DEVELOPMENT OF TECHNICAL AND ECONOMICAL PARAMETERS
AFFECTING PROCESS SELECTION TO TREAT
NICKEL LATERITE ORES

DESENVOLVIMENTO DE MODELO TÉCNICO-ECONÔMICO
PARA A SELEÇÃO DE ROTAS DE PROCESSOS APLICÁVEIS
A MINÉRIOS LATERÍTICOS DE NÍQUEL

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Abril/2008
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Dissertação de Mestrado apresentada ao Curso de Pós-Graduação em Engenharia Metalúrgica e de Minas da Universidade Federal de Minas Gerais

Área de concentração: Hidrometalurgia

Orientadora: Profª Virgínia S. T. Ciminelli

Belo Horizonte
Escola de Engenharia da UFMG
2008
Agradecimentos

Gostaria de agradecer primeiramente à Virgínia, minha orientadora, pelo suporte, pelas discussões e pelos grandes exemplos de determinação, organização e amor ao ensino e ao trabalho.

À Mineração Onça-Puma/Vale, pelo financiamento do trabalho e em especial aos colegas: Werter Silva e Valéria Lage, pelo apoio no contrato Vale-UFMG; à Ana Rita, Jota Junior, Marcelo Costa, Oliver Gobbo, Omar Antunes, Rogério Baldoino e Waldecy Castro, pelas discussões técnicas e agradável convivência na equipe de processo; ao Eduardo Drummond, Glaysson Oliveira, Guilherme Jácome, Roberta Milagres e Wagner Giorgini, pelas inúmeras explicações sobre avaliação econômica de projetos; à Aline Salgado, pela ajuda na área de mercado de níquel; ao Luis Renato, pelo apoio inicial, pelas discussões e revisão da dissertação e ao Renato Costa, pelos ensinamentos na área metalúrgica e pela revisão dos resultados.

À Vanessa Torres, pelo apoio dado na parceria Vale-UFMG e pela co-orientação em grande parte deste trabalho.

To Sunil Jakasera, for all the explanation on the nickel laterite field and the comments on the final revision

Aos funcionários e colegas do Departamento de Engenharia Metalúrgica e de Materiais, em especial a Christina, que sempre ajuda a todos os alunos, com muita boa vontade e um sorriso no rosto.

Aos meus pais, por me educarem desde pequena quanto ao valor do ensino e da busca pela instrução. Aos meus irmãos, familiares e amigos, pelo apoio sempre, e por me ajudar a tornar meus dias muito melhores.

E finalmente a DEUS, por nos permitir avançar em direção ao conhecimento técnico.
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ABSTRACT

Nickel laterite ores have increasingly become an important source of nickel due to depletion of sulfide ores. As a consequence, significant progress has been made developing novel process routes for treating nickel laterite ores. There are currently three main options: ferro nickel (FeNi), Caron (ammonia leaching) and high pressure acid leaching (HPAL) processes. Heap leaching (HL) is currently under development and many efforts are being directed to evaluate this process route. In order to select the process for a given project, initially a batch test work programme is carried out; nickel recovery and consumption of the main reagents are evaluated. The results of test work are used to conceptually assess the overall viability of the proposed project. The present work is aimed at developing a computer model that allows, at a very early stage, a preliminary evaluation of ore behavior and the technical and economical feasibility of the project for a given deposit. The model was developed considering HPAL, FeNi and HL processes and involved a detailed investigation of the operational conditions of these three process routes. The simulation, using the software MetSim, was carried out with technical and economic variables selected according to data available from industrial operations, projects under development and the literature. An economic model was built in order to compare the three process routes. An Excel file was created, linking selected MetSim model outputs to the economic model. The economic model consisted of evaluation of capital and operational costs and the calculation of the final viability (a preliminary Net Present Value - NPV) of the project. Therefore, the main outcome of the present work is a process model integrated with an economic model. The Sample A, from a project under development, was used in order to demonstrate the practical application of the model. The selected sample shows features of both typical limonite and saprolite ores. The behavior of this ore was evaluated considering the three processes. For this deposit, FeNi route was discarded and the main costs analysed. HPAL and HL were evaluated in more detail and the impact of some parameters, such as iron content, deposit size, nickel grade and product price were discussed.
RESUMO

Os minérios lateríticos de níquel têm sido alvo de muitos estudos devido ao esgotamento dos depósitos contendo sulfetos. Existem atualmente três importantes processos no tratamento dos minérios lateríticos: ferro níquel (FeNi), Caron (lixiviação amoniacal) e HPAL (lixiviação ácida sob pressão). A lixiviação em pilhas (HL) é uma rota ainda em desenvolvimento. Com o objetivo de selecionar a rota de processo para um dado projeto, ensaios em escala de bancada são realizados para avaliação da recuperação de níquel, consumo dos principais reagentes e de energia elétrica. Estas informações são utilizadas para avaliação conceitual da atratividade do projeto. O presente trabalho tem como objetivo desenvolver um modelo que permita uma avaliação do comportamento do minério e da atratividade do projeto, em um estágio inicial. O modelo de processo foi desenvolvido para três rotas de processo: HPAL, FeNi e HL e envolveu uma investigação detalhada das condições de operação de cada uma dessas rotas. A simulação, utilizando o software MetSim, foi realizada considerando variáveis técnicas e econômicas selecionadas de acordo com informações disponíveis em literatura, projetos em desenvolvimento e operações industriais. Um modelo econômico foi construído com a função de comparar as rotas. Um documento em Excel foi criado para alimentar o modelo econômico com dados de saída selecionados do MetSim. O modelo econômico permite avaliar os custos de capital, de operação e calcular a atratividade final (Valor presente preliminar) do projeto. Portanto, o produto principal do presente trabalho é um modelo de processo interligado a um modelo econômico. Uma amostra A, de um projeto em desenvolvimento, foi utilizada com o objetivo de demonstrar a aplicação prática do modelo desenvolvido. Este minério apresenta características tanto de minério saprolítico quanto de limonítico. Para este dado depósito, a rota FeNi foi descartada e os principais custos associados a ela foram discutidos. Os processos HPAL e HL foram avaliados de forma mais detalhada. O impacto de alguns parâmetros (tais como teor da impureza ferro, teor e preço do níquel, tamanho do depósito) na atratividade relativa dos processos foi analisado.
1 Introduction

Nickel ranks 24th in abundance, with elements such as iron, silicon, oxygen and magnesium, which, account for over 90% of the earth’s composition (MOSKALYK and ALFANTAZI, 2002). Main nickel minerals are sulfides, arsenides and silicates but the metal is also found disseminated in iron oxide minerals. Nickel is mainly used for alloy fabrication, such as stainless or other special steels.

Nickel occurs within two distinct ore types: sulfide and laterite. Details of each ore will be presented in Section 3. Approximately 60% of the world’s nickel production originates from sulfide ores, as presented in Figure 1.1 (million tonnes per year). However, laterite ores account for about 55% of the world’s nickel reserves, and are expected to gain a greater share of total nickel production in the future, as presented in Figure 1.2 (HEAP and TONKS, 2007). Nickel production from nickel sulfide and laterite ores are summarized in Table I.1.

Nickel laterite ores have increasingly become an important source of nickel. As a consequence, significant progress has been made developing new process routes for treating laterite nickel ores. Currently there are three different processes in operation for treating laterite laterite ores: ferro nickel (FeNi), Caron and high pressure acid leaching (HPAL) processes.

The FeNi process comprises mainly of crushing, drying, reducing and smelting. Examples of these operations are PT Inco (Indonesia), SLN Doniambo (New Caledonia), Cerro Matoso (Colombia), Codemin (Brazil) and Onça-Puma project (Brazil), which is now under construction. The Caron process consists mainly of drying, reduction, ammonia leaching and metal recovery. Examples of application of this process are found in Nicaro (Cuba), Punta-Gorda (Cuba), Companhia Níquel Tocantins (Brazil) and Yabulu (Australia). HPAL consists of pressure acid leaching, followed by several precipitation steps to produce a
nickel concentrate, either mixed hydroxyl sulphate or mixed sulphide. The concentrate is often treated by leaching and the metal recovered. Examples of current application of HPAL are found in Moa Bay (Cuba), Coral Bay (Philippines), Cawse (Australia), Murrin Murrin (Australia), Ravensthorpe (Australia) and some other projects under engineering/construction - such as Goro and Niquel do Vermelho - both owned by Vale Inco.

Figure 1.1: Nickel laterite versus sulfide production (HEAP and TONKS, 2007).

Figure 1.2: Nickel reserves – Sulfide and Laterite ores (HEAP and TONKS, 2007).
Table I.1: Global Nickel production from various sources (JIANKANG, et al., 2006)

<table>
<thead>
<tr>
<th>Ores</th>
<th>Production (tonnes/year Ni)</th>
<th>Processes</th>
</tr>
</thead>
<tbody>
<tr>
<td>Laterites</td>
<td>300,000</td>
<td>Smelting, Fe/Ni</td>
</tr>
<tr>
<td>Limonite</td>
<td>180,000</td>
<td>Ammonia leach (Caron)</td>
</tr>
<tr>
<td></td>
<td></td>
<td>HPAL¹</td>
</tr>
<tr>
<td>Nickel sulfide</td>
<td>600,000</td>
<td>Smelting –</td>
</tr>
<tr>
<td></td>
<td></td>
<td>Converting -Refining</td>
</tr>
</tbody>
</table>

¹ HPAL – High Pressure Acid Leaching

A forth route, which now is being studied in detail, is the heap leaching (HL). Some companies, such as Vale, BHP Billiton, Minara and European Nickel, are now testing it. The concept is similar to a copper heap leaching, a process well established and applied in many operations around the world. The nickel ore is heap leached, using sulfuric acid, after which the liquor passes through precipitation steps, to produce a nickel precipitate. The processes used to treat the liquor, to produce the precipitate and to obtain the metal are similar to those considered for the HPAL process. Details of each process are presented in Section 3. Besides these four process routes, there are some others under development, such as atmospheric leaching (this is already in commercial operation at Ravensthorpe – currently at ramp up stage - hybrid with HPAL), chloride leaching and sulfation-atmospheric leach process, the latter developed by Skye Resources.

The selection of the process route to be employed depends on several factors, including physical and chemical nature of the orebody, weather conditions and geographical location, which is related to supply and cost of required reagents and energy. Once the deposit is under evaluation, the project life cycle takes many years to complete. The main phases of a project life cycle are presented in Figure 1.3. In the first step (e.g. Concept or Front End Loading (FEL) I), the study is performed and decisions are made based on industrial practice database conciliated with preliminary batch test results. No capital and
engineering are spent at that level. At this preliminary stage, the overall feasibility of the proposed project is evaluated. Depending on a positive outcome it will be advanced to the next level (e.g. Pre-feasibility or FEL II), in which 10% of the engineering is done and 2% of the capital is spent. The following level is the Feasibility study or FEL 3, followed by Implementation or FEL IV and Operation or FEL V.

![Figure 1.3: Typical project life cycle (adapted from TWIGGE-MOLECEY and CLARRY, 2007).](image)

The aim of the work described in this document was undertaken to develop a model to predict the ore behavior and to evaluate the viability of a given deposit for a selected process route from FeNi, HPAL and HL. Caron process was not evaluated once new applications are hindered by the reported low nickel and cobalt recoveries. Emphasis is given to a comparison of HL with HPAL, although the former is a process still under development, and its scope of application is not well defined yet. With a process model linked with an economic model, any important parameter can be simulated and its impact on the project is predicted, which is helpful for decision-making, in a stage where many detailed information are not yet available. The simulation will be carried
out with technical and economic variables selected according to data available from current industrial operations, projects and the literature. Combined with the test work results, the model can be continuously calibrated in order to scrutinize the project under evaluation.
2 Relevance and Objective

At the present time, mining industry is growing at an unprecedented rate. The global metals consumption and hence their prices have increased to record levels. As a result, many efforts are being made to develop new ore deposits. In order to accelerate the project development new tools are required. In this context, it is anticipated that the present work will be useful for the decision-making process when evaluating new nickel laterite deposits at the conceptual level.

2.1 Specific objectives:

i) To study two processes that are currently applied to treat nickel laterites ores (FeNi and HPAL) as well as HL that is still under evaluation, in order to select the main technical parameters affecting the processes’ performance. The sources of information will be the published literature and Vale Inco projects under development.

ii) To build process models using the computer software programme, MetSim, with main metallurgical parameters as input values.

iii) To compare capital and operational costs for each process, under the selected operation conditions, based on published database information and Vale Inco projects.
3 Literature Review


World consumption of nickel is expected to grow by 4.3% a year over the next few years, rising from 1.4 million tonnes in 2006 to 1.7 million tonnes in 2011. In the non-stainless uses, nickel consumption will defy price-sensitivity by growing 4.2% in 2006 and 5.2% in 2007. Table III.1 shows nickel consumption, by end-use.

Nickel supply is usually more predictable than demand, with the expansion in finished production lifting from 1.6% in 2005 to 4.9% in 2006. Production from new projects in pipeline, such as Voisey’s Bay in Canada (60,000tpy), Ravensthorpe in Australia (currently at ramp up stage) (45,000tpy), Goro in New Caledonia (60,000tpy), Vermelho (46,000tpy) and Onça-Puma (57,000tpy) in Brazil, and will reach the market requirement or will be close to do so. As a result, meeting the growth in nickel demand will be increasingly dependent on new producers treating laterite ores using new technology.

Table III.1: Consumption, by end-use (HEAP and TONKS, 2007)

<table>
<thead>
<tr>
<th>End-use</th>
<th>%</th>
</tr>
</thead>
<tbody>
<tr>
<td>Stainless steel</td>
<td>64</td>
</tr>
<tr>
<td>Alloy steels</td>
<td>5</td>
</tr>
<tr>
<td>Non-ferrous alloys</td>
<td>14</td>
</tr>
<tr>
<td>Plating</td>
<td>7</td>
</tr>
<tr>
<td>Batteries</td>
<td>3</td>
</tr>
<tr>
<td>Other, including foundry</td>
<td>8</td>
</tr>
</tbody>
</table>

3.2 Nickel Geology

Nickel sulfide deposits are the primary nickel deposits, formed from magmatic process. Laterite deposits are formed from the weathering of nickel-bearing rocks, nickel being in silicate or oxide form. Individual chemistry of each deposit
is a product of its weathering history and of the composition of the parent rock. All the laterites are formed by weathering of the parent rock in areas of abundant rainfall (RICHARDSON et al., 1981). The extraction methods of nickel from these two ore types (e.g. nickel sulfides and laterites) are varied, at least at initial up-stream stages, because of the fundamental difference of the two ore types.

Nickel can also be found in sea nodules, as oxides and oxyhydroxides, at considerable depths on vast areas of the ocean floor. Processes to treat polymetallic sea nodules are still being evaluated. (MOSKALYK and ALFANTAZI, 2002).

Table III.2 shows the main nickel-bearing minerals found in the nickel deposits, and their composition. The nickel content shown in this table considers the typical nickel grade in those minerals. It might vary in the minerals, due to substitution. Pentlandite is the most common sulfide mineral of nickel. Millerite, Heazlewoodite and the Linnaeite series usually occur as minor constituents. The arsenides series appears in some sulfide ores and cobalt ores. Garnierite (saprolite material) is a hydrous-nickel-magnesium silicate, the most abundant nickel carrier in silicate ores. Nickeliferous limonite is a popular term used in referring to poorly crystalline to non-crystalline nickel-bearing ferric oxides. The water content of limonite varies widely (JOSEPH and BOLDT, 1967).

3.2.1 Nickel Laterite Ores

Rocks are decomposed when they experience chemical and mechanical action of tropical weather. Acid solutions are formed by the action of rainwater on decomposing organic surface material. These solutions attack and leach nickel out of the parent rock, carrying it downward. Reprecipitation of the nickel deeper in the ore deposit causes an enrichment of nickel values that range from 5 to 20 times the content in the ultramafic rock (RICHARDSON et al., 1981). Iron is the
first to precipitate and, as the ground water is further neutralized, the nickel, magnesium and silica start to precipitate. Nickel is less soluble than magnesium and, as a result, precipitates more readily.

As a consequence, laterite nickel deposits are covered by a layer rich in iron. Complete separation of iron and nickel into distinct zones is never observed. Depending on prevailing weathering factors, some or even most of the nickel may be retained in the upper zone, which is thus enriched in both iron and nickel through removal of magnesium and silicon. Since the iron in this material is mainly in the form of ferric oxide, usually lumped in the field under the name of limonite, such a nickel deposit is referred to as the limonitic type, or as a nickeliferous iron ore. This is in contrast to the silicate type of nickel deposit, often called saprolite ore, where the separation of the nickel from iron is more complete and the nickel is present as a constituent of silicate minerals formed during the lateritization process (JOSEPH and BOLDT, 1967). As it can be seen in Figure 3.1a, laterite deposits occur in layers. Typically, the top layer is made of iron-rich material, containing mainly of goethite. The next layer is composed of the limonitic ore, followed by a transition zone and the saprolite ore body. The difference in the profile for dryer and humid climates is also included in Figure 3.1a. The typical process applied to each zone is indicated in Figure 3.1b. Saprolite zone is usually treated by FeNi process, while limonite zone is treated by HPAL. Caron process usually treats ores from the limonite and the transition zones.

The limonitic fraction is comprised mainly of goethite, gibbsite and chromite. The saprolite layer contains talc, quartz, serpentine, olivine and garnierite. The oxidized lateritic ores normally contain impurities such as chromium, magnesium, manganese, iron and aluminum (MOSKALYK and ALFANTAZI, 2002).
Table III.2: Nickel-bearing minerals and theoretical formula (JOSEPH and BOLDT, 1967)

<table>
<thead>
<tr>
<th>Minerals</th>
<th>Formula</th>
<th>Nickel Content%</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Sulfides</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Pentlandite</td>
<td>(Ni, Fe)$_6$S$_8$</td>
<td>34.22</td>
</tr>
<tr>
<td>Millerite</td>
<td>NiS</td>
<td>64.67</td>
</tr>
<tr>
<td>Heazlewoodite</td>
<td>Ni$_3$S$_2$</td>
<td>73.30</td>
</tr>
<tr>
<td>Linnaeite series</td>
<td>(Fe, Co, Ni)$_3$S$_4$</td>
<td>Variable</td>
</tr>
<tr>
<td><strong>Polydymite</strong></td>
<td>Ni$_3$S$_4$</td>
<td>57.86</td>
</tr>
<tr>
<td><strong>Violarite</strong></td>
<td>Ni$_2$FeS$_4$</td>
<td>38.94</td>
</tr>
<tr>
<td><strong>Siegenite</strong></td>
<td>(Co, Ni)$_3$S$_4$</td>
<td>28.89</td>
</tr>
<tr>
<td><strong>Arsenides</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Niccolite</td>
<td>NiAs</td>
<td>43.92</td>
</tr>
<tr>
<td>Maucherite</td>
<td>Ni$_{11}$As$_8$</td>
<td>51.85</td>
</tr>
<tr>
<td>Rammelsbergite</td>
<td>NiAs$_2$</td>
<td>28.15</td>
</tr>
<tr>
<td>Gersdorffite</td>
<td>NiAsS</td>
<td>35.53</td>
</tr>
<tr>
<td><strong>Antimonide</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Breithauptite</td>
<td>NiSb</td>
<td>32.53</td>
</tr>
<tr>
<td><strong>Arsenate</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Annabergite</td>
<td>Ni$_3$As$_2$O$_8$.8H$_2$O</td>
<td>29.40</td>
</tr>
<tr>
<td><strong>Silicate and oxide</strong></td>
<td></td>
<td></td>
</tr>
<tr>
<td>Garnierite</td>
<td>(Ni, Mg)$_6$Si$<em>4$O$</em>{10}$(OH)$_8$</td>
<td>Variable up to 47%</td>
</tr>
<tr>
<td>Nickeliforous</td>
<td>(Fe, Ni)O (OH).nH$_2$O</td>
<td>Low but variable (1 to 1.7%)*</td>
</tr>
<tr>
<td>Limonite</td>
<td>(Fe, Ni)O (OH).nH$_2$O</td>
<td></td>
</tr>
</tbody>
</table>

* (DALVI et al., 2004).
Figure 3.1a: Typical-laterite profile (DALVI et al., 2004).

Figure 3.1b: Typical-laterite profile with the correspondent extraction process (ELIAS, 2001).

Figure 3.2 shows world nickel resources. As can be seen, Brazilian deposits are mainly laterite. The other countries important as sources of laterite ores are:
Australia, Indonesia, New Caledonia and Philippines. In Asia there is a project under development by European Nickel, known as Çaldag project. Sulfide deposits are mainly in Africa (e.g. South Africa and Botswana), Canada, Finland, China, Australia and Russia.

Figure 3.2: World nickel resources (Brook Hunt, 2005).

3.3 Extractive Metallurgy of Sulfide Nickel

Nickel is recovered from sulfide ores mostly by conventional crushing, grinding and flotation. The nickel concentrate is processed in a pyrometallurgical route, comprised of smelting in furnaces to separate and discard an iron-rich slag and to produce a nickel matte. This matte is comprised predominately of nickel and iron with minor constitutes of copper and cobalt. The matte is then subjected to further processing to eventually produce nickel metal and other metal byproducts. Figure 3.3 indicates commercial and new technologies available for production of nickel from sulfide ores. The commercial applications often involve smelting to produce a nickel matte and acid or ammonia leaching followed by nickel recovery by eletrorefining, electrowinning or hydrogen reduction. In the carbonyl process, the molten metal is granulated and reacts with carbon
monoxide to form nickel carbonyl. Some new technologies indicated in Figure 3.3, mainly classed as hydrometallurgical processes, are still under development, and none has been tested at commercial scale yet. Some examples are the Activox, CESL, bioheapleaching and Voisey’s Bay processes. The Activox process, owned by Norilsk, is now under evaluation in a pilot plant scale and commercial plant is now under construction, in Botswana, and comprises of fine milling and low temperature (e.g. 95 – 100°C) low pressure (e.g. 1000 kPa) oxidative leach. The nickel and cobalt in solution are then recovered by solvent extraction followed by eletrowinning. The CESL process, owned by Teck Cominco, is being evaluated in a pilot plant in Canada and comprises of a medium temperature (e.g. 150°C) and medium pressure (e.g. 1500 kPa) oxidation and leaching circuit, followed by conventional steps (e.g. solvent extraction and eletrowinning) to recover the metals. Bioheapleaching consists of ambient temperature and pressure heap leaching in the presence of microorganisms. Voisey’s Bay process is being tested in a demonstration scale in Canada and comprises of a chloride pre-leach, followed by pressure leaching, solvent extraction and eletrowinning to recover the nickel and cobalt.

3.4 Extractive Metallurgy of Laterite Nickel

As discussed in Section 3.2, typically nickel laterite deposits are comprised of an upper limonite zone and a lower saprolite zone. These two zones must be treated differently to efficiently recover the contained nickel, due to their different proportions of iron, magnesium, aluminium and silica. In general, the limonite zone is richer in iron and poorer in nickel, magnesium and silica than the underneath saprolite zone, which has higher nickel, magnesium and silica, but lower iron content.

Several factors influence on the choice of a process route, such as physical and chemical nature of the ore body, geographical location and marketability of the products. As outlined before, there are three processes currently used to recover nickel from laterite deposits. These are conventional ferro-nickel (FeNi),
reduction roast/ammonia leach processing (Caron) and high pressure acid leaching (HPAL). Figure 3.1b shows the relation between the deposit geology and the process route. Heap leaching is now being tested at pilot scale at several locations and is also considered in the present study.

3.4.1 Pyrometallurgical treatment

The pyrometallurgical treatment is used for the high nickel, low iron, saprolite ores and involves furnace smelting to yield an iron-nickel melt from which an iron-silica slag is skimmed off to leave an iron/nickel product (ferronickel). This typically has 20% - 40% nickel content. A high-grade nickel matte (up to 78% nickel) may be produced by refining of the initially smelted material. Pyrometallurgical treatment is less complex than the other two laterite treatment processes, and is a well proven technology operating at the PT Inco mine in Indonesia, the Falcondo mine in the Dominican Republic, the Cerro Matoso mine in Colombia and Codemin in Brazil, amongst others. Typical feed compositions for two pyrometallurgical operations (Cerro Mataso and PT Inco) are provided in Table III.3.

