into a hopper with live bottom screw feeders. The feeders reduce the lump size and move the MHP forward into a collection screw conveyor that in turn discharges into the repulper. An attrition scrubber is used to repulp the MHP with a small flow of PL, giving slurry of consistent composition, without any lumps. The repulped slurry is stored in two 350 m^3 agitated surge tanks.

Primary leach

In the primary leach stage, the remaining PL is combined with slurry from the MHP slurry surge tanks in a series of four 475 m^3 aerated, agitated reactors. The available ammonia and ammonium carbonate in PL is utilised to solubilise approximately 90% of the nickel and cobalt. Overflow from the secondary MHP leach thickener is also added to the first reactor so that the configuration becomes counter-current. Overflow slurry from the last reactor gravitates into the 17.5 m diameter primary leach thickener, and the resultant overflow called concentrated product liquor (CPL, typically 23 g/L Ni and 1.1 g/L Co) is forwarded to the preboil stills for further treatment.

Reductant preparation

A reductant, in the form of a cobalt-nickel sulfide (CoNiS, typically Co:Ni >2), is used in the secondary leach stage to enhance the dissolution of nickel and cobalt. CoNiS reductant serves to undo MHP aging effects due to oxidation and any refractory association of cobalt and nickel with oxidised manganese and cobalt compounds. The required amount of CoNiS reductant is produced from partial sulfiding of the soluble Co and Ni in the overflow stream from thickener No 2 of the CCD washing circuit in the ore treatment section. Ammonium hydrosulfide is metered in-line to the thickener No 2 overflow solution to precipitate a cobalt-nickel sulfide. The resulting slurry is thickened in a 19.5 m diameter clarifier. The cobalt-depleted overflow returns to the distribution boxes that feed the reduced ore leaching trains, while the underflow is used as the reductant for MHP leaching.

Secondary leach

Underflow from the primary MHP leach thickener, Fresh Leach Liquor (FLL, typically >120 g/L NH_3 and >60 g/L CO_2, sourced from the gas recovery section and at a temperature of approximately 65°C) and CoNiS slurry from the CoNiS clarifier are combined to form the secondary leach feed slurry. Secondary MHP leach comprises a series of five 84 m^3 agitated reactors. FLL is introduced into the secondary leach as lixiviant to give higher ammonia concentrations than in the primary leach. The CoNiS reductant is added in excess to that amount required for reaction with the oxidised compounds. The portion of the reductant that reacts in this process is solubilised. The last three reactors in the MHP secondary leach are sparged with air to oxidise and solubilise the remaining CoNiS reductant. Secondary leaching is followed by liquid-solid separation in a 15.5 m diameter secondary thickener. Secondary thickener overflow is returned to the primary MHP leach.

CCD washing

The MHP leach tailings from the secondary thickener underflow are combined with the reduced ore tailings in the CCD circuit in the ore treatment section. Final MHP leach dissolutions are obtained by washing the residue in the CCD circuit where metal concentrations are comparatively low and the ammonia concentration is comparatively high, both favouring the enhancement of nickel and cobalt dissolutions from MHP. Final metal dissolutions from MHP after the CCD washing are typically 99.5% Ni and 97.5% Co.

NICKEL-COBALT SEPARATION

Preboil

CPL is pumped from the MHP primary leach thickener overflow at typically 400 m^3/h and stored in two 2400 m^3 storage tanks. CPL is subjected to controlled partial distillation ('preboil') in four multi-tray distillation columns. The CPL feed to the preboil stills is preheated by the preboil still discharge in a bank of recuperative plate heat exchangers. Low-pressure steam at 70 kPag is injected at the bottom of these stills, driving off part of the dissolved ammonia and carbon

dioxide to provide an appropriate ammonia concentration in the solution for subsequent efficient nickel solvent extraction. As a result, impurities such as iron, manganese and some magnesium precipitate from the solution, as well as a small amount of nickel and cobalt. The reduction in ammonia concentration serves to weaken the nickel ammine complex, which enhances the subsequent solvent extraction of nickel from the liquor. The internals of each preboil still are cleaned using high pressure water blasting every 6 weeks to remove scale.

The preboil still discharge slurry containing approximately 0.3 g/L of solids is fed to a 15 m diameter clarifier. Further precipitation of solids occurs in the clarifier. The clarifier overflow solution, called special product liquor (SPL, typically 21.5 g/L Ni, 1.0 g/L Co, 45 g/L NH_3 and 25 g/L CO_2) and at a temperature of about 90°C, is used to preheat the preboil still feed. The SPL clarifier underflow slurry is filtered and washed on a horizontal belt vacuum filter. The filtrate is recycled to the SPL clarifier, while the filter cake is transferred to the incoming ore stockpile and hence returned to the roasters for recovery of the contained nickel and cobalt.

SPL oxidation and filtration

SPL is pumped through a series of three 2400 m^3 storage tanks. Prior to entering each tank, SPL is injected with air and mixed in an in-line mixer, to oxidise Co^{2+} ions to Co^{3+}. The oxidised SPL is filtered using three Kelly filters in order to remove trace solids. Precoat is used to aid filtration and prevent ultra-fine solids from passing through the filter cloths. The filtrate is transferred into the filtered SPL surge tank. In the SPL feed line to the ammoniacal solvent extraction circuit, the SPL temperature is controlled using a steam heated plate heat exchanger and a small amount of hydrogen peroxide solution is injected to oxidise any residual Co^{2+} ions to Co^{3+}.

Ammoniacal solvent extraction plants

Equipment

Two ammoniacal solvent extraction (ASX) circuits operate in series (with respect to the feed aqueous stream) to selectively separate nickel from cobalt. The first 'scalp' ASX circuit consists of two extraction stages in series and three stripping stages in series. The second 'follow' ASX circuit consists of three extraction stages in series and four stripping stages in series.
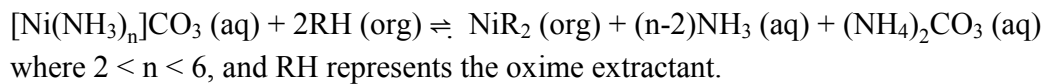
All of the extract and strip stages are conventional mixer-settler units. All mixer-settler cells in the scalp ASX circuit have three mixers in series with the third mixer overflowing into the settler. In the follow ASX circuit, only E3 and S4 cells have three mixers in series; the remaining cells have two mixers in series. The operating temperature range is 50 to 60°C, and all cells are sealed to contain ammonia vapors.

The cells are fabricated in stainless steel on elevated steel platforms with a bunded area below for spillage, with sufficient height for pipe work and maintenance access below the cells. Cell to cell pipe work is run below the settler support platform.

Nickel extraction

The ASX organic has a typical loading capacity of >14 g/L Ni; it consists of approximately 15% v/v 2-hydroxyl-5-nonyl-acetophenone oxime extractant (the active component of LIX 84) and 8% v/v isotridecanol modifier in an aliphatic diluent. Typical flow rates are 560 m^3/h scalp organic and 400 m^3/h follow organic.

The extraction and stripping of nickel in the ASX process can be represented as follows:

$[Ni(NH_3)_n]CO_3$ (aq) + 2RH (org) $\rightleftharpoons$ NiR_2 (org) + (n-2)NH_3 (aq) + $(NH_4)_2CO_3$ (aq)
where 2 < n < 6, and RH represents the oxime extractant.

