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BIOLEACHING OF NICKEL AND COBALT – THE PROGRESS AND THE POTENTIAL

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ABSTRACT

The emergence of bioleaching for the processing of sulphide ores and concentrates has occurred over the past 35 or so years. The use of this technology for the extraction of nickel and cobalt has been extensively investigated over this period, and it has found commercial application in several forms, ranging from the treatment in mechanically-agitated tanks of cobaltiferous pyrite tailings just north of the equator in central Africa, to the extraction of nickel from a polymetallic black schist orebody in heaps situated in boreal conditions just outside the Arctic Circle.

In between those extremes, several other pilot, demonstration-scale and industrial operations have been implemented across a variety of dimensions, including those already mentioned (using mechanically-agitated versus heap reactors, and in widely diverse climatic conditions), but also encompassing vastly differing mineralogies and chemistries, varied microbial communities and their associated operating temperature ranges, and with a wide-ranging selection of downstream processing flowsheets for purification and recovery.

This review traces the emergence and development of nickel and cobalt bioleaching technology, focusing on the impact of mineralogy on the chemistry of the leaching processes, the important microbial factors, and how these parameters, combined with economic considerations, influence flowsheet selection. Progress is assessed by reviewing the significant advances that have occurred on a case-by-case basis, with an emphasis on those projects that have proceeded to, or are close to, industrial application.

The future potential for bioleaching of nickel and cobalt is inextricably linked to these metals' respective markets. Both are battery metals, and future demand is likely to be closely correlated with the emergence of a low-carbon economy, and the impact this is expected to have on the energy sector. Supply-side factors will also impinge on the future application of nickel and cobalt bioleaching. Most notable amongst these are the supply from non-sulphidic resources, and geopolitical issues around responsible production (particularly for cobalt).

These factors are assessed, drawing on recent reviews and up-to-date market projections, to sketch the potential for the future use of bioleaching as a primary extraction process for these two metals.

Keywords: Bioleaching, Nickel, Cobalt

INTRODUCTION

Background and Scope

The story of the commercial application of bioleaching over the past 35 years is dominated by two metals: gold and copper. The process has been applied extensively across the globe in mechanically-agitated reactors for the pre-oxidation of refractory gold-bearing concentrates[1,2,3], and also for the extraction of copper from secondary sulphide ores, mainly in Chile[2,4]. Despite this dominance, some of the more interesting applications of both the tank- and heap-leaching processes have been for the recovery of nickel and cobalt. These applications include the treatment in mechanically-agitated tanks of cobaltiferous pyrite tailings just north of the equator in central Africa, and the extraction of nickel from a polymetallic black schist orebody in heaps situated in boreal conditions just outside the Arctic Circle.

The aim of this paper is twofold: to look back at the progress that has been made in the commercial application of bioleaching for the recovery of these two metals, and to look forward and speculate what the future might hold. In looking back, particular attention will be paid to several important factors associated with the mineralogy of the orebodies and the selection of microbes, which influence flowsheet selection. The focus will be on several case studies, using projects that have proceeded to, or are close to, industrial application. In looking forward, the respective markets for these two metals will be examined, taking cognisance of the fact that they are both battery metals, which is expected to dictate future demand as a low-carbon economy emerges in the energy sector. Supply-side factors and geopolitical issues around responsible production (particularly for cobalt) will also be assessed, to sketch the potential for the future use of bioleaching as a primary extraction process for these two metals.

Landmarks in Nickel and Cobalt Bioleaching

Interestingly, some of the earliest work conducted at Mintek in the field of bioleaching was focused on the heap leaching of nickel sulphide ores, and one of the first technical papers published in this field by Mintek (at a conference commemorating the organisation's 50th anniversary in 1984) described the heap bioleaching of low-grade nickel ores, including the results of column-leach tests conducted in 0.3-metre, 1-metre and 2-metre tall columns[5]. However, Mintek's focus moved to the development of bioleach processes for refractory gold concentrates, and later for copper sulphide concentrates and ores, and it was many years before nickel (and to a lesser extent, cobalt) featured once more in Mintek's bioleach process development.

The first application of a heap bioleach process was at the Lo Aguirre copper mine in Chile in 1980[6], and mechanically-agitated bioleaching for a refractory gold concentrate was implemented for the first time at the Fairview gold mine in South Africa in 1986[1]. However, the first serious attempt to develop a process for treating a nickel sulphide concentrate occurred somewhat later, between 1997 and 2000, with the development by Billiton of the BioNIC® process, which progressed to pilot-scale demonstration of an integrated process at the Yabulu Nickel Refinery in Queensland, Australia[7,8,9,10,11,12,13,14]. During this same period, a process for the bioleaching of cobaltiferous pyrite tailings was commercialised at the site of the Kilembe Copper Mine in Kasese, Uganda, in 1999[15,16,17,18,19,20,21,22]. This large-scale operation utilised technology developed by the French company, Bureau de Recherches Géologiques et Minières (BRGM).

Heap bioleaching of the Mt Sholl nickel-copper ore was taken to demonstration scale at the Radio Hill Mine in Western Australia between 1999 and 2002[12,23,24]. A heap bioleaching process for nickel achieved commercial application at the Talvivaara mine in Sotkamo, Finland, in 2008. The operation suffered major technical, environmental and economic problems, and closed in late 2014. It was restarted by new owners, Terrafame, in August 2015[12,25,26,27,28,29,30,31,32,33].

Mintek investigated bioleaching of a nickel-copper concentrate from the Aguablanca mine is Spain in the period between 2004 and 2008[34]. The integrated piloting work was conducted during Mintek's participation in the BioMinE project, and laid the basis for the first commercial bioleach process treating a nickel sulphide concentrate, in which the Mintek technology was implemented at the Mondo Minerals talc mine located at Vuonos, Finland, in 2015[35,36,37,38].

There are a couple of potential projects in the pipeline, including Western Areas' Cosmic Boy nickel project in Western Australia[39,40,41], and Aeon Metals Limited's Walford Creek copper-cobalt project in north-west Queensland[42,43,44].

A Note About Iron

Nickel-bearing ores and concentrates typically contain large quantities of iron sulphide, often as pyrrhotite, and sometimes with some pyrite present. Pyrrhotite is highly reactive in a bioleach process, and adds to the oxidative load, often without much benefit, since it does not contain any value. It also means that the control

of iron, either in heaps or in the post-leach unit operations for tank-based processes, is an important aspect of the metallurgical circuit. In the case of the Kasese project, the cobalt is intimately associated with pyrite, so the oxidation of pyrite is a pre-requisite for the extraction and recovery of the cobalt, and iron control is equally important.

THE PROGRESS – CASE STUDIES OF THE LANDMARK PROJECTS

The BioNIC® Process

The BioNIC® process was the first attempt to develop an integrated process incorporating the bioleaching of nickel sulphide concentrates that could compete with the conventional smelting process[12]. It was developed by Billiton as an extension of the by-then well-established BIOX® process for refractory gold concentrates. The integrated process was developed by means of a collaboration between Billiton Process Research (BPR) and Mintek. BPR developed the bioleach process, and Mintek was responsible for the subsequent processes, incorporating iron removal and downstream metals recovery by either ion exchange (IX) or solvent extraction (SX), and electrowinning (EW)[9].

The initial development work involved bench-scale batch bioleach test work, using the same mesophilic bioleach bacteria that were used in the BIOX® process, after adapting them to nickel. The batch tests established that nickel extractions ranging between 90 % and 98 % could be achieved on a variety of nickel-bearing ores and concentrates[7,8].

Continuous mini-plant testing in 120-litre reactors followed, using a concentrate that contained 8 % nickel and 27 % iron, with most of the iron sulphide in the form of pyrrhotite. At a 15 % feed solids concentration, a nickel extraction of 95 % was achieved in the bioleach process at a residence time of 8 days[7].

The next stage in the development of the process was the operation of a continuous demonstration-scale plant, designed to produce 20 kg of nickel cathode per day. The concentrate used in this campaign contained 11 % nickel and 31 % iron, with most of the iron sulphide in the form of pyrite. A nickel extraction of 92 % was obtained in the bioleach process, at a residence time of between 7 and 10 days, and a feed solids concentration of up to 17.5 %[7,9].

The bioleach liquor contained up to 40 g/L of ferric iron, and both the mini-plant and demonstration plant test work showed that very high levels of iron removal could be achieved with no measurable loss of nickel. In fact, it was suspected that nickel recovery may have been enhanced owing to the liberation of nickel through the destruction of small quantities of jarosite (which is generated in the bioleach process) in the neutralisation process[8,9].

At that stage of the process development, several options for downstream solution purification and nickel recovery were being considered, as illustrated in Figure 1. Following iron removal and solid-liquid separation, the options under consideration were upgrading (by purification and increasing the nickel tenor) of the pregnant liquor to one suitable for nickel metal production, either directly via SX or IX, or indirectly via precipitation and re-dissolution of a mixed sulphide[8]. The IX route offered the possibility of making either Class 1 nickel or an electrowon high nickel-low iron ferronickel product.

The demonstration plant flowsheet that was tested at BPR's research facility in Johannesburg during 1997 incorporated IX and EW for nickel recovery, with SX being used for cobalt recovery, as shown in Figure 2. The demonstration plant was operated for six months, and produced 700 kg of Class 1 nickel[8]. Despite the technical success of this flowsheet, it was noted that the initial cost of the resin would be significant, as would the ongoing cost of replacement of resin lost through either physical damage or loss of activity, and that the same would apply to the alternative IX route leading to the production of ferronickel. Additionally, it was also reported that the EW of ferronickel would be costly owing to low current efficiencies. The proposed SX route, using Versatic acid and Cyanex 272, was considered more economically promising[8].

The final stage of the development of the BioNIC® process was the operation of an integrated circuit at the site of Queensland Nickel Industries' Yabulu Nickel Refinery, between August 1998 and February 1999. The flowsheet differed from that of the previously-reported demonstration plant, in that SX and EW replaced IX in the downstream circuit, as shown in Figure 3[9]. Clearly, the economic factors referred to above had swayed the process development away from IX and towards SX.
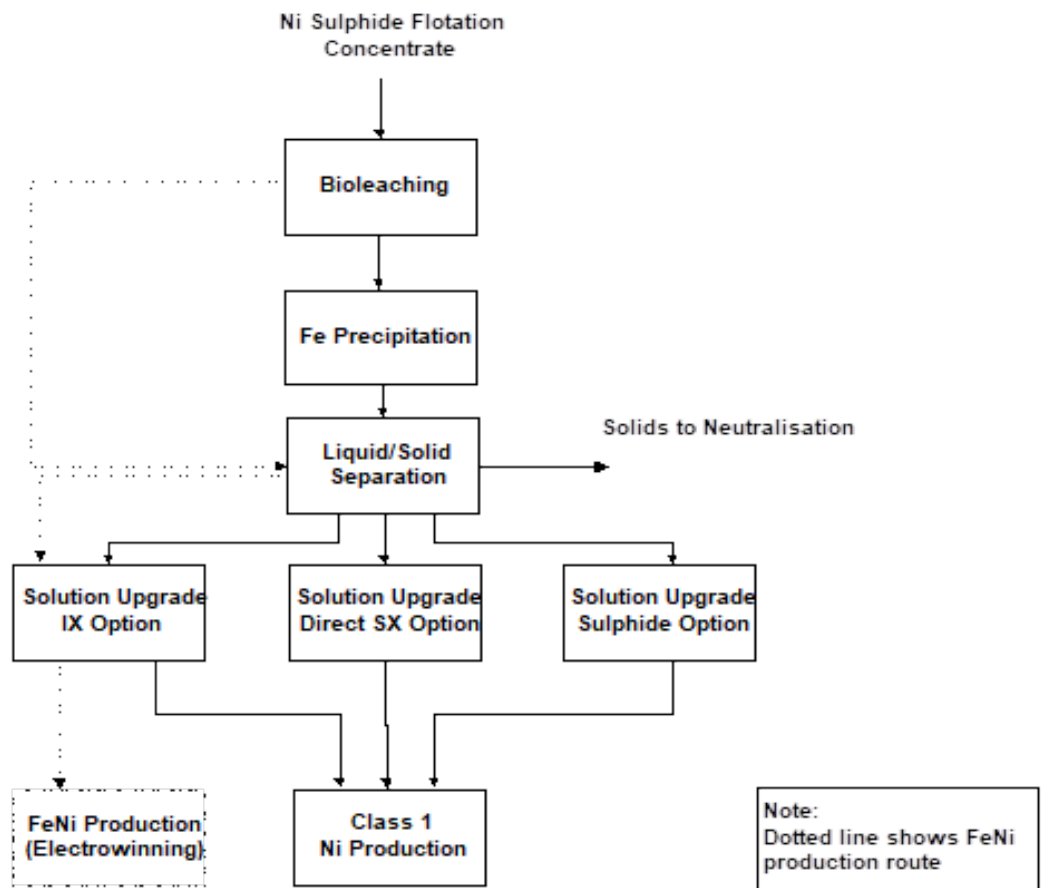
Ni Sulphide Flotation
Concentrate

Bioleaching

Fe Precipitation

Liquid/Solid Separation

Solids to Neutralisation

Solution Upgrade IX Option

Solution Upgrade Direct SX Option

Solution Upgrade Sulphide Option

FeNi Production (Electrowinning)

Class 1 Ni Production

Note:
Dotted line shows FeNi production route

Figure 1: BioNIC® flowsheet options[8]