Nickel-bearing minerals in the saprolite ore cannot be liberated and separated to form a concentrate by conventional mineral processing operations, as in
sulfide ore processing. As a result, the conventional milling and flotation operations are not applied to this type of ore. One consequence of this is that the ore is fed to the processing plant at almost the same nickel concentration as the run-of-mine material. As a result, pyrometallurgical processing of nickel laterites is much more energy intensive than sulfide smelting. The objective in the pyrometallurgical process of saprolite ores is to transform nickel oxide to metal. Iron is always associated with nickel and due to their similar free energies of formation; iron is co-reduced with nickel resulting significant amount in the end product. (HATCH ASSOCIATES LTD, 2004).

Table III.3: Typical feed composition of two different pyrometallurgical operations (DALVI et al., 2004)

<table>
<thead>
<tr>
<th>Analyses (%)</th>
<th>Cerro Matoso</th>
<th>P.T. Inco</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ni</td>
<td>2.9</td>
<td>1.8</td>
</tr>
<tr>
<td>Co</td>
<td>0.07</td>
<td>0.07</td>
</tr>
<tr>
<td>Fe</td>
<td>14</td>
<td>18</td>
</tr>
<tr>
<td>Mg</td>
<td>9</td>
<td>10</td>
</tr>
<tr>
<td>SiO₂</td>
<td>46</td>
<td>34</td>
</tr>
</tbody>
</table>

The conventional pyrometallurgical process consists of crushing, drying, reduction, smelting and refining. A simplified block diagram is shown in Figure 3.4. The first step, crushing, prepares the run-of-mine ore to feed the metallurgical plant. The crushed material is fed to the dryer. The objective is to remove water in an energy-efficient way. Nickel laterite ores can contain up to 45% water, as moisture and also as chemically combined water in hydrated minerals. As a result, the energy requirements for the dehydration of the oxide ores can be substantial (PICKLES, 2003). The energy is provided from coal or any other available fuel. The dryer uses air for fuel combustion and to control the combustion gas temperature. The dryer product contains 10-20% moisture.

The dried material feeds the kiln. The objective at this stage is to remove free and crystalline moisture from the ore and to reduce the trivalent iron to divalent
iron. As the feed progresses along the length of the kiln, the remaining free moisture is driven off, and as the temperature rises crystalline water starts to dissociate.

As the temperature is increased, the water associated with goethite followed by magnesium-silicates are dissociated. Removal of crystalline water is essentially complete once the ore has reached 700°C - 800°C. The reducing agent (e.g. coal) is fed to the kiln at the point where the temperature exceeds 600°C. Under these conditions, approximately 85% of the trivalent iron oxide has been reduced to the divalent state. There is also some reduction of both iron and nickel to the metallic state, but in terms of energy optimization, the advantage of a pre-reduction stage in the kiln lies in the conversion of Fe$^{3+}$ to Fe$^{2+}$, which would otherwise require a significant amount of relatively expensive electrical energy in the smelter. Not all the coal is consumed in the kiln and the residual material entered into the electric furnace. From an energy point of view, it is desirable to raise the temperature as high as possible, limited by the onset of sintering, before transferring it to the smelting step (HATCH ASSOCIATES LTD, 2004).

![Block diagram for the FeNi process](image)

Figure 3.4: Block diagram for the FeNi process

The kiln product is then fed to the electric furnace. In this unit operation, the remaining material is smelted and separated into slag and metal phase. The nickel and iron are reduced in this step, as can be seen in the equations (3.1) and (3.2). Since the presence of iron dilutes the nickel grade of the product, the
amount of reduced iron is an important factor, and this is determined by the quantity of reductant employed in the kiln (JOSEPH and BOLDT, 1967).

\[
\text{NiO}_\text{s} + C_\text{s} = \text{Ni}_\text{s} + \text{CO}_\text{g} \quad (3.1)
\]

\[
\text{FeO}_\text{s} + C_\text{s} = \text{Fe}_\text{s} + \text{CO}_\text{g} \quad (3.2)
\]

One of the most important metallurgical aspects in a ferronickel plant is the SiO\textsubscript{2}/MgO ratio. Figure 3.5 shows slag liquidus and tapping temperatures for a range of slag SiO\textsubscript{2}/MgO ratios and slag FeO contents. Liquidus temperature is that for the kiln product fusion, while the tapping temperature is the process one. Operations will target superheats of about 50°C – 100°C. The only exception is Cerro Matoso operation, which has a higher SiO\textsubscript{2}/MgO ratio and a higher superheating (200°C). The superheat is necessary to produce a slag fluid enough for proper settling, convection and skimming, but not hot and fluid enough to corrode the sidewall build-up and refractory (HATCH ASSOCIATES LTD, 2004).

The required metal temperature for tapping is typically about 25°C above the metal liquidus and 25°C below slag liquidus. Slag viscosity, in addition to slag temperature, is an important factor affecting furnace temperature. The lower the slag viscosity, the thinner the stagnant layer for convective heat transfer between the slag and the metal, and the higher the heat transfer from slag to metal (VOERMANN et al., 2005).

High chromium in the calcine might affect the slag liquidus temperature and viscosity. Besides that, primary chromite grains tend to accumulate on the surface of ferronickel prills settling through the slag and form a mush layer at the slag/metal interface. These chromite layers impede the coalescence of the ferro nickel prills and slow the heat transfer between the slag and metal phases. No significant operating problem should be encountered for ores at 3% Cr\textsubscript{2}O\textsubscript{3} or less (HATCH ASSOCIATES LTD, 2004). This impact will not be evaluated in
The sidewall heat flux has a relationship between electrode current and SiO$_2$/MgO ratio and is represented by the equation (3.3). This is an important parameter for the furnace heat balance. A, B, C and D are scalars numbers and “I” is the current (A) (HATCH ASSOCIATES LTD, 2004).

\[ Q(\text{W/m}^2) = A + [B (\text{SiO}_2/\text{MgO})^2 - C \text{SiO}_2/\text{MgO} + D] \text{I} \]  

(3.3)

The ferronickel produced in the smelting step may contain impurities, most notably sulfur, chromium, carbon, silicon and phosphorus at levels that should be removed to provide a refined ferronickel in the specification required by the
market. The refining is a well-established technology, and is usually offered as a turn-key package from specialized engineering companies.

The first step in the refining is the elimination of sulfur (de-sulfurization), under reducing conditions. This is achieved by adding Na₂O or CaO. A source of silica is used to react with the oxygen, driving the reaction to the right. The reaction between sulfur and the soda ash is presented below (JOSEPH and BOLDT, 1967):

$$3\text{Na}_2\text{O}(s) + 2\text{S}(s) + \text{Si}(s) = 2\text{Na}_2\text{S}_2(s) + \text{Na}_2\text{SiO}_3(s)$$  \hspace{1cm} (3.4)

The second step in the refining process is operated under oxidizing conditions. Air is blown through the metal, which preferentially oxidizes most of the chromium and the remaining silicon to produce a chromium-rich slag, which is skimmed off. Then the lime is added to reduce phosphorus concentration, as indicated by the equation (3.5). Carbon is eliminated mainly as carbon monoxide (JOSEPH and BOLDT, 1967).

$$2\text{P}(s) + 5\text{FeO}(s) + 4\text{CaO}(s) = \text{Ca}_4\text{P}_2\text{O}_6(s) + 5\text{Fe}(s) + 3/2 \text{O}_2(g)$$  \hspace{1cm} (3.5)

3.4.2 High Pressure Acid Leaching process

High-pressure acid leaching (HPAL) was firstly applied in Moa Bay, Cuba in the early 1960’s. However, it is only since in the late 1990’s that this process received broader attention. Three new plants were built in Western Australia in the late 1990’s and another new plant commenced operation in the Philippines in 2004. There are several other projects under engineering/construction at present including Goro (New Caledonia) and Níquel do Vermelho (Brazil). A typical simplified HPAL block diagram is presented in the Figure 3.6:
The HPAL process is used more commonly for limonitic ores with high iron, containing low magnesium and silica. The process starts with crushing, followed by screening. In this step, it is commonly evaluated if the ore can be concentrated (beneficiated) by screening, where the enriched-silica fraction is separated from the ore.

The product from the beneficiation step is fed to the autoclave. Sulfuric acid leaching of limonite is performed at high temperatures (240-270°C) in acid resistant titanium lined autoclaves (or in Pachuka tanks as in Moa Bay). At this temperature, equilibrium vapor pressure reaches 33-35 atm. Typical operating parameters for autoclave leaching are: pulp density of 25-35% solids, 30-90 minutes residence time, 246°C (e.g., Moa) to 270°C (e.g., Goro), 200-520 kg/t of acid consumption per dry ore and 3800-5400 kPa pressure. (MOSKALIK and ALFANTAZI, 2002). Iron (in the trivalent state) and aluminum follow a dissolution-precipitation path, forming solid products. Iron in the form of goethite (α FeOOH) and aluminum in the form of boehmite (AlOOH) dissolve initially to soluble ferric and aluminum sulfates respectively, according to equations (3.6) and (3.7). Gibbsite (Al(OH)_3), the major aluminium phase in the limonite ore, is transformed to boehmite at around 135-155°C. Nickel and cobalt in the assumed form of oxides, dissolve according to the reactions (3.8) and (3.9) respectively and remain in the aqueous phase as sulfates (GEORGIOU and PAPANGELAKIS, 1998).
FeOOH$_{(s)}$ + 3H$_{(aq)}^+$ = Fe$_{3+}^{3+}$(aq) + 2H$_2$O$_{(aq)}$.

(3.6)

AlOOH$_{(s)}$ + 3H$_{(aq)}^+$ = Al$_{3+}^{3+}$(aq) + 2H$_2$O$_{(aq)}$.

(3.7)

NiO$_{(s)}$ + 2H$_{(aq)}^+$ = Ni$_{2+}^{2+}$(aq) + H$_2$O$_{(aq)}$.

(3.8)

CoO$_{(s)}$ + 2H$_{(aq)}^+$ = Co$_{2+}^{2+}$(aq) + H$_2$O$_{(aq)}$.

(3.9)

2Fe$_{3+}^{3+}$(aq) + 3H$_2$O$_{(aq)}$ = Fe$_2$O$_3$(s) + 6H$_{(aq)}^+$

(3.10)

Fe$_{3+}^{3+}$(aq) + SO$_{4}^{2-} +$ H$_2$O$_{(aq)}$ = FeOH$SO$_4$(s) + H$_{(aq)}^+$

(3.11)

2FeOH$SO_4$(s) + H$_2$O$_{(aq)}$ = Fe$_2$O$_3$(s) + 2SO$_{4}^{2-} +$ 4H$_{(aq)}^+$

(3.12)

3Al$_{3+}^{3+}$(aq) + 2SO$_{4}^{2-} +$ 7H$_2$O$_{(aq)}$ = (H$_3$O)Al$_3$SO$_4$$_2$(OH)$_6$(s) + 5H$_{(aq)}^+$

(3.13)

Al$_{3+}^{3+}$(aq) + SO$_{4}^{2-} +$ H$_2$O$_{(aq)}$ = AlOH$SO_4$(s) + H$_{(aq)}^+$

(3.14)

Ferric cations hydrolyze after the dissolution of goethite, forming hematite or basic ferric sulfate, as indicated by the reaction (3.10) and (3.11). Basic ferric sulfates transform to hematite (3.12). Basic ferric sulfates formation depends on leaching conditions and it is favored by highly acidic environment. Higher temperature, on the other hand, favors the formation of hematite. These reactions regenerate the acid consumed by goethite dissolution, which is one of the important advantages of the HPAL process. Aluminum cations also hydrolyze, leading to the formation of solid products. Again, most of the acid consumed by boehmite dissolution is regenerated, as shown in equations (3.13) and (3.14) (GEORGIOU and PAPANGELAKIS, 1998). The autoclave product is a hot slurry, typically at 20-25% solids, containing typically 4-6g/L nickel, 30-60 g/L sulfuric acid, <8 g/L iron and <3g/L aluminum.

Autoclave discharge is fed to a solid/liquid separation step. Counter-current decantation is employed in most plants. It usually involves six to seven stages of thickeners. The majority of HPAL plants burn elemental sulfur to produce sulfuric acid and the by-product steam is utilized in various unit operations. (MOSKALYK and ALFANTAZI, 2002). Primary neutralization can be run before CCD (e.g. at Cawse and Vermelho project).

The slurry from the leaching stage or the CCD overflow, depending on the flowsheet adopted, feeds a downstream treatment. The two routes already
proven in commercial operations are the mixed hydroxide precipitation (MHP) and mixed sulfide precipitation (MSP), where the nickel is precipitated as hydroxyl sulfate or sulfide, respectively. Bulong, now closed, had adopted the direct solvent extraction flowsheet, with no nickel and cobalt precipitation step. MHP is produced commercially from Norlisk’s Cawse and BHP’s Ravensthorpe operations. MSP is an intermediate product at Minara’s Murrin Murrin plant and Sumitomo’s Coral Bay facility (LONGWORTH, 2007). Both processes are discussed in the following sections.

3.4.2.1 Mixed Hydroxide Precipitate (MHP) route (WILLIS, 2007)

MHP lacks selectivity for nickel and cobalt over iron, aluminum, chromium and manganese. Iron, aluminum and chromium can be removed prior to nickel and cobalt precipitation, although some collateral loss of nickel and cobalt with these elements, by co precipitation, is unavoidable. Copper and zinc will report to the MHP product; however these are usually very minor contaminants that can be tolerated. Manganese cannot be removed prior to nickel and cobalt precipitation, and some manganese contamination of the MHP product must be tolerated. As a consequence the MHP flowsheet is not suited to ores with high manganese content.

MHP flowsheet usually comprises of the following stages: recycle leach, iron removal stage 1, iron removal stage 2 (optional), mixed hydroxide precipitate stage 1, mixed hydroxide precipitate stage 2 and manganese removal (if required).

The autoclave slurry discharge feeds the recycle leach, where the residual acid from the leaching phase is used to recover, by re-dissolution, any nickel and cobalt co-precipitated as hydroxide in the second neutralization and second mixed hydroxide precipitate steps. This stage is operated usually at higher than 70°C, and at ambient pressure.
In the primary iron removal stage, the excess acid is neutralized using limestonne slurry. Iron and other impurities are removed from solution as a precipitate. Iron, as Fe$^{3+}$, can be removed from leach liquors to a residual concentration of < 200 mg/L at pH 2.5-3.0. Under these conditions, chromium is also removed to very low levels, however aluminum is only removed to 2 g/L. Figure 3.7 shows an illustrative profile for each of the major elements for a specific plant. This profile may vary under different conditions, but it gives a typical sequence of metal precipitation. It shows that, in the iron removal circuit, the first element to precipitate is Fe$^{3+}$, followed by Cu$^{2+}$. Operating temperatures of 70-90°C favor fast reaction kinetics and lead to a low slurry viscosity. Operation at ambient temperature is possible but the reaction rate is slow. Co-precipitation of nickel and cobalt in this pH range (e.g. 2.5-3.0) is minimal, typically <0.5%, hence the precipitates formed may be washed to recover soluble nickel and cobalt and then discarded. The residual aluminum, when reported to the MHP product, has deleterious effects in down-stream processing. Its presence at levels as low as 0.5% has been demonstrated to reduce nickel/cobalt dissolution in subsequent ammonia leaching by 10 to 30%. As a result it is desirable to precipitate aluminum prior to MHP. Removal of aluminum to <50 mg/L requires a pH of the order of 4.4 - 4.8. In this range co-precipitation of nickel and cobalt is significant, typically 4 - 10%. In Figure 3.7, it reaches up to 20%. If a pH of 4.4 - 4.8 is maintained in the secondary iron removal stage, the quantity of precipitate is relatively small (typically less than 10% of the solids precipitated in stage 1) and it can be contacted with acidic autoclave discharge slurry to re-dissolve the co-precipitated nickel and cobalt hydroxides. If the iron removal circuit is performed in a single stage the quantity of precipitates is such that there is no practical means to recover the co-precipitated nickel and cobalt, and these values are lost to the tailings.
The solution from the secondary iron removal circuit is introduced to the nickel/cobalt precipitation stage, where the pH is adjusted to precipitate nickel and cobalt. Typical feed solutions contain about 3.5 - 4.0 g/L nickel. Many nickel laterite projects adopt a flowsheet where the nickel and cobalt are precipitated by addition of magnesia for reasons discussed later in this section. Stoichiometric amounts of magnesia, based on nickel and cobalt in the feed, is added as dry powder. This flowsheet is used by plants/projects such as Cawse, Ravensthorpe and Níquel do Vermelho. The mixed hydroxide precipitate, typically contains an average of 38-40% nickel and Ni:Mn ratio of about 12:1. Typically 95% nickel and cobalt are recovered. Equation (3.15) is the simplified nickel precipitation reaction with MgO (JAYASEKERA, 2007).

$$
\text{MgO}_\text{s} + \text{NiSO}_4\text{(aq)} + \text{H}_2\text{O}_\text{l} = \text{Ni(OH)}_2\text{(s)} + \text{MgSO}_4\text{(aq)}
$$

(3.15)

Magnesia may be replaced by caustic soda (sodium hydroxide), hydrated lime (calcium hydroxide) or soda ash (sodium carbonate). Caustic soda eliminates
the magnesium content of the MHP product that is associated with unreacted magnesia and offers slightly improved selectivity for nickel over manganese, through faster mixing and more precise pH control, yielding a higher quality product. However it is generally very expensive and produces slurry difficult to settle and filter. Lime introduces calcium, which results in gypsum precipitation and therefore contamination of the MHP product. The use of soda ash results in the precipitation of basic carbonates along with hydroxides, however the solubility of manganese carbonate is lower than that of nickel carbonate, thus the selectivity for nickel over manganese is further reduced.

The MHP 1 product slurry is sent to a solid/liquid separation that comprises of a thickener and a filter. The overflow is sent to the MHP stage 2, where residual nickel and cobalt are recovered by adjusting the pH to 7.5-8.0 with lime slurry. Some manganese is co-precipitated. The precipitate typically contains 10 to 20% manganese. Solids product from stage 2 can be re-dissolved in the releach circuit. A low-grade mixed hydroxide product can be produced by employing a single stage precipitation at pH 7.5-8.0. This approach will lower the project capital expenditure however the product may contain >10% manganese, limiting marketing options.

The MHP 1 product is washed and sold as an intermediate product or leached with ammonia and further processed on site to produce nickel metal. The solution from MHP 2 circuit is treated with lime slurry to remove manganese and recovered water is recycled upstream. The requirement for a manganese removal circuit is project specific. Generally more than 50% of the manganese remains in solution following two stages of nickel-cobalt precipitation. When it is necessary for any of the barren solution to be discarded, environmental regulations will probably stipulate removal of aqueous manganese, which then requires inclusion of a manganese removal stage. When any of the barren solution is recycled within the process, manganese will accumulate in the circuit until the losses in spent ore moisture and iron removal filter cake balance the
inputs. If the equilibrium concentration of manganese in the circuit does not adversely affect product quality, then manganese removal can be excluded.

3.4.2.2 Mixed Sulfide Precipitate (MSP) route (WILLIS, 2007)

Precipitation of nickel and cobalt as a mixed sulfide is highly selective for nickel and cobalt over iron, aluminum, chromium and manganese. This flowsheet has been used at Moa Bay, Murrin Murrin and Coral Bay operations. The reaction of Fe$^{3+}$ with hydrogen sulfide produces undesirable elemental sulfur, and it must therefore be removed or reduced prior to sulfide precipitation. Because of the excellent selectivity, the removal of aluminum, chromium and manganese prior to nickel and cobalt precipitation is unnecessary.

MSP flowsheet usually comprises of the following stages: pre-reduction (optional), partial neutralization, nickel and cobalt precipitation, iron removal or final neutralization (if required). Pre-reduction may be employed if exclusion of copper from the mixed sulfide product is desirable. Selective precipitation of copper as copper sulfide can be achieved by contacting the PLS with hydrogen sulfide gas under carefully controlled redox conditions (Eh about 380 mV). Figure 3.8 shows the CuS formation condition, for a system at 25°C. Simultaneously, approximately 75% of the aqueous Fe$^{3+}$ present is reduced to Fe$^{2+}$. Figure 3.9 shows that within the stability region of CuS, Fe$^{2+}$ may be formed. Any Cr$^{6+}$ present is quantitatively reduced to Cr$^{3+}$.

The function of the partial neutralization is to adjust the feed liquor pH ahead of sulfide precipitation to establish the optimum conditions for nickel and cobalt precipitation. Figure 3.10 shows the Eh-pH diagram for the Ni-S-H$_2$O system. Ni$^{2+}$ region is larger than those of Fe$^{2+}$ and CuS. In the pre-reduction and pre-neutralization conditions nickel remains in solution as Ni$^{2+}$. Figure 3.11 shows the residual nickel concentration as a function of MSP feed pH. The reduction of Fe$^{3+}$ to Fe$^{2+}$ during sulfide precipitation could potentially consume significantly more hydrogen sulfide than the actual nickel and cobalt precipitation reactions.
adding to the capital and operating costs for hydrogen sulfide production. The recommended approach is therefore to neutralize the PLS to pH 2.5-3.0, using limestone in order to precipitate most of the iron present prior to nickel and cobalt precipitation.

Figure 3.8: Eh-pH diagram, for Cu-S-H2O system, at 25°C, 1 molal system (HARRIS and KRAUSE, 1994).

Nickel can be precipitated from feed liquor with pH over 2.5 to a residual concentration in barren solution of over 30 mg/L by reaction with hydrogen sulfide gas. Hydrogen sulfide injection is controlled to maintain a predetermined hydrogen sulfide overpressure in the reactor vessel. Under these conditions, nickel and cobalt precipitate as sulfides, along with any copper and zinc present. A small amount of Fe\(^{2+}\) also precipitates as sulfides. Fe\(^{3+}\), aluminum, chromium, manganese and magnesium are not precipitated. Alternative precipitants are sodium sulfide (Na\(_2\)S) or sodium hydrosulfide (NaHS). The product generally contains over 55% nickel (dry basis) and less than 15% moisture. Precipitation takes place predominantly on the surface of
existing particles. Provision of seed surface area in the form of fine mixed sulfide particles is therefore essential.

![Eh-pH diagram](image)

Figure 3.9: Eh-pH diagram, for Fe-S-H2O system, at 25°C, 1 molal system (HARRIS and KRAUSE, 1994).

Final neutralization is necessary if the barren solution to be recycled contains a significant amount of Fe\(^{2+}\). If it is to be discarded, environmental regulations will probably stipulate removal of aqueous manganese. Air/oxidant injection helps to oxidize the Fe\(^{2+}\). The ferric ion (Fe\(^{3+}\)) can be removed to residual concentrations of less than 500 mg/L at pH 2.0-2.5. If manganese removal is necessary, a residual manganese concentration below 100 mg/L can be achieved by adjusting the solution pH to 8.5-9.0 thus precipitating manganese hydroxide. If a lower residual concentration is required, oxidation by air injection to form manganese dioxide may be included.
Figure 3.10: Eh-pH diagram, for Ni-S-H2O system, at 25°C, 1 molal system (HARRIS and KRAUSE, 1994).

Figure 3.11: Residual nickel concentration as a function of pH (WILLIS, 2007)
3.4.3 Caron Process

The Caron process can be used to treat limonite ores or a mixture of limonite and saprolite, with high iron and high magnesium content. It is a combined pyrometallurgical and hydrometallurgical process, and some examples of its applications are at Niquel Tocantins, Nicaro, Punta-Gorda and Yabulu. The standard reductive roast/ammoniacal leach process, which has proven satisfactory in treating most oxide ores, was developed by M. H. Caron during the period of 1920’s to the 1950’s. The Caron process is summarized by RICHARDSON et al. (1981).

A typical composition of an oxide ore to be treated by Caron process is presented in table III.4. The ore is initially crushed to increase the specific surface area to successfully respond to roast and leach. Usually after mining, the wet ore contains 20 to 50% moisture. It is crushed to less than 1 inch in size and then dried to less than 1 to 3% moisture.