Any Co^{2+}, copper, manganese and iron ions present in the SPL feed are also extracted into the organic, but zinc in the SPL feed is not extracted and remains with the relatively inert Co^{3+} ammine complexes in the follow raffinate.

The feed SPL to the scalp ASX plant typically contains 21.5 g/L Ni and 1.0 g/L Co. In the scalp ASX circuit, nickel is partially extracted from the SPL into the scalp ASX organic stream. The scalp aqueous raffinate stream containing approximately 9 g/L Ni is pumped via the scalp raffinate transfer tank to feed the follow ASX circuit. The temperature is controlled by a steam heated plate heat exchanger. The aqueous raffinate stream exiting the final follow extraction stage typically contains 1.0 g/L Co, <0.05 g/L Ni and <0.1 g/L Zn.

Nickel stripping

The scalp nickel-loaded organic is pumped directly to the scalp strip circuit where it is contacted counter-currently with a concentrated ammoniacal ammonium carbonate strip liquor (SL, typically 250 to 300 g/L NH_3). Nickel is transferred from the loaded organic into SL producing the scalp nickel loaded strip liquor (Ni LSL, typically 75 g/L Ni). The stripped organic is recycled back to the scalp extraction circuit.

Similarly, the nickel-loaded organic in the follow ASX circuit is pumped directly to the follow stripping circuit where it is contacted counter-currently with SL. Nickel is transferred from the loaded organic into SL producing the follow Ni LSL. The stripped organic is recycled back to the follow extraction circuit.

Organic cleaning and regeneration

Oxidation of the cobalt in SPL feed to the ASX process is never 100% and some reduction of Co^{3+} ions back to the Co^{2+} state also takes place due to reaction with thiosulfate. Any Co^{2+} ions in the SPL are extracted from aqueous solution into the organic phase by a similar reaction to nickel ions. Some then undergo oxidation to Co^{3+}, which bind strongly with the oxime extractant and are then not stripped in the nickel stripping stages using concentrated ammonia ammonium carbonate solution.

The oxime-based organic extractant in the ASX organic is slowly hydrolysed to the corresponding ketone by the strong ammonia solutions used in the process. The ketone is retained in the organic phase.

Separate organic cleaning and regeneration plants treat small bleed streams of the stripped organic from the scalp and follow ASX circuits, to maintain the organic solutions in the two circuits at typically <3 g/L Co and <2% v/v ketone.

COBALT PRECIPITATION AND REFINING

Cobalt sulfide precipitation and washing

Cobalt bearing raffinate from the follow ASX extraction circuit is contacted with ammonium hydrosulfide to precipitate the cobalt, zinc and remaining nickel as a mixed sulfide, which is thickened in a 33 m diameter thickener.

The sulfided liquor containing negligible nickel, cobalt and zinc overflows from the thickener into an aeration tank to oxidise any residual ammonium hydrosulfide. About two-thirds of the barren liquor is returned to thickener No 8 in the ore washing CCD thickener circuit and the remainder sent to the tailings stills.

The thickened cobalt sulfide pulp is washed using water in two more counter-current stages to partially remove dissolved calcium and magnesium. The slurry is then transferred to the two cobalt sulfide stripping stills (1.2 m diameter by 21.5 m high) where 70 kPag steam is injected to recover ammonia and carbon dioxide. The off-gas reports to the gas recovery section. The still bottoms report to a thickener for a further stage of solids washing and concentration.

The majority of the cobalt sulfide slurry is transferred to the stage 1 leach circuit in the cobalt refinery, which has a capacity of approximately 2000 t/a cobalt. Impure cobalt sulfide feedstock in excess of the cobalt refining capacity is filtered, packaged in bulk bags and sold as an intermediate product.

Table 3.   Typical composition of cobalt hydroxy hydrosulfide solids

Stage 1 – Cobalt sulfide leaching and clarification

Cobalt sulfide feed slurry (typically 10% w/w solids) is first leached in a series of six agitated aeration tanks at 80°C for approximately 24 h, which typically results in >90% cobalt dissolution. Leach slurry exiting the atmospheric leach is further leached using oxygen in a 16 m^3 horizontal brick-lined autoclave with four compartments, at 500 kPag and 110°C for approximately 2 h. Typically >99% extent of cobalt dissolution is achieved and the pH of the final slurry is <2.

Leach slurry exiting the autoclave is flashed to atmospheric pressure and clarified in a 2 m diameter high-rate thickener to separate residual solids. The thickened underflow pulp is pumped to a two-stage vacuum belt filter and the solids are washed in two counter-current steps with water. The residue from the belt filter (predominantly elemental sulfur) is discharged directly into bulk bags for disposal. The thickener overflow stream is polished using bag filters, then indirectly water-cooled to ~60°C. Hydrogen peroxide solution is injected to oxidise Fe^{2+} ions to Fe^{3+}.

Stage 2 - Zinc and iron removal by solvent extraction

The crude cobalt sulfate solution feed to stage 2 typically contains >50 g/L Co, 2.5 g/L Ni, 5 g/L Zn and 0.1 g/L Fe; the flow rate is typically 6 m^3/h. The stage 2 solvent extraction circuit is comprised of ten conventional mixer-settlers. The organic contains approximately 12% v/v bis(2,4,4-trimethylpentyl) phosphinic acid extractant (the active component of Cyanex 272) in an aliphatic diluent and the flow rate is typically 6 m^3/h.

Zinc and iron are selectively extracted in four extract cells. The pH is controlled between 2.5 in the first cell, E1 and 3.5 in the last cell, E4 using aqueous ammonia addition. Zinc in the raffinate stream is typically <0.001 g/L.

Zinc- (and iron) loaded solvent is scrubbed in two cells using dilute sulfuric acid to minimise transfer of cobalt. Spent scrub solution enters E1. Scrubbed loaded organic is stripped in four cells using dilute sulfuric acid to produce the acidic zinc loaded strip liquor (Zn LSL, typically >30 g/L Zn).

Stage 3 – Cobalt transfer to ammine solution by solvent extraction

The objective of this stage is to transfer the cobalt (and nickel) to an organic phase and convert the cobalt to ammine complexes during stripping. This transfer eliminates sulfate and other anions such as chloride from the final product, and allows the removal of nickel by QNI's patented ASX process in stage 4.

The stage 3 solvent extraction circuit is comprised of nine conventional mixer-settlers. The organic contains approximately 8% v/v di-2-ethyl hexyl phosphoric acid extractant (DEHPA) and 10% v/v isotridecanol modifier in an aliphatic diluent, and the flow rate is typically 60 m^3/h.

The iron and zinc - free cobalt sulfate raffinate solution from stage 2 is contacted counter-currently with the organic in three extraction cells, to extract >99% of the cobalt. The stripped organic entering the last extraction cell, E3 is in the ammonium form and the pH in each extraction cell is >7. The ammonium sulfate raffinate solution containing <0.1 g/L Co is recycled to the ASX raffinate stream to recover the residual cobalt by precipitation with ammonium hydrosulfide.

The cobalt-loaded organic is scrubbed in two cells using hot water to minimise transfer of entrained sulfate ions. Spent scrub solution enters E1.