Concentrate Feed

REPULPER

PRIMARY BIOLEACH

SECONDARY BIOLEACH

IRON PRECIPITATION

Acid

Air

Air

SOLID/LIQUID SEPARATION

TAILINGS

ION-EXCHANGE Cu GUARD CCT

ION-EXCHANGE Ni RECOVERY

Spent

NICKEL PLATE PRODUCT

ELECTROWINNING
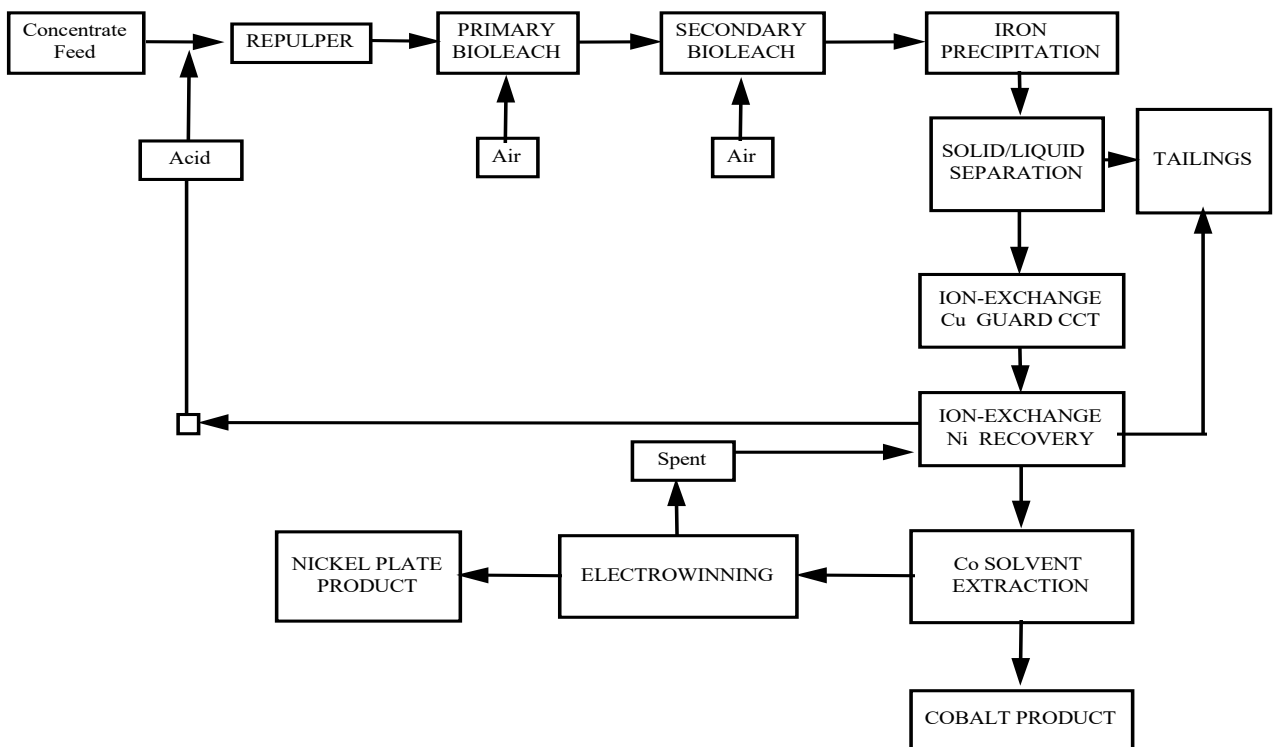
Co SOLVENT EXTRACTION

COBALT PRODUCT

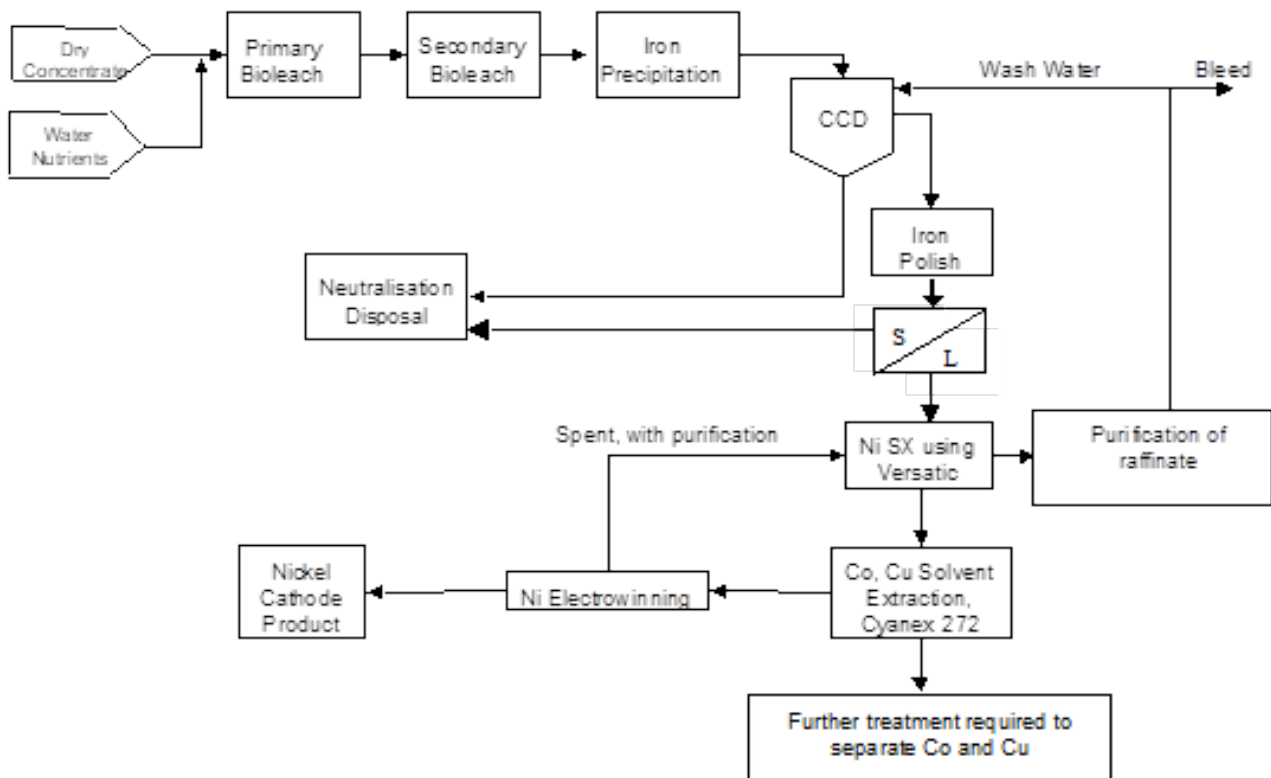**Figure 2: BioNIC® demonstration plant flow sheet[7]**

Figure 3: Integrated BioNIC® circuit[9]

The concentrate used in this campaign contained 12 % nickel and 31 % iron, which is very similar to that of the concentrate used in the previous demonstration plant campaign. Under optimised conditions, of a 7-day residence time and a 17.5 % feed solids concentration, nickel and cobalt extractions in the bioleach process were 94 % and 98 %, respectively. The final solution nickel and iron tenors under those operating conditions were 23.8 g/L and 38.8 g/L, respectively, demonstrating the robustness of the mesophilic bacterial culture at high metal concentrations. At a longer residence time of 9 days, with a feed solids concentration of 15 %, the nickel and cobalt extractions were 97 % and 99 %, respectively.

The downstream unit operations included iron precipitation, solid-liquid separation and washing, iron polishing, nickel SX, cobalt SX and nickel EW[9,10]. The iron precipitation process succeeded in removing over 99 % of the iron in solution. Solid-liquid separation could be effected by either pressure filtration (with a washing stage) or a counter-current decantation circuit, with recovery of over 99 % of the dissolved metal, producing a solution containing 7.0 to 7.5 g/L of nickel.

The iron polishing circuit aimed to reduce the iron content to below 2 mg/L by removal of the residual ferrous iron. This was achieved by the addition of limestone in a four-stage, agitated and aerated process, operating at a temperature of between 50 and 70 °C. Nickel losses were limited to below 0.1 % when the operating pH level was below 5.0.

The objective of the nickel SX circuit, using Versatic™ Acid 10 as the extractant, was to maximise the nickel and cobalt recoveries, while minimising the co-extraction of magnesium, manganese and calcium. The target specifications for this circuit were a raffinate stream containing less than 100 mg/L nickel, greater than 80 % cobalt recovery and less than 1 % calcium extraction. These targets were achieved by optimising the pH profile across the circuit, operating a scrubbing circuit, and optimising the number of extraction stages.

The loaded strip liquor from the nickel SX plant was passed through an activated carbon column to remove dissolved and entrained Versatic acid from the solution. The pH level of this stream was then raised from 2.8 to approximately 6.0 using sodium hydroxide, before being treated in the cobalt SX plant, where the aim, using Cyanex 272, was to ensure adequate removal of cobalt and copper with minimal co-extraction of nickel. An overall cobalt extraction of over 98 % was achieved. The product solution contained 38 g/L of cobalt. Further purification to remove copper, zinc and iron would be necessary, prior to cobalt metal production.

The raffinate solution (advance nickel electrolyte) leaving the cobalt SX plant was again passed through a carbon column to remove any dissolved or entrained organic phase, before moving to the EW circuit, with the

spent electrolyte being returned to the nickel SX circuit for use as strip liquor. The objectives of the nickel EW circuit were the consistent production of ASTM-grade nickel cathode with a suitable deposit morphology, and to determine the current efficiency and energy consumption required to produce the high-quality nickel cathode. Three of the four cathodes produced during the campaign met the American Society for Testing and Materials (ASTM) requirements, and all four were flexible and ductile, and stripped easily from the titanium cathodes. An average current efficiency of 99 % was obtained, with an average energy consumption of 3.12 kWh/kg of nickel metal produced, at a current density of 200 A/m².

The averaged overall recoveries of nickel and cobalt, for the whole flowsheet as tested, for the two bioleach conditions mentioned previously, are summarised in Table 1.

Table 1: Overall metal recoveries achieved in the integrated BioNIC® circuit[9]

Overall bioleach residence time, d	7	9
Feed solids concentration, %	17.5	15.0
Nickel recovery	91.9	95.2
Cobalt recovery	86.6	88.0

It was concluded that the overall recoveries achievable would depend on a trade-off between the capital cost of the bioleach plant and the recoveries achieved, as determined by the residence time and the feed solids concentration.

In the course of the BioNIC® process development, it was observed that the unleached sulphides in the bioleach residues occurred mainly as simple unleached particles, rather than being contained within unliberated complex particles, and it was suggested that a modest regrind of the concentrate prior to bioleaching may have improved the metal dissolutions[8]. However, the authors did not appear to consider that a regrind may also have resulted in a reduction in the bioleach residence time, which had been found to be considerably longer than the 4 to 4.5 days usually required for the refractory gold BIOX® process.

Further development of the BioNIC® process involved an evaluation of thermophilic microorganisms. In a continuous test using a microbial culture operating at a temperature of 68 °C, a nickel extraction of 98 % was achieved in a retention time of 4 days[11]. It was noted, however, that the feed solids concentration was limited to a maximum of 10 %. Further continuous test work using a microbial culture operating at 78 °C attained a similar nickel extraction in a retention time of 3 days[13]. Despite extensive development work, and the commissioning and subsequent operation of a commercial demonstration plant treating a copper sulphide concentrate at Chuquicamata in Chile in 2003[14], the commercial application of thermophilic bioleaching in mechanically-agitated vessels has not been realised. Questions remain regarding the robustness of thermophilic microbial cultures (which limits the feed solids concentration and raises concerns regarding the viability of the process at a large scale), mass transfer and oxygen (O_2) supply, and materials of construction, amongst others.

Despite the technical successes in the BioNIC® process development, and claims that the process was economically competitive[8] and a viable process for the treatment of nickel sulphide concentrates[9], it did not achieve commercial implementation. Nevertheless, the extensive development work undertaken to prove the technical viability of the process laid the groundwork for the future application of bioleaching for the treatment of nickel sulphide concentrates.

The Kasese Cobalt Project, Uganda

In 1998, the world's first commercial bioleaching plant to treat a base metal concentrate was established at the site of the Kilembe Copper Mine in Uganda. The plant, owned and operated by the Kasese Cobalt Company (KCC), was designed to produce about 1,000 t/a of cobalt cathode from a cobaltiferous pyrite concentrate, using mesophilic bacteria. The technology was supplied by the French company, BRGM.

The Kilembe sulphide deposit was mined for 26 years, between 1956 and 1982, producing 16 million tonnes of copper[19,21]. In the copper recovery process, the tailings was a cobalt-rich pyrite concentrate, which was stockpiled at the Kasese railway terminal, 12 km downhill from Kilembe Mine. The stockpile comprised about 900,000 tonnes of material, containing approximately 80 % pyrite and 1.38 % cobalt. Most of the cobalt occurred in a disseminated ionic form within the pyrite lattice. The pyrite concentrate also contained small amounts of copper, nickel and zinc. The non-sulphide components were mainly quartz, and there were no carbonates in the concentrate. The particle size distribution of the concentrate was relatively coarse, with 95 % less than 150 µm and 80 % smaller than 90 µm. Being mostly pyrite, the specific gravity of the concentrate was about 4,000 kg/m³.

Over time, the pyrite stockpile became dispersed over a wide area by the heavy rainfall that occurs during certain months in the tropical savanna climate that prevails. This led to the production of acid and the release of metals, which threatened the nearby Queen Elizabeth National Park, the large nearby Lake George and the local communities, as shown in Figure 4.

Figure 4: An aerial view of the Kasese pyrite concentrate stockpile before the implementation of the bioleach plant, showing the de-vegetated areas downhill from the stockpile caused by acidic run-off[20]

Process Development

BRGM undertook a 10-year-long programme of process development that began at its facility in at Orléans in France in 1988, with the objective of demonstrating the technical and economic feasibility of applying a biohydrometallurgical process for the recovery of cobalt from the Kasese pyrite stockpile[19]. Some of the factors that favoured the feasibility of the project included an abundance of water at the site, a nearby and abundant supply of limestone, reasonable infrastructure (including electricity supply), a strong political will to clean the environment at the border of the nearby natural reserve and lake, and favourable site topology.

The metallurgical development programme included microbial evaluations, optimisation of the bioleach process parameters in batch and continuous tests at varying scales, gas-liquid mass transfer studies in tanks up to 65 m^3 in volume, and the evaluation of various flowsheets for downstream solution purification and metals recovery[19].