Table III.4: Typical composition of an oxide ore (RICHARDSON et al., 1981)

<table>
<thead>
<tr>
<th>Ni (%)</th>
<th>Co (%)</th>
<th>Fe (%)</th>
<th>Mg (%)</th>
<th>SiO₂ (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.26</td>
<td>0.1</td>
<td>46.5</td>
<td>0.1</td>
<td>1.9</td>
</tr>
</tbody>
</table>

A typical block diagram of Caron process is presented in Figure 3.13. The dried ore is fed to reduction roasting stage, where it is gradually heated to a specified temperature under a reducing atmosphere. The aim is to reduce the nickel oxide to metal, making the nickel amenable to ammoniacal leaching. The reduction of nickel and cobalt is a critical step in the Caron process since the co-reduction of iron has to be avoided. A critical evaluation of the efficiency of the process with respect to metal reduction is not straightforward. Most of the iron in the limonitic fraction is converted to magnetite, Fe₃O₄, whilst the iron in the serpentine fraction is mostly unaffected. The dryer fuel is oil or any other available fuel (JOSEPH and BOLDT, 1967). The primary reactions taking place during the roasting are those affecting nickel and iron (equations (3.16) to
(3.24)). Similar reactions occur for cobalt and copper. Reduction of metal oxides mainly takes place at temperatures in the range of 700-900°C. Temperature has an effect on reaction kinetics, with higher temperatures giving faster reactions and therefore a practical lower limit of about 540°C is used in most operations. An upper limit of about 900°C is observed, primarily to avoid the sintering that occurs at high temperatures.

\[
\begin{align*}
NiO(s) + H_2(g) &= Ni(s) + H_2O(g) & (3.16) \\
3 Fe_2O_3(s) + H_2(g) &= 2 Fe_3O_4(s) + H_2O(g) & (3.17) \\
Fe_3O_4(s) + H_2(g) &= 3 FeO(s) + H_2O(g) & (3.18) \\
FeO(s) + H_2(g) &= Fe(s) + H_2O(g) & (3.19) \\
NiO(s) + CO(g) &= Ni(s) + CO_2(g) & (3.20) \\
3 Fe_2O_3(s) + CO(g) &= 2 Fe_3O_4(s) + CO_2(g) & (3.21) \\
Fe_3O_4(s) + CO(g) &= 3 FeO(s) + CO_2(g) & (3.22) \\
FeO(s) + CO(g) &= Fe(s) + CO_2(g) & (3.23) \\
Ni(s) + Fe(s) &= FeNi(s) & (3.24)
\end{align*}
\]

In order to obtain a maximum reduction of nickel oxide to metallic nickel, most of the iron must also be reduced, once a substantially large fraction of the nickel in laterite ores is tied up with iron oxides. However, the complete reduction of iron to metallic form is unnecessary and undesirable. After reduction, the resulting material is cooled at 120°C -150°C under inert atmosphere to prevent re-oxidation of the nickel and other desired metals. A typical composition of a reduced oxide ore after losing 20% of mass is presented in table III.5.
After cooling, the roasted ore is sent to the ammoniacal leaching, where metallic nickel is oxidized and dissolved. The amount of metal solubilized is greatly dependent on the reduction roast conditions, which in turn depends on the type of ore. Equations (3.25) to (3.27) show the typical overall leach reactions. Iron-ammamine complex is stable under very limited conditions. In oxidizing conditions, this complex is decomposed to form the insoluble hydroxide, as shown in Equation (3.28). Air is sparged into the leach tanks to provide the required oxygen. Figures 3.13 and 3.14 show that at pH between 9 and 10, and under oxidative environment, Ni(NH$_3$)$_6^{++}$ is the thermodynamically predominant nickel species whilst that for iron is Fe(OH)$_3$(s).

Figure 3.13: Eh-pH diagram for Ni-NH$_3$-H$_2$O system at 298 K and 1 atm total pressure. The activities of NH$_3$ above pH = 9.25 and NH$_4^+$ below 9.25 are unity. The ionic nickel species is 10$^{-2}$ mol/L (RICHARDSON et al., 1981)
Once leached, the dissolved metal values are separated and recovered from the pulp in a solid/liquid separation. The final washed slurry is then introduced to a steam-stripping tower to remove the ammonia and carbon dioxide. The off-gasses are vented to a reagent recovery system, whilst the tailings slurry is pumped to a disposal pond. On the other hand, the solution is treated to produce a nickel-rich product. Several options are available for final product recovery from pregnant liquor, such as steam stripping, sulfide precipitation, solvent extraction and electrowinning. The selection of a given option will depend on economic considerations such as price and availability of raw materials and marketability of the final product.

\[
\begin{align*}
Ni(s) + 8NH_3(g) + 1/2O_2(g) + 2CO_2(g) + H_2O(l) & = Ni(NH_3)_6^{++}(aq) + 2NH_4^+(aq) + 2CO_3^{--}(aq) & (3.25) \\
FeNi(s) + 12NH_3(g) + O_2(g) + 3CO_2(g) + H_2O(l) & = Ni(NH_3)_6^{++}(aq) + Fe(NH_3)_4^{++}(aq) + 2NH_4^+(aq) + 3CO_3^{--}(aq) & (3.26) \\
Fe(s) + 6NH_3(g) + 1/2O_2(g) + 2CO_2(g) + H_2O(l) & = Fe(NH_3)_4^{++}(aq) + 2NH_4^+(aq) + 2CO_3^{--}(aq) & (3.27) \\
4Fe^{++}(aq) + O_2(g) + 2H_2O(l) + 8OH^-(aq) & = 4Fe(OH)_3(s) & (3.28)
\end{align*}
\]

Figure 3.14: Eh-pH diagram for Fe-NH₃-H₂O system at 298 K and 1 atm total pressure. The activities of NH₃ above pH = 9.25 and NH₄⁺ below 9.25 are unity. The ionic nickel concentration is 10⁻²M (RICHARDSON et al., 1981)
3.4.4 Heap Leaching process

Nickel laterite heap leaching is an innovative technology that is now being tested by several companies in many parts of the world. Some examples of pilot plant testing of this technology include European Nickel PLC and BHP Billiton’s, joint technology development, in Çaldağ – Turkey; Vale Inco’s, Niquel do Piauí – Brazil and Minara Resources’ operation at Murrin Murrin site – Australia. A block diagram is presented in Figure 3.15 to illustrate a conventional heap leaching circuit.

Several patents related to heap leaching using sulfuric acid are available and summarized in Appendix A. Table III.6 show a consolidation of the main features of each patent.

Following mining, ore is crushed to the desired particle size. Usually the crushed ore, prior to stacking in the heap, is homogeneously wetted, in an agglomeration drum, to raise the total moisture. This treatment was found to significantly improve the permeability of the heap (AGATZINI-LEONARDOU et al., 1997).

![Block diagram for nickel heap leaching](image)

In the heap, acid contained solution is sprinkled on top. The nickeliferous serpentine is dissolved in preference to the hematite and goethite minerals under the mild conditions prevailing during heap leaching. Calcite contained in
the ore adversely affects the permeability of the heap and the sulfuric acid consumption (AGATZINI-LEONARDOU and ZAFIRATOS, 2004). A high acid concentration in the heap feed solution inevitably leads to faster extraction of iron and others metals, which must be removed from solution in the iron precipitation step prior to production of nickel precipitates for sale (LONGWORTH, 2007).

Heap leaching produces an essentially solid-free solution, at ambient temperature, containing typically 2-4g/L nickel, 15-30g/L iron, 2.5g/L aluminum and 30-50g/L sulfuric acid. The down-stream processes to treat this solution are similar to the ones used in the HPAL process, and are explained in section 3.4.2. Some key differences that have to be considered when treating a heap leach PLS, instead of a HPAL PLS, are: lower temperature, lower nickel and higher iron concentrations (WILLIS, 2007).

The addition of a heap leach to existing HPAL plant could imply in a much lower capital intensity given existing acid plant facilities. If heap leaching is used to supplement HPAL, limonite is usually processed in HPAL and saprolite in heap leaching. Separation of ore types and their processing can improve overall recoveries and reduce iron contamination. Separation of ore particle size can be also undertaken, with the coarse material being treated with HL and the fines with HPAL or atmospheric leaching (HEAP AND TONKS, 2007). Ravensthorpe is an example of hybrid atmospheric leaching (not heap leaching) and HPAL.
<table>
<thead>
<tr>
<th>Patent n°</th>
<th>Main characteristics</th>
</tr>
</thead>
</table>
| GR 1001555   | i) First to describe the heap leaching process for Ni laterite  
               ii) Describes two different techniques for feeding solution to the heap:  
                   - leach solution recycles inside the heap/column until no nickel leaching is noticed  
                   - solution is recycled after pH adjustment |
| GR1003569    | This is an addition to GR 1001555, which considers the use of seawater or water of various chemical composition or municipal effluent streams for the production of a dilute sulfuric acid solution for heap leaching. |
| US 63125000  | Describes the method of heap leaching nickel containing laterite ore that have tangible clay content. Considers forming at least one heap from a mixture of the ore and concentrated sulfuric acid, at least one heap of saprolite ore to be used as an iron precipitation reactor and/or at least one heap includes bedrock material for final pH control and iron removal. |
| AU2003/9003632 | Considers beneficiation and separation of the coarse (silicious low grade) and the fine (upgrade ore) fractions. High pressure acid leaching or atmospheric leaching can treat the upgraded ore, while the low-grade material is treated by heap leaching. |
| AU2005/902462 | Leach liquor is prepared from saline sea water or hypersaline underground water. |
| WO2004/031422 | Irrigation of the heap with a leach solution containing both sulfuric acid and dissolved sulfur dioxide. The usage of the latter in addition to sulfuric acid leads in increased recoveries of both nickel and cobalt, with the most marked improvements being found with respect to cobalt recovery. |
| WO2006/000020 | Separation of fine and coarse fractions. The coarse material is leached while the fine ore fraction or part of it is used to neutralize or partially neutralize the PLS produced in the coarse material heap leaching. |
| EP 1 790 739 A1 | The heap leaching step comprises of dynamic heaps, counter-current process, with at least 2 stages, where a heap section or a heap represents each stage. The final solution, from the first heap, the PLS (Pregnant leaching solution) has acidity of 10-30 g/L. This results in a reduced acid consumption, reaching 350 kg/t of ore depending on mineralogy, and in a reduced leaching cycle. |
| WO2007/016737 | Separate fine and coarse material. The fines feed a HPAL process. The coarse material feeds a heap leaching. Portion of leach solution from HPAL is passed directly to the heap leaching. The rest is further processed. |
3.4.4.1 Çaldag heap leaching pilot plant

All the process information described below is provided by PURKISS (2006). European Nickel has demonstrated a nickel laterite heap leach process at its Çaldag nickel deposit in Western Turkey. A large scale demonstration plant has been constructed at Çaldag and three heaps are operational. A small precipitation plant has been producing a nickel cobalt hydroxide. Nickel recovery in the first heap has exceeded 75%.

All material is screened and crushed to minus 30 mm and used in the first and second heaps after agglomeration. The coarse material, over 30 mm, is discarded. The water is added through spray nozzles in the agglomeration drum and all agglomerated ore is stacked after 24 hours air curing.

In Çaldag nickel is associated with the ferruginous phase, which explains the longer leach cycles required to extract the metal compared to other laterites that have been tested. The pregnant solution from the leach is collected in ponds and recirculated through the heaps to maximize the metal content before being pumped to the precipitation plant. The acid strength in the ponds during leach operation has been adjusted to about 75 g/L. Demonstration of the atmospheric heap leach and leach kinetics of various types of ores were performed by Heap 1 and Heap 3. Heap 2 was used to demonstrate the neutralization step to reduce acid content of the leach solution to be fed to the precipitation plant. Figure 3.16 shows a schematic drawing of the leach circuit.

![Heap leaching circuit at Çaldag.](image)

Figure 3.16: Heap leaching circuit at Çaldag.
In the precipitation plant, iron is precipitated by raising the pH, and the solids thickened, filtered and disposed of as the waste product from the process. The liquor from the iron thickener is then treated by raising the pH further with soda ash to produce a nickel-cobalt carbonate with a nickel content of above 30% that is filtered and packaged for shipment to refineries in Europe, China and Australia.

3.4.4.2 *Niquel do Piauí* heap leaching pilot plant

Vale has built a pilot plant to evaluate a deposit in Piauí state, Brazil. The ore was crushed and screened and the minus 2 inch material was stockpiled. In the agglomeration process, sulfuric acid and water were added and the material was cured for about 24 hours.

The counter-current heap leaching process is being tested in leach columns, where the leach solution is fed in the top of the column. A schematic drawing of the counter current process is shown in Figure 3.17. The PLS is being stored and a downstream circuit is being constructed.

![Figure 3.17: Heap leaching circuit at Piauí.](image)
4 Modeling

This work began with a detailed investigation of the operational conditions of the following process routes: i) FeNi, ii) HPAL and iii) HL

In the first option, process evaluation and economic analysis considered the FeNi alloy as the final product. A variation of this process involves sulfur addition into the kiln to produce matte. As the production of matte or FeNi alloy does not affect the process selection, only the FeNi product was considered in the present study. The HPAL and HL processes may involve, as intermediate products, nickel/cobalt hydroxides or nickel/cobalt sulfides. Again, this decision does not affect the process selection and therefore only the production of a mixed hydroxide precipitate as intermediate product was considered. The process routes were considered independently, that is, no combined process routes were evaluated in the present work.

The CRU Report - Nickel Quarterly January 2007 provides a table showing the firm, probable and possible nickel laterite projects up to 2015 (Appendix B). Based on this table, from over one hundred (100) nickel projects; fifty four (54) will be applied to laterite ores. From these fifty four projects, one (1) opted for acid chloride leaching process (ACLP), one (1) for Caron process, six (11%) for heap leaching process, twenty five (46%) for pressure acid leaching (PAL) and seventeen (31%) for smelting. Two projects restricted to mining operations and two were not declared. Figure 4.1 shows the contribution of each process technology for the total of fifty-four projects. Acid chloride leaching (ACLP) is not a consolidated technology and has been selected for just one project. Caron process is established technology, but new applications are hindered by the reported low nickel and cobalt recoveries. Northern Province is the only new project opting for this technology, as showed in Appendix B. Therefore the trend for new developments is clearly towards PAL, smelting and heap leaching technologies.
As discussed earlier in Section 3.4, different geological lithotypes determine different process routes. Based on this, a process model, using the computer software MetSim, was built for each route. The choice of which process to employ depends on several factors, including physical and chemical nature of the ore body, weather conditions and geographical location, which is related to supply and cost of required reagents and energy. Different processes also result in different end products and the marketability of such products may depend on geographical location. For the present work, only one product type for each route was considered, that is, mixed hydroxide precipitate (MHP) for heap leaching and HPAL and FeNi for the smelter. The choice of MHP was due to available information to the process and economical models. The statement of MHP as a product is very dependent on its marketability at the time the project is being developed.

The methodology adopted to develop this study for each selected process involved: i) development of the process model, ii) estimation of the capital cost, iii) assessment of operational costs and iv) economic evaluation. With a process model linked with the economic model, some case studies were evaluated.
4.1 Process model

MetSim is a computer program developed for chemical, metallurgical and environmental applications. It is supplied by PROWARE – Tucson, Arizona, USA. Using physical-chemical concepts, scientific and industrial database, the software program’s main applications are to evaluate test data, calculate plant material balances, develop flowsheets, assess process feasibilities, detail process designs, analyse control systems, train operators and optimize processes.

Many organizations, throughout the world, such as Anaconda, Bateman, BHP Billiton, Codelco, Cominco Engineering Services, Vale, Hazen Research, Rio Tinto, University of Minnesota, report the utilization of MetSim software. Being already tested and well accepted in the mining industry, this software was chosen to develop the process model in the present study.

The software calculates the mass and heat balances in the sequence of the unit operations; that is, from the first unit operation to the last one, in the tolerance range chosen by the client. The software also performs the chemistry of the process in the order that the reactions are listed in each unit operation. This means, reagents from some reactions may be products of the previous ones. Each reaction has an extent that indicates the percentage by mass of the first reactant in the chemical equation that will be converted to products, provided there are sufficient reagents. The heat balance calculated across each unit operation considers an internal thermodynamic database. Heat of reactions and heat losses are calculated and used to compute temperatures of discharge streams.

One MetSim process model was built for each process under evaluation, considering literature information typical to the industry and Vale Inco projects information. The details of each model will be presented in Section 5 and in
Appendix C to E. An output standard spreadsheet was created to link the MetSim model to the economic evaluation model.

4.2 Economic Evaluation

An economic model was built in order to compare the process routes. An excel file was created, linking selected MetSim model outputs to the economic model. The economic model consists of a project capital cost evaluation, an operation cost evaluation and the calculation of the final viability (a very preliminary Net Present Value – NPV) of the project.

4.2.1 Capital Cost

A list of the equipment database is comprised of the major equipment costs for each process route. These prices were based on similar Vale Inco existing projects and they were scaled up or down for the evaluated project using the traditional equation (4.1):

$$ A = B \times \left( \frac{C}{D} \right)^{6/10} $$

where $A$ is the equipment cost of the project under evaluation, $B$ is the known equipment cost, $C$ is the size of the project under development and $D$ is the size of the known project. In order to have a more precise scaling, the ratio in the parentheses was calculated using the feed (mass flowrate) for each equipment or area. Equation (4.1) was applied for all the major equipment in order to generate a list of equipments with their respective costs.

In order to calculate the direct cost, indexes based on the equipment list costs were used, as listed in Table IV.1 to IV.3, each table showing the indexes applied to each process route under evaluation. These factors were obtained from similar Vale Inco projects and may vary from one country to the other. As
an input number, they can be changed if necessary. For the present work, they were considered as fixed numbers. The final direct cost considered for each project included the overall equipment cost together with the costs of all the items shown in Table IV.1 to IV.3.

Table IV.1: Indexes applied to estimate the direct cost for FeNi route

<table>
<thead>
<tr>
<th>Item</th>
<th>Factor based on equipment cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Civil works/ Earth works/ Drainage/ Paving/Concrete</td>
<td>53%</td>
</tr>
<tr>
<td>Piping</td>
<td>11%</td>
</tr>
<tr>
<td>Installation</td>
<td>14%</td>
</tr>
<tr>
<td>Steelwork</td>
<td>33%</td>
</tr>
<tr>
<td>Platework</td>
<td>16%</td>
</tr>
<tr>
<td>Substation/ electrical/Instrumentation</td>
<td>29%</td>
</tr>
<tr>
<td>Control Systems</td>
<td>11%</td>
</tr>
<tr>
<td>Freight</td>
<td>7%</td>
</tr>
<tr>
<td>Buildings</td>
<td>13%</td>
</tr>
<tr>
<td>Capital Spares and First Fill</td>
<td>5%</td>
</tr>
</tbody>
</table>

Table IV.2: Indexes applied to estimate the direct cost for HPAL route

<table>
<thead>
<tr>
<th>Item</th>
<th>Factor based on equipment cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Civil works/ Earth works/ Drainage/ Paving/Concrete</td>
<td>20%</td>
</tr>
<tr>
<td>Piping</td>
<td>35%</td>
</tr>
<tr>
<td>Installation</td>
<td>20%</td>
</tr>
<tr>
<td>Steelwork</td>
<td>10%</td>
</tr>
<tr>
<td>Platework</td>
<td>20%</td>
</tr>
<tr>
<td>Substation/ electrical/Instrumentation</td>
<td>30%</td>
</tr>
<tr>
<td>Control Systems</td>
<td>5%</td>
</tr>
<tr>
<td>Freight</td>
<td>6%</td>
</tr>
<tr>
<td>Buildings</td>
<td>15%</td>
</tr>
<tr>
<td>Capital Spares and First Fill</td>
<td>5%</td>
</tr>
</tbody>
</table>

In order to calculate the total project capital cost, an index based on the direct cost was used to calculate contingency (a budget for an event that might occur in the future, usually causing problems and requiring alterations to original design), as listed in Table IV.4. Costs such as environmental licensing, temporary facilities, energy transmission line, mine development, owner’s cost and EPCM (engineering, procurement, construction management) were not
considered in the evaluation since these costs are not directly related with the selection of the process route.

Table IV.3: Indexes applied to estimate the direct cost for heap leaching route

<table>
<thead>
<tr>
<th>Item</th>
<th>Factor based on equipment cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Civil works/ Earth works/ Drainage/ Paving/Concrete</td>
<td>13%</td>
</tr>
<tr>
<td>Piping</td>
<td>30%</td>
</tr>
<tr>
<td>Installation</td>
<td>10%</td>
</tr>
<tr>
<td>Steelwork</td>
<td>8%</td>
</tr>
<tr>
<td>Platework</td>
<td>15%</td>
</tr>
<tr>
<td>Substation/ electrical/Instrumentation</td>
<td>25%</td>
</tr>
<tr>
<td>Control Systems</td>
<td>5%</td>
</tr>
<tr>
<td>Freight</td>
<td>5%</td>
</tr>
<tr>
<td>Buildings</td>
<td>8%</td>
</tr>
<tr>
<td>Capital Spares and First Fill</td>
<td>5%</td>
</tr>
</tbody>
</table>

Table IV.4: Contingency index applied to estimate the total capital cost

<table>
<thead>
<tr>
<th>Item</th>
<th>Factor based on direct cost</th>
</tr>
</thead>
<tbody>
<tr>
<td>Contingency</td>
<td>30%</td>
</tr>
</tbody>
</table>

4.2.2 Operational cost (OPEX)

The operational cost considered is comprised of: mining, administrative, labor, maintenance, reagents, power, consumables, product contract transportation and contingency. The same reagents prices and power prices were adopted for all process routes. Mining and product transportation costs were calculated based on a fixed price per tonnes of ore and ton of product, respectively. Costs with administrative, labor, maintenance, consumables and contingency were calculated based on the indexes shown in Table IV.5. Preliminary OPEX is the operational cost calculated without the variables: administration, labor and contingency. The final one includes these three variables.
Table IV.5: Indexes for operational cost

<table>
<thead>
<tr>
<th>Item</th>
<th>%</th>
<th>Base</th>
</tr>
</thead>
<tbody>
<tr>
<td>Maintenance</td>
<td>2</td>
<td>Equipment cost</td>
</tr>
<tr>
<td>Consumables</td>
<td>11</td>
<td>Reagent cost</td>
</tr>
<tr>
<td>Administrative</td>
<td>3</td>
<td>Preliminary opex</td>
</tr>
<tr>
<td>Labor</td>
<td>5</td>
<td>Preliminary opex</td>
</tr>
<tr>
<td>Contingency</td>
<td>3</td>
<td>Preliminary opex</td>
</tr>
</tbody>
</table>

4.2.3 Economic Analysis

In order to compare the process routes, an economic analysis was performed. The costs related in the item 4.2.1 were considered to be spent 50% in the Year 1 and 50% in the Year 2 of the project. Years 1 and 2 were considered as construction years. The operation cost was spent starting on Year 3. All the projects were considered to last 20 years. Some input parameters considered in the economic analysis are shown in Tables IV.6 and IV.7.

A present value, considering nickel and cobalt production, capital cost, operational cost and the discount rate was calculated for each simulation. To compare two different processes for a given project, an adimensional number was created as a ratio of the present values of the two evaluated processes. When all these steps had been taken into account, it was possible to evaluate technical and economic parameters that influence the process selection and to quantify their impact on the economic feasibility of the project.

Table IV.6: Parameters considered in the economic evaluation

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
<th>Unit</th>
</tr>
</thead>
<tbody>
<tr>
<td>Nickel price</td>
<td>$12000</td>
<td>US$/ton</td>
</tr>
<tr>
<td>Cobalt price</td>
<td>$30000</td>
<td>US$/ton</td>
</tr>
<tr>
<td>MHP terms of sale</td>
<td>80%</td>
<td>LME°</td>
</tr>
<tr>
<td>FeNi terms of sale</td>
<td>95%</td>
<td>LME°</td>
</tr>
<tr>
<td>Availability</td>
<td>85%</td>
<td>%</td>
</tr>
<tr>
<td>Discount Rate</td>
<td>12%</td>
<td>%/Year</td>
</tr>
</tbody>
</table>

LME° - London Metal Exchange

° same percentage for all processes
Table IV.7: Start up curve for each route

<table>
<thead>
<tr>
<th></th>
<th>FeNi</th>
<th>Heap Leaching</th>
<th>HPAL</th>
</tr>
</thead>
<tbody>
<tr>
<td>Year 1</td>
<td>50%</td>
<td>25%</td>
<td>50%</td>
</tr>
<tr>
<td>Year 2</td>
<td>75%</td>
<td>50%</td>
<td>75%</td>
</tr>
<tr>
<td>Year 3</td>
<td>100%</td>
<td>100%</td>
<td>100%</td>
</tr>
</tbody>
</table>
5 Results

The resulting process model for each route is presented in the sections below. Before running each model, there are several parameters that need to be known. They do not depend on the process route, hence, they were regarded as project inputs. They will vary from project to project and as any variable in the model, they can be changed to evaluate their impact on the process route evaluation. Figure 5.1 presents the project inputs for all the models.