Scrubbed loaded organic is stripped in four cells using concentrated ammonia ammonium carbonate strip solution (SL, typically 250 to 300 g/L NH$_3$) to form the cobalt ammine complexes. Cobalt loaded strip liquor (Co LSL, typically 75 g/L Co and 3 g/L Ni) is oxidised with air in two agitated tanks at atmospheric pressure to oxidise >99% of Co^{2+} ions to Co^{3+}. A small amount of hydrogen peroxide solution is then injected to oxidise any residual Co^{2+} ions to Co^{3+}.

Stage 4 - Nickel removal by solvent extraction

The oxidised Co LSL passes through an activated carbon column, which absorbs residual traces of the stage 3 organic. Oxidised Co LSL is then subjected to controlled partial distillation (preboil) using steam in a multi-tray column. This lowers the uncomplexed ammonia concentration without destabilising the cobalt ammines and causing cobalt precipitation in the column. The purpose of lowering the ammonia content is to improve the nickel extraction equilibrium in the following solvent extraction process.

The stage 4 solvent extraction circuit is comprised of six conventional mixer-settlers. The organic used is a small side stream (typically 6 m^3/h) of stripped organic from the scalp ASX circuit. The organic stream is first stripped using SL in three cells in order to lower the nickel content to <0.1

g/L. The organic then extracts nickel (and traces of copper, manganese and iron) from the oxidised, preboiled Co LSL in three cells. Nickel in the raffinate stream is <0.001 g/L. Both the nickel loaded strip liquor stream and nickel-loaded organic stream are recycled to the scalp ASX circuit.

Stage 5 – Calcium and magnesium removal by ion exchange

The final purification step is the removal of magnesium and calcium by ion exchange. The nickel-free Co LSL feed solution typically contains 70 g/L Co, <0.1 g/L Mg and <0.01 g/L Ca. The cobalt solution is first passed through an activated carbon column that absorbs residual traces of the organic used in nickel removal, preventing fouling of the ion exchange resin or contamination of the final product. Calcium and magnesium are removed in two ion exchange columns in series. There are a total of four 2 m^3 columns, allowing one column to be regenerated and one column to be on standby. Cross-linked macroreticular cation resins bonded to a styrene/divinylbenzene support, with an iminodiacetic functional group are used. The resin is changed to the ammonium form prior to loading. The regeneration sequence comprises the following steps:
1. Rinse with ammonia solution, to remove cobalt liquor from the column;
2. Strip cobalt, magnesium and calcium using dilute sulfuric acid;
3. Regenerate to ammonium form using ammonia solution.
The product solution from the ion exchange circuit is discharged to the precipitation circuit.

Stage 6 - Cobalt precipitation

The purified cobalt ammine liquor is stripped using steam in a 22 m^3 horizontal stainless steel autoclave with five agitated compartments. The autoclave discharge slurry is flashed to atmospheric pressure and fed into a 2 m diameter high-rate thickener. The thickener underflow containing 35% w/w solids is filtered in a continuous belt pressure filter. The filtrate is returned to the thickener while the filter cake is discharged for drying and packaging.

Stage 7 - Drying and packing

The filter cake has a residual moisture content in the range of 15 to 22% w/w. A portion is packaged for sale and the remainder fed to a fluid bed dryer to produce a free flowing powder

with 8 to 11% w/w moisture content. The cobalt oxide/hydroxide product is packaged in bulk bags and containerised for shipping.

Stage 8 - Zinc and iron sulfidation

Zn LSL, the zinc and iron laden sulfuric acid solution from stage 2 stripping, is treated with ammonium hydrosulfide solution in a stirred tank reactor to precipitate the zinc and iron present. The solids are filtered and washed on a batch basis and the zinc sulfide cake (typically 65% Zn dry basis) is packaged for sale.

Cobalt product specification

The final purified product, QN ChemGrade Cobalt, is a moisture-containing powder with a mean particle size of 9 to 15 microns. It consists of a mixture of $CoOOH$, Co_3O_4, $CoCO_3$ and free water. QN ChemGrade Cobalt is ideally suited for production of various inorganic and organic cobalt chemicals used in many applications including the production of rechargeable batteries, video tapes, ceramics and glassware (pigments), cutting tools (extra fine cobalt powder) and synthetic fibres. QN ChemGrade Cobalt can be used to produce cobalt chemicals such as cobalt sulfate, cobalt chloride, cobalt nitrate, cobalt hydroxide, cobalt acetate and cobalt oxide.

Table 4.   Typical composition of QN ChemGrade Cobalt

NICKEL PRECIPITATION, CALCINATION AND REDUCTION

Nickel loaded strip liquor treatment

Ammonium hydrosulfide solution is metered in-line to the separate scalp and follow Ni LSL streams to selectively precipitate the small amount of copper impurity (typically <0.05 g/L). The sulfided follow Ni LSL is directed to a 15 m diameter clarifier. The clarifier overflow is combined with the sulfided scalp Ni LSL in the filter feed tank. Polishing filters are used to remove trace solids and thus prevent copper and sulfur contamination of the final product. The mixed copper-nickel sulfide precipitate (CuNiS, typically 25% Cu and 20% Ni dry basis) is pumped from the clarifier underflow to a plate and frame filter press with diaphragm pressing. The filter cake is washed and then collected in bulk bags for sale.

Basic nickel carbonate precipitation

Sulfided and filtered Ni LSL is fed to four parallel stills that completely strip off ammonia and ammonium carbonate to precipitate the nickel ions as basic nickel carbonate (BNC, typically 52% Ni dry basis). The feed liquor is first preheated using hot overflow from the BNC product thickeners in recuperative plate heat exchangers. BNC slurry from the downstream thickeners is recycled as a seeding material to promote particle growth in order to improve the dewatering properties. The 2.7 m diameter by 24 m high stills have fourteen single bubble cap plates, which gradually silt due to the precipitating nickel, resulting in increasing pressure drop across a still. Each still is chemically cleaned using ammonia leach solution every four days.

The BNC slurry is processed in one of two independent parallel nickel reduction plants, Line 1 and Line 2. Line 2, which was commissioned in 2007 as part of the brownfield expansion to process MHP from the Ravensthorpe Nickel project, has a nominal capacity of 45 kt/a Ni compared with 33 kt/a Ni for Line 1. In each nickel reduction plant the basic nickel carbonate slurry is dewatered prior to reporting to a rotary kiln – reduction furnace – sinter furnace circuit for calcination, reduction and sintering into saleable nickel compacts.

Basic nickel carbonate dewatering

Dewatering is accomplished in two stages: thickening to 35-40 %w/w solids in two high-rate thickeners, and filtering to >60 %w/w solids. The Line 1 thickener is 18 m diameter and the Line 2 thickener is 23 m diameter. More than one half of the total thickener underflow is returned to the Ni LSL stills as seedings. The remainder of the thickener underflow is further dewatered in two Larox pressure filters. A portion of the BNC filter cake is packaged in bulk bags for sale. In the event of either calciner being off-line, BNC slurry from the thickener underflows is stored in two 2700 m^3 agitated slurry storage tanks. The overflows from both thickeners are combined and pumped to settling ponds via the recuperative plate heat exchangers that preheat the Ni LSL still feed.