Commercial Implementation

Construction of the full-scale bioleaching operation at Kasese began in 1998 and was completed in 1999[19]. A simplified process flowsheet is shown in Figure 5.
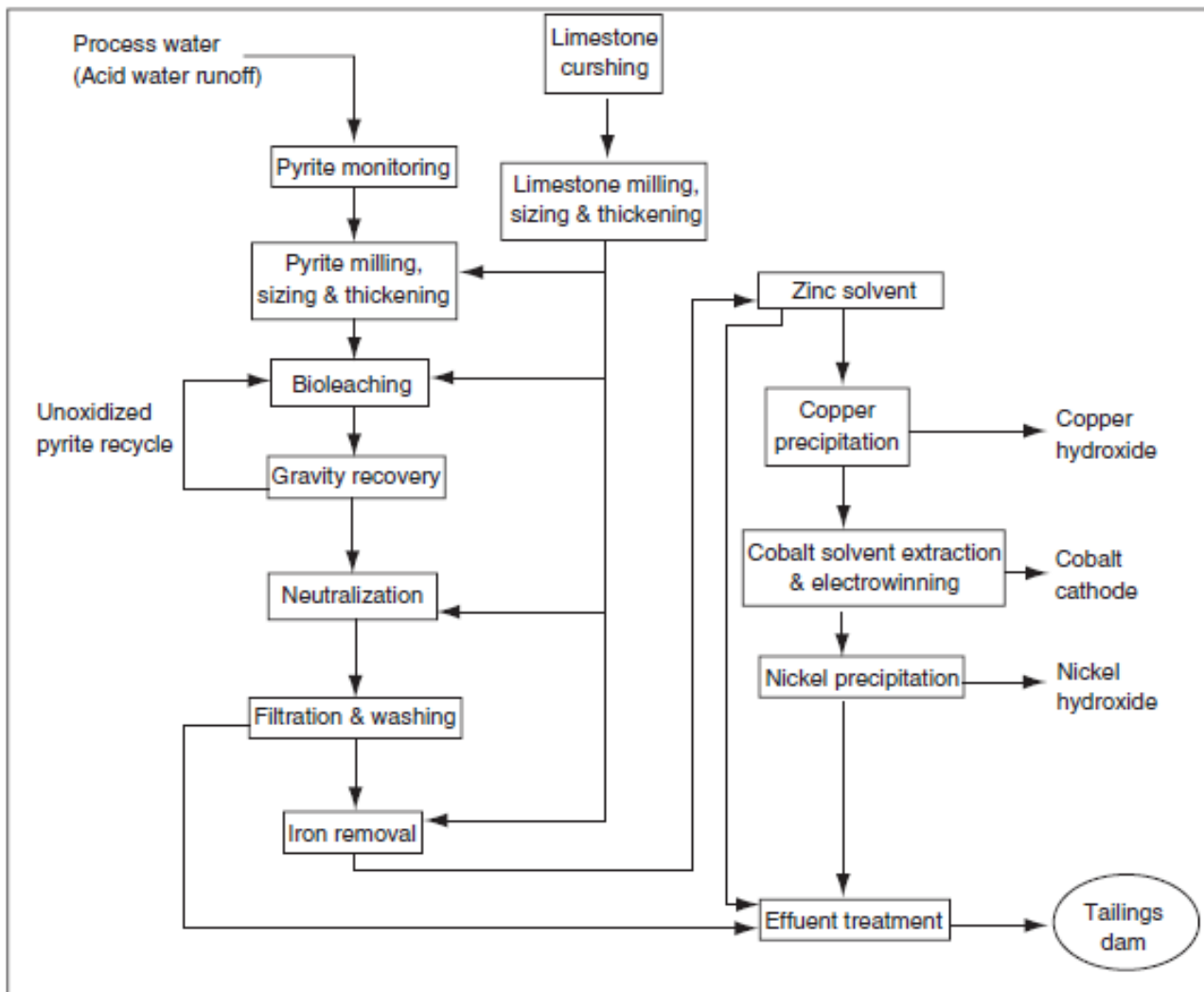

Figure 5: Simplified process flowsheet of the Kasese Cobalt Company plant[19]

The process comprised seven separate stages[19]:

1. Hydraulic reclamation of the pyrite concentrate.

2. Physical preparation of the pyrite concentrate and the limestone.

3. Bioleaching.

4. Iron removal.

5. Solution purification and SX.

6. Cobalt EW and conditioning.

7. Effluent treatment and waste management.

Construction of the plant was being managed by Bateman Minerals & Industrial Limited (BMI) of South Africa. BMI, as the overall Engineering, Procurement, Construction, and Management (EPCM) contractor for the project, used the input from several different technology consultants[17]:

• Signet (Australia) was responsible for pyrite recovery from the stockpile by monitoring, pyrite milling, bioleaching (using BRGM's technology), partial neutralisation and residue filtration.

- Krebs (France) was the process consultant for the package comprising iron removal by hydrolysis and precipitation, zinc removal by SX, copper recovery by precipitation, cobalt recovery into spent electrolyte by SX and nickel hydroxide recovery.

- The process consultant for the package incorporating cobalt EW and cathode conditioning was BMI, in conjunction with Mintek (South Africa), which provided process design input, including a process guarantee.

- BMI, as the EPCM manager, also had responsibility for the limestone quarry, the tailings dam for the final plant residue, plant utilities and infrastructure.

Pyrite concentrate was reclaimed from the stockpiles at a rate of around 245 t/d using high-pressure monitor guns, and pumped to the process plant. The slurry was thickened, and the thickened underflow was pumped to the regrind circuit, where the particle size was reduced to 80 % passing 35 μm. The reground pyrite, classified by hydrocyclones, was thickened to 50 % solids prior to bioleaching. A fraction of the pregnant solution after filtration of the bioleach slurry was recycled to increase the cobalt tenor in the circuit. A concentrated solution of nutrients was introduced to the feed stream, and the pulp density of the feed to the bioleach circuit was reduced to 20 %[19].

Limestone from a nearby quarry was transported to the site, where it was crushed and then milled to 80 % passing 44 μm. The ground limestone was stored in a tank as a slurry at 50 % solids by mass, and was distributed to the bioleach and neutralisation circuits by means of a ring main[19].

The bioleach circuit, depicted in Figure 6, comprised five reactors, each with a volume of 1,380 m^3, arranged as three primary reactors operated in parallel, with one secondary reactor and one tertiary reactor operated in series. The reactors were constructed from 304L grade stainless steel, and pulp transfer was by gravity flow through launders[15,18,19].

Figure 6: The bioleach reactors at the Kasese Cobalt Company plant

The agitation system used in these bioleach reactors was quite different from those encountered in virtually all other commercial-scale bioleach applications, past and present. The industry standard is to utilise high-solidity-ratio axial flow impellers, but the Kasese plant employed a radial flow turbine that was designed by the French mixing company, Robin Industries (now known as Milton Roy Mixing). The agitation system, shown in Figure 7,

was known as BROGIM®, and comprised two upper axial flow impellers and one lower turbine. The turbine was a disc with eighteen vertical flat blades attached to the lower face. Air was injected directly beneath the turbine through four open pipes[15,18,19].

Figure 7: Lower part of the BROGIM® agitation system utilised in the Kasese bioleach reactors[19]

Air was provided by five blowers, which could supply up to 20,000 Nm^3/h of air to each tank. Under normal operating conditions, the air flow rates were between 10,000 and 15,000 Nm^3/h in the primary reactors, and between 5,000 and 10,000 Nm^3/h in the secondary and tertiary reactors[15,18,19]. The operation made use of off-gas analysis to measure both O_2 and carbon dioxide (CO_2) uptake rates in the bioleach reactors, and to determine the O_2 utilisations. The volumetric O_2 uptake rates measured in one of the primary bioleach reactors are shown in Figure 8. Under normal operating conditions, the O_2 uptake rate was typically around 1,375 mg/(L.h), which translates to 33.0 kg/(m^3.d). This is a high value, indicative of the O_2 demand required

for the oxidation of the high-grade pyrite concentrate that formed the feed to the process, combined with the relatively high feed solids concentration of 20 %. Under these operating conditions, the O_2 utilisation in the primary bioleach reactors was greater than 50 %, and the dissolved O_2 concentrations measured near the surfaces of the primary bioleach reactors was between 1.5 and 2.0 mg/L[18,19].
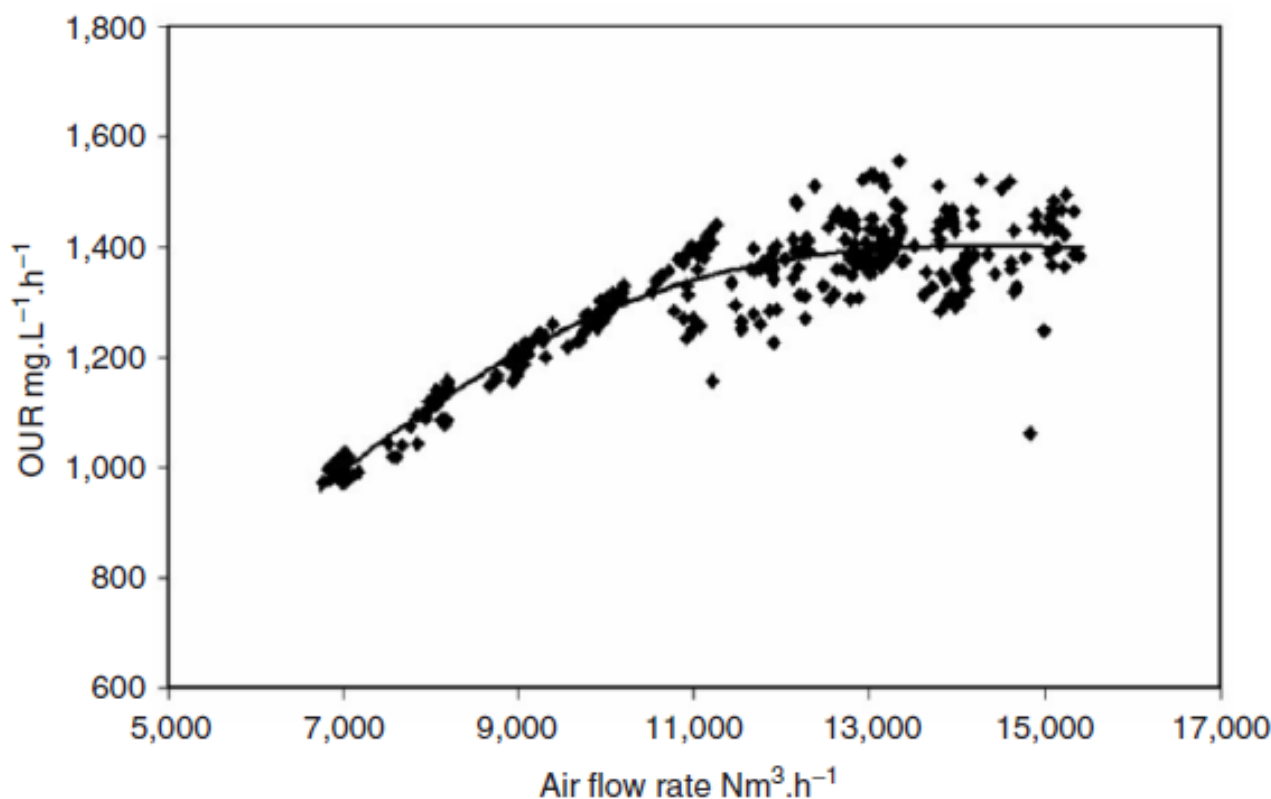

Figure 8: Oxygen uptake rates measured in a primary bioleach reactor at the Kasese plant[19]

The operating temperature in the bioleach plant was 42 °C. Temperature control was effected via stainless steel cooling coils connected to a cooling tower. Each primary tank contained more than 1.5 km of cooling coils for heat removal[15,18,19]. The oxidation of pyrite is highly exothermic, with a heat of reaction of -12,481 kJ/kg, resulting in an overall heat load in the Kasese bioleach plant of around 25 MW[17].

The pH levels in the bioleach reactors were controlled by the addition of limestone slurry at a controlled rate. The pH values were maintained at between 1.4 and 1.5 in the primary reactors, and between 1.5 and 1.7 in secondary and tertiary reactors. The limestone addition served two purposes: to neutralise some of the sulphuric acid produced by the oxidation of the pyrite, and to generate the CO_2 required by the autotrophic bioleach bacteria[15,18,19]. The gas analysis equipment proved valuable in diagnosing the need for a constant supply of limestone to the primary bioleach reactors. In many industrial bioleaching applications, carbonate minerals are usually present in the feed concentrate, which dissolve and liberate CO_2 which is then utilised by the autotrophic bacteria to support their growth. However, the Kasese pyrite contained no carbonates, and a change in the limestone addition procedure improved bacterial growth and stabilised the process[19].

Another interesting aspect of the Kasese bioleach plant was the use of a gravity circuit to recover and recycle unleached coarse sulphide particles from the bioleach residue (Figure 5). The primary objective of this was to improve cobalt recovery by extending the residence time of the coarse unleached fraction of the bioleach residue. A simulation exercise showed that this was possible[22], and so the gravity circuit was included in the flowsheet. The circuit included a hydrocyclone which removed fine particles of iron hydroxide, and a series of gravimetric spirals that produced a concentrate, which was recycled to the feed of the bioleach circuit[19].

Approximately 60 % of the pyrite was oxidised in the primary bioleach stage, with around 20 to 30 % of the oxidation occurring in the secondary and tertiary reactors. The total residence time, accounting for the addition of limestone slurry and the gas hold-up, was about 6 days, and a cobalt dissolution of around 80 % was achieved[19].