![Figure 5.1: Project inputs.](image)

5.1 FeNi model

The FeNi pyrometallurgy process model was built considering seven main areas. These were: i) Coal Milling, ii) Ore drying, iii) Kiln, iv) Dust Conveying; v) Electric Furnace, vi) Off-gas handling, vii) Reagents and viii) Final discharge. Figure 5.2 shows an illustrative flowsheet of FeNi process. Appendix C presents the MetSim model flowsheets and all the chemical reactions and controllers considered to build the model.

The first step in building the model is to evaluate all the necessary chemical components. For the FeNi model, the chemical/mineralogical composition considered in the feed is comprised of: NiO, FeO, Fe₃O₄, Fe₂O₃, Fe₂O₃·H₂O, Al₂O₃, CaO, Co₃O₄, Cr₂O₃, MnO, MgO and Mg(OH)₂. If the desired deposit does not present any of these components, it will be considered as zero in the model.
After calculating the feed (oxide basis) considering the elemental composition, the difference will be reported as SiO$_2$.

Figure 5.2: FeNi process illustrative flowsheet

5.1.1 Coal Milling

Coal is a reagent used as reductant and as fuel in the dryer and kiln area. Coal for the dryer is sent to the roller mill and then to the combustion chamber. Coal for the kiln is split in two streams. One stream is sent directly to the kiln as reductant. The coal used as fuel is also sent the roller mill to be grinded and heated. Diesel is used as fuel to the kiln, and is added based on the amount of coal that feeds the mill. The air to the kiln air heater is added based on the amount of diesel that feeds the mill; the dilution air is added in order to achieve 170°C in the mill air feed.
5.1.2 Ore drying

The crushed ore feeds the dryer, together with heated air. The aim of this step is to reduce the moisture in the ore, using coal as fuel. The ore fed to the plant is back calculated based on the reserve, mine lifetime and moisture content in the ore. As explained earlier in Section 5, three important inputs for all the models are: i) reserve (dry basis), ii) mine lifetime and iii) moisture of the deposit. They vary from project to project. Considering these variables, the feed flowrate to the plant, in tonnes/h, is calculated, considering the equation (5.1). The decision of considering the reserve as an input, instead of the annual nickel production, was due to the fact that each process has a different nickel recovery and this would incorporate another variable to be analysed. In this way, nickel production is a process model output.

\[
\text{Feed (tonnes/h)} = \frac{(\text{Reserve (tonnes)} / \text{mine lifetime (year)})/8760 \text{ (h/year)})}{1 - \text{moisture}}
\]  

(5.1)

The dryer discharge is the dried solid material to feed the kiln and gases that are sent to a gas handling system, in order to be cleaned. The primary combustion air added to the dryer is based on the amount of coal that feeds the dryer. The gas added to the off gas system is proportional to the amount of gas from the dryer.

Some process parameters that are necessary at this stage are shown on Table V.1. These values are typical numbers or numbers from Vale Inco projects, and can be changed if required.

5.1.3 Ore reduction

Dried material feeds the kiln, where the free and crystalline moisture is removed. The trivalent iron oxide is reduced to divalent state. There is also
some reduction of both iron and nickel to the metallic state. Equations (5.2) to (5.10) are the main reactions taking place during ore reduction:

Table V.1: Ore drying process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Product and dust moisture content</td>
<td>10%</td>
</tr>
<tr>
<td>Dust loss (fraction of solids reporting to the off gas)</td>
<td>8%</td>
</tr>
<tr>
<td>Dryer air feed temperature</td>
<td>1100°C</td>
</tr>
<tr>
<td>Dryer temperature</td>
<td>160°C</td>
</tr>
<tr>
<td>Heat loss in the dryer</td>
<td>5%</td>
</tr>
<tr>
<td>Heat loss in the dryer combustion chamber</td>
<td>6%</td>
</tr>
</tbody>
</table>

\[
2 \text{C}_\text{(s)} + \text{O}_\text{2(g)} = 2 \text{CO}_\text{(g)} \quad \text{(5.2)}
\]

\[
\text{Mg(OH)}_\text{2(s)} = \text{MgO}_\text{(s)} + \text{H}_\text{2O}_\text{(g)} \quad \text{(5.3)}
\]

\[
\text{NiO}_\text{(s)} + \text{CO}_\text{(g)} = \text{Ni}_\text{(s)} + \text{CO}_\text{2(g)} \quad \text{(5.4)}
\]

\[
\text{Co}_\text{3O}_\text{4(s)} + 4 \text{CO}_\text{(g)} = \text{Co}_\text{(s)} + 4 \text{CO}_\text{2(g)} \quad \text{(5.5)}
\]

\[
\text{Fe}_\text{2O}_\text{3}\cdot\text{H}_\text{2O}_\text{(s)} = \text{Fe}_\text{2O}_\text{3(s)} + \text{H}_\text{2O}_\text{(g)} \quad \text{(5.6)}
\]

\[
3 \text{Fe}_\text{2O}_\text{3(s)} + \text{CO}_\text{(g)} = 2 \text{Fe}_\text{3O}_\text{4(s)} + \text{CO}_\text{2(g)} \quad \text{(5.7)}
\]

\[
\text{Fe}_\text{3O}_\text{4(s)} + \text{CO}_\text{(g)} = 3 \text{FeO}_\text{(s)} + \text{CO}_\text{2(g)} \quad \text{(5.8)}
\]

\[
\text{FeO}_\text{(s)} + \text{CO}_\text{(g)} = \text{Fe}_\text{(s)} + \text{CO}_\text{2(g)} \quad \text{(5.9)}
\]

\[
2 \text{CO}_\text{(g)} + \text{O}_\text{2(g)} = 2 \text{CO}_\text{2(g)} \quad \text{(5.10)}
\]

Primary and secondary combustion air is added to the furnace based on the amount of fuel that feeds the kiln, while the tertiary combustion air is added to satisfy reduction conditions and it is controlled by the amount of O₂ in the kiln gas discharge. The reductant coal is added in order to satisfy reduction requirements, and it is controlled by the amount of carbon in the kiln discharge. The gas added to the off gas system is proportional to the amount of gas feeding the system and the air added to the dust conveying system is proportional to the amount of dust. Table V.2 shows the ore reduction parameters and the values considered in this modeling.
Table V.2: Ore reduction process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Dust loss (fraction of solids reporting to the off gas)</td>
<td>17%</td>
</tr>
<tr>
<td>Kiln discharge temperature</td>
<td>900°C</td>
</tr>
<tr>
<td>Heat loss in the kiln</td>
<td>12%</td>
</tr>
<tr>
<td>Heat loss in the kiln combustion chamber</td>
<td>12%</td>
</tr>
</tbody>
</table>

5.1.4 Dust Conveying

The MetSim model for the dust conveying is a simple flowsheet, where dust from the kiln and furnace feed the conveying system and then cooled. The cooled dust feeds an extruder. Extruder product feeds the kiln, to recovery the nickel in the dust.

5.1.5 Smelting Process

The kiln product feeds the electric furnace. In this unit operation, the material is smelted and separated into slag and metal phase. Almost all the nickel and iron are reduced. Equations (5.11) to (5.25) represent the main reactions that take place in this unit operation. Carbon electrodes are controlled to obtain the desired carbon in the furnace feed. The electric furnace power is calculated based on the furnace heat balance, in order to achieve the desired slag liquidus temperature. Table V.3 shows the main electric furnace process parameters.

5.1.6 Off gas system

Off gas from the furnace is cooled and treated in order to recover the dust that is sent to the dust conveying system. Infiltration air is added proportionally to the amount of gas feeding the off gas system.

\[
\begin{align*}
\text{Al}_2\text{O}_3(s) & = \text{Al}_2\text{O}_3(\text{slag}) \\
\text{SiO}_2(s) & = \text{SiO}_2(\text{slag})
\end{align*}
\]  

(5.11)  
(5.12)
\[
\begin{align*}
\text{Mg(OH)}_2(s) &= \text{MgO}(s) + \text{H}_2\text{O}(s) \\
\text{Fe}_2\text{O}_3\cdot\text{H}_2\text{O}(s) &= \text{Fe}_2\text{O}_3(s) + \text{H}_2\text{O}(g) \\
\text{MgO}(s) + \text{SiO}_2(\text{slag}) &= \text{MgSiO}_3(\text{slag}) \\
\text{Fe}(s) &= \text{Fe}(m) \\
3\ \text{Fe}_2\text{O}_3(s) + \text{C}(s) &= 2\ \text{Fe}_3\text{O}_4(s) + \text{CO}(g) \\
\text{Fe}_3\text{O}_4(s) + \text{C}(s) &= 3\ \text{FeO}(s) + \text{CO}(g) \\
\text{Ni}(s) &= \text{Ni}(m) \\
2\ \text{NiO}(s) + \text{C}(s) &= 2\ \text{Ni}(m) + \text{CO}_2(g) \\
\text{FeO}(s) + \text{C}(s) &= \text{Fe}(m) + \text{CO}_2(g) \\
2\ \text{FeO}(s) + \text{SiO}_2(\text{slag}) &= \text{Fe}_2\text{SiO}_4(\text{slag}) \\
\text{Co}(s) &= \text{Co}(m) \\
\text{Co}_3\text{O}_4(s) + 2\ \text{C}(s) &= 3\ \text{Co}(m) + 2\ \text{CO}_2(g) \\
2\ \text{Cr}_2\text{O}_3(s) + 3\ \text{O}_2(s) &= 4\ \text{CrO}_3(\text{slag})
\end{align*}
\] (5.13-5.25)

Table V.3: Electric furnace process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Dust loss (fraction of solids reporting to the off gas)</td>
<td>1%</td>
</tr>
<tr>
<td>Heat loss</td>
<td>Equation (3.3)</td>
</tr>
<tr>
<td>Matte %Ni</td>
<td>25%</td>
</tr>
</tbody>
</table>

5.1.7 Reagents/ Final discharge

In the reagents and final discharge sections, all the reagents and discharge streams included in the model are presented. Reagents are added on demand to the process.

The results (output) from the metallurgical simulation comprise major reagent consumption calculated according to the mass and energy balance. A schematic diagram showing the output from FeNi pyrometallurgy is presented in Figure 5.3. Due to its major impact on overall cost, power should be highlighted as one of the major output.
By changing any of the project input and process parameters indicated in Section 5.1, various reagents/power consumptions and different stream flowrate is obtained. A standard output spreadsheet feeds the excel model. Various output spreadsheets will lead to varying results in the economic analysis.

5.2 High Pressure Acid Leaching (HPAL)

The HPAL process model was built considering nine major areas. These are: i) beneficiation and acid leaching; ii) primary iron precipitation, iii) counter current decantation (CCD), iv) secondary iron precipitation, v) mixed hydroxide precipitation (MHP) circuit 1, vi) MHP circuit 2, vii) impurities removal, viii) reagents and ix) products/tailings. Figure 5.4 shows a schematic flow diagram of this process. Appendix D presents the MetSim model flowsheets and all the chemical reactions and controllers considered to build the model.

The feed to the HPAL model are all the major elements in the oxide form, namely; NiO, FeO, MgO, MnO, CaO, CoO, CuO, Cr$_2$O$_3$, Al$_2$O$_3$, Fe$_2$O$_3$ and FeOOH. The balance (100% basis) considering the chemical analyses of the major elements is reported as SiO$_2$, an inert component.
5.2.1 Beneficiation/Acid Leaching

In the beneficiation circuit, the ore is crushed to the target particle size to feed the autoclave and screened to remove the coarse high silica, low nickel material. Ore fed to the plant is back calculated based on the reserve, mine lifetime and moisture content in the ore, as explained in Section 5.1.2. After beneficiation, the slurry is fed to the autoclave, together with concentrated sulfuric acid. Equations (5.26) to (5.34) are the main reactions that occur in the autoclave at high temperature and high-pressure environment. The discharge solution is depressurized and sent to the neutralization step. The heat from the let down is used to heat the feed to the autoclave. The process parameters that are fed to the model are presented on Table V.4.

\[
\begin{align*}
\text{NiO}_\text{(s)} + \text{H}_2\text{SO}_4\text{(aq)} & = \text{NiSO}_4\text{(aq)} + \text{H}_2\text{O}(l) & (5.26) \\
\text{CoO}_\text{(s)} + \text{H}_2\text{SO}_4\text{(aq)} & = \text{CoSO}_4\text{(aq)} + \text{H}_2\text{O}(l) & (5.27) \\
\text{MnO}_\text{(s)} + \text{H}_2\text{SO}_4\text{(aq)} & = \text{MnSO}_4\text{(aq)} + \text{H}_2\text{O}(l) & (5.28) \\
\text{MgO}_\text{(s)} + \text{H}_2\text{SO}_4\text{(aq)} & = \text{MgSO}_4\text{(aq)} + \text{H}_2\text{O}(l) & (5.29)
\end{align*}
\]
FeO(s) + 3H2SO4(aq) = FeSO4(aq) + 3 H2O(l) \hspace{1cm} (5.30)

Al2O3(s) + 3H2SO4(aq) = Al2(SO4)3(aq) + 3 H2O(l) \hspace{1cm} (5.31)

Al2O3(s) + 4H2SO4(aq) + 5 H2O(l) = 2 Al3(SO4)2 H2O (OH)6(s) \hspace{1cm} (5.32)

2 FeOOH(s) + 3 H2SO4(aq) = Fe2(SO4)3(aq) + 4 H2O(l) \hspace{1cm} (5.33)

Fe2(SO4)3(aq) + 3H2O(l) = 3H2SO4(aq) + Fe2O3(s) \hspace{1cm} (5.34)

Table V.4: Beneficiation/ Acid leaching process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Screen cut-off</td>
<td>800 mm</td>
</tr>
<tr>
<td>Feed to the autoclave % solids</td>
<td>40%</td>
</tr>
<tr>
<td>Pre-heater pressure</td>
<td>1340 kPa</td>
</tr>
<tr>
<td>Autoclave pressure</td>
<td>4545 kPa</td>
</tr>
<tr>
<td>Autoclave temperature</td>
<td>250°C</td>
</tr>
<tr>
<td>Free acid in the autoclave discharge</td>
<td>50 g/L</td>
</tr>
<tr>
<td>Flash pressure</td>
<td>111.3 kPa</td>
</tr>
</tbody>
</table>

5.2.2 Primary iron precipitation, CCD and secondary iron precipitation

The autoclave discharge is sent to the primary neutralization step to remove the excess acid, majority of iron, and chromium and some aluminum. The first step is a releach tank, in which the pulp from the autoclave is contacted with the residues from the secondary neutralization and the MHP 2 circuits. The goal is to dissolve any precipitated nickel, to reduce nickel losses. After the releach tank, the pulp is introduced to the primary neutralization circuit. The neutralization agent used in the present model is limestone. The pulp from the first tank is sent to the CCD (counter-current decantation) circuit, in where the solid liquid separation takes place. The CCD underflow is sent to the tailings or an effluent treatment plant and the overflow (Pregnant Leach Solution – PLS) is sent to the secondary neutralization circuit, where the remaining iron, aluminum and chromium are precipitated using limestone. Some nickel may also be precipitated. To recover the nickel values from this residue, it is also sent back to the releach tank. The overflow of the thickener following secondary

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neutralization, with low iron, aluminum and chromium is sent to the nickel recovery circuit. Equations (5.35) to (5.44) show the main reactions that take place in these areas. Flocculants are added proportionally to the amount of dry solids feeding the thickener of each circuit. Table V.5 shows some process parameters that have been considered in modeling this step. Again, as all the others process parameters, these values are typical numbers or numbers from a Vale Inco projects, and can be changed as necessary.

Table V.5: First neutralization, CCD and second neutralization process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>First neutralization final pH</td>
<td>2.5</td>
</tr>
<tr>
<td>CCD’s thickener underflow % solids</td>
<td>41%</td>
</tr>
<tr>
<td>Wash water to the CCD (water/mass of dry ore)</td>
<td>2.3</td>
</tr>
<tr>
<td>Wash water pH</td>
<td>4</td>
</tr>
<tr>
<td>Mixing efficiency(^1) in the CCD</td>
<td>95%</td>
</tr>
<tr>
<td>Second neutralization thickener underflow % solids</td>
<td>15%</td>
</tr>
<tr>
<td>Secondary neutralization pH</td>
<td>4.5</td>
</tr>
</tbody>
</table>

\(^1\)The mixing efficiency means that 95% of the original liquid phase in the pulp is replaced by wash water (pH=4).

**Releach tank**

\[
\text{Ni(OH)}_2(s) + \text{H}_2\text{SO}_4(aq) = \text{NiSO}_4(aq) + \text{H}_2\text{O}(l) \tag{5.35}
\]

**First Neutralization**

\[
\begin{align*}
\text{Fe}_2(\text{SO}_4)_3(aq) + 3\text{CaCO}_3(s) + 9\text{H}_2\text{O}(l) &= 3\text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2\text{Fe(OH)}_3(s) \tag{5.36} \\
\text{Al}_2(\text{SO}_4)_3(aq) + 3\text{CaCO}_3(s) + 9\text{H}_2\text{O}(l) &= 3\text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2\text{Al(OH)}_3(s) \tag{5.37} \\
\text{Cr}_2(\text{SO}_4)_3(aq) + 3\text{CaCO}_3(s) + 9\text{H}_2\text{O}(l) &= 3\text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2\text{Cr(OH)}_3(s) \tag{5.38} \\
\text{H}_2\text{SO}_4(aq) + \text{CaCO}_3(s) + \text{H}_2\text{O}(l) &= \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) \tag{5.39}
\end{align*}
\]

**Second Neutralization**

\[
\begin{align*}
\text{Fe}_2(\text{SO}_4)_3(aq) + 3\text{CaCO}_3(s) + 9\text{H}_2\text{O}(l) &= 3\text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2\text{Fe(OH)}_3(s) \tag{5.40} \\
\text{Al}_2(\text{SO}_4)_3(aq) + 3\text{CaCO}_3(s) + 9\text{H}_2\text{O}(l) &= 3\text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2\text{Al(OH)}_3(s) \tag{5.41} \\
\text{Cr}_2(\text{SO}_4)_3(aq) + 3\text{CaCO}_3(s) + 9\text{H}_2\text{O}(l) &= 3\text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2\text{Cr(OH)}_3(s) \tag{5.42} \\
\text{NiSO}_4(aq) + \text{CaCO}_3(s) + 3\text{H}_2\text{O}(l) &= \text{CO}_2(g) + \text{CaSO}_4.2\text{H}_2\text{O}(s) + \text{Ni(OH)}_2(s) \tag{5.43} \\
\text{H}_2\text{SO}_4(aq) + \text{CaCO}_3(s) + \text{H}_2\text{O}(l) &= \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) \tag{5.44}
\end{align*}
\]
5.2.3 Mixed Hydroxide Precipitate (MHP) 1 and 2

The secondary neutralization product is sent to the MHP (mixed hydroxide precipitate) 1 circuit, where stoichiometric amount of dry MgO based on nickel and cobalt in the feed is added to neutralize the solution and to precipitate the cobalt and nickel as hydroxides. The pulp is sent to a thickener, where flocculants are added proportionally to the amount of solids, and filtered. This solid is the final product considered in this model. The solution is sent to the MHP 2 circuit, where the remaining nickel and cobalt are precipitated, using lime. Some manganese, considered as an impurity, may contaminate the precipitate thus the solid from this circuit is sent back to the releach tank to recover nickel and cobalt. The solution is sent to the impurity removal circuit. Equations (5.45) to (5.51) show the main reactions that take place in these circuits. Table V.6 shows some process parameters that are considered in modeling this stage.

MHP 1

\[
\begin{align*}
\text{CoSO}_4^{aq} + \text{MgO}^{s} + \text{H}_2\text{O}^{l} &= \text{MgSO}_4^{aq}. + \text{Co(OH)}_2^{s} \quad (5.45) \\
\text{NiSO}_4^{aq} + \text{MgO}^{s} + \text{H}_2\text{O}^{l} &= \text{MgSO}_4^{aq}. + \text{Ni(OH)}_2^{s} \quad (5.46) \\
\text{H}_2\text{SO}_4^{aq} + \text{MgO}^{s} &= \text{MgSO}_4^{aq}. + 2\text{H}_2\text{O}^{l} \quad (5.47)
\end{align*}
\]

MHP 2

\[
\begin{align*}
\text{H}_2\text{SO}_4^{aq} + \text{Ca(OH)}_2^{s} &= \text{CaSO}_4.2\text{H}_2\text{O}^{s} \quad (5.48) \\
\text{CoSO}_4^{aq} + 2\text{H}_2\text{O}^{l} + \text{Ca(OH)}_2^{s} &= \text{CaSO}_4.2\text{H}_2\text{O}^{s}. + \text{Co(OH)}_2^{s} \quad (5.49) \\
\text{NiSO}_4^{aq} + \text{MgO}^{s} + \text{H}_2\text{O}^{l} &= \text{MgSO}_4^{aq}. + \text{Ni(OH)}_2^{s} \quad (5.50) \\
\text{MnSO}_4^{aq} + 2\text{H}_2\text{O}^{l} + \text{Ca(OH)}_2^{s} &= \text{CaSO}_4.2\text{H}_2\text{O}^{s} + \text{Mn(OH)}_2^{s} \quad (5.51)
\end{align*}
\]

5.2.4 Impurities Removal

An impurity removal step is considered in order to obtain good quality water for recycling or to prepare the effluent for discharge. Lime is used as neutralization agent, added three times the stoichiometric amount of manganese, usually, the
main impurity to be removed. As a result, this circuit is commonly named as Manganese Removal. Flocculant is added proportionally to the amount of solids in the reactors. The main reactions that occur in this circuit are represented by equations (5.52) and (5.53). Table V.7 shows the process parameters.

Table V.6: MHP 1 and 2 process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>MHP 1 thickener underflow % solids</td>
<td>30%</td>
</tr>
<tr>
<td>Washing efficiency in the MHP filter(^1)</td>
<td>70%</td>
</tr>
<tr>
<td>Water to filter cake washing (water/t of dry solids in the feed)</td>
<td>4.5</td>
</tr>
<tr>
<td>Final cake % solids</td>
<td>54%</td>
</tr>
<tr>
<td>Final pH in the MHP 2 circuit</td>
<td>8.7</td>
</tr>
<tr>
<td>MHP 2 thickener underflow % solids</td>
<td>13%</td>
</tr>
</tbody>
</table>

\(^1\)Washing efficiency means that after one displacement wash, 70% of the initial liquor is taken out and 30% of it remains in the cake.

\[
\text{MnSO}_4(aq) + 2\text{H}_2\text{O}(l) + \text{Ca(OH)}_2(s) = \text{CaSO}_4.2\text{H}_2\text{O}(s) + \text{Mn(OH)}_2(s) \tag{5.52}
\]

\[
\text{MgSO}_4(aq) + 2\text{H}_2\text{O}(l) + \text{Ca(OH)}_2(s) = \text{CaSO}_4.2\text{H}_2\text{O}(s) + \text{Mg(OH)}_2(s) \tag{5.53}
\]

Table V.7: Impurity removal’s process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Impurities removal thickener underflow % solids</td>
<td>44%</td>
</tr>
</tbody>
</table>

5.2.5 Reagents and Final discharge

In the reagents and final discharge sections, all the reagents and discharge streams included in the model are presented. Reagents are added on demand to the process. Table V.8 shows the process parameters considered in the analyses of these stages.

Table V.8: Reagents process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flocculant supply % solids</td>
<td>0.5%</td>
</tr>
<tr>
<td>Lime supply % solids</td>
<td>10%</td>
</tr>
<tr>
<td>Limestonne supply % solids</td>
<td>30%</td>
</tr>
</tbody>
</table>
The results from the metallurgical simulation are all the model outputs, including major reagents, calculated based on the mass and heat balances. The output from HPAL modeling is presented in Figure 5.5.

![Figure 5.5: HPAL output](image)

5.3 Heap leaching

The heap leaching process model was built considering eight main process areas. These are: i) agglomeration and heap leaching, ii) primary iron precipitation, iii) secondary iron precipitation, iv) MHP 1, v) MHP 2, vi) impurities removal, vii) reagents and viii) products/tailings. Figure 5.6 illustrates a flow diagram of this process. Appendix E presents the MetSim model flowsheet as well as all the chemical reactions and controllers considered to build the model.