Basic nickel carbonate calcination

The filtered BNC is fed to two parallel rotary kilns; the Line 1 kiln is 3.8 m diameter by 55 m long and the Line 2 kiln is 4.4 m diameter by 70 m long. The kilns are direct fired, using coal seam gas, counter-current to the BNC feed. The Line 1 calciner has the ability to be fired in either oxidising or reducing mode; the Line 2 calciner is only operated in reducing mode. The solids residence time is approximately 3 h and the burner zone temperature is approximately 1200°C.

Under oxidising conditions, the BNC feed is dried and decomposed into water, carbon dioxide and nickel oxide. The Line 1 calcine produced under oxidising conditions is sold as granular or powder nickel oxide products (NiO, typically 78% Ni).

Under reducing conditions, the BNC feed is dried, decomposed, and the nickel oxide is partially reduced to nickel metal. The Line 1 and Line 2 calcine produced under reducing conditions is a mixture of approximately 60% nickel metal and 40% nickel oxide (nickel metal/oxide calcine, typically 90 to 92% Ni).

The calciner off-gas cleaning in Line 1 is accomplished using a multiclone and a dry electrostatic precipitator. In Line 2, the off-gas passes through a baghouse before exhausting through the stack. In both Lines, the captured dust is blended directly with their respective calciner feeds and returned to the main process stream. In reducing mode, sulfur impurity tends to concentrate in the dust. To meet the product sulfur specification, a portion of the dust is bled out and stored until it can be processed through the Line 1 kiln in oxidising mode.

Each calciner discharges into a water-cooled rotary cooler that cools the calcine to <100°C. These discharge through trommel screens into oversize crushing and milling systems, which protect the downstream processes from blockage by clinker or other large material.

Calcine reduction

The nickel metal/oxide calcine produced in reduction mode is blended with sawdust and then fed to three parallel travelling-belt reduction furnaces. The calcine/sawdust mix is deposited at a depth of typically 3 cm across the full width of a stainless steel strip flat belt, which travels on rollers at a uniform speed (typically 0.5 to 1.1 m/min), carrying the feed material through successive heat up, reduction and then cooling sections. The two 34 m long Drever furnaces in

Line 1 have 1.2 m wide belts and are electrically heated, while the single 80 m long Line 2 furnace has a 2 m wide belt and is fired with coal seam gas. The furnaces are gas tight, with atmosphere seals at both ends, allowing a controlled reducing atmosphere to be maintained inside the furnace. Haywood (2003) described the development of a computational fluid dynamic model of QNI's nickel reduction furnaces.

Up to 5% w/w sawdust is added to the feed to increase the permeability of the bed and act as a supplementary reductant in the ensuing reduction step. Ammonia synthesis gas (syngas, a mixture of hydrogen and nitrogen in the ratio of three to one by volume) is injected into each reduction furnace near the discharge end and flows counter-current to the travelling bed of nickel metal/oxide calcine. Off-gas is vented and flared at the feed end of the furnace.

The calcine is heated to approximately 1000°C for 1 h and further reduction of the nickel oxide takes place by the action of hydrogen and carbon monoxide. The nickel content increases from typically 92% to 97.5% in the reduction furnaces. To prevent re-oxidation of the nickel metal, the reduced calcine is cooled indirectly at the end of the furnace to <100°C before passing through an inert gas purge chamber and then being discharged off the travelling belt.

Reduced calcine handling and compaction

The reduced calcine partially agglomerates at the high furnace temperatures and must be crushed before being stored in the reduced calcine bins. Fused material is sized and milled to ensure that only <2 mm material is fed to the compactors. Stearic acid is mixed with the reduced calcine to act as a binder (and provide lubrication) for the downstream compactors, and act as a reductant in the sintering furnace. In Line 1, mixing is accomplished in a 30 t batch blender while Line 2 uses small high-intensity solid/solid rotary mixers. Courtoy rotary type tabletting presses compact the nickel powder and discharge the approximately 100 g nickel compacts onto mesh belt conveyors feeding the two sintering furnaces.

Nickel compact sintering

The two travelling-belt compact sintering furnaces are similar to the reduction furnaces. The 58 m long furnace in Line 1 has a 1.2 m wide belt and is electrically heated, while the 61 m long Line 2 furnace has a 2 m wide belt and is heated using coal seam gas-fired radiant elements.

Compacts are continuously fed onto the welded stainless steel mesh furnace belts as a single layer through an atmosphere seal section and then pass into the heat-up, burn-off and furnace sections that heat and sinter the compacts at approximately 900°C for 1 h. Syngas is injected into each sintering furnace near the discharge end and flows counter-current to the travelling bed of nickel compacts, to accomplish the final reduction to nickel metal. Off-gas is vented and flared at the feed end of the furnace. The nickel content increases from typically 97.5% to >99% in the sintering furnaces.

After the heating sections, the compacts travel through a slow cooling section where they exchange heat with the counter-current syngas stream. The compacts then enter the Convecool section where high velocity impingement of a cooled recirculating gas ensures rapid cooling. Finally, the product exits through the discharge purge chamber before being discharged from the travelling belt at typically <100°C.

Packing

In Line 1, the compacts discharge from the sinter furnace into a single 10 t hopper. In Line 2, the sintered compacts are transferred to two 72 t live capacity storage hoppers, each on load cells. The product is discharged into 2 t bags or directly into 25 t containers. The bags and containers are weighed before sale.

Nickel product specifications

QN Nickel Carbonate, produced as a moist filter cake consisting of a mixture of amorphous nickel hydroxide and nickel carbonate, is ideally suited for production of various inorganic and organic nickel chemicals used in many applications.

QN Nickel Oxide, produced in both fine powder and granular forms, is suitable for production of soft ferrite powders, pigments and catalysts.

QN Nickel Compacts and QN HiGrade Nickel metal products, produced in the form of 30 mm cylindrical compacts with two flat sides to eliminate rolling on conveyors, are designed for use by the nickel alloy, stainless steel and specialty steel industries.

Table 5.   Typical compositions of QNI nickel products

Figure 5.   QNI Nickel Compacts

GAS RECOVERY SECTION

Steam stripping stills

Ammonia and carbon dioxide gases are recovered from various streams: -

- Ni LSL is steam stripped in four parallel stills to precipitate a purified slurry stream containing nickel as BNC for further processing in the final nickel products section.
- Ore tailings slurry underflow from thickener No 8 in the CCD circuit and some other waste streams are steam stripped to produce ammonia-free streams suitable for discharging to the tailings storage facility.
- CPL is partially steam stripped in four preboil stills to produce SPL for further processing in the ASX plants.
- Cobalt sulfide slurry is steam stripped in two stills to produce an ammonia-free stream suitable for acid leaching in the cobalt refinery.
- Weak leach liquor (WLL, typically 10 to 30 g/L NH_3) from scrubbers in the ore and MHP leaching sections, along with WLL from scrubbers in various other areas of the plant, is steam stripped in two stills to produce ammonia-free water suitable for scrubbing ammonia rich off-gas from the leaching circuits.