Iron was removed in two stages. The first stage comprised a series of three 300 m³ agitated tanks, where limestone slurry was used to increase the pH of the bioleach slurry stepwise to a level of 3. The neutralised slurry was filtered on a 73 m² belt filter. The primary filtrate was pumped to the next iron removal area and the wash filtrate was used to dilute the bioreactor feed to the required density. The washed solid residue from the

filter was re-pulped with raw water and transferred to the effluent treatment area. A second stage of neutralisation of the primary filtrate raised the pH to a level of 5.7, aiming to achieve a final concentration iron of less than 1 mg/L to prevent iron poisoning of the zinc extractant[19].

Zinc was extracted at a pH level of 3 with di-(2-ethylhexyl) phosphoric acid (D2EHPA) diluted in high-flash-point kerosene. Cobalt, copper and nickel that were co-extracted into the organic phase with the zinc were removed by scrubbing, and zinc was then stripped from the scrubbed organic phase using a 50 g/L sulphuric acid solution. Copper was removed from the zinc raffinate by neutralisation in two stages to a pH level of 5 to 7 using a caustic soda solution, to prevent copper poisoning of the extractant in the downstream cobalt SX plant. Cobalt was extracted using Cyanex 272, diluted in high-flash-point kerosene. Nickel co-extracted into the organic phase with the cobalt was scrubbed, and cobalt was stripped from the scrubbed organic phase using a 130 g/L sulphuric acid solution. Finally, precipitation of nickel from the cobalt raffinate at a pH level of 9-10 using caustic soda solution produced a saleable nickel hydroxide product[19].

Cobalt was recovered by electrolysis from the pure solution generated by SX (which contained between 40 and 50 g/L cobalt), and produced a 99.9 % pure product. The cobalt cathodes were crushed, burnished and drummed for shipment[19].

All plant effluents were treated in a single operation. This included the re-pulped cake from the belt filter, strip solution from zinc SX and the overflow solution from nickel precipitation. The effluent was mixed with slaked lime slurry to a pH level of 9 before being transported to the tailings dam. In the tailings dam, the final solid waste products settled, and the water was discharged at a controlled rate to the nearby Rukoki River. Up to 10 % of the outflow was sent to a reed bed for further polishing. This constructed wetland used locally available reeds (*Phragmites mauritianus*). It was observed that a significant amount of the heavy metals and other chemical components were retained in the reed bed (as much as 90 % at low flow rates). The effluent from the wetland was observed to encourage growth of vegetation in the arid 150 ha trail area that had resulted from the runoffs of pyrite prior to implementation of this process on the site[19].

By any measure, the Kasese bioleach plant represented a major step forward in the development of base metal bioleaching outside of copper. At the time of its construction, the 1,380 m³ bioreactors were the largest individual reactor volumes for a plant of this type[2]. By 2002, the Kasese bioleach plant was producing 67 tonnes of cobalt a month, representing 2 % of the world's production at the time[20]. In the words of the bioleach technology developer, BRGM, it was "the first industrial installation incorporating bioleaching into a sophisticated hydrometallurgical flowsheet allowing the selective extractions of various metals", and "as such, in the history of biohydrometallurgy, this operation can be considered as the milestone opening an era of the complex application of bioleaching"[19].

By 2014, the pyrite stockpiles had been treated, the site environment had been restored, and the plant was closed.

BioHeap™: The Mt Sholl/Radio Hill Project, Western Australia

At around the same time that mechanically-agitated bioleaching processes for the treatment of nickel and cobalt concentrates were being developed and commercialised, a process for heap leaching of a nickel-copper sulphide ore was being investigated in Australia. The project proceeded to on-site demonstration scale, marking it as a significant landmark in the unfolding story of nickel and cobalt bioleaching.

The bioleaching technology in question is known as BioHeap™, and it was developed by Titan Resources NL through its 100 % owned subsidiary, Pacific Ore Technology Ltd[23]. The basis of the BioHeap™ technology is a "proprietary" moderately thermophilic culture of predominantly sulphur-oxidising bacteria[12]. The Mt Sholl nickel-copper deposit is located in the Pilbara region of Western Australia. It grades 0.92 % copper, 0.67 % nickel, 11.1 % iron, 4.05 % sulphur, 2.3 % aluminium and 4.0 % magnesium. Sulphide minerals make up about 15 % of the ore, but they are finely disseminated within the host rock. The predominant iron sulphide is pyrrhotite. Pentlandite particles occur in the range of 30 µm to 2 mm, and the chalcopyrite is finer, with 60 to 70 % less than 100 µm. More than half of the chalcopyrite is encapsulated within silicate minerals[12,23].

A substantial test work programme was conducted, which included batch bioleach tests on milled samples to investigate the impact of temperature and pH level on nickel and copper extraction, and a column bioleach test in a 5-metre tall aerated column. In the column test, the redox potential remained in the range of 350 to 420 mV vs Ag|AgCl, which meant that the majority of the soluble iron was in the ferrous form. This necessitated a process to oxidise the iron to the ferric form, prior to its removal. To this end, a 1-metre column was packed with an inert material and inoculated with an iron-oxidising bacterial culture, to evaluate the application of a separate, inert heap for this purpose[23].

Following the completion of the development programme, two 5,000 t ore heaps, each 5 m high, were constructed at the Radio Hill Mine, as shown in Figure 9. The first heap was stacked with Mt Sholl disseminated

ore (with a crush size of 100 % passing 7.5 mm) and equipped with aeration and irrigation lines. The heap was aerated at a rate of 1,000 m³/h. The second, auxiliary heap was stacked with barren rock with a crush size of 100 % passing 50 mm, also equipped with irrigation and aeration lines. This heap was also aerated at a rate of 1,000 m³/h[12,23].

Figure 9: The 5,000 tonne BioHeap™ test heaps at the Mt Sholl/Radio Hill project[24]

A pregnant leach solution (PLS) pond was provided from which the leach solutions could be circulated. The pH level in the pond was maintained between 1.4 and 1.8, and a small liquor treatment plant was built to precipitate ferric iron and a mixed nickel-copper hydroxide or carbonate. A bank of thirty EMEW® copper EW cells was also installed to enable the investigation of copper cathode production from a bleed stream from the liquor after iron removal. The flowsheet is shown in Figure 10[23].

The heaps were each irrigated with acidified pond water until the pH level of the exit solutions was below 2.2. The heaps were each then inoculated via the irrigation systems with batches of 45 m³ of inoculum, which had been cultivated on site at a temperature of between 50 and 55 °C. The heap containing the Mt Sholl ore was inoculated with moderately thermophilic microorganisms, and the auxiliary heap was inoculated with iron-oxidising bacteria to create an iron oxidation reactor. Irrigation of the first heap was then started, and it was subsequently inoculated with a second batch of moderately thermophilic microorganisms. Aeration of the first heap was then initiated, and the solution was circulated between the heap and the pond, to enable build-up of the solution tenors[12,23].

It was planned to control both the iron and nickel concentrations in the pond at 10 g/L. Liquor from the pond was circulated to the waste heap for oxidation of ferrous iron to ferric iron, and the liquor from the base of the waste heap was sent to the iron precipitation tanks, where iron was removed by the addition of limestone. The resultant slurry was filtered, and the filtrate, containing nickel, copper and cobalt and some aluminium was either returned to the pond (if the nickel level was below 10 g/L), or sent for nickel/cobalt precipitation. In practice, the leach rates of both nickel and iron were considerably higher than anticipated, and the precipitation plant was not able to cope. For this reason, irrigation of the first heap was stopped on several occasions for long periods, to enable the precipitation section to catch up[12,23].
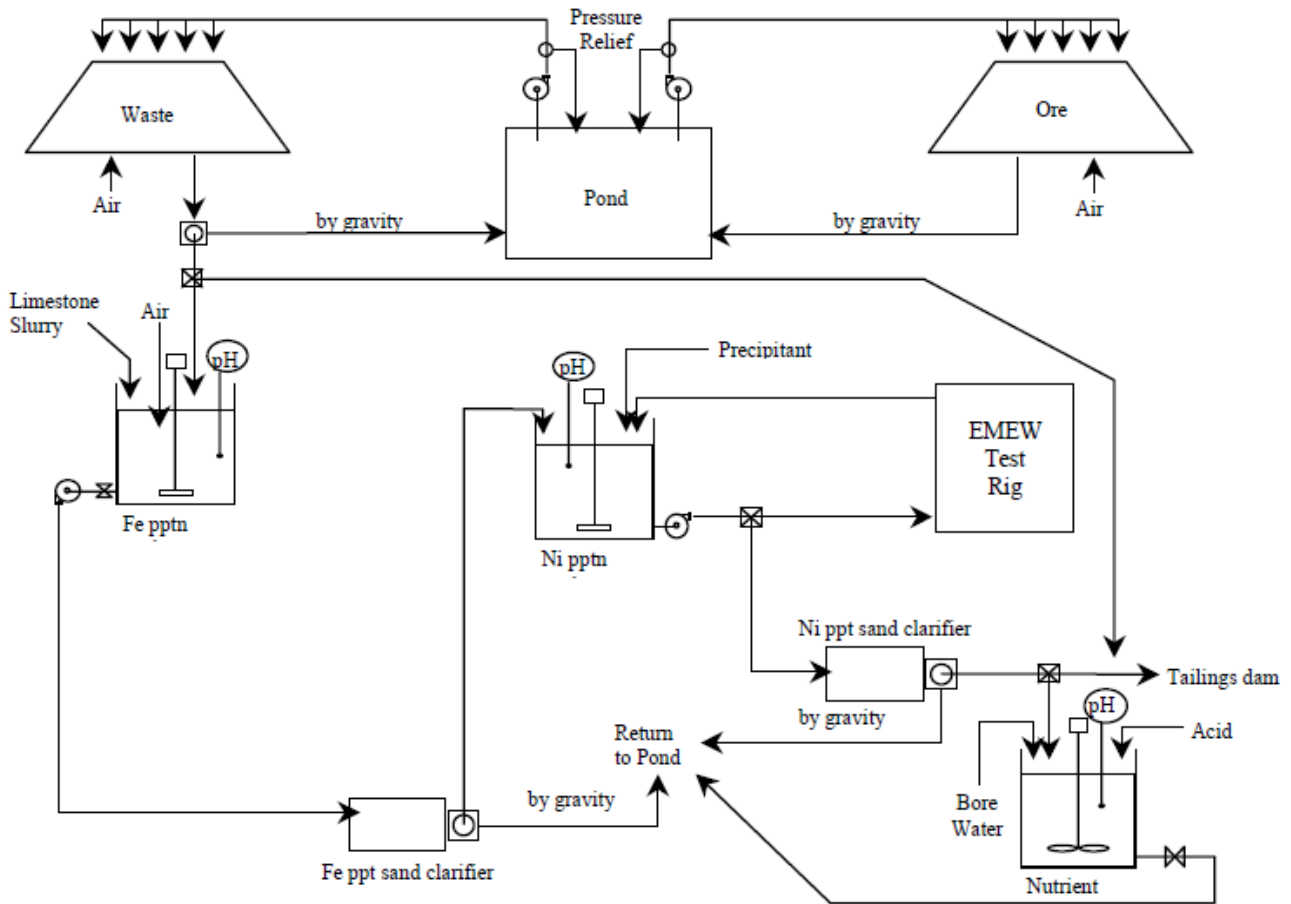
Figure 10: The BioHeap™ pilot plant flowsheet at the Mt Sholl/Radio Hill project [23]

The nickel and copper recoveries obtained from the Mt Sholl ore in the BioHeap™ pilot plant trial at the Radio Hill mine are shown in Figure 11. In a period of about 12 months, nickel and copper recoveries of 90 % and 50 %, respectively, were achieved[23].
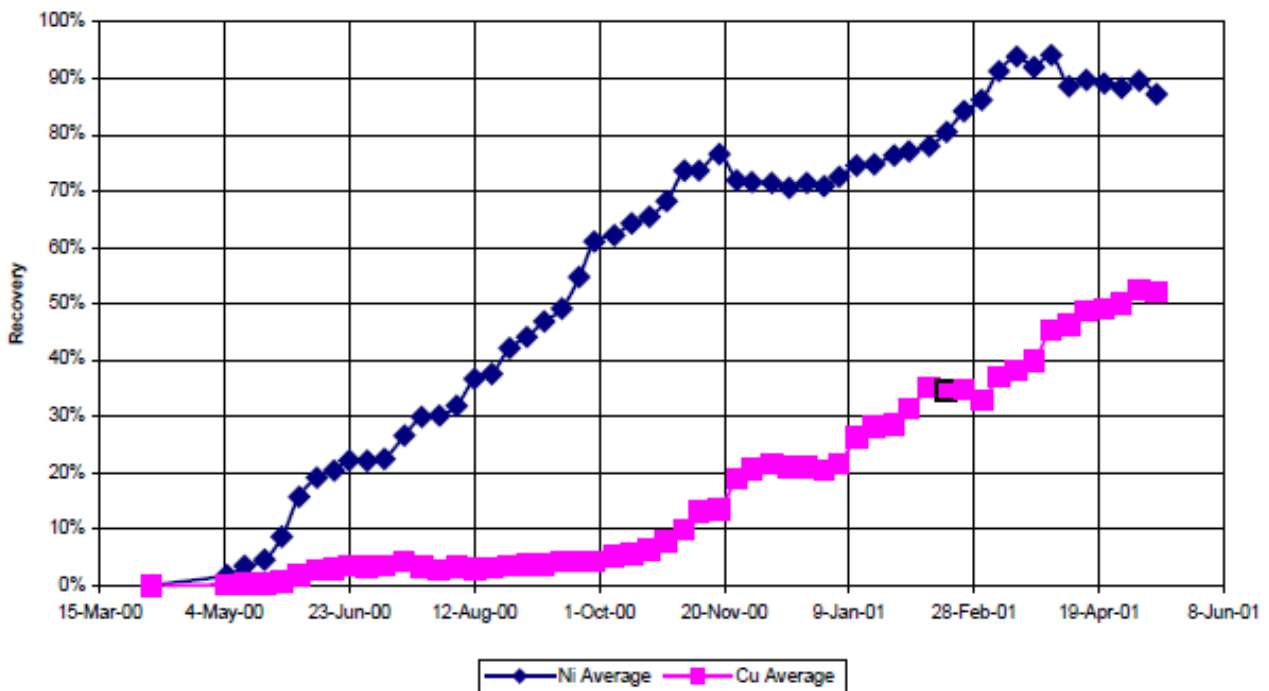