The feed to the heap leaching model are all the major elements in their respective oxide forms: NiO, FeO, MgO, MnO, CaO, CoO, CuO, Cr₂O₃, Al₂O₃, Fe₂O₃, and FeOOH. Some of these elements can occur in the saprolite or the ferruginous form, and this mineralogy affects the atmospheric leaching kinetics. As a result these elements were considered in two stages (e.g. NiO\textsubscript{sap} and NiO\textsubscript{fer}, CoO\textsubscript{sap} and CoO\textsubscript{fer}, MgO\textsubscript{sap} and MgO\textsubscript{fer}, MnO\textsubscript{sap} and MnO\textsubscript{fer}, Fe₂O₃\textsubscript{sap}, Fe₂O₃\textsubscript{fer}) where “sap” means saprolite and “fer” means ferruginous. After the balance based on the chemical analysis of the major elements, the remaining is reported as SiO₂, an inert component.
5.3.1 Agglomeration/ Heap leaching

The ore is crushed to the target particle size to the heap, and screened to remove the coarse high silica, low nickel material. The ore feed flowrate is back calculated based on the reserve, mine lifetime and moisture content in the ore, as explained in Section 5.1.2, equation (5.1). The screened ore feeds the agglomeration drum, where acid is added on a stoichiometric base aimed at 50% neutralization of the magnesium content of the ore. Water is added in the desired ratio. The agglomerated ore is stockpiled in the heap leaching pad, where the leaching cycle takes place. Acid solution is added as reagent.

Figure 5.6: Process flow diagram for heap leaching

Equations (5.54) to (5.60) represent the main leaching reactions. Pereira and Gobbo (2006) obtained the extraction curves for the saprolite and the ferruginous fraction of a given ore, at different acid concentrations. These curves are shown in Figures 5.7 and 5.8. Equation (5.61) was fitted to the experimental data. The parameter X, for each acid concentration and a given
leach cycle (time), was obtained using Figures 5.7 and 5.8 and Equation (5.61). Table V.10 shows the values of parameter X for the corresponding mineralogy and acid concentration.

\[
\begin{align*}
\text{NiO} & (s) + \text{H}_2\text{SO}_4(aq) = \text{NiSO}_4(aq) + \text{H}_2\text{O}(l) \quad (5.54) \\
\text{CoO} & (s) + \text{H}_2\text{SO}_4(aq) = \text{CoSO}_4(aq) + \text{H}_2\text{O}(l) \quad (5.55) \\
\text{MnO} & (s) + \text{H}_2\text{SO}_4(aq) = \text{MnSO}_4(aq) + \text{H}_2\text{O}(l) \quad (5.56) \\
\text{MgO} & (s) + \text{H}_2\text{SO}_4(aq) = \text{MgSO}_4(aq) + \text{H}_2\text{O}(l) \quad (5.57) \\
\text{Fe}_2\text{O}_3(s) + 3\text{H}_2\text{SO}_4(aq) & = \text{Fe}_2(\text{SO}_4)_3(aq) + 3\text{H}_2\text{O}(l) \quad (5.58) \\
\text{Al}_2\text{O}_3(s) + 3\text{H}_2\text{SO}_4(aq) & = \text{Al}_2(\text{SO}_4)_3(aq) + 3\text{H}_2\text{O}(l) \quad (5.59) \\
\text{CaO} & (s) + \text{H}_2\text{SO}_4(aq) + \text{H}_2\text{O}(l) = \text{CaSO}_4.2\text{H}_2\text{O}(s) \quad (5.60)
\end{align*}
\]

![Graph](image.png)

Figure 5.7: Nickel extraction at different acid concentration for a ferruginous ore

Metal Extraction (%) = \(100 \times (1 - (1 - X)^{\text{time}})\) \hspace{1cm} (5.61)

63
Figure 5.8: Nickel extraction at different acid concentration for a saprolite ore.

Table V.9: Parameter X for the saprolite and ferruginous fractions of a selected ore, at different acid concentration

<table>
<thead>
<tr>
<th>Acid concentration</th>
<th>Saprolite</th>
<th>Ferruginous</th>
</tr>
</thead>
<tbody>
<tr>
<td>20 g/L</td>
<td>0.015713</td>
<td>0.000783</td>
</tr>
<tr>
<td>50 g/L</td>
<td>0.027697</td>
<td>0.001474</td>
</tr>
<tr>
<td>100 g/L</td>
<td>0.039395</td>
<td>0.002979</td>
</tr>
<tr>
<td>150 g/L</td>
<td>0.044943</td>
<td>0.004686</td>
</tr>
<tr>
<td>200 g/L</td>
<td>0.06763</td>
<td>0.006237</td>
</tr>
</tbody>
</table>

Based on the numbers shown on Table V.9, a program was written in order to calculate the rate of reaction for the saprolite and ferruginous materials by using equation (5.61), for the desired acid concentration and leach cycle time.

The amount of water added to the heap is calculated in order to obtain the desired solution flowrate (equation (5.62)). The heap area used only for water balance purpose is calculated based on the equation (5.63).

Solution to the heap \((\text{t/h}) = (\text{Heap area (m}^2) \times \text{irrigation rate (m}^3/\text{h/m}^2))/3\) (5.62)
Heap area (m²) = Reserve (t) x total number of cycles (y/cycle)  
mine lifetime (y) x bulk density (t/m³) x Heap height (m)  

(5.63)

Irrigation rate, reserve, mine lifetime, total leaching cycle, bulk density of ore and heap height, were input parameters, as defined in Table V.10. Inclusion of the denominator (3) in equation (5.62) is consistent with the equation derived by Pereira and Gobbo (2006) patent developed for a counter current operation, in which the heap is divided in sections. Three sections have been considered in the present work, thus implying that only one third of the heap will be fed by the new solution.

The washing of the heap following the leaching is to recover entrained nickel. In the model, the amount of water added to wash is enough to recover 90% of the dissolved nickel, that is, the ore is washed in order to decrease nickel concentration lost with the moisture in the discharged tailings. The evaporation from the heap and ponds are calculated based on the open area and the net evaporation rate (equation (5.64)). The area, calculated by Equation (5.65), considers as open area for evaporation purposes the heap area combined with the pond area. In order to calculate the pond area, the total pond volume was determined considering the flowrate that feeds the pond (one third of the irrigation rate multiplied by the heap area), surge of 15 days and a pond 10 meters high. After the cycle, the heap is washed. Table V.10 summarizes the selected process parameters for agglomeration and heap leaching.

Evaporation flowrate (m³/h) = Open area (m²) x net evaporation rate (m/y)  
8760 (h/y)  

(5.64)

Open area (m²) = Heap area (m²) x (1 + (irrigation rate (m³/d/m²)/3)  
(15(d) x 10(m))  

(5.65)

5.3.2 Primary and secondary neutralization

Primary and secondary neutralization circuits are the same as considered for the HPAL process, with two exceptions: (i) in heap leaching, clean solution (with
no solids) feeds the neutralization step, while in the HPAL, all the slurry from the autoclave feeds the primary neutralization and (ii) the solid/liquid separation for the HPAL process was a CCD circuit, while for heap leaching, a belt filter. Table V.11 shows some process parameters’ values adopted in the neutralization stages.

Table V.10: Agglomeration and heap leaching process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ore bulk density</td>
<td>1 t/m³</td>
</tr>
<tr>
<td>Screen cut-off</td>
<td>800 mm</td>
</tr>
<tr>
<td>Agglomeration drum discharge % solids</td>
<td>65%</td>
</tr>
<tr>
<td>Heap Height</td>
<td>4 m</td>
</tr>
<tr>
<td>Leaching cycle (without washing)¹</td>
<td>20 weeks</td>
</tr>
<tr>
<td>Total heap leaching cycle with washing¹</td>
<td>½ year</td>
</tr>
<tr>
<td>Irrigation rate</td>
<td>10 l/h/m²</td>
</tr>
<tr>
<td>Net evaporation rate²</td>
<td>2000 mm/y</td>
</tr>
<tr>
<td>Remaining ore moisture after leaching¹</td>
<td>35%</td>
</tr>
<tr>
<td>Remaining ore moisture after washing¹</td>
<td>40%</td>
</tr>
<tr>
<td>Free acid in the heap discharge</td>
<td>8 g/L</td>
</tr>
<tr>
<td>Wash efficiency in the heap²</td>
<td>80%</td>
</tr>
<tr>
<td>Pond autonomy</td>
<td>15 days</td>
</tr>
<tr>
<td>Pond height</td>
<td>10 m</td>
</tr>
</tbody>
</table>

¹The heap leaching cycle comprises of the leaching with acid during a period of time and after that, water is added in order to wash the solids for disposal.
²Washing efficiency means that after one displacement wash, 80% of the initial liquor is taken out and 20% of it remains in the cake.
³Once the water balance is an important factor in the heap leaching, due to the bigger open area, the net evaporation rate is an input and will impact in the amount of process water to feed the plant.

Table V.11: First neutralization and second neutralization process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Primary neutralization thickener underflow % solids</td>
<td>25%</td>
</tr>
<tr>
<td>Second neutralization thickener underflow % solids</td>
<td>15%</td>
</tr>
<tr>
<td>Washing efficiency in the iron removal filter¹</td>
<td>99.5%</td>
</tr>
<tr>
<td>Final cake % solids</td>
<td>40%</td>
</tr>
</tbody>
</table>

¹Washing efficiency means that after one displacement wash, 99.5% of the initial liquor is taken out.

5.3.3 Mixed Hydroxide Precipitate (MHP) 1 and 2

MHP 1 and 2 are the same as considered for the HPAL process. Table V.12 shows the selected process parameters.
Table V.12: MHP 1 and 2 process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>MHP 1 thickener underflow % solids</td>
<td>30%</td>
</tr>
<tr>
<td>Washing efficiency in the MHP filter(^1)</td>
<td>70%</td>
</tr>
<tr>
<td>Final cake % solids</td>
<td>54%</td>
</tr>
<tr>
<td>MHP 2 thickener underflow % solids</td>
<td>13%</td>
</tr>
</tbody>
</table>

\(^1\)Washing efficiency means that after one displacement wash, 70% of the initial liquor is taken out and 30% of it remains in the cake.

5.3.4 Impurities Removal

Impurities removal in the heap leaching circuit is the same as the one considered for the HPAL process. Table V.13 shows the values adopted for the process parameters.

Table V.13: Impurity removal process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Impurities removal thickener underflow % solids</td>
<td>44%</td>
</tr>
</tbody>
</table>

5.3.5 Reagents and final discharge

In the reagents and final discharge sections, all the reagents and discharge streams included in the model are shown in Table V.14. Reagents are added on demand to the process.

Table V.14: Reagents process parameters

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>Flocculant supply % solids</td>
<td>0.5%</td>
</tr>
<tr>
<td>Lime supply % solids</td>
<td>10%</td>
</tr>
<tr>
<td>Limestonne supply % solids</td>
<td>30%</td>
</tr>
</tbody>
</table>

The results from the metallurgical simulation are all the model outputs, including the major reagent consumption, calculated according to the mass and heat balance. The outputs from the heap leaching modeling are indicated in Figure 5.5.
5.4 Case Studies

In order to show a practical application of the developed models, some case studies were investigated using Sample A, from a project under development. This sample shows an atypical ore chemical composition (Table V.15). Comparing this ore’s composition with the data shown in Figure 3.1b, one may observe that the magnesium is in the range of a limonite ore, while iron is in the range of saprolite ore.

Table V.15: Ore type A chemical analysis

<table>
<thead>
<tr>
<th>Ni (%)</th>
<th>Co (%)</th>
<th>Mg (%)</th>
<th>Al (%)</th>
<th>Mn (%)</th>
<th>Ca (%)</th>
<th>Si (%)</th>
<th>Fe (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>1.2</td>
<td>0.06</td>
<td>2.8</td>
<td>2.8</td>
<td>0.3</td>
<td>0.3</td>
<td>28</td>
<td>16</td>
</tr>
</tbody>
</table>

For the HPAL model, all the iron was considered in the form of FeOOH. In the heap-leaching model all the manganese, cobalt, magnesium and 62% of the nickel were considered in the saprolite form, while all the iron was considered in the ferruginous form. In the pyrometallurgical model, all the nickel and cobalt were considered in the oxide form. Iron was considered as hematite.

5.4.1 Case 1

The behavior of ore type A was evaluated for the three process routes considered in this study. The input to all the three models were 70 million tonnes of reserve, 20 years life time, 10% ore moisture and nickel price of US$ 12,000/t. Appendix B shows the reserve/resources for projects to be developed until 2015. Figure 5.9 indicates the percentage of projects for each reserve interval and shows that 46% are deposits with less than 100 million t. As a result, 70 million t is within the interval of the majority of nickel projects to be developed until 2015.
Based only on chemical analysis, it can be concluded that the Sample A is not suitable for FeNi process, since the SiO$_2$/MgO ratio is very high (approximately 12) comparing to the normal industrial practice (e.g. ratio of 1.5 to 2.5). This value is not presented in Figure 5.10. The ratio given in Fig 5.10 for an ore with approximately 15% FeO, is about 3.
In order to evaluate the output from the model, the three different process models were run using the sample A ore and the results are shown in Figure 5.11. The vertical line (axis y) is the ratio of present values for each route relatively to the HPAL route. It is confirmed that this specific ore is not favourable for the FeNi process. The HPAL and HL results were similar, heap leaching is slightly more attractive than the HPAL. The FeNi operation cost distribution obtained from the MetSim model is presented in Figure 5.12. It is shown that the two major costs are power and reagents, 34% and 48%, respectively, of the total cost. The consumables distribution shows that this cost is mainly related to carbon consumption in furnace electrodes (Figure 5.13) as a result of the very high SiO$_2$/MgO ratio in the ore and the consequent high temperature in the furnace.

Figure 5.10 shows the slag liquidus temperature for different SiO$_2$/MgO ratios. Ratios higher than 4 should be extrapolated from the data in the diagram. Metal tapping temperature (a process input) considered was 1700°C, a higher number than those shown in the chart. This temperature would require confirmation by testwork. In order to evaluate Mg impact on the pyrometallurgical process, a simulation was done considering the same ore, but a SiO$_2$/MgO value of 2.5. This number was taken from one random point in Figure 5.10 of an ore with 15% FeO. The result from this simulation is shown in Figure 5.11. An increase in the Mg content causes a significant improvement in the relative NPV but the FeNi route is still not attractive. As shown in Appendix B, most of the prospective projects considering this route will be applied to relatively high grade ores and a high production throughput.

The conclusion from case 1 is that low grade, small to medium-size deposits with low magnesium are not to be treated by the FeNi pyrometallurgical process. The similar attractiveness factors obtained with the HPAL and heap leaching routes for this type of deposit suggests that a more detailed investigation should be conducted with the two remaining options.
Figure 5.11: Comparison of the behavior of sample A for the three selected process routes

Figure 5.12: Operation cost distribution
5.4.2 Case 2

Type A ore was demonstrated to be suitable for being treated by either HPAL or heap leaching (Case 1). The aims of Case study 2 were to evaluate the effect of the size of the deposit and its nickel content, both important variables, on the relative performance of HPAL and HL processes. In this case, the models were run considering the ore body size varying from 40 Million tonnes to 160 Million tonnes, for projects of 20 years lifetime. Figure 5.14 shows the distribution of projects in each nickel grade interval, considering the spreadsheet in Appendix B. The selection of 1.5% and 1.8% Ni grades for the present simulation was based on the data shown in Figure 5.14.

Figures 5.15 to 5.17 show the results from Case study 2, for the selected nickel grades. In order to compare the economic viability of the project, the simulation of HPAL, with 160 Million tonnes orebody and 1.2% nickel grade were selected as a base case. The vertical axis in the following figures is the ratio of the NPV for each simulation and the base case.
Figure 5.14: Percentage of projects for each nickel content interval

Figure 5.15 shows the results obtained for deposits with 1.2% nickel. From this specific deposit (low grade ore), and considering all the economical inputs discussed in section 4, the heap leaching route is more favorable than the HPAL process, regardless the size of the deposit.

Figure 5.15: Comparison of HPAL and HL for sample A, with Ni grade of 1.2% and different ore body sizes.
Figure 5.16 shows the simulation for Sample A, when nickel grade was increased to 1.5%. The suitability of HL has increased with the increased nickel grade but with increasing deposit size, HPAL compares similar to HL. The results also indicate that the relatively high sensitivity (slope) of the HPAL process with respect to the size of the deposit. This effect can be due to the lower operational cost per unit of nickel of HPAL route when comparing to heap leaching. For the Sample A ore, with 1.5% nickel, a deposit over 180 Mt would favor the HPAL route, whilst HL would be suitable for deposit sizes below that figure with same nickel grade.

Figure 5.17 shows the simulation for Sample A, with the nickel grade increased to 1.8%. The HPAL route has an even better performance than that shown in Figure 5.16. It can be concluded that for Sample A ore with 1.8% nickel, both process routes can be used which would deliver similar NPVs.

The point where the NPV of heap leaching and HPAL are equal, for the ores with 1.5% and 1.8% Ni, are shown in Figure 5.18. This graph can be populated with more number, once more simulations are performed. The ore with 1.2% nickel was neglected once the lines are parallel to each other. Projects above the line, points to the HPAL process and to HL below the line.

Figure 5.16: Behavior comparison of sample A, with Ni grade of 1.5% and different ore body sizes.
The results shown in Figures 5.15 to 5.17 indicate that the HPAL is more attractive to larger deposits with higher nickel content.

Figure 5.17: Behavior comparison of sample A, with Ni grade of 1.8% and different ore body sizes.

Figure 5.18: Sample A orebody size, showing the region where Heap leaching is more attractive than HPAL, as a function of % nickel

5.4.3 Case 3

It is common to undertake beneficiation tests with an unknown ore sample in order to evaluate the possibility to concentrate the nickel in the finer fraction. An
example of this evaluation is an investigation carried out with one Vale Inco ore. The main objective of this study was to evaluate the behavior of nickel minerals in the beneficiation circuit. Different samples have been processed. It was found that the nickel enrichment could be related with the silica content in the sample. Equation (5.66) relates nickel enrichment with silica content in the feed, for blended samples, considering screening in 200 Mesh Tyler (0.74mm) and silica concentration in the 0 – 45% interval. Cobalt, aluminum, magnesium and iron enrichments are lower than the nickel, and are described by equations (5.67) to (5.70). Equation (5.71) relates the average mass recovery with silica content in the feed, for the aforementioned range of silica and screen opening. Equation (5.72) relates Ni average metallurgical recovery with silica content. A, B, C and D are scalars numbers, obtained based on test work results (MENDES et. al., 2003).

Equations (5.66) to (5.71) have been applied to the present case study in order to evaluate the impact of screening circuit on the attractiveness of the HL/HPAL processes. The equations evaluate nickel and cobalt enrichments and the major impurities based on silica concentration. These are important parameters which should be adjusted to each project (ore) under consideration.

\[
\text{Ni Enriq.} = A_1 (\%\text{Si})^3 - B_1 (\%\text{Si})^2 + C_1 (\%\text{Si}) + D_1 \tag{5.66}
\]

\[
\text{Co Enriq.} = A_2 (\%\text{Si})^3 - B_2 (\%\text{Si})^2 + C_2 (\%\text{Si}) + D_2 \tag{5.67}
\]

\[
\text{Al Enriq.} = A_3 (\%\text{Si})^3 - B_3 (\%\text{Si})^2 + C_3 (\%\text{Si}) + D_3 \tag{5.68}
\]

\[
\text{Mg Enriq.} = A_4 (\%\text{Si})^3 + B_4 (\%\text{Si})^2 - C_4 (\%\text{Si}) + D_4 \tag{5.69}
\]

\[
\text{Fe Enriq.} = A_5 (\%\text{Si})^3 - B_5 (\%\text{Si})^2 + C_5 (\%\text{Si}) + D_5 \tag{5.70}
\]

\[
\text{Mass Recovery (\%)} = A_6 (\%\text{Si}) + B_6 \tag{5.71}
\]

\[
\text{Ni Met. Rec.} = A_7 (\%\text{Si})^3 + B_7 (\%\text{Si})^2 - C_7 (\%\text{Si}) + D_7 \tag{5.72}
\]

Figure 5.19 and 5.20 shows the results for the case study 3. Nickel recovery obtained in all the simulations was approximately 75% and the mass recovery was around 70%. The explanation for this behavior is that the beneficiation circuit reduced the HPAL attractiveness, for both nickel content tested (e.g.
1.2% and 1.5%), due to nickel losses in the coarse fractions which would be discarded. In this case, an attractive process route would be to integrate the HPAL and HL, with HPAL treating high-grade fraction (e.g. enriched) and HL treating the coarse material (e.g. low grade material).

![Figure 5.19: Comparison of HL and HPAL routes, the latter with and without beneficiation, for different ore body sizes and 1.2%Ni.](image)

![Figure 5.20: Comparison of HL and HPAL routes, the latter with and without beneficiation, for different ore body sizes and 1.5%Ni.](image)
5.4.4 Case 4

The goal of case 4 is to evaluate the impact of nickel price on the selection of heap leaching and HPAL routes. Figure 5.21 and 5.22 summarize the results of the simulation with 1.2 and 1.5% Ni ores. A comparison was carried out with US$ 20,000/t and US$ 12,000/t.

Figure 5.21: Evaluation of the impact of nickel prices on process selection for a 1.2% nickel ore.

Figure 5.22: Evaluation of the impact of nickel prices on process selection for a 1.5% nickel ore.

Figure 5.21 shows that for lower nickel price and ore nickel grade at 1.2%, heap leaching route is favorable for Sample A, considering all ore body sizes evaluated. For nickel price of US$ 20,000/t, deposits over 100 million tons
favors the HPAL route. Figure 5.22 indicates that for lower nickel price, HPAL and heap leaching results are similar, whilst at nickel price of US$ 20,000/t, and ore body size over 70 million tons favors HPAL. In general, higher nickel prices favors HPAL route. Smaller deposits and low-grade deposits better suited for heap leaching process, even with higher nickel prices.

5.4.5 Case 5

As explained in section 3.4.2, iron minerals in the HPAL process react with sulfuric acid to form mainly iron (III) sulfate. But a great amount of this metal precipitates as hematite and jarosite, causing the regeneration of the acid consumed. Therefore, an increase in iron in the HPAL feed does not affect the attractiveness of this process. Conversely, in heap leaching process iron minerals react with sulfuric acid to form iron sulfate, with eventual significant impacts on the acid and neutralization reagent consumption. The objective of the Case study 5 was to evaluate the impact of iron grade on the performance of heap leaching and HPAL. Sample A composition was considered, with an increase in the iron content from 16% to 21% and then to 40%. These iron grades were selected from data given in Figure 3.1b.

Figure 5.23 presents the results of this simulation, for nickel content of 1.2%. It shows that for ores with 1.2% Ni and high iron grade (e.g. 40%), the NPV of the project using heap leaching route is decreased with increasing orebody size. This means that the operational cost becomes increasingly higher than the profit. Therefore, for this type of ores, HPAL is more attractive than HL for full range of orebody sizes studied. For an illustrative purpose, Figures 5.24 to 5.27 depicted the operation costs and reagents distribution for the simulation considering ore body size of 160 Mt, with nickel content of 1.5% and iron content of 16% and 40%.
Figures 5.24 and 5.25 present the results considering iron content of 16%. It can be seen that 59% of the operational cost is related to reagents and from this cost 73% account for sulfuric acid and limestone. Figure 5.26 and 5.27 show the results for the same simulation, but with 40% iron in the ore. Reagents cost increased up to 64%, and 86% of that is related to sulfuric acid and limestone. Operational cost from the simulation with 40% iron is 63% higher than the ore with 16% iron.

HPAL curves on Figure 5.23 are quite similar for the three iron grades. Heap leaching is more attractive for all the orebody sizes studied, for 16% iron. For 21% iron, at higher ore body sizes, both the HL and and HPAL curves give similar NPVs. This is clearly evident from data summarized in Figure 5.28. It can be concluded that deposits over 170 Mt favor HPAL route.

Summarised in Figure 5.29 are the overall results for the simulation considering 1.5% nickel. It shows that HL is not suitable for ores with high iron grade (e.g. 40%). HPAL is shown to be attractive for Sample A with high iron content for the full range of ore body sizes studied. HPAL curves on Figure 5.29 are the same for the three iron grades. For iron levels of 16%, at higher ore body sizes, both the HL and HPAL show the same response. For the iron level of 21%, a higher ore body size favors HPAL. Figures 5.30 and 5.31 show the equations for both processes and both iron contents. Based on them, it can be concluded that larger deposits with 16% iron content favor the HPAL route whilst the attractiveness of both process routes are similar, when iron content of the ore is increased to 21%.
Figure 5.23: Evaluation of iron concentration impact on the ore with 1.2% nickel, for the HPAL and heap leaching processes.

Figure 5.24: Operational cost distribution for the heap leaching process for the ore body size of 160 Mt, 1.5% nickel and 16% iron.
Figure 5.25: Reagents cost distribution for the heap leaching process for the ore body size of 160 Mt, 1.5% nickel and 16% iron.