Figure 4.   Steam stripping columns used for ammonia and carbon dioxide recovery

Gas cooler condensers

The first stage of gas recovery from off-gases from the various stills is gas cooling and condensing. This is performed in the water-cooled gas cooler condensers (GCC). These heat exchangers are either plate and frame or shell and tube. The condensate is a concentrated ammonia ammonium carbonate solution (typically 250 to 300 g/L NH_3 and 180 to 240 g/L CO_2), most of which is forwarded to the various solvent extraction circuits for use as strip liquor. The remainder is added to the FLL stream.

Absorbers

In the second stage of gas recovery, the off-gases from the GCCs are contacted counter-currently in packed column absorbers with part of the overflow of thickener No 6 in the CCD circuit. The absorbers are filled with Pall rings in order to increase the contact surface area. The absorption is exothermic; therefore the absorber solution is continuously recycled through external water-cooled plate heat exchangers. A bleed of the absorber solution is returned to thickener No 5 in the CCD circuit as FLL. A portion of the FLL is transferred without cooling to the MHP secondary leach.

The final off-gases from the gas absorption area are scrubbed using water in order to minimise atmospheric emissions.

AUXILIARY PLANT

Most reagents and utilities required by the process are prepared on site. These include process gases, power, steam, several water grades, and compressed air. Reagents of note include ammonia synthesis gas, carbon dioxide gas, aqua-ammonia solution and ammonium hydrosulfide solution.

Power and steam

The original plant design utilised three oil-fired boilers (No 1, 2 & 3), each with a maximum continuous rating of 114 t/h, for both power and steam generation. A new, large, pulverised coal-fired boiler (No 4) was commissioned in 1982 in response to the oil crisis. Boiler No 4, with a maximum continuous rating of 249 t/h, satisfied the total steam demand until the site expansion

in 2007. Boiler No 3 was converted from oil to coal seam gas in 2006 and operated in parallel to boiler No 4 to meet total process steam demand. The two oil-fired boilers (No 1 & 2) are available to provide backup.

Three 12.5 MW turbo-generators supply most of the plant requirement for power and 70 kPag turbine back pressure steam is used for ammonia and carbon dioxide distillation from process liquor and tailings streams. The refinery is synchronised with the state power grid and purchased power is available to meet peak demand of approximately 40 MW or while a turbine is out of service.

Steam at three different pressures; high pressure (HP), medium pressure (MP) and low pressure (LP) is sourced as follows and distributed plant wide via an integrated piping network:

- HP steam (at 4.1 MPag) is generated by the boilers and is primarily used in the on-site generation of power. The remainder is distributed to the plant as high and medium pressure steam.

- MP steam is supplied at a range of pressures (ie 205, 345, 460, 1040 & 2770 kPag) and is generated from HP steam by letdown stations in the boiler house and in local areas of the plant.

- LP steam (at 70 kPag) is power turbine exhaust steam and is distributed along major LP steam headers for use at the various stills (i.e. preboil, Ni LSL, tailings, cobalt sulfide and water management). Supplementary LP steam demand is met by de-superheating HP steam in the boiler house. Any excess due to power generation / process steam demand imbalance is vented.

Gas plant

The plant consists of several units, namely, the two coal seam gas reformers, the carbon dioxide recovery system (Benfield system using potassium carbonate), an ammonia plant, an ammonium hydrosulfide production unit and a carbon dioxide absorption plant. Reformed coal seam gas is used to produce ammonia synthesis gas (syngas, a mixture of hydrogen and nitrogen in the ratio of three to one by volume). Syngas is used for production of hydrogen sulfide and ammonia, and as reductant for nickel reduction and sinter furnace operations. The plant supplies aqueous ammonia solution (typically 25% w/w NH_3) to the process for make up and ammonium

hydrosulfide solution (typically 12 % w/w H_2S) for precipitation of cobalt-nickel sulfide, cobalt sulfide, copper-nickel sulfide and zinc sulfide. Additional anhydrous ammonia is purchased to meet the total plant ammonia requirement of approximately 15 kt/a. Carbon dioxide from the reforming plant is purified and used to make up losses in the process and for inert gas purging needs. Additional carbon dioxide requirements are provided by recovery from boiler No 3 off-gases.

Water supply

The Yabulu Refinery consumes on average 20 000 m^3/d of water, which is supplied by:

- borefield make-up to the main site reservoir
- municipal water primarily from the Mt. Spec reservoir.

Water is treated in several steps and a range of water types is distributed plant wide via an integrated piping network - boiler feed water, process water, demineralised water, cooling water, fire water, potable water, service water and recycle water.

Reverse osmosis plant

The Yabulu plant site borders the coast of Halifax Bay. Historically ocean discharge was the accepted method to manage excess tailings water volumes, within the license conditions set by the Environmental Protection Agency. Scott and Richardson (2001) described the commissioning of a reverse osmosis (RO) water regeneration plant in 2000, which has allowed up to 12 000 m^3/d of tailings water to be recycled for re-use in the plant. This has the dual benefits of eliminating the practice of effluent discharge to the sea and reducing the quantity of bore water that is pumped to the plant. Clean water is combined with service water in the reservoir. The RO brine is pumped to saline evaporation ponds.

Cooling water

Cooling water is supplied to the plant through two independent sets of cooling towers at an average temperature of 33°C. Cooling water is treated in the cooling tower basins with various chemicals to inhibit corrosion, disperse sludge and to control the pH. Sodium chlorite is also

injected into the cooling water to inhibit algae growth. Water is constantly blown down to waste to control suspended solids, TDS and chlorides.

Compressed air

Compressors at the power plant produce plant air at a delivery pressure between 620 and 690 kPag. Instrument air is produced at similar pressures by removing water and oil from plant air. Compressed air is distributed plant wide.

Energy

Heavy fuel oil is shipped into the Townsville Port and pumped 25 km through a dedicated buried pipeline to two 25 kt storage tanks at the Yabulu plant site. About 160 kt/a of oil is used in the ore roasters and through intermittent use of the standby boilers No 1 & 2.

Coal from the Collinsville Mine (300 km south of Townsville) is railed to Yabulu and stored on a 20 kt stockpile. About 330 kt/a of coal is used in the ore dryers and boiler No 4.

Coal seam gas is supplied by pipeline from coalfields in the Bowen Basin. Approximately 5 PJ/a of gas is used in one ore dryer, the gas production plant, the nickel production plant and boiler No 3.

PROCESS CONTROL AND ANALYSIS

The Yabulu Refinery has eight separate Control Rooms across the plant, which are being progressively connected to the ABB System6 Distributed Control System (DCS). Most equipment is controlled remotely from the Control Rooms and the plant is generally controlled automatically during normal operation.

A Production Information Management System (PIMS) provides real-time and historical data to assist with process analysis, management and reporting as well as on-going improvement. All personnel have access to the PIMS through their desktop PC's connected to the plant IT network. The PIMS is based on an OSIsoft PI data historian, which collates data from the ABB System6

DCS, several Process Logic Controllers in the production areas, electronic logbooks used by plant operators and maintainers, and the Laboratory Information Management System in the central analytical laboratory. The analytical laboratory provides all chemical analyses required for quality control of the nickel and cobalt refining processes, and analyses of the ore shipment samples, MHP shipment samples and final nickel and cobalt product lot samples.