Figure 11: Nickel and copper recoveries achieved in the BioHeap™ pilot plant at the Mt Sholl/Radio Hill project[23]

Figure 12 shows the temperature profiles measured at the top and bottom of the first heap during the pilot plant trial. The early readings show the condition of the heap before acidification, inoculation and aeration. Acidification and inoculation resulted in a sharp decline in the temperatures, and the onset of aeration had little impact on heat generation in the heap. During the first few weeks of irrigation, the heap temperature remained relatively low, although the nickel leach rate was quite rapid. The spike in the temperatures that took place in June 2000 occurred when irrigation was stopped and aeration was continued. During this period, the temperature at the top of the heap reached a maximum of 87 °C. Irrigation was restarted, and the application rate was gradually increased to 10 L/(m^2.h), but the high temperatures persisted. The irrigation rate was temporarily increased to 30 L/(m^2.h), which resulted in the temperature at the top of the heap decreasing during July 2000. Interestingly, the temperature at the bottom of the heap increased at this time, as the heat from the top of the heap was transported downwards through the heap by the higher irrigation rate. The irrigation rate was lowered once ponding was observed on the top surface of the heap. The heap went through a second cycle of raised temperature at the top of the heap, but once the irrigation rate was stabilised, the temperatures also stabilised at about 40-50 °C. The generation of heat, particularly during the periods when irrigation was suspended, was almost certainly primarily caused by the exothermic oxidation of pyrrhotite, although the warm climate at the site probably also contributed to the favourable temperatures attained in the heap[12,23].
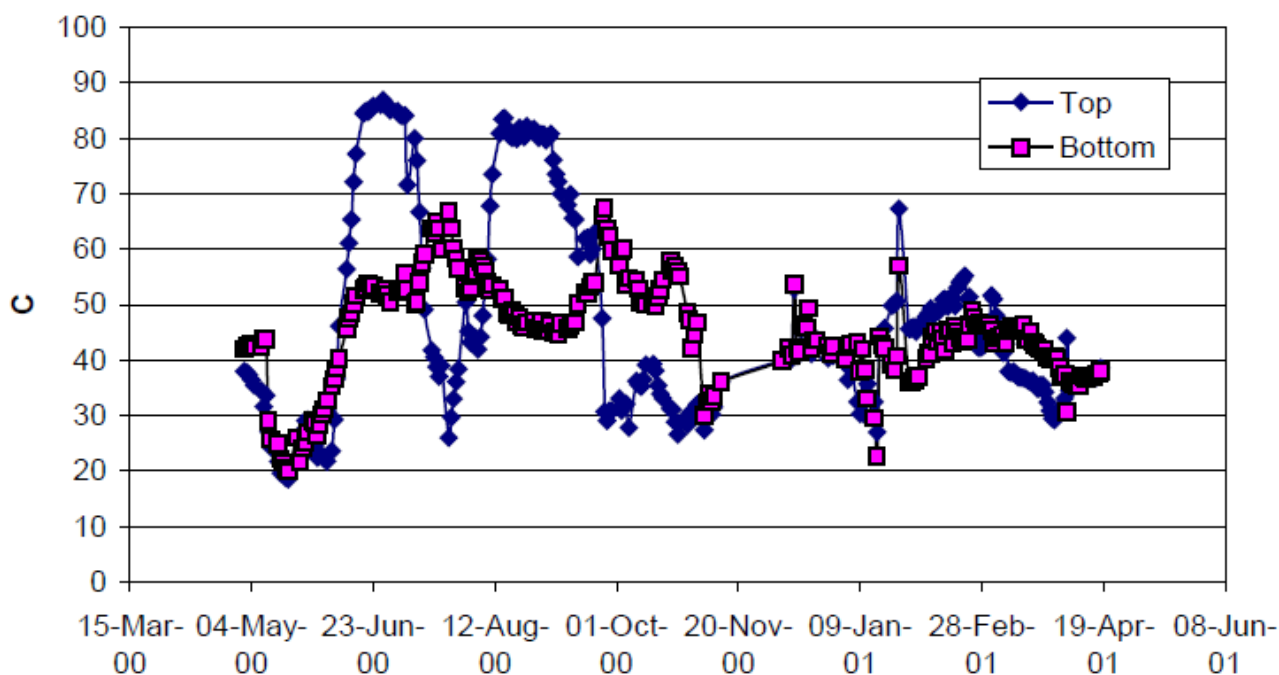

Figure 12: Temperature profiles measured in the BioHeap™ pilot plant at the Mt Sholl/Radio Hill project[23]

The field trial was considered "a great success, yielding results beyond expectation"[23]. This project was the first demonstration at a significant scale of the application of heap bioleaching for the extraction of nickel and copper from an ore containing base metal sulphides such as pentlandite and chalcopyrite. The Mt Sholl ore reportedly had a relatively low acid consumption (although the acid consumption value was not reported). However, the process did not proceed to commercialisation owing to a lack of available suitable deposits with sufficiently low acid consumption[24], but it certainly laid the groundwork for future endeavours in a very different location on the other side of the planet.

The Talvivaara/Terrafame Project, Sotkamo, Finland

The tale of Talvivaara which became Terrafame is one of two halves. It is a fascinating saga that encompasses science, technology and engineering, politics, law and even the arts.

The Talvivaara deposits in Sotkamo, Finland are the largest nickel sulphide deposits in Europe[12,25]. In 2007, the resource was estimated to be 340 million tonnes[25]; by 2009 it had grown to 642 million tonnes[26], by 2011 it was 1,121 million tonnes[27] and in 2019 it was estimated to be 1,550 million tonnes[33]. The deposit comprises two different polymetallic orebodies hosted by a black schist, Kuusilampi and Kolmisoppi[26,33]. The complex ore has average grades of 0.27 % nickel, 0.56 % zinc, 0.14 % copper, 0.02 % cobalt, 10.3 % iron, 8.4 % sulphur, 7.2 % carbon and about 50 % silicon. The main minerals are pyrrhotite, pyrite, pentlandite, sphalerite, violarite, chalcopyrite and graphite[25].

Sotkamo is located in eastern Finland, at a latitude of 64° north, 350 km south of the Arctic Circle. The temperature in the winter months ranges between 0 and -20 °C, with occasional short cold spells of below -30 °C in January and February. The average annual snowfall is 0.7 m[27].

Demonstration Heap Trial

Owing to the complex and low-grade nature of the Talvivaara ores, it was concluded that exploitation of the deposits was not economically viable using conventional techniques[27]. The application of bioleaching for the treatment of the Talvivaara ore was studied extensively over a period of two decades. This included laboratory-scale test work and column leaching, and culminated in the construction by the Talvivaara Mining Company of a 17,000 tonne demonstration heap on the site in 2005, as shown in Figure 13. The heap was 30 m wide, 60 m long and 8 m high. It was placed on an impermeable 2 mm thick high-density polyethylene (HDPE) liner. The liner was protected on both sides with a 100 mm layer of sand. The HDPE liner and the protecting layer were covered with a 500 mm thick drainage layer, comprised of material between 4 and 100 mm. Drainage pipes were installed in the layer, and aeration lines were installed above the drainage layer[25].
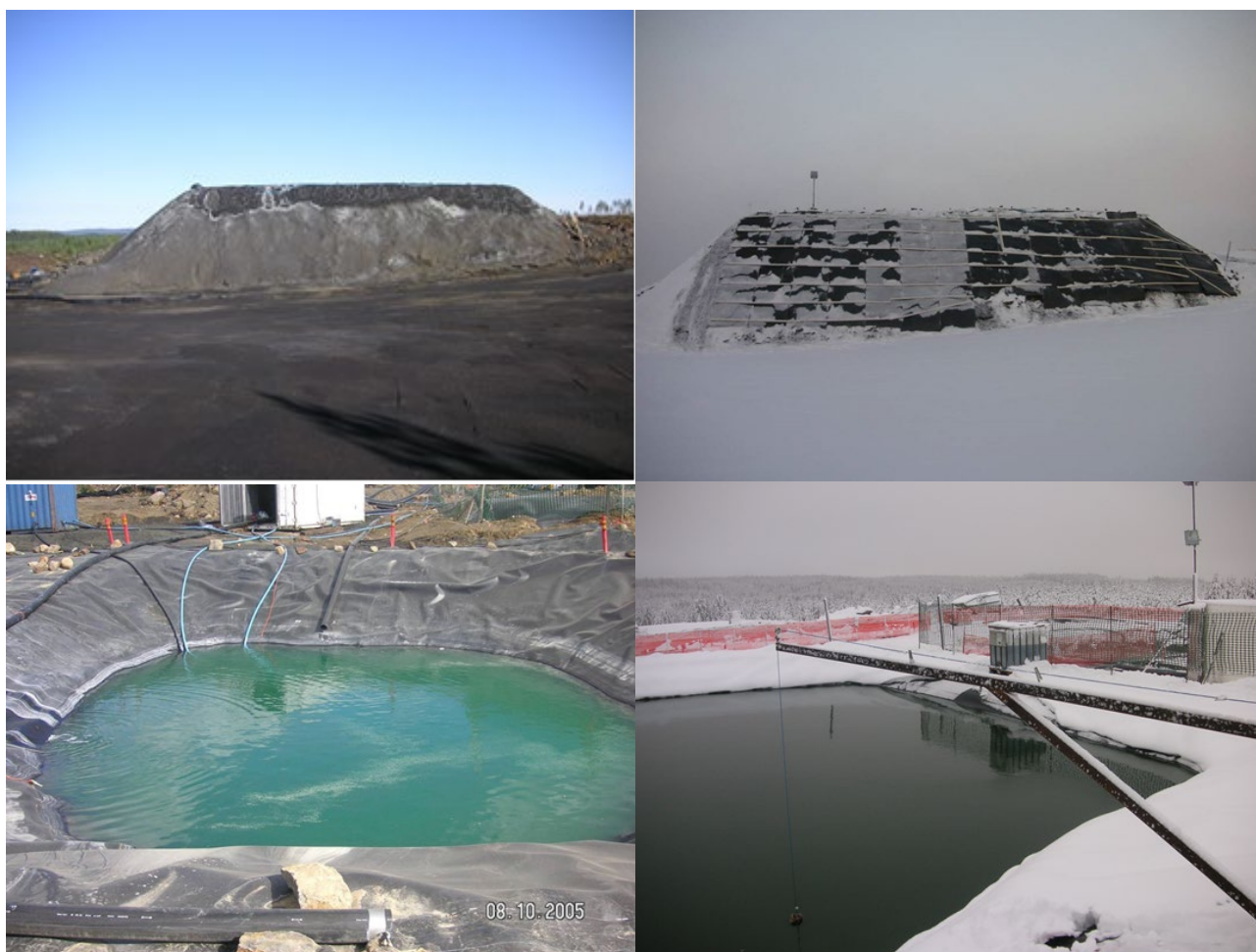

Figure 13: Heap bioleaching demonstration plant at Talvivaara, showing the extremes of climate experienced at the site[25]

HDPE-lined PLS ponds were also provided, and a network of HDPE piping was installed allowing the PLS to be pumped between the ponds and the heap. The ore was crushed to 80 % passing 8 mm, with less than 10 % finer than 250 μm. Before the demonstration heap was constructed, column experiments were performed at particle sizes of 6, 8, and 12 mm. The 8 mm particle size was selected because nickel recovery with the 12 mm particle size was 45 % less than that attained at a crush size of 8 mm. The ore was agglomerated in an agglomeration drum, using sulphuric acid and a site-prepared inoculum solution[25,28].

The heap was irrigated at a rate of 5 L/(m2.h) with a solution at a pH level of 1.8. The temperature of the irrigant solution ranged between 4 and 20 °C. The temperature within the heap varied between 30 and 90 °C, and the PLS temperature was between 40 and 50 °C. The elevated temperatures were maintained during the winter months, driven by the exothermic oxidation of the pyrrhotite in the ore. Over a period of 500 days, the following recoveries were achieved: nickel, 92 %; zinc, 82 %; cobalt, 14 % and copper, 2.5 %. The low copper recovery was attributed to the fact that the copper mineral is chalcopyrite. Sulphuric acid consumption was 16 kg/t, and the rate of acid consumption decreased towards the end of the primary leaching phase[25].

In February 2007, the primary demonstration heap was reclaimed and restacked to a secondary heap, and the bioheapleaching process was started again. Secondary leaching was continued until November 2008, when the pilot heap had to be reclaimed as the Kuusilampi open pit was expanding to the site of the heap. The final recoveries after 21 months of secondary leaching were: nickel, 99 %; zinc, 99 %; cobalt, 35 % and copper, 22 %[28].

The main reason for the secondary leaching phase is to obtain better recoveries of copper and cobalt. Copper occurs as chalcopyrite, and most of the cobalt occurs in pyrite. Sulphide minerals have varying semiconductor properties, and galvanic interactions occur when there is contact between mineralogical phases. During dissolution of a mineral assembly of different sulphides, the minerals with the highest rest potentials are galvanically protected and their leaching is hindered until the minerals with lower rest potentials have been leached. The electrochemical rest potentials of chalcopyrite and pyrite are higher than those of pyrrhotite, pentlandite, and sphalerite. Placing the ore on a secondary heap provides sufficient time for the almost complete removal of the latter sulphides, allowing the rate of chalcopyrite and pyrite leaching to increase. Transferring the material to a secondary leaching stage also enhances the recovery of metals from those parts of the primary heaps where the leaching solution has poor contact with the ore particles. Such areas include the slopes of the heaps, and areas within the bulk of the heap where solution flow was not ideal[28].

The successful execution of the demonstration heap trial paved the way for the commercial implementation of the process.