Figure 5.26: Operational cost distribution for the heap leaching process for the ore body size of 160 Mt, 1.5% nickel and 40% iron.
Figure 5.27: Reagents cost distribution for the heap leaching process for the ore body size of 160 Mt, 1.5% nickel and 40% iron.

Figure 5.28: Evaluation of heap leaching and HPAL attractiveness on the ore with 1.2% nickel and 21% iron.
Figure 5.29: Evaluation of iron concentration impact on the ore with 1.5% nickel, for the HPAL and heap leaching processes.

Figure 5.30: Evaluation of heap leaching and HPAL attractiveness on the ore with 1.5% nickel and 16% iron.
Figure 5.31: Evaluation of heap leaching and HPAL attractiveness on the ore with 1.5% nickel and 21% iron.
6 Conclusions

The major outcome of the present study was a computer model which can predicts the suitability of a process route to treat a given nickel laterite orebody. The model was integrated with economic considerations, which enables it to determine the NPV of the project with various input parameters such as nickel grade, iron content, the size of the ore body, beneficiation of the ore or nickel price. The process routes considered were FeNi, HPAL and HL. Five case studies were undertaken considering an ore from a project under evaluation. From the study undertaken, following conclusions and general comments can be made:

Case Study 1: Composition of ore

- FeNi process is not suitable for low grade, small to medium-size deposits with low magnesium.
- HPAL and HL deliver similar NPVs for this orebody. HL is slightly attractive than HPAL. However, the model outcome suggests that a more detailed study need to be undertaken to validate the initial predictions.

Case Study 2: Nickel Grade and size of the deposit

- From this specific deposit (1.2% nickel), the HL is more favorable than the HPAL process regardless the size of the deposit.
- Increasing nickel content for 1.5%, HL shows an even higher attractiveness factors than for the 1.2% Ni ore but the superiority with respect with the HPAL is smaller and both processes reach similar performance for the 160 Mt deposit.
- The results also indicate a higher sensitivity of the HPAL process with respect to the size of the deposit, compared to HL, thus indicating that the former is more affected by scale.
• The higher the nickel content and the higher the deposit size, more favorable is for the HPAL route.

Case Study 3: Impact of a screening cut-off

• For this specific ore, considering some theoretical equations, the concentration of the ore by screening was not an advantage.

Case Study 4: Impact of nickel price

• In general, higher nickel prices favor the HPAL route.
• Smaller deposits and low-grade deposits demonstrate better attractiveness of the HL process, even considering a high nickel price.

Case Study 5: Impact of iron concentration

• HPAL curves were similar for the three Fe grades (e.g. 16%, 21% and 40%).
• HL is more attractive for all the orebody sizes studied, at Fe levels of 16% and Ni 1.2%.
• For 21% Fe and Ni 1.2%, at higher orebody sizes, heap leaching and HPAL curves are very close.
• For ores with high Fe grade (e.g. 40%) and Ni 1.2%, the NPV of HL is reduced as orebody size increases. This means that the operational cost becomes increasingly higher than the profit. As a result, for this type of ores HPAL is more attractive for all range of orebody size studied.
• For 40% Fe and 1.5% Ni, heap leaching is still not the best extraction process. In this case, HPAL is also more attractive for the majority of orebody size studied.
7 References


PEREIRA, G., GOBBO, O. (2006) Process for extraction of nickel, cobalt, and other base metals from laterite ores by using heap leaching and product containing nickel, cobalt, and other metals from laterite ores PATENT EP 1 790 739 A1


STYLIANI, A., DIMITRA, D. (1991) *Method for extraction of nickel and/or cobalt from nickel and/or cobalt oxide ores by heap leaching with a dilute sulfuric acid solution, prepared from sea water at ambient temperature* PATENT GR 1003569


8 APPENDIXES
APPENDIX A – Revision on the existing nickel heap leaching patents

A revision on the existing patents was done and the main identified patents related to the heap leaching process using sulfuric acid as leaching agent are summarized below. The first one is a Greek patent (GR 1001555 – 22/march/1994) named “Nickel and Cobalt recovery from low-grade nickel oxide ores by the technique of heap leaching using dilute sulfuric acid at ambient temperature”. This invention describes a hydrometallurgical method to treat low-grade nickel and cobalt oxide ores, using sulfuric acid as leaching agent, at ambient temperature, in order to extract nickel and cobalt as soluble cations. Some parameters for the ore feeding the columns or heap were considered. The ore considered in the patent requires an approximately 10% moisture. This can be accomplished by wetting the ore with water or leach solution. This has the goal to agglomerate the fines and to permit the clay constituents to swell before the heap starts to operate, once it can compromise the established porosity. Particle size and size distribution is another important factor once the fines are good for the leaching due to the greater surface area. But on the other hand, it impedes the percolation of the leach solution through the column or heap by blocking the solution pathway. It is reported that ores with particle size even as large as 3 cm can be used without creating any problem to the percolation and, at the same time, the efficiency of the extraction is very good.

The inventors studied two different techniques. The first one considers the leach solution recycled to the column/heap as many times as it is required so that no increase in nickel concentration is noticed. A fresh solution then replaces the former solution and the process is repeated. The pregnant solution in general has low nickel concentration. To evaluate this technique, some experiments were carried out. It was found that nickel extraction occurs at pH values lower than 1 to 1.3. At pH values higher than 1.3 nickel extraction is negligible. Interesting results from this testwork are that the iron and nickel extraction rates
differ significantly. The Fe/Ni ratios in the leach liquors obtained ranges from 1/1 to 2/1, for nickel recoveries up to approximately 70%. Beyond this value of nickel recovery, intense iron dissolution takes place, while nickel dissolution is very slow. All the elements extraction starts after the extraction of all the calcium in the ore is completed. The calcium in solution is then gradually precipitated as gypsum, due to an increase of the sulfate anions concentration in the leach liquor. In the case there is no calcium in the ore, the leaching of the other metal cations start directly and simultaneously.

The second technique considers the leach solution is recycled to the ore after pH adjustment to a predetermined value. This technique reduces the amount of water required and allows the utilization of the free acid remaining in the final leach liquor. Some experiments were undertaken and the conclusions are: the Fe/Ni ration the leach liquor does not exceed 5/1, even for nickel recoveries higher than 75%; the acid consumption is less and the extraction time is shorter by using the second technique. A lower acidity limit was 0.25M (pH ~1.2), a similar value to the one that was found for the first technique.

During the experimental investigation, some findings led to some suggestions provided by the inventors for commercial applications, such as: the increase in acidity significantly shortens leaching time; the ratio of the ore weight in the heap to the total volume of the leach solution is important to the final chemical composition of the leach liquor. The value of this ratio is chosen based on the water resources, weather conditions, the acid concentration used and the desired leaching time; the final nickel concentration depends on the Aluminum concentration in the same solution. Aluminum concentrations higher than 11-12g/L induces nickel precipitation, thus reducing final nickel recovery and it was found that washing the heaps with water is only required after completion of leaching while it does not affect leaching if it takes place at an intermediate stage.
The second Greek patent (GR 1003569 – 23/april/2001) is an addition to the first one and is named “Method for extraction of nickel and/or cobalt from nickel and/or cobalt oxide ores by heap leaching with a dilute sulfuric acid solution, prepared from sea water at ambient temperature”. This invention concerns an improvement to the method described in Patent GR 1001555, once it refers to the use of sea water or water of various chemical compositions or municipal effluent streams, free of residual solids for the production of a dilute sulfuric acid solution required for the application of the nickel heap leaching method. The use of seawater aims at making possible the usage of the heap leaching process also in regions where there is no fresh water or the use of fresh water renders the method very expensive. For this study, some experiments were undertaken in the following conditions: 10% moisture in the ore fed to the column, leach solution consisting of sea water or industrial effluent water and sulfuric acid with an acid normality of 2N and a flowrate of 653 l/m² day. The ration of the leach solution volume to the ore weight was selected as 1 l/kg. The conclusions of the testwork are that (i) the final concentration of nickel, cobalt, iron, aluminum and chromium in the pregnant solution as well as the percent extraction values are not differentiated when sea water, water of various compositions and effluent water are used, (ii) the final concentration of magnesium and sodium are relatively higher due to the higher initial concentration, (iii) in general the final calcium concentration is low due to the precipitation of gypsum (although its equilibrium concentration can be slightly higher in the presence of anions, such as chloride), and (iv) no differentiation on the Fe/Ni ratio, the extraction rate or on the leaching mechanism of the ore leaching is noticed when using different sources of water.

BHP Billiton wrote the patent US 6.312.500 – 06/november/2001 - named “Heap leaching of nickel containing ore.” This invention describes an economical method of heap leaching of nickel-containing laterite ores that have tangible clay content. In this invention, clay is considered as material less then 325 mesh Tyler (<44 microns) in size. The method considered in this invention includes reducing the ore to an average particle size of less than about one
inch. Then, the reduced ore is pre-treated with concentrated acid to physically and/or chemically agglomerate the fine clay particles. It is believed that the acid breaks down the clay minerals and metal silicates and solubilize the silica gel so the reprecipitated silica and the precipitated metal sulfates will act as binding agents, forming strong pellets. These pellets allow a high percolation flux rate into the heap. The amount of acid used to agglomerate the pellets is usually the amount required to neutralize the readily available MgO. In general, it ranges from approximately 20 to 100 kg of acid per ton of ore, having a concentration of at least 100g/l acid. The agglomerated pellets are generally cured for an amount of time that may vary from an hour to several days, depending on the completion of the “chemical bonding process” that takes place. Preferably, the curing time ranges from about 4 to about 10 hours. During the curing process, excess water evaporates and some of the water is incorporated as crystal water causing the reprecipitation of a magnesium-iron-silica gel “glue” that acts as a chemical bonding agent between the solid particles. The curing time will be a function of the quantity of excess, or free, water that needs to be removed from the “glue” phase. The heap made with these pellets usually has an apparent density of 0.9 to 1.1 g/cm³ and a porosity of 0.3 to 0.6. This allows the use of a high percolation flux (from 10 to 100 l/hr/m²), thereby resulting in significantly increasing leaching kinetics.

Optimum recovery of nickel is observed when approximately 300 to 700 kg sulfuric acid per ton of dry ore is consumed. The acid is typically added to the heap recycled solution with an acidity of about 2 g/l to about 500 g/l, preferably at least about 10 g/l, and likely from approximately 20 g/l to 100 g/l to minimize the overall acid consumption. A low acid gradient throughout the heap increases the selectivity of nickel leaching compared to iron. Iron, when present as ferric iron, precipitates as hydroxides or basic sulfates within the heap if the pH of leachate is over 3. As a result of MgO content, saprolite ores are very basic and are considered to have acid neutralization potential. If this ore is mixed with nickel containing acidic solution, (pH from 0.5 to 2), or if such a solution is passed through an ore pile of saprolite ore, the residual acid will be
neutralized, raising the pH a range of 3 to 4. If there is any ferric iron present in
the acid nickel solution, it will precipitate as the pH is raised, since ferric
hydroxide starts to precipitate at a pH of about 2.5.

This invention considers forming at least one heap from a mixture of the ore and
concentrated sulfuric acid, at least one heap of saprolite ore to be used as an
iron precipitation reactor and at least one heap that includes bedrock material
for final pH control and iron removal. To ensure maximum recovery of nickel at
maximum efficiency, each heap is leached in the following manner. The first
one is leached with fresh lixiviant having a concentration of at least 10gpl. The
leach liquor is collected and analyzed for the presence of free acid. If the pH is
greater than 2 or the free sulfuric acid content is less than about 1g/l then the
liquor can be directed to a nickel recovery operation. In practice, the majority of
the leach liquor is directed to heap 2. In addition to applying the leach liquor
from heap 1, fresh sulfuric acid (having a concentration of at least 10gpl) is
applied to heap 2. The leach liquor from heap 2 is analyzed for the presence of
free acid. The process may continue depending on the number of heaps.

Another BHP Billiton patent (AU 2003/9003632 – 14/july/2003) (WO
2005/005671 – 20/january/2005) is named “Process for recovery of nickel and
cobalt by heap leaching of low-grade nickel or cobalt containing material”. This
invention refers to a method for improving the recovery of nickel and cobalt from
laterite ores, by beneficiating and separating the course and fine material. The
beneficiation step goals to separate the ore into an upgraded ore fraction and a
coarse, silicious low grade tailings fraction which is substantially free from fines
and clay materials. The upgraded ore can be treated by high pressure acid
leaching process or atmospheric leaching process, while the low-grade material
is treated in a heap leaching. The inventors have found that where the low
grade siliceous tailings are substantially free of fines and clay materials, they
have a high permeability which makes them suitable for heap leaching without
the requirement for the pelletisation step, as reported in US patent 6.312.500.
Another way of treating the coarse material is to separate the saprolite and
limonite ores in two separates heaps. Forming separates heaps has the advantage that leaching the limonite provides for maximum nickel recovery and the saprolite leaching provides for acid neutralization and iron removal. The resulting heap leachate may be blended with the leach solution from the acid leaching of the upgraded fraction to produce a blended leachate. As an alternative, the resultant leachate from the low grade ore fractions may be further processed for nickel and cobalt recovery independently from the leach solution from the upgraded ore fraction. The low grades of nickel and cobalt, in the tailings fraction may be approximately 0.3% to 0.7% nickel and 0.01% to 0.03% cobalt.

The other BHP Billiton patent (AU 2005/902462 – 13/may/2005) (WO 2006/119559 – 16/november/2006) is named “An improved process for heap leaching of nickeliferous oxidic ores”. This invention is related to a process of extracting nickel and cobalt from laterite ores by heap leaching of the ore with a leach liquor prepared from saline or hypersaline water. In many areas of the world where laterite deposits occurs, good quality water is in short supply and is a costly resource. Once hydrometallurgical processes require large quantity of water, this invention is related to the usage of hypersaline underground water or seawater. The hypersaline water useful in the process is generally from surface and/or underground brines, concentrated seawater or the effluent from desalination processes. Typically water sourced from brines has total solids higher than seawater and the inventors have found that hypersaline waters with total solids in excess provide improved nickel and cobalt recovery in heap leaching processes. Total dissolved solids that are useful in the process are in the range of 5-200 g/l, most preferably from 50-150 g/l. Saline and/or hypersaline water used for the leach solution is also used for the agglomeration process.

European Nickel PLS has applied for the patent WO 2004/031422 – 15/april/2004 – named “Heap leaching base metals from oxides ores”. It is related to the method of heap leaching, considering irrigating the heap with a
leach solution containing both sulfuric acid and dissolved sulfur dioxide. The sulfur dioxide is preferably provided in the form of either sodium metabisulphite (SMS) or burned elemental sulfur. Preferably, SMS is added regularly to the leach solution flowing throughout the leach. Its concentration in the leach solution is preferably maintained at or about 5g/l. The sulfuric acid solution is provided preferably at a concentration of 75g/l. The usage of SMS in addition to sulfuric acid results in increased recoveries of both nickel and cobalt, with the most marked improvement being found with respect to cobalt recovery. The mechanism by which the SMS contributes to increase metal recovery is largely uncertain.

European Nickel PLS has applied for another patent, WO 2006/000020 – 05/january/2006 – named “Improved leaching of base metals”. The improvement of this patent, compared to WO 2004/031422 is a step of separating the ore to be leached into fine and coarse fractions, and treat them separately. The coarse material is leached with the leach agent and the fine ore fraction or part of it is used to neutralize or partially neutralize the pregnant leach solution produced in the coarse material heap. In this invention, the ore may be agglomerated with sulfuric acid prior to forming the heap and SMS or sulfur dioxide is added to a sulfuric acid based leach solution.

Companhia Vale do Rio Doce, now Vale, patent (PI 06 05643v1 – 30/october/2006) (EP 1 790 739 A1 – 24/november/2006) is named “Process for extraction of nickel, cobalt, and other base metals from laterite ores by using heap leaching and product containing nickel, cobalt, and other metals from laterite ores.” It describes a method to remove nickel and cobalt, using heap leaching, comprising of the following steps: i) crushing, ii) agglomeration and iii) heap leaching. Crushing has as many stages as necessary to reach the desired particle size. In the agglomeration step, sulfuric acid is added in relation to the amount of fines in the ore and to the amount of elements that has a fast kinetic with the acid (e.g. magnesium, aluminum, calcium). Water is added enough to reach the desired moisture content for the agglomeration
The heap leaching step comprises of dynamic heaps, counter-current operation, with at least 2 stages, where a heap section or a heap represents each stage. The two phases are the ore and the leaching solution that are fed from opposite sides in the circuit. New leaching solution, containing from 50 to 200g/l sulfuric acid, feeds the last heap (e.g. third heap, for a system with three stages). The solution from the third stage goes to a pond and feeds the second stages. The final solution, from the first heap, is called PLS (Pregnant leaching solution) and has acidity of 10-30 g/L. This results in a reduced acid consumption, reaching 350 kg/t of ore, depending on the mineralogy, and in a reduced leaching cycle.

Murrin Murrin Operations PTY LTD has applied for the patent WO 2007/016737 – 15/february/2007 – named “Hydrometallurgical method for the extraction of nickel and cobalt from laterite ores”. It relates to a method for the extraction of nickel and cobalt from laterite ores, considering the steps: (i) beneficiation of a run of mine laterite ore, (ii) pressure acid leaching of the fine ore fraction with sulfuric acid solution, (iii) agglomeration of at least a fraction of the coarse ore fraction by wetting the ore to 20% to 30% moisture, (iv) forming at least one heap, (v) passing the leach solution from the pressure acid leaching step to a solid/liquid separation, and a portion of the liquor is passed to further processing for metals recovery and a further portion of the liquor is passed directly to the heap formed in step (ii), (vi) leaching the coarse material in the heaps with liquor separated in step (v), wherein the leaching agent is the free acid remaining in the acid pressure leaching discharge and (vii) passing pregnant leach solution from the heap to metal recovery. The effect of passing the liquor from the pressure acid leaching, containing about 10-50 g/L of sulfuric acid, to the heap is both the recovery of nickel an cobalt from the coarse material and the neutralization of free acid. The leach solution from the heap contains about 0-20g/L free acid, about 3-5g/L nickel and 1-15 g/L of iron.
APPENDIX B

Table B.1: Firm, Probable and Possible Nickel Projects, 2006 – 2015
<table>
<thead>
<tr>
<th>Country</th>
<th>Company</th>
<th>Product</th>
<th>Process Tech'gy</th>
<th>Capacity Increase tpy</th>
<th>Reserves/ Resources in tonnes</th>
<th>Grade %Ni</th>
<th>Category</th>
<th>Timing</th>
<th>Status</th>
<th>Capex US$/tonne</th>
<th>Capex US$/lb Ni</th>
</tr>
</thead>
<tbody>
<tr>
<td>Australia</td>
<td>Allegiance Mining</td>
<td>Avebury</td>
<td>G S Flotation</td>
<td>4.4</td>
<td>1.16%</td>
<td>PP</td>
<td>Q3 2007</td>
<td>Firm</td>
<td>57</td>
<td>3.2</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Australian Mines</td>
<td>Marriott</td>
<td>G S Flotation</td>
<td>0.6</td>
<td>1.40%</td>
<td>Inf.</td>
<td>After 2010</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>BHP Billiton</td>
<td>Mt Keith</td>
<td>Conc B S LPL</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>600</td>
<td>10.9</td>
</tr>
<tr>
<td>Australia</td>
<td>BHP Billiton</td>
<td>Ravensthorpe</td>
<td>G L PAL</td>
<td>263.3</td>
<td>0.65%</td>
<td>-</td>
<td>Q1 2008</td>
<td>Firm</td>
<td>2200</td>
<td>22.2</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>BHP Billiton</td>
<td>Yakabindie</td>
<td>Conc G S Flotation</td>
<td>289.0</td>
<td>0.58%</td>
<td>-</td>
<td>2012</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Compass Resources NL</td>
<td>Browns</td>
<td>G L</td>
<td>2.8</td>
<td>0.17%</td>
<td>-</td>
<td>2007</td>
<td>Possible</td>
<td>24.8</td>
<td>15.0</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Consolidated Minerals</td>
<td>Widgiemooltha (N&amp;C)</td>
<td>Conc B S Flotation</td>
<td>10.4</td>
<td>1.49%</td>
<td>-</td>
<td>2010</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Fox Resources</td>
<td>Mt Sholl B2</td>
<td>Conc G S Flotation</td>
<td>1.0</td>
<td>0.70%</td>
<td>-</td>
<td>2008</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Gladstone Pacific Nickel</td>
<td>Marlborough Stage 1</td>
<td>Metal G L PAL</td>
<td>125.0</td>
<td>0.80%</td>
<td>-</td>
<td>After 2010</td>
<td>Possible</td>
<td>900</td>
<td>11.3</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>GME Resources</td>
<td>NWest</td>
<td>G L Heap Leach</td>
<td>48.8</td>
<td>1.25%</td>
<td>M, Inf.</td>
<td>After 2010</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Heron Resources</td>
<td>Jump-Up Dam</td>
<td>Inter'diates G L Heap Leach</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>After 2010</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Heron Resources / Inco</td>
<td>Kalgoorlie Nickel</td>
<td>Metal G L PAL</td>
<td>324.8</td>
<td>0.83%</td>
<td>Ind.</td>
<td>After 2012</td>
<td>Possible</td>
<td>1400</td>
<td>12.7</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Ivanhoe Nickel &amp; Platinum</td>
<td>Syerston</td>
<td>Metal G L PAL</td>
<td>96.0</td>
<td>0.69%</td>
<td>-</td>
<td>After 2010</td>
<td>Possible</td>
<td>640</td>
<td>16.6</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
<td>Jervois Resources</td>
<td>Young</td>
<td>Metal G L ACLP</td>
<td>167.0</td>
<td>0.72%</td>
<td>-</td>
<td>After 2010</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
<td></td>
</tr>
<tr>
<td>Australia</td>
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<td>Q1 2008</td>
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<td>-</td>
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<td>4.80%</td>
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<td>Q4 2006</td>
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<td>Possible</td>
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<td>Diggers South</td>
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<td>1.54%</td>
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<td>Product</td>
<td>B/G</td>
<td>Type of Orebody</td>
<td>Process Tech'gy</td>
<td>Reserves/Resources in tonnes</td>
<td>Grade</td>
<td>Ni</td>
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<td>Xstrata Nickel</td>
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<td>FeNi</td>
<td>G</td>
<td>L</td>
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<td>L</td>
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<td>Inf.</td>
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<td>14.784</td>
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<td>S</td>
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<td>G</td>
<td>S</td>
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<td>L</td>
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<td>S</td>
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<td>S</td>
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<td>S</td>
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<td>Nil Hyd.</td>
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<td>PAL</td>
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<td>1.3</td>
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<td>Inf.</td>
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<td>Jilin</td>
<td>15</td>
<td>Metal</td>
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<td>S</td>
<td>Flot/Sim/Ref</td>
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<td>L</td>
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<td>G</td>
<td>L</td>
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<td>1.00%</td>
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<td>After 2011</td>
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<td>Las Camarinas</td>
<td>23</td>
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<td>G</td>
<td>L</td>
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<td>0.40%</td>
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<td>G</td>
<td>L</td>
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<td>22</td>
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<td>L</td>
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<td>G</td>
<td>L</td>
<td>PAL</td>
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<td>1.33%</td>
<td>M</td>
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<td>Chevchenko</td>
<td>21</td>
<td>FeNi</td>
<td>G</td>
<td>L</td>
<td>Smelting</td>
<td>164.4</td>
<td>0.0079</td>
<td>PP</td>
<td>2008</td>
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</table>
### Firm, Probable and Possible Nickel Projects, 2006Q4 - 2015