PRODUCTION COSTS

The Yabulu Nickel Refinery has been in operation since 1974, and has undergone many changes that have affected the production costs of the nickel and cobalt products. In addition to the costs of purchasing the imported laterite ore and MHP feed stocks, there are significant production costs associated with operation of the complex metallurgical refinery including energy, maintenance, labour and reagents. Reid (1996) and Reid and Barnett (2002) compared the production costs of the Yabulu Nickel Refinery with those of other nickel operations. Some of the major changes that have made Yabulu a more cost competitive nickel and cobalt producer are described below.

Feed stocks

The Yabulu plant was built to process ore from the Greenvale ore-body, which never fulfilled the original expectations in terms of either nickel grade (from original drilling results) or metal recovery (from piloting results). It made good sense to locate the refinery on the coast rather than adjacent to the ore-body, to allow for future ore importation. (The bulk of the world's nickel laterite deposits lie in the Asia Pacific region.) Reid (1994a) described the change of nickel laterite ore source from the domestic Greenvale Mine to imports from New Caledonia, Indonesia and the Philippines. Reid (1996) noted that as a consequence, the delivered ore cost to Yabulu is relatively high, being above US$1.00/lb contained Ni (varies with currency exchange rates, etc), and is a major unit cost. Reid and Fittock (2004) noted that a most significant impact of the progressive change from Greenvale to imported ore was an increase in nickel and cobalt production due to the higher grades and recoveries associated with the imported ore. In the ten years between 1979 and 1989, nickel production from Greenvale ore had been about 20 kt/a and cobalt production averaged about 900 t/a. During the five-year transition to processing 100%

imported ore in 1995, nickel production progressively increased by nearly 2 kt/a to 28.7 kt/a and cobalt by more than 100 t/a to 1480 t/a.

Rather than increase ore imports as a means to expand production further, the recent brownfield expansion has been based on the treatment of a mixed nickel-cobalt hydroxide concentrate. This required an expansion of the nickel and cobalt refining sections of the Yabulu operation, while the ore treatment section was not affected. As a consequence of the economies of scale achieved through this expansion, the production cost (US$/lb Ni) and labour efficiency (t Ni/employee/year) have been significantly lowered.

Energy

The Yabulu Refinery was originally designed with oil as the primary fuel. In the years immediately following commissioning in 1974, the project suffered from the four-fold increase in the oil price. Consequently there has been a progressive shift to cheaper energy sources. Fries (1987) described the changeover of two ore dryers to coal-fed spreader grate stoker air heaters (commissioned in 1981) and the installation of the pulverised coal-fired boiler No 4 to replace the three oil-fired boilers (commissioned in 1982). Following the change from domestic Greenvale ore to the higher moisture imported ore in the 1990's, solar drying was chosen in preference to building a fourth coal-fired dryer.

A pipeline to supply coal seam gas (predominantly methane) to Yabulu was completed in 2004, prior to the recent major brownfield expansion. Thus the Line 2 nickel calciner, reduction furnace and sinter furnace are designed with coal seam gas-firing. The remaining oil-fired dryer, naphtha-fired Line 1 calciner, one of the three oil-fired boilers and the naphtha-fired syngas plant were converted to coal seam gas in 2006. Conversion of the oil-fired burners on the ore roasters to coal seam gas is currently being studied, however it involves some technical risk to the process and is costly.

Refining technology

The original design of the Yabulu Refinery, based on the Caron process of reduction roasting and ammonia ammonium carbonate leaching to dissolve nickel and cobalt as ammine complexes, is

described by Reid (1978, 1979 and 1980). Two major innovations in nickel and cobalt refining have been implemented at the Yabulu Refinery over the last 30 years.

The most significant is the installation of the ASX process. In 1989 QNI commissioned a solvent extraction plant capable of separating nickel from cobalt, both present as ammines in Caron type ammoniacal liquors. Development and installation of the first ASX plant is described by Price and Reid (1993), Reid and Price (1993) and Reid (1994b). Reid (1995) described the improvements in nickel product quality that resulted from the introduction of solvent extraction. Skepper and Fittock (1996) described some subsequent operational changes to the first ASX plant. A second ASX plant was installed in series with the first plant, as part of the major brownfield expansion.

Reid and Fittock (2004) discussed the significant benefits of the ASX process, which include:
- Efficient separation of nickel and cobalt.
- Reduction of impurity levels in both nickel and cobalt products.
- Ability to increase metal production - nickel by 50%, cobalt by 100%.
- Reduction in energy consumption.
- Reduction in hydrogen sulfide consumption.
- Reduction in product shipping costs.

The installation of the ASX circuit in 1989 opened the opportunity for the second major innovation in refining technology - value adding to the cobalt product through refining the cobalt sulfide intermediate to a high purity cobalt chemical product. Gray, Price and Fittock (1996) and Fittock (1997) described the development of QNI's cobalt refining process, from laboratory bench scale studies initiated in 1990 through to full-scale production of QN ChemGrade Cobalt containing low levels of impurities in 1997.

Reid and Fittock (2004) noted that the operating cost of the cobalt refinery turned out to be about equivalent to that of the previously operating cobalt sulfide product circuit consisting of spray dryer, compactors and packaging plant. This meant that the entire cash flow benefit of eliminating the cobalt sulfide refining charge flowed to the project, resulting in a short payback time on the A\$35M refinery construction cost.

REFERENCES

Fittock, J E, 1997. QNI Limited cobalt refinery: process development, installation and operation, in *Nickel-Cobalt 97 Volume 1 Hydrometallurgy and Refining of Nickel and Cobalt (Proceedings of 36th Conference of Metallurgists of CIM)* (eds: W C Cooper and I Mihaylov), pp 329-338 (Canadian Institute of Mining, Metallurgy and Petroleum: Montreal).

Fries, F A, 1987. Coal drying of Greenvale nickel lateritic ore, in *Proceedings The Australasian Institute of Mining and Metallurgy 1987 Annual Conference*, pp 87-93 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Gordon, A, 2003. Solar drying of laterite ores at QNI Yabulu refinery, in *Proceedings Eighth Mill Operators Conference*, pp 235-244 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Gray, J, Price, M J and Fittock, J E, 1996. Operation of a cobalt purification pilot plant, in *Value Adding Through Solvent Extraction Volume 1 (Proceedings of International Solvent Extraction Conference 1996)* (eds: D C Shallcross, R Paimin and L M Prvcic), pp 703-708 (The University of Melbourne: Melbourne).

Haywood, R, 2003. Process optimisation and design of a belt furnace for nickel oxide reduction, in *Proceedings Third International Conference on CFD in the Minerals and Process Industries*, pp 325-330 (CSIRO: Melbourne).

Hobson, R and Piper, R, 2002. Plant and process design improvements to increase the efficiency of nickel reduction at QNI, in *Proceedings Metallurgical Plant Design and Operating Strategies*, pp 221-247 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Mackenzie, J and Lumsdaine, I, 2002. Characterising cobalt losses for improved recoveries, in *Proceedings Metallurgical Plant Design and Operating Strategies*, pp 367-372 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Price, M J and Reid, J G, 1993. Separation and recovery of nickel and cobalt in ammoniacal systems – process development, in *Solvent Extraction in the Process Industries Volume 1 (Proceedings of International Solvent Extraction Conference 1993)* (eds: D H Logsdail and M J Slater), pp 159-166 (Elsevier Applied Science: London and New York).