Commercial Implementation

The environmental permit for the commercial operation was granted in March 2007, and the construction phase of the project commenced shortly after that. In April 2008, mining began at the Kuusilampi open pit, and in July 2008 the bioheapleaching process was initiated. The first metal sulphides were produced at the plant in October 2008.

The process flowsheet for the Talvivaara operation is shown in Figure 14, and a view of the entire processing facility is shown in Figure 15. The mining method is large-scale open pit mining. Materials handling covers all the physical ore processing steps from the primary crusher to the heaps. After 13 to 14 months of bioleaching on the primary heap leach pad, the leached ore is reclaimed, conveyed and re-stacked onto the secondary heap leach pad. After secondary leaching, the barren ore remains permanently in the secondary heaps. In the metals recovery process, the metals are precipitated from the PLS using gaseous hydrogen sulphide. The resulting products are intermediates which are transported for further processing in various refineries[27].
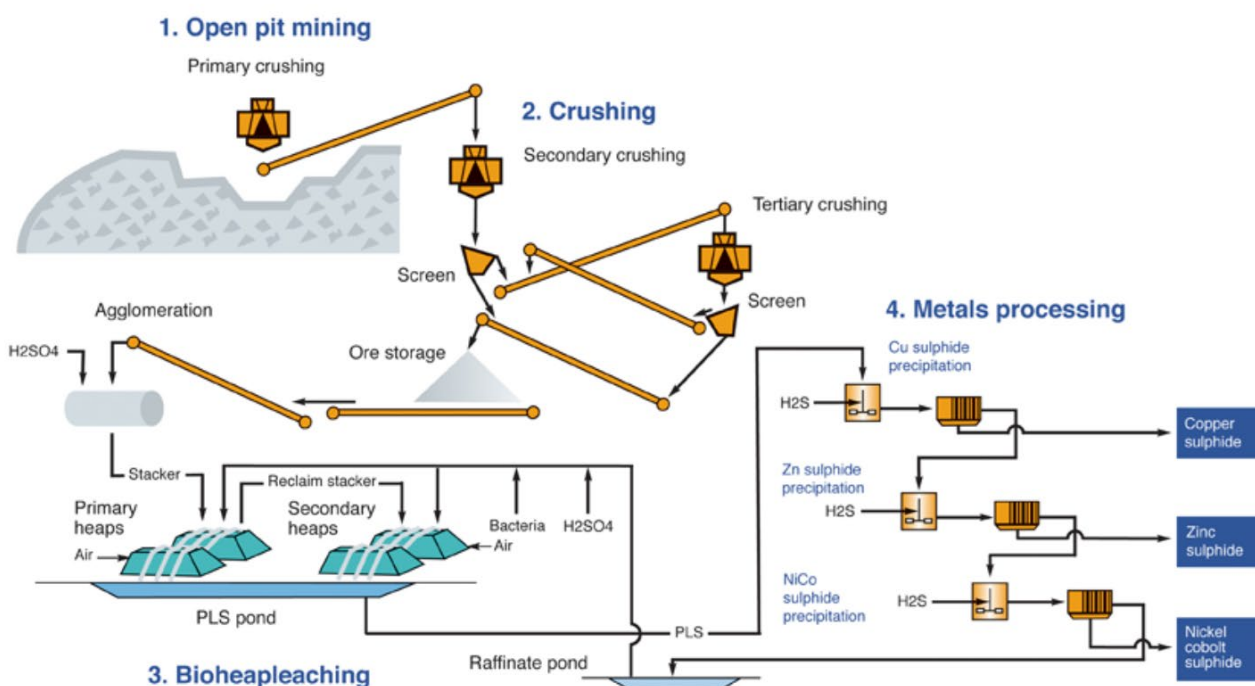

Figure 14: Process flowsheet for the Talvivaara heap bioleaching operation[25,27,28,29]

The plant was designed for an annual ore production of 24 to 25 million t/a[27,28,29]. After primary crushing, the ore is conveyed to a fine crushing station, where it is crushed and screened in three further stages to 80 % passing 8 mm. The crushed ore is then agglomerated in a rotating drum, where PLS is added to the ore in

order to consolidate the fine ore particles with the coarser ones. After agglomeration, the ore is conveyed and stacked as 8 m high heaps on the primary heap pad (Figure 16, Figure 17). The primary heap comprises of two heap pads, each 2,400 m long and 400 m wide. The heap pad sections are further divided in four sectors. The heap pad is equipped with piping, laid on the bottom of the pad, through which 32 low-pressure fans supply air to the stacked ore at a rate of 0.08 $m^3/(t.h)$. From the top, the heap is irrigated using drip irrigation at a rate of 5 $L/(m^2.h)$ with leaching solution, which is collected from the bottom of the heap via a drainage system and discharged to the PLS collection ponds. Once a sufficient metal tenor is attained, a side stream of about 10-20 % of PLS is pumped to the metals recovery plant, while the rest of the solution is circulated back to the heap. Either fresh or purified process water is added to the circulation to maintain the process water balance and appropriate leaching conditions. The pH value of the irrigation solution is adjusted with sulphuric acid[25,27,28].

Figure 15: The Talvivaara/Terrafame processing facility near Sotkamo in Finland

Figure 16: A view of the primary heap leach pad and the PLS ponds at the Talvivaara heap bioleaching plant[28]

Figure 17: A panoramic view of the primary heap leach pad at the Talvivaara heap bioleaching plant

After 13 to 14 months of primary leaching, the leached ore is reclaimed, conveyed, and re-stacked onto the secondary heap pad, where it is leached further. After the first stacking cycle, the ore is reclaimed using excavators and conveyed to the secondary leach pad for further leaching. While the excavators move ahead reclaiming the ore, the stacker follows behind them, stacking new ore from the open pit. This cycle is continuous. The irrigation rate for the secondary heaps is 2 L/(m².h). At the completion of the secondary leaching phase, which is about 3½ years, the barren ore is intended to remain permanently in the secondary heap area. The secondary leaching pads are constructed on top of waste rock dumps, which reduces earthwork quantities, the final footprint of the operation and the rehabilitation costs. The secondary pads are stacked with four 15 m lifts, and the 60 m high heap will eventually be covered and re-vegetated[25,27,28].

The PLS from the heaps is analysed for pH level, redox potential, acidity, total iron concentration, ferrous iron level and metal concentrations. Solid samples are collected at different depths and locations throughout the heap leach period and are assayed for metal concentrations and mineralogy. Temperature measurements are made at various depths and locations throughout the heap. The rates of irrigation, PLS removal and aeration are continuously monitored, and modified as needed[25,27,28].

In the metals recovery process, the metals are precipitated from the PLS using gaseous hydrogen sulphide and pH adjustment. The resulting products are intermediates: copper sulphide, zinc sulphide and a mixed nickel-cobalt sulphide. These intermediates are transported to various refineries for further processing[27,29].

The metal recovery flowsheet is shown in Figure 18, and includes the following unit operations:

- Precipitation of copper sulphide.
- Precipitation of zinc sulphide.
- Pre-neutralisation and aluminium removal.
- Precipitation of mixed nickel/cobalt sulphide.
- Iron removal.
- Final precipitation.

The sulphide precipitation lines for the different metals are very similar. The process flow is divided into two parallel working trains of equal size[27].

Solid wastes are produced in the pre-neutralisation, iron removal and final (total) precipitation steps. In pre-neutralisation, the acidity in the feed and the acid formed during copper and zinc sulphide precipitation are neutralised. A considerable proportion of the aluminium in the solution is also precipitated. The underflow from the pre-neutralisation thickeners is filtered on belt filters, and the inert precipitate is transported to the waste rock area as a dry filter product. Iron is removed from the solution as goethite and/or hydroxide, which is pumped to the gypsum pond. After iron removal the solution goes to the final precipitation where the pH is raised to a value of 10 with slaked lime. The final precipitation is the outlet for manganese and magnesium in the process. Remaining iron and gypsum also precipitate in this step. The remaining metals precipitate as hydroxides. The underflow from the final precipitation thickeners is pumped to the gypsum pond. The overflow from the thickener is recycled back to the heap leaching circuit[29].

Some of the reagents used are produced on site, and the circuit includes two hydrogen plants, two hydrogen sulphide plants, a grinding and elutriation plant for lime and slaked lime, and an O₂ plant[27].
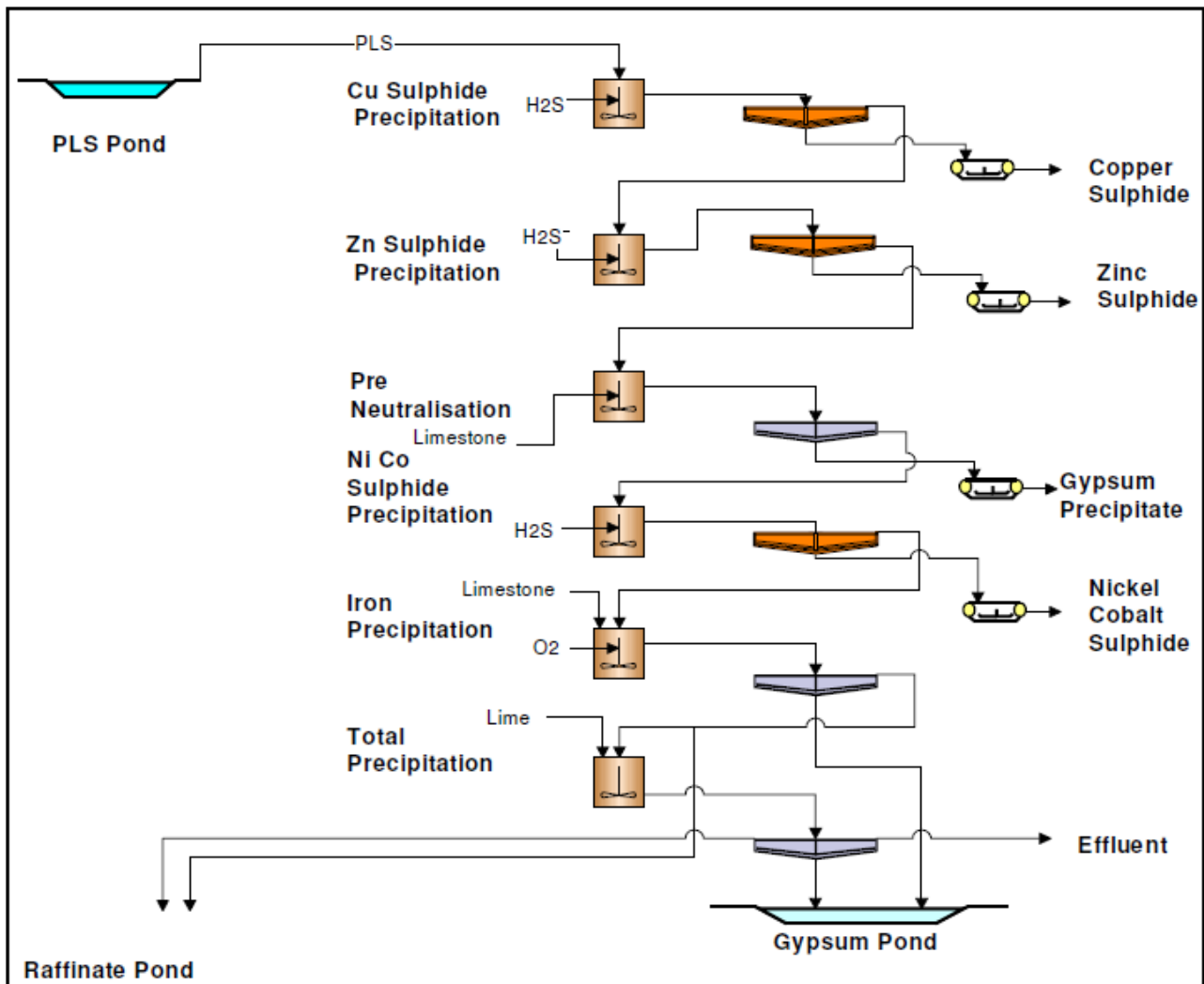
Figure 18: Metal recovery flowsheet at the Talvivaara heap bioleaching plant[27,28]

Production

The planned nickel production of 50,000 t/a was anticipated to be reached in 2012, with additional production targets of 90,000 t/a of zinc, 15,000 t/a of copper and 1,800 t/a of cobalt. Several challenges were experienced which resulted in considerably lower production. By 2011, the production volumes achieved were 16,087 t of nickel (2010: 10,382 t, 2009: 735 t) and 31,815 t of zinc (2010: 26,462 t, 2009: 3,133 t)[29].

The crushing processes caused the biggest problems in in materials handling area of the process plant. The primary crusher had to be fitted with a new mantle. The fine crushing circuit had to be redesigned and more crushers had to be added[29].

The installation and commissioning of the primary heap reclaiming and secondary heap stacking systems presented a major challenge. Both systems were started up in autumn 2010, and the secondary stacker worked well. Commissioning of the primary heap reclaiming equipment was slower, resulting in reduced overall crushing and stacking output[29].

The establishment of the first heap sector proceeded slowly due to the difficulties in the crushing plant. Irrigation and aeration could only be started when about 25 m of heap had been built, and the agglomerates started to break apart during that time. The particle size was also smaller than planned. When irrigation and aeration were started, there were difficulties in getting an even flow of air through the heap. With improved crushing systems, the bioheapleaching started to progress according to expectations. The primary heap was fully stacked for the first time in November 2010, and secondary leaching started with good results[29].

Following commissioning, the metal plant operations faced a series of technical challenges, ranging from a hydrogen plant failure to insufficient hydrogen sulphide capacity caused by installation faults in the hydrogen sulphide generator. Also, hydrogen sulphide emissions forced production levels to be restricted for several months because of odour discharges. That was solved through the use of hydrogen peroxide as an odour controlling chemical, and in 2013 a catalytic scrubber for the hydrogen sulphide gas was installed[29].