<table>
<thead>
<tr>
<th>Country</th>
<th>Company</th>
<th>Project</th>
<th>Product</th>
<th>Type of Orebody</th>
<th>Process Tech'gy</th>
<th>Reserves/ Resources in tonnes</th>
<th>Grade %Ni</th>
<th>Category</th>
<th>Timing</th>
<th>Status</th>
<th>Capex US$m</th>
<th>Capex US$/lb Ni</th>
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</thead>
<tbody>
<tr>
<td>Madagascar</td>
<td>Dynatec / Sumitomo / KORES</td>
<td>Ambatovy</td>
<td>60 Metal G L PAL</td>
<td>PAL</td>
<td>125.0</td>
<td>1.04%</td>
<td>PP</td>
<td>2010</td>
<td>Probable</td>
<td>2800</td>
<td>21.2</td>
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<tr>
<td>Myanmar</td>
<td>CNMC / Myanmar Govt.</td>
<td>Tagaung Taung</td>
<td>32 FeNi G L Smelting</td>
<td>Smelting</td>
<td>40.0</td>
<td>2.00%</td>
<td>-</td>
<td>After 2010</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
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<tr>
<td>New Caledonia</td>
<td>Angloky Minerals</td>
<td>Nakalay-Boagila</td>
<td>52 Metal G L PAL</td>
<td>PAL</td>
<td>228.0</td>
<td>0.0149%</td>
<td>-</td>
<td>After 2012</td>
<td>Possible</td>
<td>-</td>
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<tr>
<td>New Caledonia</td>
<td>Eramet / SLN / BHP Billiton</td>
<td>Northern Province</td>
<td>25 Metal G L Caron</td>
<td>Caron</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>After 2015</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
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<tr>
<td>New Caledonia</td>
<td>CVRD - Inco</td>
<td>Goro</td>
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<td>200.0</td>
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<td>Prov. 2008</td>
<td>Firm</td>
<td>2750</td>
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<td>Cap Bocage</td>
<td>28 Metal G L PAL</td>
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<td>1.30%</td>
<td>M.ind. After 2012</td>
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<td>-</td>
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<tr>
<td>New Caledonia</td>
<td>SLN/Eramet</td>
<td>Doniambo</td>
<td>15 Mattle B L Smelting</td>
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<td>-</td>
<td>-</td>
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<td>New Caledonia</td>
<td>SMSF/Xstrata Nickel</td>
<td>Koniambo</td>
<td>60 FeNi G L Smelting</td>
<td>Smelting</td>
<td>142.1</td>
<td>2.13%</td>
<td>M.ind. 2010</td>
<td>Possible</td>
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<tr>
<td>Papua N. Guinea</td>
<td>CMCC/Highlands Pacific</td>
<td>Ramu River</td>
<td>33 Metal G L PAL</td>
<td>PAL</td>
<td>75.7</td>
<td>0.91%</td>
<td>PP 2010</td>
<td>Firm</td>
<td>838</td>
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<td>Papua N. Guinea</td>
<td>Nuiglin Ni (Resource Mining)</td>
<td>Woon Gap</td>
<td>46 FeNi G L Smelting</td>
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<td>120.0</td>
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<td>Possible</td>
<td>-</td>
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<tr>
<td>Philippines</td>
<td>Benguet Group</td>
<td>Santa Cruz</td>
<td>8.021664 - G L -</td>
<td>-</td>
<td>11.9</td>
<td>1.68%</td>
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<td>After 2010</td>
<td>Possible</td>
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<td>Philippines</td>
<td>BHP Billiton</td>
<td>Adlay</td>
<td>5 Ore G L Mining</td>
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<td>5.8</td>
<td>1.61%</td>
<td>-</td>
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<td>Sablayan</td>
<td>40 Metal G L PAL</td>
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<td>-</td>
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<td>After 2012</td>
<td>Possible</td>
<td>665</td>
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<tr>
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<td>Phlnico/Inchuan/Bac Steel</td>
<td>Nonoc</td>
<td>40 Ni Oxide G L PAL</td>
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<td>1.10%</td>
<td>-</td>
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<tr>
<td>Philippines</td>
<td>Sumitomo MM</td>
<td>Rio Tuba II</td>
<td>10 Ni-Co Sul. G L PAL</td>
<td>PAL</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>Q2 2009</td>
<td>Firm</td>
<td>285</td>
<td>12.9</td>
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<tr>
<td>Philippines</td>
<td>Toledo / Atlas / Involdika</td>
<td>Benong</td>
<td>10 Ore G L Mining</td>
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<td>275.0</td>
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<td>Celestial (Palawan)</td>
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<td>Amur Minerals Corp.</td>
<td>Kun-Manie</td>
<td>8 Conc G G S Floation</td>
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<td>46.1</td>
<td>0.45%</td>
<td>-</td>
<td>After 2010</td>
<td>Possible</td>
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<td>Russia</td>
<td>Norilsk Nickel</td>
<td>Talymr and Koia</td>
<td>17 Metal B S Flo/Sm/Ref</td>
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<td>-</td>
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<td>Serbia</td>
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<td>Feronikeli</td>
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<td>-</td>
<td>-</td>
<td>2010</td>
<td>Firm</td>
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<td>South Africa</td>
<td>ARMLonene</td>
<td>Nkomati I</td>
<td>5 Conc B S Floation</td>
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<tr>
<td>South Africa</td>
<td>ARMLonene</td>
<td>Nkomati II</td>
<td>24 Metal G S Activox</td>
<td>Activox</td>
<td>139.0</td>
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<td>-</td>
<td>2010</td>
<td>Probable</td>
<td>620</td>
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<td>Possible</td>
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<td>South Africa</td>
<td>Ridge Mining / Anglo Platinum</td>
<td>Shaba's Ridge</td>
<td>24 Matte G G S Flo/Sm</td>
<td>Flo/Sm</td>
<td>775.0</td>
<td>0.18%</td>
<td>M, Ind, Inf. Mid-2008</td>
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<tr>
<td>South Korea</td>
<td>Nickel Mining (Posco/SMSP)</td>
<td>Gwangyang</td>
<td>30 FeNi G L Smelting</td>
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<td>-</td>
<td>-</td>
<td>-</td>
<td>2009</td>
<td>Firm</td>
<td>352</td>
<td>5.3</td>
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<tr>
<td>Tanzania</td>
<td>Barrick / Xanrata Nickel</td>
<td>Kabanga</td>
<td>35 Conc G G S Floation</td>
<td>Floation</td>
<td>26.4</td>
<td>2.60%</td>
<td>Inf. 2009</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
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<tr>
<td>Turkey</td>
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<td>Catalag</td>
<td>20 Ni-Co Hyd. G G S Heap Leach</td>
<td>Heap Leach</td>
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<td>Northmet</td>
<td>30 Conc G G S Floation</td>
<td>Floation</td>
<td>808.0</td>
<td>0.11%</td>
<td>-</td>
<td>2008</td>
<td>Possible</td>
<td>250</td>
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<tr>
<td>USA</td>
<td>Rio Tinto</td>
<td>Eagle</td>
<td>24 Conc G S Floation</td>
<td>Floation</td>
<td>5.0</td>
<td>3.70%</td>
<td>-</td>
<td>H2 2008</td>
<td>Possible</td>
<td>-</td>
<td>-</td>
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<tr>
<td>USA</td>
<td>Teck Cominco</td>
<td>Mesaba</td>
<td>20 Ni-Co Hyd. G G G LPL</td>
<td>LPL</td>
<td>1000.0</td>
<td>0.14%</td>
<td>-</td>
<td>After 2010</td>
<td>Possible</td>
<td>500</td>
<td>11.3399256</td>
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<tr>
<td>Vietnam</td>
<td>Asian Mineral Resources</td>
<td>Ban Phuc</td>
<td>4 Conc G G S Floation</td>
<td>Floation</td>
<td>1.0</td>
<td>2.40%</td>
<td>PP 2008</td>
<td>Possible</td>
<td>33.6</td>
<td>4.1261615</td>
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<tr>
<td>Zambia</td>
<td>Albidon</td>
<td>Enterprise</td>
<td>8 Conc G G S Floation</td>
<td>Floation</td>
<td>6.9</td>
<td>1.40%</td>
<td>Inf.</td>
<td>Q2 2008</td>
<td>Probable</td>
<td>65</td>
<td>3.8</td>
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<td>Zimbabwe</td>
<td>Bindura</td>
<td>Hunters Road</td>
<td>8 Conc G G S Floation</td>
<td>Floation</td>
<td>-</td>
<td>-</td>
<td>-</td>
<td>Possible</td>
<td>100</td>
<td>5.7</td>
<td></td>
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</tr>
</tbody>
</table>

**Notes:**
1. **G** = Greenfield project, **B** = Brownfield expansion, **S** = Sulphide, **L** = Laterite.
2. **LPL** = Low pressure leach, **PAL** = Pressure acid leach, **ACLP** = Atmospheric chloride leach process.
3. **PP** = Proven & probable, **M** = Measured, **Ind.** = Indicated, **Inf.** = Inferred, **Prov.** = Proven, **Prob.** = Probable

* São João do Piauí project, now called Niquel do Piauí, is a project that is now being studied by Vale under HL process technology
APPENDIX C – FeNi model data

Coal Milling

Figure C1 shows MetSim flowsheet for the coal milling. It will present different colors for different streams state, that is, all the aqueous streams are in blue, solid in black, organic in dark green, metal in red, matte in dark blue, slag in orange and gas in pink. The controllers numbers will be shown in each figure.

Reactions:

In the Kiln roller mill air heater (unit operation 12)  
2 diesel + 2161 O$_2$(g) = 2330 CO(g) 1988 H$_2$O(g) + 2 SO$_2$(g)  
2 CO(g) + 1 O$_2$(g) = 2 CO$_2$(g)  
Extent*  
100%

Second chamber (unit operation 13)  
2 diesel + 2161 O$_2$(g) = 2330 CO(g) 1988 H$_2$O(g) + 2 SO$_2$(g)  
2 CO(g) + 1 O$_2$(g) = 2 CO$_2$(g)  
Extent*  
100%

* The extent of reaction is fixed based on Vale Inco projects.

Controllers:

Controls are used to simulate process control loops, where some constrains are applied. These controls function in a similar way to the ones in the process plant. For this area, they are:

1010 – Diesel to the kiln mill
Based on the amount of coal that feeds the mill.

1020 – Air to the kiln air heater
Based on the amount of diesel that feeds the mill.

1030 – Dilution air
Air is added in order to get 170°C in the mill air feed.

**Ore drying**

Figure C.2 shows the MetSim flowsheet for this area.

### Reactions:

<table>
<thead>
<tr>
<th>Dryer combustion Chamber (unit operation 22)</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>$\text{H}<em>2\text{O}</em>{(aq)} = \text{H}<em>2\text{O}</em>{(g)}$</td>
<td>100%</td>
</tr>
<tr>
<td>$2\text{C}_{(s)} + 1 \text{O}<em>2(g) = 2 \text{CO}</em>{(g)}$</td>
<td>100%</td>
</tr>
<tr>
<td>$2 \text{C}<em>6\text{H}</em>{14}(s) + 13 \text{O}<em>2(g) = 12 \text{CO}</em>{(g)} + 14 \text{H}<em>2\text{O}</em>{(g)}$</td>
<td>100%</td>
</tr>
<tr>
<td>$4 \text{C}_3\text{H}<em>7\text{N}</em>{(s)} + 15 \text{O}<em>2(g) = 20 \text{CO}</em>{(g)} + 10 \text{H}<em>2\text{O}</em>{(g)} + 2\text{N}_2(g)$</td>
<td>100%</td>
</tr>
<tr>
<td>$1 \text{C}_4\text{H}<em>8\text{S}</em>{(s)} + 4 \text{O}<em>2(g) = 4 \text{CO}</em>{(g)} + 2 \text{H}<em>2\text{O}</em>{(g)} + \text{SO}_2(g)$</td>
<td>100%</td>
</tr>
<tr>
<td>$1 \text{C}_6\text{H}<em>6\text{O}</em>{(s)} + 4 \text{O}<em>2(g) = 6 \text{CO}</em>{(g)} + 3 \text{H}<em>2\text{O}</em>{(g)}$</td>
<td>100%</td>
</tr>
<tr>
<td>$2 \text{CO}_{(g)} + 1 \text{O}_2(g) = 2 \text{CO}_2(g)$</td>
<td>100%</td>
</tr>
</tbody>
</table>

* The extent of reaction is fixed based on Vale Inco projects.

### Controllers:

2010 – Ore feed to the plant

Ore feed flowrate is back calculated based on the reserve, mine lifetime and moisture content in the ore. As explained in the beginning of section 5, three important inputs for all the models are: i) reserve (dry basis), ii) mine lifetime and iii) moisture of the deposit. They vary from deposit to deposit. Considering these variables, the feed flowrate to the plant, in t/h, is calculated, considering the equation (5.1).

2020 – Dryer primary combustion air

Based on the amount of fuel that feeds the dryer

2030 – Dryer secondary combustion air
Air is added in order to get 1100°C in the dryer air feed.

2040 – Dryer burner coal
Fuel is added to get the dryer running at 160°C

2050 – Dryer offgas infiltration air
Infiltration air is added proportionally to the amount of gas feeding the offgas system.

Ore reduction

Figure C.3 shows the MetSim flowsheet for this area and Figure C.4 shows the one for the dust conveying area.

Reactions:

**Kiln combustion Chamber (unit operation 35)**

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>(1 \text{ H}<em>2\text{O}</em>{(aq)} = 1 \text{ H}<em>2\text{O}</em>{(g)})</td>
<td>100%</td>
</tr>
<tr>
<td>(2 \text{ diesel} + 2161 \text{ O}<em>{2(g)} = 2330 \text{ CO}</em>{(g)} + 1988 \text{ H}<em>2\text{O}</em>{(g)} + 2 \text{ SO}_2(g))</td>
<td>100%</td>
</tr>
<tr>
<td>(2\text{C}<em>6\text{H}</em>{14(s)} + 13 \text{ O}<em>{2(g)} = 12 \text{ CO}</em>{(g)} + 14 \text{ H}<em>2\text{O}</em>{(g)})</td>
<td>100%</td>
</tr>
<tr>
<td>(4\text{C}<em>5\text{H}<em>5\text{N}</em>{(s)} + 15 \text{ O}</em>{2(g)} = 20 \text{ CO}_{(g)} + 10 \text{ H}<em>2\text{O}</em>{(g)} + 2\text{N}_2(g))</td>
<td>100%</td>
</tr>
<tr>
<td>(1\text{C}<em>3\text{H}<em>3\text{S}</em>{(s)} + 4 \text{ O}</em>{2(g)} = 4 \text{ CO}_{(g)} + 2 \text{ H}<em>2\text{O}</em>{(g)} + 1\text{SO}_2(g))</td>
<td>100%</td>
</tr>
<tr>
<td>(1\text{C}<em>6\text{H}<em>6\text{O}</em>{(s)} + 4 \text{ O}</em>{2(g)} = 6 \text{ CO}_{(g)} + 3 \text{ H}<em>2\text{O}</em>{(g)})</td>
<td>100%</td>
</tr>
<tr>
<td>(2\text{CO}<em>{(g)} + 1 \text{ O}</em>{2(g)} = 2 \text{ CO}_2(g))</td>
<td>100%</td>
</tr>
</tbody>
</table>

**Second chamber (unit operation 37)**

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>(1 \text{ H}<em>2\text{O}</em>{(aq)} = 1 \text{ H}<em>2\text{O}</em>{(g)})</td>
<td>100%</td>
</tr>
<tr>
<td>(2 \text{ diesel} + 2161 \text{ O}<em>{2(g)} = 2330 \text{ CO}</em>{(g)} + 1988 \text{ H}<em>2\text{O}</em>{(g)} + 2 \text{ SO}_2(g))</td>
<td>100%</td>
</tr>
<tr>
<td>(2\text{C}<em>6\text{H}</em>{14(s)} + 13 \text{ O}<em>{2(g)} = 12 \text{ CO}</em>{(g)} + 14 \text{ H}<em>2\text{O}</em>{(g)})</td>
<td>100%</td>
</tr>
<tr>
<td>(4\text{C}<em>5\text{H}<em>5\text{N}</em>{(s)} + 15 \text{ O}</em>{2(g)} = 20 \text{ CO}_{(g)} + 10 \text{ H}<em>2\text{O}</em>{(g)} + 2\text{N}_2(g))</td>
<td>100%</td>
</tr>
<tr>
<td>(1\text{C}<em>3\text{H}<em>3\text{S}</em>{(s)} + 4 \text{ O}</em>{2(g)} = 4 \text{ CO}_{(g)} + 2 \text{ H}<em>2\text{O}</em>{(g)} + 1\text{SO}_2(g))</td>
<td>100%</td>
</tr>
<tr>
<td>(1\text{C}<em>6\text{H}<em>6\text{O}</em>{(s)} + 4 \text{ O}</em>{2(g)} = 6 \text{ CO}_{(g)} + 3 \text{ H}<em>2\text{O}</em>{(g)})</td>
<td>100%</td>
</tr>
<tr>
<td>(2\text{CO}<em>{(g)} + 1 \text{ O}</em>{2(g)} = 2 \text{ CO}_2(g))</td>
<td>100%</td>
</tr>
</tbody>
</table>
Rotary Kiln (unit operation 44)

1 H₂O(aq) = 1 H₂O(g) 100%
1 C₆H₁₄(g) = 1 C₆H₁₄(g) 100%
1 C₆H₆N(s) = 1 C₆H₆N(g) 100%
1 C₄H₄S(s) = 1 C₄H₄S(g) 100%
1 C₆H₆O(s) = 1 C₆H₆O(g) 100%
2 C₆H₁₄(g) + 13 O₂(g) = 12 CO(g) + 14 H₂O(g) 100%
4 C₆H₆N(g) + 15 O₂(g) = 20 CO(g) + 10 H₂O(g) + 2N₂(g) 100%
1 C₄H₄S(g) + 4 O₂(g) = 4 CO(g) + 2 H₂O(g) + 1SO₂(g) 100%
1 C₆H₆O(g) + 4 O₂(g) = 6 CO(g) + 3 H₂O(g) 100%
2 C(g) + 1 O₂(g) = 2 CO(g) 10%
1 Mg(OH)₂(s) = 1 MgO(s) + 1 H₂O(g) **
1 NiO(s) + 1 CO₂(g) = 1 Ni(s) + 1 CO₂(g) **
1 Co₃O₄(s) + 4 CO₂(g) = 3 Co(s) + 4 CO₂(g) **
1 Fe₂O₃·H₂O(s) = 1 Fe₂O₃(s) + 1 H₂O(g) **
3 Fe₂O₃(s) + 1CO(g) = 2 Fe₃O₄(s) + 1 CO₂(g) **
1 Fe₂O₃(s) + 1CO(g) = 3 FeO(s) + 1 CO₂(g) **
1 FeO(s) + 1CO₂(g) = 1 Fe(s) + 1 CO₂(g) **
2 CO(g) + 1 O₂(g) = 2 CO₂(g) 35%

Off gas system (unit operation 46)

2 CO(g) + 1 O₂(g) = 2 CO₂(g) 100%

* The extent of reaction is fixed based on Vale Inco projects.
** The extent of reaction is based on temperature profile for each metal.

Controllers:

3010 – Kiln primary combustion air
Based on the amount of fuel that feeds the kiln

3020 – Kiln secondary combustion air
Based on the amount of fuel that feeds the kiln

3030 – Kiln reductant coal
Reductant is added to satisfy reduction requirements. It is controlled by the amount of C in the kiln discharge
3040 – Kiln burner coal
Fuel is added to get the solid discharge temperature of 900°C

3060 – Kiln tertiary air
Tertiary air is added to satisfy reduction conditions. It is controlled by the amount of O₂ in the kiln gas discharge

3070 – Kiln offgas infiltration air
Infiltration air is added proportionally to the amount of gas feeding the offgas system.

3080 – Air to the dust conveying
Air is added proportionally to the amount of dust.

Smelting process

Figure C.5 presents electric furnace area.

Reactions:

Electric Furnace (unit operation 58)

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>1 H₂Oₐq = 1 H₂Oₐg</td>
<td>100%</td>
</tr>
<tr>
<td>1 Al₂O₃(s) = 1 Al₂O₃(slag)</td>
<td>100%</td>
</tr>
<tr>
<td>1 SiO₂(s) = 1 SiO₂(slag)</td>
<td>100%</td>
</tr>
<tr>
<td>1 Mg(OH)₂(s) = 1 MgO (s) + 1 H₂Oₐq</td>
<td>100%</td>
</tr>
<tr>
<td>1 Fe₂O₃·H₂O(s) = 1 Fe₂O₃(s) + 1 H₂Oₐq</td>
<td>100%</td>
</tr>
<tr>
<td>1 MgO (s) + 1 SiO₂(slag) = 1 MgSiO₃(slag)</td>
<td>100%</td>
</tr>
<tr>
<td>1 Fe₉(m) = 1 Fe₉(m)</td>
<td>100%</td>
</tr>
<tr>
<td>3 Fe₂O₃(s) + 1 C(s) = 2 Fe₃O₄ (s) + 1 CO(g)</td>
<td>100%</td>
</tr>
<tr>
<td>1 Fe₃O₄(s) + 1 C(s) = 3 FeO (s) + 1 CO₂(g)</td>
<td>100%</td>
</tr>
<tr>
<td>1 Ni(s) = 1 Ni(m)</td>
<td>100%</td>
</tr>
<tr>
<td>2 NiO(s) + 1 C(s) = 2 Ni(m) + 1 CO₂(g)</td>
<td>Controlled</td>
</tr>
<tr>
<td>1 FeO₉(s) + 1 C(s) = 1 Fe₉(m) + 1 CO₂(g)</td>
<td>Controlled</td>
</tr>
<tr>
<td>2 FeO₉(s) + 1 SiO₂(slag) = 1 Fe₂SiO₄(slag)</td>
<td>100%</td>
</tr>
</tbody>
</table>
$1 \text{Co(s)} = 1 \text{Co(m)}$ 100%

$1 \text{Co}_3\text{O}_4(s) + 2 \text{C(s)} = 3 \text{Co(m)} + 2 \text{CO}_2(g)$ Controlled

$2 \text{Cr}_2\text{O}_3(s) + 3 \text{O}_2(g) = 4 \text{CrO}_3(\text{slag})$ 2%

* The extent of reaction is fixed based on Vale Inco projects. The reaction in which the extent is written controlled means that there are some constrains applied. The model, using controllers, calculates the extent.

**Controllers:**

5010 – Electric Furnace power
Power is calculated based on the furnace heat balance, in order to get the desired slag liquidus temperature.

5020 – Controls FeO reduction reaction
The reaction is controlled to get 25% nickel in the matte.

5030 – Controls Ni partition coefficient. Partition coefficient is the ration between the amount of metal in the product by the amount in the slag.
Controls NiO reduction extent of reaction in order to active the desired Ni partition coefficient

5040 – Controls Co partition coefficient
Controls Co reduction extent of reaction in order to active the desired Co partition coefficient

5050 – Controls carbon electrodes feed
Carbon electrodes are controlled to get the desired C in the furnace feed.

**Off-gas system/ Reagents/ Final discharge**

Figure C.6 presents off-gas system area, while Figure C.7 and C.8 present reagents and final discharge areas.
Reactions:

**Furnace Offgas system (unit operation 61)**

- \[ C(s) + O_2(g) = CO_2(g) \] (100%)
- \[ Al_2O_3(slqg) = Al_2O_3 (s) \] (100%)
- \[ Fe_2SiO_4(slqg) = 2 FeO (s) + SiO_2(s) \] (100%)
- \[ MgSiO_3(slqg) = MgO (s) + SiO_2(s) \] (100%)

* The extent of reaction is fixed based on Vale Inco projects.

Controllers:

5060 – Total air infiltration in the furnace area

Infiltration air is added proportionally to the amount of gas feeding the off gas system.
Figure C.1: Coal milling area flowsheet.
Figure C.2: Ore drying area flowsheet.
Figure C.3: Kiln area flowsheet.
Figure C.4: Dust Conveying area
Figure C.6: Off-gas system area
Figure C.7: Reagents area flowsheet.
Figure C.8: Final discharge area flowsheet
APPENDIX D – HPAL model data

Beneficiation/ Acid Leaching

Figure D.1 presents MetSim model flowsheet for the beneficiation and acid leaching area.