Reid, J G, 1978. Metallurgical operations at the Yabulu nickel refinery of Queensland Nickel Pty Ltd, in *Proceedings The Australasian Institute of Mining and Metallurgy 1978 Annual Conference*, pp 347-354 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Reid, J G, 1979. Operations at the Greenvale nickel project mine and refinery, in *International Laterite Symposium* (eds: D J I Evans, R S Shoemaker and H Veltman), pp 368-381 (Society of Mining Engineers of the American Institute of Mining, Metallurgical, and Petroleum Engineers Inc: New York, NY).

Reid, J G, 1980. Nickel-cobalt ore leaching by Queensland Nickel Pty Ltd, Yabulu, Qld, in *Australasian Mining and Metallurgy* (ed: J T Woodcock), pp 585-590 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Reid, J G and Price, M J, 1993. Ammoniacal solvent extraction at Queensland Nickel: process installation and operation, in *Solvent Extraction in the Process Industries Volume 1 (Proceedings of International Solvent Extraction Conference 1993)* (eds: D H Logsdail and M J Slater), pp 225-231 (Elsevier Applied Science: London and New York).

Reid, J G, 1994a. The Yabulu nickel refinery transition from Greenvale ore to imported ore, in *Proceedings The Australasian Institute of Mining and Metallurgy 1994 Annual Conference*, pp 261-263 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Reid, J G, 1994b. Hydrometallurgy at work – progression from concept to reality, in *Proceedings 6th Australasian Institute of Mining and Metallurgy Extractive Metallurgy Conference*, pp 11-15 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Reid, J G, 1995. Queensland Nickel – value added products from technological advances, in *Proceedings The Australasian Institute of Mining and Metallurgy 1995 Annual Conference*, pp 131-135 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Reid, J G, 1996. Laterite ores – nickel and cobalt resources for the future, in *Proceedings Nickel '96 Mineral to Market* (eds: E J Grimsey and I Neuss), pp 11-16 (The Australasian Institute of Mining and Metallurgy: Melbourne).

Reid, J G, and Fittock, J E, 2004. Yabulu 25 years on, in *Proceedings International Laterite Nickel Symposium 2004* (eds: W P Imrie and D M Lane), pp 599-618 (The Minerals, Metals, and Materials Society: Pennsylvania).

Reid, J G and Barnett, S C C, 2002. Nickel laterite hydrometallurgical processing update, in *Proceedings ALTA 2002 Nickel/Cobalt 8* (ALTA Metallurgical Services).

Scott, J and Richardson, D, 2001. QNI Yabulu refinery water re-use project, in *Proceedings ALTA 2001 Nickel/Cobalt 7* (ALTA Metallurgical Services).

Skepper, I G and Fittock, J E, 1996. Nickel cobalt separation by ammoniacal solvent extraction: the operating experience, in *Value Adding Through Solvent Extraction Volume 1 (Proceedings of International Solvent Extraction Conference 1996)* (eds: D C Shallcross, R Paimin and L M Prvcic), pp 777-782 (The University of Melbourne: Melbourne).

Willenborg, J R, 1985. Improvements in roaster operation at Queensland Nickel Pty Ltd, in *Smelting and Refining Operators' Symposium*, pp 155-159 (The Australasian Institute of Mining and Metallurgy: Melbourne).

List of figure captions

Figure 1. Aerial view of the Yabulu Refinery prior to the major brownfield expansion

Figure 2. Yabulu Refinery flowsheet in 2007 after the major brownfield expansion

Figure 3. Herreshoff roasters used for reduction of laterite ore

Figure 4. Steam stripping columns used for ammonia and carbon dioxide recovery

Figure 5. QNI Nickel Compacts

List of table captions

Table 1   Typical composition of imported nickel laterite ore

Table 2.   Typical composition of Ravensthorpe MHP

Table 3.   Typical composition of cobalt hydroxy hydrosulfide solids

Table 4.   Typical composition of QN ChemGrade Cobalt

Table 5.   Typical compositions of QNI nickel products

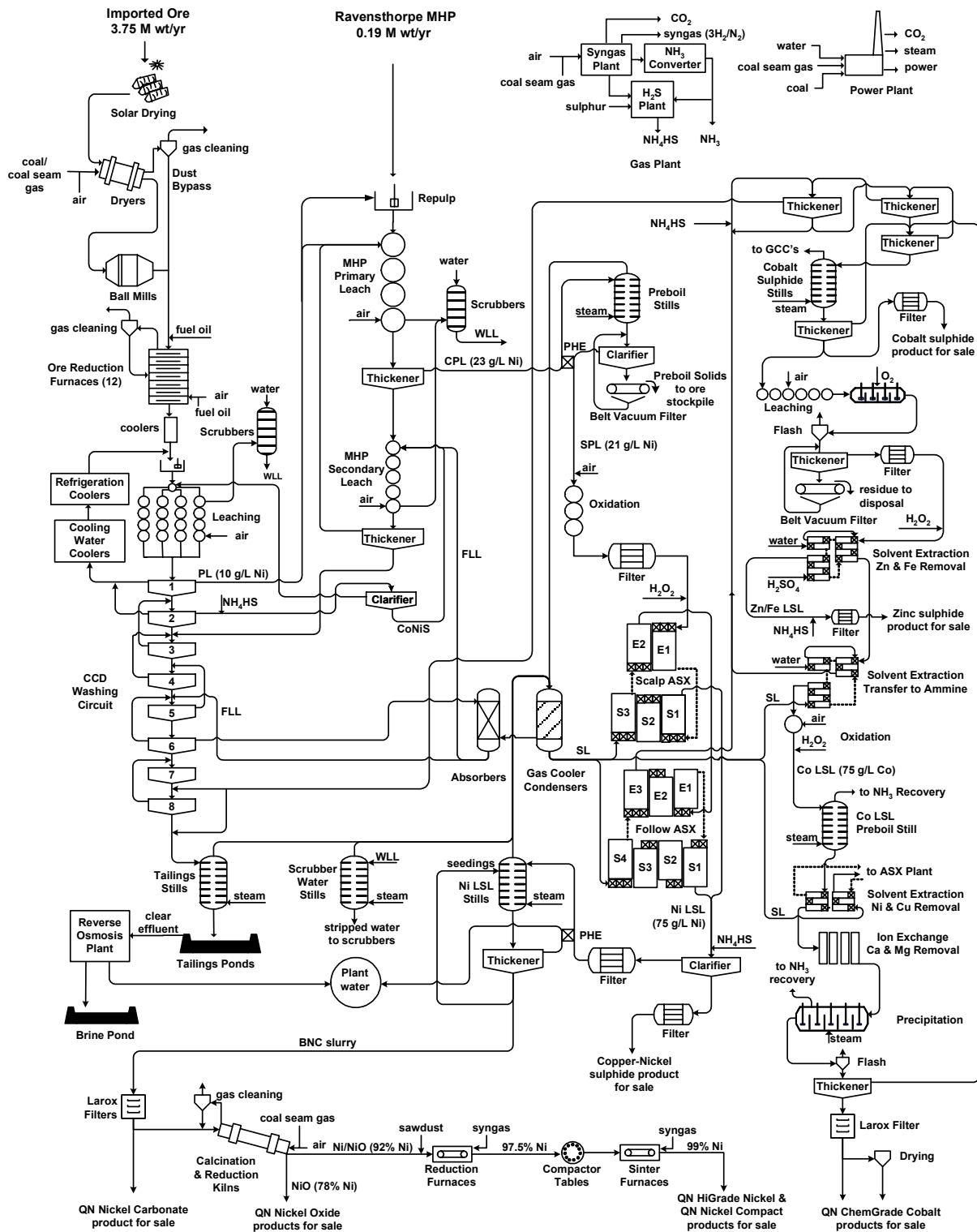Figure 1. Aerial view of the Yabulu Refinery prior to the major brownfield expansion