The quality of the zinc and nickel/cobalt sulphides was good. The nickel/cobalt sulphide contained about 50 % nickel and 1.5 % cobalt, and the moisture content was below 20 %. The zinc sulphide contained about 64 % zinc, and its moisture content was about 10 %. The impurity levels in both sulphide products were very low[29].

Bankruptcy and Beyond: Talvivaara Becomes Terrafame

By 2014, the Talvivaara operation was facing serious operational and environmental challenges. The production delays, a requirement for further construction that resulted in a major cessation of ore mining between 2012 and 2013, combined with several leaks of metal-contaminated tailings, which threatened local waterways, drove the company to bankruptcy[30,32]. Members of the management were charged with criminal environmental offenses. The story of Talvivaara even became the subject of a fictionalised film (Jättiläinen in Finnish, The Mine in English) that was released in 2016[45]. There was considerable government involvement, and the asset was acquired by Terrafame in 2015, after which it went into a new commissioning phase[30].

The fortunes of the operation have improved substantially in the years after the acquisition by Terrafame. Several practical factors have been addressed to improve the operability of the plant. One of the major complications is that the material hardens significantly during primary leaching, so much so that it fuses together and loses its granularity entirely (Figure 19), making re-mining of the ore after primary leaching difficult.

Figure 19: The Terrafame ore after primary leaching, illustrating how it fuses and loses granularity

The original single bucket wheel excavation system was found to be inadequate for the re-mining task, as the wear on the buckets was substantial, the force needed to break the material was too high, and the material formed large boulders, creating issues for the belts and stacking on the secondary leaching pad. The first solution was to install several additional excavators, and then load the partially liberated lumps with a front end loader to a crushing feeder, from where the material continued to the conveyor system. Although this worked, a large fleet of excavators was needed, which was costly. Also, the size of the particles sent to secondary leaching remained much larger than desired[30].

To solve this issue, Terrafame developed mobile surface mining as a more feasible solution for the second round of reclaim. The surface miners used are similar to asphalt breakers, with a rotating spiked drum crushing the surface of the heap. This allows the material to be broken into sufficiently small chunks before it is fed to the conveyors, removing the need for intermediate crushing feeders. A disadvantage of this approach is that an area of the active heap is sacrificed as a work area[30].

Several other practical aspects of the operation were optimised and improved. These included repositioning of the aeration and drainage pipes, and optimising the movement and stacking of the ore after re-mining of the primary leach pad, to minimise downtime.

As the operation of the heaps has improved, the heat formation inside the heaps has also increased considerably. As seen in Figure 20, the outlet temperatures of the primary heap solutions were on average 15 to 30 °C higher than the ambient temperature in 2016, and there was considerable variation in the heat formation between different heaps. In 2017, the temperature difference between ambient temperature and the outlet solutions increased to 40 to 50 °C. This trend continued in 2018, with the heap outlet temperatures reaching 50 to 60 °C higher than the ambient temperature. Therefore, even though the extreme winter conditions result in ambient temperatures as low as -35 °C, the heaps do not freeze.
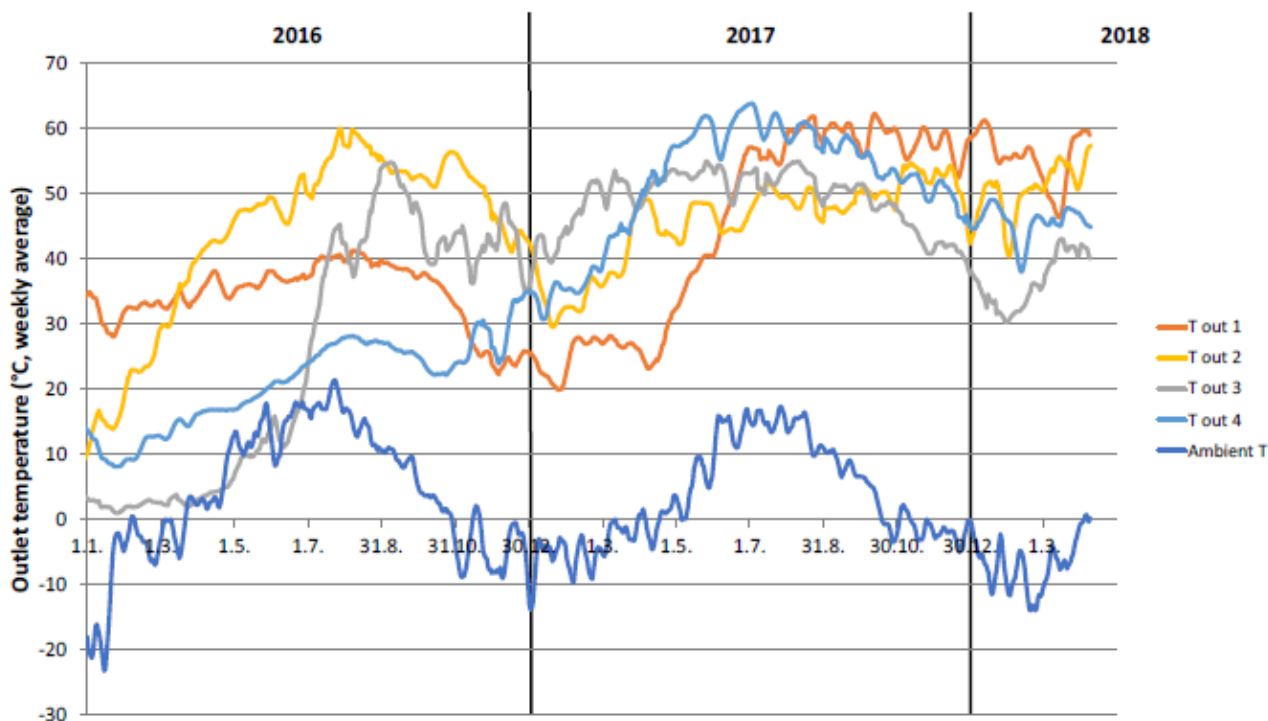

Figure 20: Primary heap outlet temperatures at the Terrafame heap bioleaching plant between 2016 and 2018[32]

In the period between September 2015 and early 2017, the third complete dynamic cycle of re-mining and re-stacking of the leach pads was completed[30]. As a result, in 2017, 20,864 tonnes of nickel and 47,205 tonnes of zinc were produced, and the net sales of Terrafame more than doubled from the previous year to EUR 218.8 million. In addition, a new quarterly record of 13,772 tonnes was achieved in zinc production and the first commercial deliveries of copper were made at the end of 2017.

By 2018, the mine's production targets for nickel and zinc were revised downwards to 35,000 and 75,000 t/a, respectively, with the aim of achieving those targets in 2020 or 2021[32]. By 2018, net sales increased to EUR 325.8 million, with nickel production of 27,377 tonnes, and zinc production of 61,608 tonnes[46].

In 2019, the annual production targets were again decreased to 30,000 tonnes of nickel and 60,000 tonnes of zinc. Net sales in the year decreased slightly to EUR 310.4 million. Nickel production increased to 27,468 tonnes, while zinc production decreased to 55,222 tonnes[47].

Future Plans

There are two strategic developments under way at the Terrafame operation: uranium production and the establishment of a battery chemicals plant[32]. Terrafame submitted a permit application to the Finnish Government in October 2017 for uranium recovery. In February 2020 the Finnish government granted a uranium extraction permit to Terrafame, but the decision has been appealed, and it is expected that it will take a further two years before a final decision is made. After that, the process of getting the uranium extraction plant ready for production will take approximately one year[48].

In November 2017, Terrafame announced a plan to invest in a battery chemicals plant to produce nickel and cobalt chemicals to be used in the electric vehicle (EV) industry. The decision was based on forecasts of growing demand for EV batteries, coupled with an indication that the share of nickel in battery applications is also increasing. The company estimated that by 2030, approximately 25 % of new cars world-wide will either be electric or hybrid vehicles. The plant is intended to have an annual production capacity of about 150,000 tonnes of nickel sulphate and 5,000 tonnes of cobalt sulphate, making Terrafame one of the largest nickel sulphate producers in the world[32]. The plant is now under construction, and commercial production is

planned to commence in 2021. The plant's nickel sulphate capacity will be enough for approximately one million electric vehicle batteries per year, with a cobalt sulphate capacity for approximately 300,000 electric vehicle batteries per year[47].

The remarkable story of Talvivaara and Terrafame, a tale of two halves, from apparent failure to commercial success, highlights one of the most notable application of biohydrometallurgy for the treatment of nickel and cobalt sulphides.

The Mondo Minerals Project, Vuonos, Finland

The journey through the landmarks of nickel-cobalt bioleaching began in the southern hemisphere, describing several projects in Australia and Africa, and then jumped to the far northern hemisphere in Finland. The journey ends just a little less than 200 km south of the Terrafame heap bioleaching operation, at a place known as Vuonos, not far from the more well-known town of Outokumpu. Vuonos is home to a talc mine that produces a nickel-rich pyrrhotite concentrate as a by-product of a flotation process. Previously owned by Mondo Minerals (Mondo), but more recently acquired by Elementis plc, the Vuonos mine is one of two mines that produce such concentrates, the other located very close to the Terrafame operation, in Sotkamo.

The high-grade sulphide flotation concentrates from these operations contain pyrrhotite, pentlandite, pyrite, gersdorffite and magnesite. Nickel is the main metal of interest, but the concentrates also contain a small amount of cobalt, and a significant quantity of arsenic. Previously, the concentrates had been sold to toll smelters, but the arsenic content has made this option less attractive. Mondo therefore elected to produce a value-added nickel product to enhance its revenue and profitability streams, and avoid environmental liabilities. Mondo selected Mintek's proprietary bioleaching technology as the most suitable for the recovery of nickel and cobalt from the side streams[36,37].

Metallurgical Test Work

A two-year long test work programme, which developed and successfully demonstrated the application of Mintek's technology to treat the by-product from Mondo's talc production process, formed the basis for a feasibility study, which showed that bioleaching, followed by a nickel- and cobalt-precipitation process, was an economically-viable option for Mondo to derive value from the by-product. An important aspect of the process is that it includes the production of a stable arsenic-bearing waste, suitable for impoundment.

The metallurgical test work programme conducted by Mintek incorporated regrinding, bioleaching, iron and arsenic precipitation (and stability testing of the product), and nickel-cobalt (or mixed) hydroxide precipitation. Additional magnetic separation and flotation test work was conducted by the Geological Survey of Finland (GTK), and ancillary test work was conducted by various vendors. For example, milling test work was undertaken by FLSmidth, and settling and filtration test work was performed by Tenova Delkor (in some instances, using materials produced in the metallurgical test work programme, and in others, using synthetic solutions).

Three phases of test work were undertaken. In Phase 1, bioleach amenability and optimisation tests were carried out in laboratory-scale, continuously-operated reactors. The effect of process parameters such as plant residence time, feed solids concentration and grind size of the concentrates on metal extraction was evaluated. The results of these tests were used to define the bioleach operating parameters for Phase 2, which was performed in a continuously-operated bioleach mini-plant. The product from the mini-plant was collected and used to demonstrate that iron and arsenic could be removed from the bioleach slurry with minimal nickel loss, to confirm the stability of the neutralised iron- and arsenic-bearing precipitates, and to demonstrate the production of a high-grade mixed hydroxide precipitate (MHP) with a combined nickel and cobalt content of more than 40 %. A continuous mini-plant test was undertaken for the iron/arsenic precipitation step, employing a recycle for seeding and limestone ($CaCO_3$) as the neutralising agent.

It was subsequently determined that rejection of some of the pyrrhotite contained in the concentrates would result in a significant reduction in the size (and therefore the cost) of the bioleach plant. Pyrrhotite rejection was undertaken by magnetic separation, with minimal nickel losses. The rejected pyrrhotite had a low nickel content, and was considered to be a saleable product. Additional upgrading of the concentrate by flotation was also undertaken, to reject talc and magnesite from the concentrate. In Phase 3, the suitability of bioleaching to recover nickel and cobalt from the concentrate after pyrrhotite rejection was demonstrated in laboratory-scale, continuously-operated reactors and a continuously-operated bioleach mini-plant. Additional test work was also conducted on a synthetic solution to confirm the design of the iron/arsenic precipitation process, and the solid-liquid separation process that followed. This was deemed necessary because the upgraded concentrate (after magnetic separation and flotation) had a significantly higher nickel and iron content, and so the concentrations of these metals in the bioleach product were also higher.

In Phase 1 of the metallurgical test work programme, it was shown that:

- Nickel and cobalt extractions of 95 % could be obtained in a five-stage continuous plant, at an overall residence time of 7 days, a feed grind size of 80 % < 30 μm, and a feed solids concentration of 15 %.

- Increasing the feed solids concentration to 17.5 % resulted in a marginal decrease in the nickel and cobalt extractions, to 93.5 % and 94.0 %, respectively.

- Reducing the feed grind size to 80 % < 20 μm improved the nickel and cobalt recoveries to 97 % and 95 %, respectively.

- There was a linear relationship between the level of sulphide oxidation and the level of nickel dissolution obtained in the bioleach process.

In Phase 2, the following was demonstrated:

- Nickel and cobalt extractions of 94 % could be obtained in a four-stage continuous plant, at an overall residence time of 7.2 days, a feed grind size of 80 % < 20 μm, and a feed solids concentration of 17.5 %. The slightly lower extractions in this test were ascribed to the mini-plant system having four and not five stages, and a greater potential for some short-circuiting of solids in the gravity overflow system used in the mini-plant. By contrast, the laboratory-scale reactor system used in Phase 1 employed pumps to transfer the pulp between the reactors.

- Iron and arsenic removal of over 95 % could be achieved in a six-stage neutralisation plant, operated at a temperature of 35 °C, with the pH level being controlled at between 3.0 and 3.5. No nickel and cobalt losses were observed.

- The precipitated product did not release nickel, iron, cobalt or arsenic, and would be classified as regular waste according to the European Standard EN 12457-3 stability testing procedure.

- A MHP containing around 42 % nickel, 2.4 % cobalt, and between 1.8 and 2.0 % magnesium could be produced at a precipitation pH level of between 7.0 and 7.8.