Reactions:

<table>
<thead>
<tr>
<th>Reactions</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Autoclave (unit operation 8)</td>
<td></td>
</tr>
<tr>
<td>$1\text{Ni}_2\text{O}_3 + 1\text{H}_2\text{SO}_4(\text{aq}) = 1\text{NiSO}_4(\text{aq}) + 1\text{H}_2\text{O}()$</td>
<td>96%</td>
</tr>
<tr>
<td>$1\text{Co}_2\text{O}_3 + 1\text{H}_2\text{SO}_4(\text{aq}) = 1\text{CoSO}_4(\text{aq}) + 1\text{H}_2\text{O}()$</td>
<td>96%</td>
</tr>
<tr>
<td>$1\text{Mn}_2\text{O}_3 + 1\text{H}_2\text{SO}_4(\text{aq}) = 1\text{MnSO}_4(\text{aq}) + 1\text{H}_2\text{O}()$</td>
<td>96%</td>
</tr>
<tr>
<td>$1\text{Mg}_2\text{O}_3 + 1\text{H}_2\text{SO}_4(\text{aq}) = 1\text{MgSO}_4(\text{aq}) + 1\text{H}_2\text{O}()$</td>
<td>92%</td>
</tr>
<tr>
<td>$1\text{Fe}_2\text{O}_3 + 3\text{H}_2\text{SO}_4(\text{aq}) = 1\text{FeSO}_4(\text{aq}) + 3\text{H}_2\text{O}()$</td>
<td>100%</td>
</tr>
<tr>
<td>$1\text{Al}_2\text{O}_3 + 3\text{H}_2\text{SO}_4(\text{aq}) = 1\text{Al}_2(\text{SO}_4)_3(\text{aq}) + 3\text{H}_2\text{O}()$</td>
<td>35%</td>
</tr>
<tr>
<td>$1\text{Al}_2\text{O}_3 + 4\text{H}_2\text{SO}_4(\text{aq}) + 5\text{H}_2\text{O}() = 2\text{Al}_3(\text{SO}_4)_2\cdot \text{H}_2\text{O} (\text{OH})_6(\text{s})$</td>
<td>100%</td>
</tr>
<tr>
<td>$2\text{FeOOH}(\text{s}) + 3\text{H}_2\text{SO}_4(\text{aq}) = \text{Fe}_2(\text{SO}_4)_3(\text{s}) + 4\text{H}_2\text{O}()$</td>
<td>100%</td>
</tr>
<tr>
<td>$\text{Fe}_2(\text{SO}_4)_3(\text{aq}) + 3\text{H}_2\text{O}() = 3\text{H}_2\text{SO}_4(\text{aq}) + 1\text{Fe}_2\text{O}_3(\text{s})$</td>
<td>92%</td>
</tr>
</tbody>
</table>

* The extent of reaction is fixed based on Vale Inco projects.

Controllers:

20 – Ore feed to the plant
Ore feed flowrate is back calculated based on the reserve, mine lifetime and moisture content in the ore. As explained in appendix C, equation (5.1) is used in this controller.

50 – Water to the beneficiation circuit
Water is used in the process for the scrubber and screening step. After these steps, the slurry feeds a thickener. The underflow is sent to the autoclave and the overflow is used as back-up water to the beneficiation circuit (recirculation load). An important variable in this step is the beneficiation thickener underflow.
% solids. As a result, in the model, water is added to achieve 40% w/w pulp to be fed to the autoclave. The % moisture in the pulp to be fed to the autoclave is an important variable and should be confirmed with test work.

60 – Flocculant addition in the feed thickener
Flocculant is added proportionally to the amount of solids in the feed.

70 – Acid addition to the autoclave
Calculated in order to get 50g/L free acid in the autoclave discharge.

First neutralization/CCD/Second neutralization

Figures D.2 to D.4 present MetSim model flowsheet for the first neutralization, CCD and second neutralization.

Reactions:

<table>
<thead>
<tr>
<th>Releach unit operation (unit operation 12)</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ni(OH)$_2$(s) + H$_2$SO$_4$ = NiSO$_4$ + H$_2$O</td>
<td>100%</td>
</tr>
<tr>
<td>Co(OH)$_2$(s) + H$_2$SO$_4$ = CoSO$_4$ + H$_2$O</td>
<td>100%</td>
</tr>
</tbody>
</table>

Neutralization 1 (unit operation 14)

Fe$_2$(SO$_4$)$_3$(aq) + 3 CaCO$_3$(s) + 9 H$_2$O(l) = 3 CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) + 2 Fe(OH)$_3$(s) 90%
Al$_2$(SO$_4$)$_3$(aq) + 3 CaCO$_3$(s) + 9 H$_2$O(l) = 3 CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) + 2 Al(OH)$_3$(s) 10%
Cr$_2$(SO$_4$)$_3$(aq) + 3 CaCO$_3$(s) + 9 H$_2$O(l) = 3 CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) + 2 Cr(OH)$_3$(s) 50%
H$_2$SO$_4$(aq) + CaCO$_3$(s) + H$_2$O(l) = CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) 100%

Neutralization 2 (unit operation 25)

Fe$_2$(SO$_4$)$_3$(aq) + 3 CaCO$_3$(s) + 9 H$_2$O(l) = 3 CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) + 2 Fe(OH)$_3$(s) 100%
FeSO$_4$(aq) + 1 CaCO$_3$(s) + 3 H$_2$O(l) = 1 CO$_2$(g) + 1CaSO$_4$.2H$_2$O(s) + 1 Fe(OH)$_2$(s) 100%
Al$_2$(SO$_4$)$_3$(aq) + 3 CaCO$_3$(s) + 9 H$_2$O(l) = 3 CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) + 2 Al(OH)$_3$(s) 99%
Cr$_2$(SO$_4$)$_3$(aq) + 3 CaCO$_3$(s) + 9 H$_2$O(l) = 3 CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) + 2 Cr(OH)$_3$(s) 90%
NiSO$_4$(aq) + 1 CaCO$_3$(s) + 3 H$_2$O(l) = 1 CO$_2$(g) + 1CaSO$_4$.2H$_2$O(s) + 1 Ni(OH)$_2$(s) 10%
H$_2$SO$_4$(aq) + CaCO$_3$(s) + H$_2$O(l) = CO$_2$(g) + 3CaSO$_4$.2H$_2$O(s) 100%

* The extent of reaction is fixed based on Vale Inco projects.
Controllers:

110 – Limestone to the neutralization 1 circuit
Limestone is added in order to reach pH 2.5 in the final tank.

210 – Water to the CCD
Water is added to get 2.3 ratio of mass of water/mass of dry ore

220 – Sulfuric Acid to the CCD water
Acid is added to get pH = 4 in the CCD washing water

230 – 270 - Flocculant addition in the CCD.
Flocculant is added proportionally to the amount of solids feed.

310 – Recycle seed to the pre-mix tank second neutralization step
75% of the thickener underflow is recycled to the pre-mix to provide seeds for
the neutralization

320 – Limestone to the second neutralization circuit
Limestone is added in order to reach pH = 4.5 in the final tank.

330 - Flocculant addition in the thickener
Flocculant is added proportionally to the amount of solids formed in the
neutralization step.

**Mixed Hydroxide precipitate 1 and 2**

Figures D.5 and D.6 present MetSim model flowsheet for the MHP1 and MHP 2
circuits.
Reactions:

**MHP 1 (unit operation 34)**

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>CuSO(_4)(aq) + MgO(s) + H(_2)O = MgSO(_4)(aq) + Cu(OH)(_2)(s)</td>
<td>100%</td>
</tr>
<tr>
<td>Al(_2)(SO(_4))(aq) + 3 MgO(s) + 3 H(_2)O = 3 MgSO(_4)(aq) + 2 Al(OH)(_3)(s)</td>
<td>100%</td>
</tr>
<tr>
<td>CoSO(_4)(aq) + MgO(s) + H(_2)O = MgSO(_4)(aq) + Co(OH)(_2)(s)</td>
<td>95%</td>
</tr>
<tr>
<td>2 NiSO(_4)(aq) + MgO(s) + 7 H(_2)O = Ni(_2)(OH)(_2)SO(_4).6H(_2)O(s) + MgSO(_4)(aq)</td>
<td>35%</td>
</tr>
<tr>
<td>NiSO(_4)(aq) + MgO(s) + H(_2)O = MgSO(_4)(aq) + Ni(OH)(_2)(s)</td>
<td>90%</td>
</tr>
<tr>
<td>H(_2)SO(_4)(aq) + MgO(s) + 2 H(_2)O = MgSO(_4)(aq) + 2 Mg(OH)(_2)(s)</td>
<td>99%</td>
</tr>
</tbody>
</table>

**MHP 2 (unit operation 43)**

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>H(_2)SO(_4)(aq) + Ca(OH)(_2)(s) = CaSO(_4).2H(_2)O(s)</td>
<td>100%</td>
</tr>
<tr>
<td>CoSO(_4)(aq) + 2 H(_2)O + Ca(OH)(_2)(s) = CaSO(_4).2H(_2)O(s) + Co(OH)(_2)(s)</td>
<td>100%</td>
</tr>
<tr>
<td>NiSO(_4)(aq) + 2 H(_2)O + Ca(OH)(_2)(s) = CaSO(_4).2H(_2)O(s) + Ni(OH)(_2)(s)</td>
<td>95%</td>
</tr>
<tr>
<td>MnSO(_4)(aq) + 2 H(_2)O + Ca(OH)(_2)(s) = CaSO(_4).2H(_2)O(s) + Mn(OH)(_2)(s)</td>
<td>20%</td>
</tr>
<tr>
<td>MgSO(_4)(aq) + 2 H(_2)O + Ca(OH)(_2)(s) = CaSO(_4).2H(_2)O(s) + Mg(OH)(_2)(s)</td>
<td>0.9%</td>
</tr>
</tbody>
</table>

* The extent of reaction is fixed based on Vale Inco projects.

Controllers:

410 – Recycle seed to the pre-mix tank in MHP 1

60% of the thickener underflow is recycled to the pre-mix to provide seeds for the MHP.

420 – MgO to MHP 1 circuit

MgO is added stoichiometric to cobalt and nickel, to precipitate 85% of the nickel and cobalt in the solution.

430 – Flocculant addition in the thickener of MHP 1

Flocculant is added proportionally to the amount of solids formed in the MHP step.

440 – Water to filter cake washing

Water is added to get 4.5 t of water/t of dry solids in the feed.
510 – Recycle seed to the pre-mix tank in MHP 2
64% of the thickener underflow is recycled to the pre-mix to provide seeds for the neutralization

520 – Lime to the MHP 2 circuit
Lime is added in order to reach pH = 8.7 in the final tank.

530 – Flocculant addition in the thickener of MHP 2
Flocculant is added proportionally to the amount of solids formed in the MHP 2 step.

Impurities removal/ Reagents and Final discharge

Figure D.7 presents MetSim model flowsheet for impurity removal circuit. Figures D.8 to D.10 present reagents and final discharge areas.

Reactions:

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>MnSO₄(aq) + 2H₂O(l) + Ca(OH)₂(s) = CaSO₄·2H₂O(s) + 1 Mn(OH)₂(s)</td>
<td>50%</td>
</tr>
<tr>
<td>MgSO₄(aq) + 2H₂O(l) + Ca(OH)₂(s) = CaSO₄·2H₂O(s) + 1 Mg(OH)₂(s)</td>
<td>30%</td>
</tr>
</tbody>
</table>

* The extent of reaction is fixed based on Vale Inco projects.

Controllers:

610 – Lime to impurities removal circuit
Lime is added three times stoichiometric on the manganese content.

620 – Recycle seed to the pre-mix tank
50% of the thickener underflow is recycled to the pre-mix to provide seeds for the neutralization

630 – Flocculant addition in the thickener of impurities removal 1
Flocculant is added proportionally to the amount of solids formed in the impurities removal step.
Figure D.1: Beneficiation and acid leaching areas flowsheet
Figure D.2: First neutralization area flowsheet
Figure D.3: CCD area flowsheet
Figure D.4: Second neutralization area flowsheet
Figure D.5: MHP 1 area flowsheet
Figure D.6: MHP 2 area flowsheet
Figure D.7: Impurities removal area flowsheet
Figure D.8: Reagents 1 area flowsheet
Figure D.9: Reagents 2 area flowsheet
High Pressure Acid Leaching model

Figure D.10: Product/Tailings area flowsheet
APPENDIX E – Heap Leaching model data

Agglomeration/Heap Leaching

Figure E.1 presents MetSim model flowsheet for the heap leaching area.

Reactions:

<table>
<thead>
<tr>
<th>Heap leach (unit operation 6)</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>$1\text{NiO}_{\text{sap(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{NiSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{RRS}$</td>
</tr>
<tr>
<td>$1\text{CoO}_{\text{sap(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{CoSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{RRS}$</td>
</tr>
<tr>
<td>$1\text{MnO}_{\text{sap(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{MnSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{RRS}$</td>
</tr>
<tr>
<td>$1\text{MgO}_{\text{sap(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{MgSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{RRS}$</td>
</tr>
<tr>
<td>$1\text{Fe}_2\text{O}_3\text{sap(s)} + 3\text{H}_2\text{SO}_4(aq) = 1\text{Fe}_2(\text{SO}_4)_3(aq) + 3\text{H}_2\text{O}(l)$</td>
<td>$\text{RRS}$</td>
</tr>
<tr>
<td>$1\text{Al}_2\text{O}_3\text{sap(s)} + 3\text{H}_2\text{SO}_4(aq) = 1\text{Al}_2(\text{SO}_4)_3(aq) + 3\text{H}_2\text{O}(l)$</td>
<td>$\text{RRS}$</td>
</tr>
<tr>
<td>$1\text{CaO}_{(s)} + 1\text{H}_2\text{SO}_4(aq) + 1\text{H}<em>2\text{O}(l) = 1\text{Gypsum}</em>{(s)}$</td>
<td>$\text{RRS}$</td>
</tr>
<tr>
<td>$1\text{NiO}_{\text{fer(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{NiSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{Rf}$</td>
</tr>
<tr>
<td>$1\text{CoO}_{\text{fer(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{CoSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{Rf}$</td>
</tr>
<tr>
<td>$1\text{MgO}_{\text{fer(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{MgSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{Rf}$</td>
</tr>
<tr>
<td>$1\text{Al}_2\text{O}_3\text{fer(s)} + 3\text{H}_2\text{SO}_4(aq) = 1\text{Al}_2(\text{SO}_4)_3(aq) + 3\text{H}_2\text{O}(l)$</td>
<td>$\text{Rf}$</td>
</tr>
<tr>
<td>$1\text{Cr}_2\text{O}_3\text{fer(s)} + 3\text{H}_2\text{SO}_4(aq) = 1\text{Cr}_2(\text{SO}_4)_3(aq) + 3\text{H}_2\text{O}(l)$</td>
<td>$\text{Rf}$</td>
</tr>
<tr>
<td>$1\text{Fe}_2\text{O}_3\text{fer(s)} + 3\text{H}_2\text{SO}_4(aq) = 1\text{Fe}_2(\text{SO}_4)_3(aq) + 3\text{H}_2\text{O}(l)$</td>
<td>$\text{Rf}$</td>
</tr>
<tr>
<td>$1\text{MnO}_{\text{fer(s)}} + 1\text{H}_2\text{SO}_4(aq) = 1\text{MnSO}_4(aq) + 1\text{H}_2\text{O}(l)$</td>
<td>$\text{Rf}$</td>
</tr>
</tbody>
</table>

$\text{Rf}$ and $\text{RRS}$ designate the extraction calculated using equation 5.61, as explained in section 5.

Controllers:

1 – Ore feed to the plant
Ore feed flowrate is back calculated based on the reserve, mine lifetime and moisture content in the ore, as explained in appendix C.

2 – Acid addition to agglomeration drum
Stoichiometric based on 50% neutralization of the Mg content in the ore.
3 – Water to agglomeration
Calculated in order to get 65% solids in the agglomeration discharge.

3216 – Acid to the heap
Calculated in order to get 8g/L free acid in the heap discharge.

3217 – Water to the heap
Calculated in order to get the desired solution to the heap. Equation 5.62 and 5.63 present the controller equation.

3219 – Water to the heap washing
The washing step has the goal of washing the solids to recover mainly the nickel. In the model, the amount of water added to wash is the enough to recover 90% of the nickel in solution, that is, the ore is washed in order to recover the moisture that contains nickel.

3225 – Evaporation
Evaporation in the heap and ponds are calculated based on the open area and the net evaporation rate, as explained in section 5.3.1.

Primary and secondary neutralization

Figure E.2 and E.3 present MetSim model flowsheet for the primary and secondary neutralization areas.

Reactions:

<table>
<thead>
<tr>
<th>Releach unit operation (unit operation 10)</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Ni(OH)$_2(s)$ + H$_2$SO$_4$ = NiSO$_4$ + H$_2$O</td>
<td>100%</td>
</tr>
<tr>
<td>Co(OH)$_2(s)$ + H$_2$SO$_4$ = CoSO$_4$ + H$_2$O</td>
<td>100%</td>
</tr>
</tbody>
</table>

Neutralization 1 (unit operation 12)

<table>
<thead>
<tr>
<th>Equation</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>Fe$_2$(SO$_4$)$_3(aq)$ + 3 CaCO$_3(s)$ + 9 H$<em>2$O$</em>(l)$ = 3 CO$_2(g)$ + 3CaSO$_4$.2H$<em>2$O$</em>(s)$ + 2 Fe(OH)$_3(s)$</td>
<td>90%</td>
</tr>
<tr>
<td>Al$_2$(SO$_4$)$_3(aq)$ + 3 CaCO$_3(s)$ + 9 H$<em>2$O$</em>(l)$ = 3 CO$_2(g)$ + 3CaSO$_4$.2H$<em>2$O$</em>(s)$ + 2 Al(OH)$_3(s)$</td>
<td>10%</td>
</tr>
</tbody>
</table>
$$\text{Cr}_2(\text{SO}_4)_3(aq) + 3 \text{CaCO}_3(s) + 9 \text{H}_2\text{O}(l) = 3 \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2 \text{Cr(OH)}_3(s) \quad 50\%$$

$$\text{H}_2\text{SO}_4(aq) + \text{CaCO}_3(s) + \text{H}_2\text{O}(l) = \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) \quad 100\%$$

**Neutralization 2 (unit operation 21)**

$$\text{Fe}_2(\text{SO}_4)_3(aq) + 3 \text{CaCO}_3(s) + 9 \text{H}_2\text{O}(l) = 3 \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2 \text{Fe(OH)}_3(s) \quad 100\%$$

$$\text{FeSO}_4(aq) + 1 \text{CaCO}_3(s) + 3 \text{H}_2\text{O}(l) = \text{CO}_2(g) + 1\text{CaSO}_4.2\text{H}_2\text{O}(s) + 1 \text{Fe(OH)}_2(s) \quad 100\%$$

$$\text{Al}_2(\text{SO}_4)_3(aq) + 3 \text{CaCO}_3(s) + 9 \text{H}_2\text{O}(l) = 3 \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2 \text{Al(OH)}_3(s) \quad 99\%$$

$$\text{Cr}_2(\text{SO}_4)_3(aq) + 3 \text{CaCO}_3(s) + 9 \text{H}_2\text{O}(l) = 3 \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) + 2 \text{Cr(OH)}_3(s) \quad 90\%$$

$$\text{NiSO}_4(aq) + 1 \text{CaCO}_3(s) + 3 \text{H}_2\text{O}(l) = \text{CO}_2(g) + 1\text{CaSO}_4.2\text{H}_2\text{O}(s) + 1 \text{Ni(OH)}_2(s) \quad 10\%$$

$$\text{H}_2\text{SO}_4(aq) + \text{CaCO}_3(s) + \text{H}_2\text{O}(l) = \text{CO}_2(g) + 3\text{CaSO}_4.2\text{H}_2\text{O}(s) \quad 100\%$$

* The extent of reaction is fixed based on Vale Inco projects.

** Controllers:**

**3205** – Limestone to the neutralization 1 circuit
Limestone is added in order to reach pH 2.5 in the final tank.

**3206** – Recycle seed to the pre-mix tank in iron removal 1
75% of the thickener underflow is recycled to the pre-mix to provide seeds for the neutralization

**3207** – Limestone to the iron removal 2 circuit
Limestone is added in order to reach pH = 4.5 in the final tank.

**3209** – Recycle seed to the pre-mix tank in secondary neutralization step
75% of the thickener underflow is recycled to the pre-mix to provide seeds for the neutralization

**3210** – Flocculant addition in the thickener of iron removal 1
Flocculant is added proportionally to the amount of solids formed in the neutralization step.

**3211** – Flocculant addition in the thickener of iron removal 2
Flocculant is added proportionally to the amount of solids formed in the neutralization step.
3220 – Water to filter cake washing
Water is added to the filter in order to get 99.5% solution recovery.

**Mixed Hydroxide Precipitate (MHP) 1 and 2**

Figure E.4 and E.5 present MetSim model flowsheet for the first and second mixed hydroxide precipitate areas.

**Reactions:**

**MHP 1 (unit operation 30)**

CuSO$_4$(aq) + 1 MgO(s) + H$_2$O = 1 MgSO$_4$(aq) + 1 Cu(OH)$_2$(s) 100%

Al$_2$(SO$_4$)$_3$(aq) + 3 MgO(s) + 3 H$_2$O = 3 MgSO$_4$(aq) + 2 Al(OH)$_3$(s) 100%

CoSO$_4$(aq) + MgO(s) + H$_2$O = 1 MgSO$_4$(aq) + 1 Co(OH)$_2$(s) 95%

2 NiSO$_4$(aq) + 1 MgO(s) + 7 H$_2$O = Ni$_2$(OH)$_2$SO$_4$.6H$_2$O(s) + 1 MgSO$_4$(aq). 35%

NiSO$_4$(aq) + MgO(s) + H$_2$O = 1 MgSO$_4$(aq) + 1 Ni(OH)$_2$(s) 90%

H$_2$SO$_4$(aq) + MgO(s) = 1 MgSO$_4$(aq) + 2H$_2$O 99%

MnSO$_4$(aq) + MgO(s) + H$_2$O = 1 MgSO$_4$(aq) + 1 Mn(OH)$_2$(s) 40%

**MHP 2 (unit operation 39)**

H$_2$SO$_4$(aq) + Ca(OH)$_2$(s) = CaSO$_4$.2H$_2$O(s) 100%

CoSO$_4$(aq) + 2H$_2$O + Ca(OH)$_2$(s) = CaSO$_4$.2H$_2$O(s) + 1 Co(OH)$_2$(s) 100%

NiSO$_4$(aq) + 2H$_2$O + Ca(OH)$_2$(s) = CaSO$_4$.2H$_2$O(s) + Ni(OH)$_2$(s) 95%

MnSO$_4$(aq) + 2H$_2$O + Ca(OH)$_2$(s) = CaSO$_4$.2H$_2$O(s) + 1 Mn(OH)$_2$(s) 20%

MgSO$_4$(aq) + 2H$_2$O + Ca(OH)$_2$(s) = CaSO$_4$.2H$_2$O(s) + 1 Mg(OH)$_2$(s) 0.9%

* The extent of reaction is fixed based on Vale Inco projects.

**Controllers:**

1100 – MgO to MHP 1 circuit
MgO is added stoichiometric to cobalt and nickel, to precipitate 85% of the nickel and cobalt in the solution.

1200 – Recycle seed to the pre-mix tank in MHP 1
60% of the thickener underflow is recycled to the pre-mix to provide seeds for the MHP.
1300 – Flocculant addition in the thickener of MHP 1
Flocculant is added proportionally to the amount of solids formed in the MHP step.

2100 – Lime to the MHP 2 circuit
Lime is added in order to reach pH = 8.7 in the final tank.

3200 – Recycle seed to the pre-mix tank in MHP 2
64% of the thickener underflow is recycled to the pre-mix to provide seeds for the neutralization

3220 – Water to filter cake washing
Water is added to get 4.5 t of water/t of dry solids in the feed.

**Impurities removal/ Reagents and Final discharge**

Figure E.6 presents MetSim model flowsheet for impurity removal circuit. Figures D.7 to D.9 present reagents and final discharge areas.

**Reactions:**

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Extent*</th>
</tr>
</thead>
<tbody>
<tr>
<td>MnSO$_4$(aq) + 2H$_2$O(l) + Ca(OH)$_2$(s) = CaSO$_4$.2H$_2$O(s) + 1 Mn(OH)$_2$(s)</td>
<td>50%</td>
</tr>
<tr>
<td>MgSO$_4$(aq) + 2H$_2$O(l) + Ca(OH)$_2$(s) = CaSO$_4$.2H$_2$O(s) + 1 Mg(OH)$_2$(s)</td>
<td>30%</td>
</tr>
</tbody>
</table>

* The extent of reaction is fixed based on Vale Inco projects.

**Controllers:**

3100 – Lime to MHP 1 circuit
Lime is added three times stoichiometric on the manganese content.

1300 – Flocculant addition in the thickener of impurities removal 1
Flocculant is added proportionally to the amount of solids formed in the impurities removal step.
Figure E.1: Agglomeration/Heap area flowsheet
Figure E.2: Primary neutralization area flowsheet
Figure E.3: Secondary neutralization area flowsheet
Figure E.4: MHP 1 area flowsheet
Figure E.5: MHP 2 area flowsheet
Figure E.6: Impurities removal area flowsheet
Figure E.7: Reagents 1 area flowsheet
Figure E.8: Reagents 2 area flowsheet
Figure E.9: Products/Tailings area flowsheet