Figure 2. Yabulu Refinery flowsheet in 2007 after the major brownfield expansion

Figure 3. Herreshoff roasters used for reduction of laterite ore

Figure 4. Steam stripping columns used for ammonia and carbon dioxide recovery

Figure 5. QNI Nickel Compacts

Table 1. Typical composition of imported nickel laterite ore

Moisture (% wet wt)	Chemical Analysis (% w/w dry basis)											
	Ni	Co	Fe	Si	Mg	Mn	Cu	Zn	Ca	Al	Cr	S
35	1.6	0.15	37	7.2	5.2	0.8	0.01	0.05	0.1	2.1	1.2	0.2

Table 2. Typical composition of Ravensthorpe MHP

Moisture (% wet wt)	Chemical Analysis (% w/w dry basis)											
	Ni	Co	Fe	Si	Mg	Mn	Cu	Zn	Ca	Al	Cr	S
41	40	1.4	0.1	0.5	1.8	2.8	0.01	0.2	0.2	0.05	0.01	3.8

Table 3. Typical composition of cobalt hydroxy hydrosulfide solids

Chemical Analysis	% w/w dry basis
Ca	0.001 - 0.018
Co	38 - 45
Cu	0.002 - 0.007
Fe	0.02 - 0.22
Mg	0.02 - 0.09
Mn	0.002 - 0.10
Ni	1 - 6
S	32 - 36
Si	0.03 - 0.25
Zn	2 - 5

Table 4. Typical composition of QN ChemGrade Cobalt

Chemical Analysis	% w/w dry basis
Al	≤ 0.002
Bi	< 0.001
C	≤ 0.25
Ca	≤ 0.006
Cd	≤ 0.001
Co	64.5 - 67
Cr	≤ 0.002
Cu	≤ 0.001
Fe	≤ 0.002
K	< 0.001
Mg	≤ 0.005
Mn	≤ 0.002
N*	≤ 1.25
Na	≤ 0.005
Ni	≤ 0.006
P	< 0.002
Pb	≤ 0.001
S	≤ 0.02
Si	≤ 0.03
Zn	≤ 0.002

* analysis on wet sample

Table 5. Typical compositions of QN nickel products

Chemical Analysis (% w/w dry basis)	Nickel Carbonate	Nickel Oxide	Nickel Compacts	Hi Grade Nickel
Al	≤ 0.01	≤ 0.02	≤ 0.02	≤ 0.02
C	≤ 5.0	≤ 0.01	≤ 0.01	≤ 0.01
Ca	≤ 0.02	≤ 0.03	≤ 0.03	≤ 0.03
Co	≤ 0.03	≤ 0.03	≤ 0.06	≤ 0.06
Cr	≤ 0.001	≤ 0.001	≤ 0.001	≤ 0.001
Cu	≤ 0.005	≤ 0.01	≤ 0.01	≤ 0.01
Fe	≤ 0.01	≤ 0.02	≤ 0.02	≤ 0.02
Mg	≤ 0.02	≤ 0.04	≤ 0.04	≤ 0.04
Mn	≤ 0.03	≤ 0.06	≤ 0.06	≤ 0.06
Ni	> 50.5	≤ 78.5	≥ 99.0	≥ 99.45
S	≤ 0.05	≤ 0.005	≤ 0.02	≤ 0.01
Si	≤ 0.03	≤ 0.04	≤ 0.05	≤ 0.05
Zn	≤ 0.005	≤ 0.01	≤ 0.01	≤ 0.01

List of Headings

ABSTRACT	**(Level 1)**
PROCESS FLOWSHEET OVERVIEW	**(Level 1)**
NICKEL LATERITE ORE TREATMENT	**(Level 1)**
Characteristics of imported ore	(level 2)
Bulk materials handling	(level 2)
Ore receival	(level 2)
Solar drying and ore blending	(level 2)
Ore drying	(level 2)
Ore grinding	(level 2)
Ore reduction	(level 2)
Reduced ore leaching and washing	(level 2)
Disposal of washed tailings	(level 2)
MIXED NICKEL-COBALT HYDROXIDE TREATMENT	**(Level 1)**
Characteristics of MHP	(level 2)
MHP receival	(level 2)
MHP leaching	(level 2)
Repulp	*(level 3)*
Primary leach	*(level 3)*
Reductant preparation	*(level 3)*
Secondary leach	*(level 3)*
CCD washing	*(level 3)*
NICKEL-COBALT SEPARATION	**(Level 1)**
Preboil	(level 2)
SPL oxidation and filtration	(level 2)
Ammoniacal solvent extraction plants	(level 2)
Equipment	*(level 3)*
Nickel extraction	*(level 3)*
Nickel stripping	*(level 3)*
Organic cleaning and regeneration	*(level 3)*
COBALT PRECIPITATION AND REFINING	**(Level 1)**
Cobalt sulfide precipitation and washing	(level 2)
Stage 1 – Cobalt sulfide leaching and clarification	(level 2)

Stage 2 - Zinc and iron removal by solvent extraction   (level 2)

Stage 3 – Cobalt transfer to ammine solution by solvent extraction   (level 2)

Stage 4 - Nickel removal by solvent extraction   (level 2)

Stage 5 – Calcium and magnesium removal by ion exchange   (level 2)

Stage 6 - Cobalt precipitation   (level 2)

Stage 7 - Drying and packing   (level 2)

Stage 8 - Zinc and iron sulfidation   (level 2)

Cobalt product specification   (level 2)

NICKEL PRECIPITATION, CALCINATION AND REDUCTION   (Level 1)

Nickel loaded strip liquor treatment   (level 2)

Basic nickel carbonate precipitation   (level 2)

Basic nickel carbonate dewatering   (level 2)

Basic nickel carbonate calcination   (level 2)

Calcine reduction   (level 2)

Reduced calcine handling and compaction   (level 2)

Nickel compact sintering   (level 2)

Packing   (level 2)

Nickel product specifications   (level 2)

GAS RECOVERY SECTION   (Level 1)

Steam stripping stills   (level 2)

Gas cooler condensers   (level 2)

Absorbers   (level 2)

AUXILIARY PLANT   (Level 1)

Power and steam   (level 2)

Gas plant   (level 2)

Water supply   (level 2)

Reverse osmosis plant   *(level 3)*

Cooling water   *(level 3)*

Compressed air   (level 2)

Energy   (level 2)

PROCESS CONTROL AND ANALYSIS (Level 1)

PRODUCTION COSTS   (Level 1)

Feed stocks   (level 2)

Energy   (level 2)

Refining technology   (level 2)

REFERENCES   (Level 1)