- A two-stage approach for MHP production was recommended. Between 80 and 90 % of the nickel and cobalt would be precipitated in the first stage, to minimise contamination of the MHP precipitate with unreacted MgO, and therefore to maximise the nickel grade. Following a solid/liquid separation step, the nickel and cobalt remaining in the barren liquor would be recovered by precipitation with lime (CaO), and the resultant solids would be recycled to the iron/arsenic precipitation process, where the valuable metals would be re-dissolved.

In Phase 3, the following was demonstrated for the upgraded concentrate blend:

- Nickel and cobalt extractions of 97 % and 98 %, respectively, were obtained at an overall residence time of 7 days, a feed grind size of 80 % < 20 μm, and a feed solids concentration of 15.0 %.

- In this test, very high soluble metal concentrations were measured – the total metal concentration (of iron, nickel, cobalt and arsenic) approached 65 g/L. Under these conditions, the redox potential in the first-stage bioleach reactor did not exceed 550 mV (vs Ag|AgCl), and so the maximum recommended feed solids concentration for the process was set at 15 %.

- At very high iron (60 g/L) and nickel (30 g/L) tenors in the feed to iron/arsenic precipitation, efficient agitation of the slurry was not possible for pH values greater than 1.5, since the slurry viscosity increased significantly. The introduction of a seed recycle was able to decrease the viscosity of the slurry, owing to the dilution of the iron tenor in the feed solution.

- Iron removal of greater than 99 % could be achieved in a five-stage neutralisation plant, with minimal nickel and cobalt losses. By implementing a seed recycle, viscosity effects were eliminated, the slurry could be agitated efficiently, and the product could be thickened and filtered.

The overall conclusion reached at the end of the metallurgical test work programme was that a process consisting of concentrate regrinding, magnetic separation, flotation, bioleaching, iron/arsenic removal by lime precipitation, metal precipitation to produce a MHP, and tailings neutralisation, had been successfully demonstrated.

Based on the outcomes of the metallurgical test work programme, a set of process design criteria, a process flowsheet and a mass balance were developed. Mintek, as the bioleach technology provider, provided the design criteria for the bioleach plant. The design criteria, flowsheet and mass balance formed the basis for a feasibility study that was commissioned by the client, and executed by Tenova Mining & Minerals[36].

Process Flowsheet and Process Description

The nickel sulphide treatment plant was constructed on the site of the existing Vuonos talc concentrator plant. The new plant was fully integrated with the existing one in terms of labour, services and utilities, and made use of available space within existing buildings where possible. The location of this plant, in eastern Finland, meant that it would experience very low temperatures in winter. The minimum ambient temperature specified in the design criteria was -15 °C, with a peak of -30 °C. The bioleach tanks, which were heat generating, were located outside of the existing buildings.

The sulphide treatment plant was designed to process a 50:50 blend of Sotkamo and Vuonos concentrates, or 100 % Sotkamo concentrate, at full throughput. The plant is modest in size, being required to treat 35 t/d of sulphide concentrate.

In the descriptions that follow, note that the flowsheets are based on the original design, whereas the process description describes the flowsheet after some modifications were made to the circuit, either during commissioning or subsequently during plant operation.

The process flowsheet for the concentrate preparation circuit, comprising stockpiling, repulping, milling, magnetic separation and flotation, is shown in Figure 21. A front end loader loads the Sotkamo concentrate into a feed bin, from where it is transferred by a pipe conveyor into a repulp tank. The repulp tank also receives fresh concentrate as slurry directly from the Vuonos plant. The combined feed slurry is pumped to the mill where the concentrate is ground to 80 % passing 20 μm. The regrind mill is an ultrafine vertical grinding mill operating in open circuit. The mill discharge is pumped to the magnetic separator.

The magnetic separator produces a concentrate which contains 3 to 5 % nickel. The concentrate is filtered on a disc filter. The non-magnetic fraction is pumped to flotation for further upgrading.

Flotation is conducted in three rougher cells and one scavenger cell, with a conditioner ahead of the first stage. Flotation tailings are returned to the existing Vuonos flotation circuit tails disposal tank and pumped to the tailings facility.

The flotation concentrate is thickened to reduce the level of flotation reagents in the bioleach feed slurry. The thickener underflow is pumped to the bioleach feed tank where the density is reduced from 65 % to 50 % solids using recycled water, and nutrients are added.
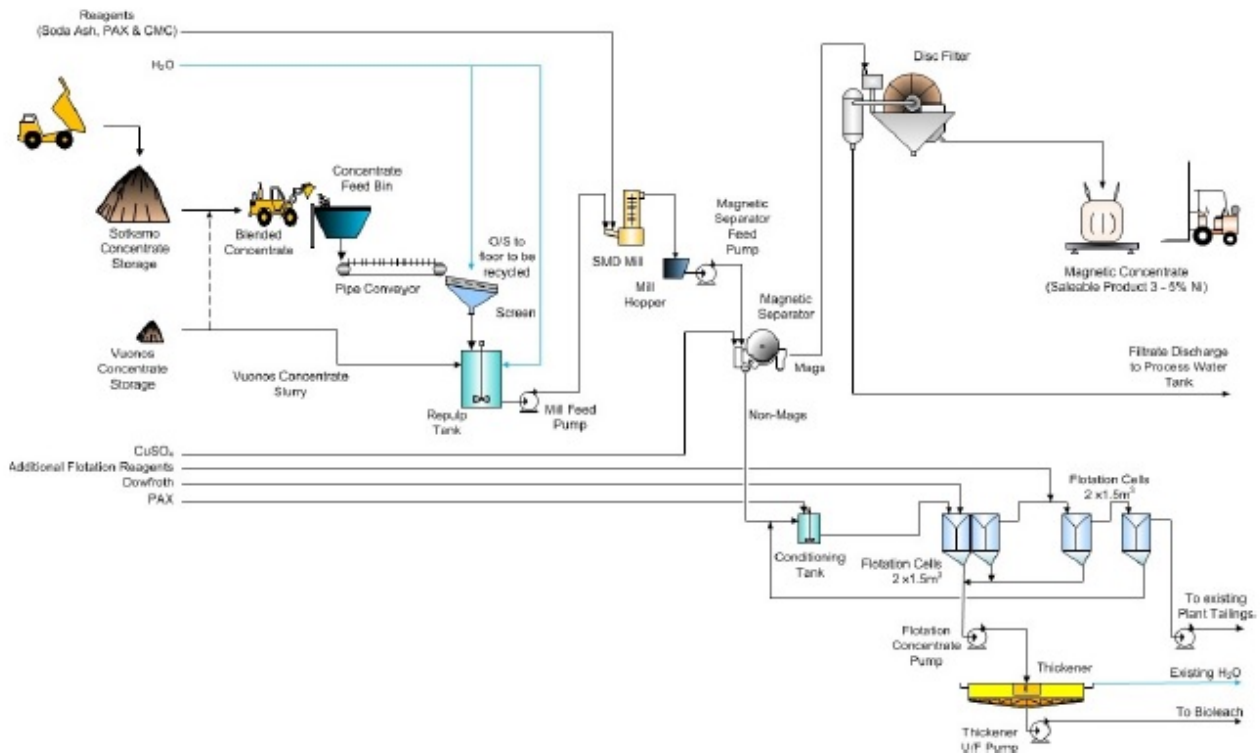

Figure 21: Mondo Minerals nickel sulphide plant process flowsheet: concentrate stockpile, screening, milling, magnetic separation and flotation[36]

The process flowsheet for the bioleach circuit is shown in Figure 22. The slurry discharge from the stock tank is diluted in-line to 15 % solids prior to being blended with nutrients in the splitter box, from where it is distributed to the three primary bioleach reactors.
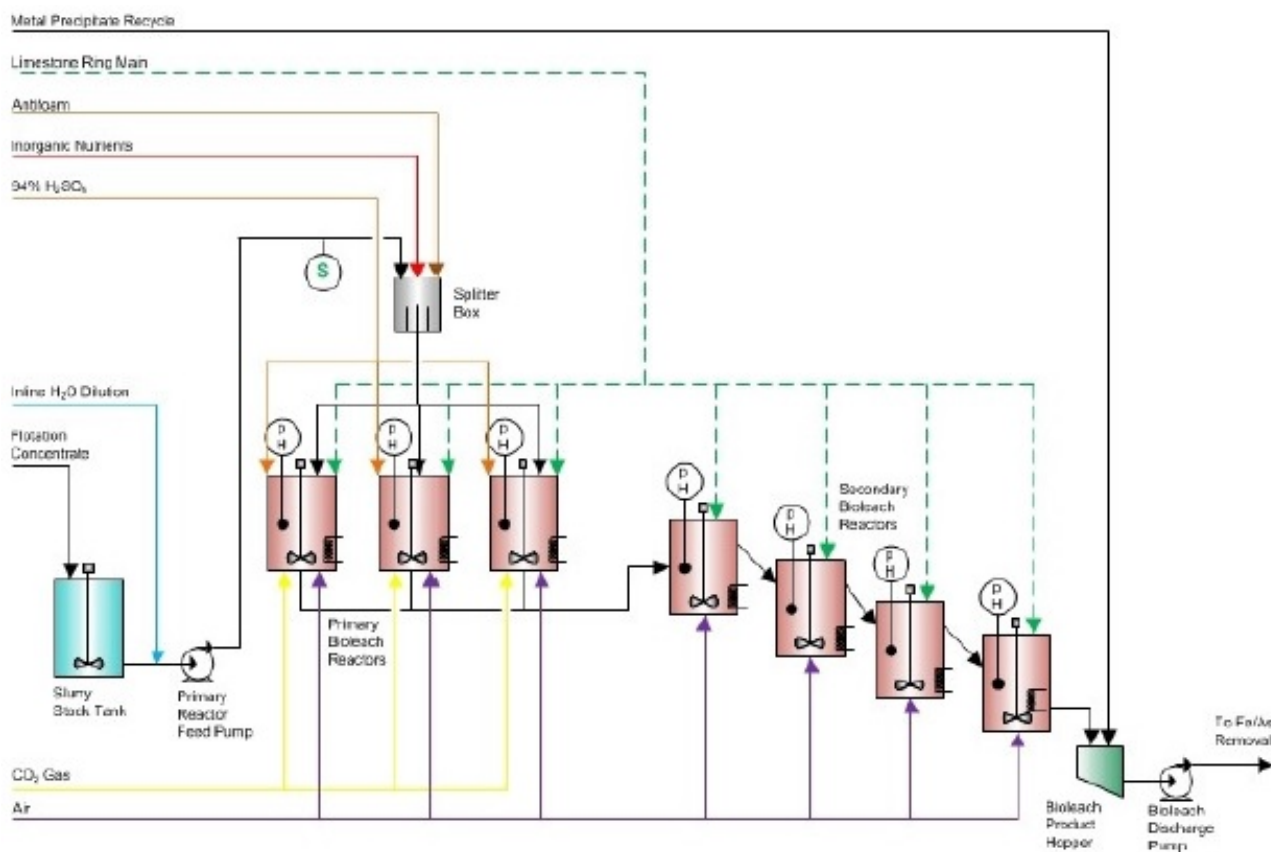

Figure 22: Mondo Minerals nickel sulphide plant process flowsheet: bioleaching[(36)]

The bioleach circuit originally consisted of seven tanks, with an overall residence time of seven days at the design flow rate. Three of the tanks were configured as primary oxidation reactors, followed by four tanks configured as secondary oxidation reactors. However, a fourth primary reactor was added at a late stage. There was space available for the vessel, and the installation of a fourth primary reactor provides additional capacity in the plant, if needed. Each tank is fitted with an agitator for dispersing air, suspending solids and maintaining homogeneity, an air sparge ring for injecting air supplied by a blower, a separate pipe for adding gaseous CO_2, and a number of vertical cooling coils which perform a dual function as baffles. The bioleach reactors are shown in Figure 23.

CO_2 is required in the bioleach process and is supplied by adding CO_2 gas directly to the primary reactors. The CO_2 is delivered by road tanker, and stored in pressurised tanks.

The pH level in the reactor bank is allowed to vary as the reaction proceeds. The pH was expected to remain within a range of 1.2 to 1.6, with the higher pH values in the primary oxidation reactors, and declining as the level of sulphide oxidation increases in the secondary oxidation stages. Provision is made for pH control, if needed, by the addition of either sulphuric acid or limestone.

The oxidation reactions are exothermic, and so the slurry temperature inside the bioleach reactors tanks is maintained at 46.5 °C by banks of cooling coils located in each tank. The operating temperature must not exceed 49 °C, as this will have a deleterious effect on microbial performance. Cooling water is supplied to each set of cooling coils by a single header fed from the main feed line. Individual coils in each tank are valved to allow for isolation of any coil in the event of a cooling coil leak. The cooling coils also act as baffles in each tank, and are supported by brackets mounted on the tank walls.

The agitators in the bioleach reactors are supplied by Afromix. Each agitator in the four primary bioleach reactors is fitted with a dual-impeller system, similar to that shown in Figure 24, which is a relatively new innovation in bioleach reactor design.

Figure 23: The bioleach reactors at the Mondo Minerals nickel sulphide plant[37]

Figure 24: The dual P4/P3 impeller system in the three primary bioleach reactors at the Mondo Minerals nickel sulphide plant[36]

The lower impeller, known as the P4, is a downward-pumping, four-blade, high-solidity-ratio hydrofoil impeller of the type that has become the standard in bioleach reactors. It is designed to disperse high volumes of air whilst also maintaining the solids in suspension and promoting heat transfer. Such impellers, which typically have a solidity ratio of more than 90 %, are characterised by their ability to operate at high gas volumes without

flooding. The upper impeller is the innovative aspect: it is an upward-pumping, three-blade medium-solidity-ratio impeller, known as the P3, which was originally designed for high-viscosity applications, but has now found application in three-phase (gas-liquid-solid) mixing applications. Improved mass-transfer performance is achieved through surface air induction created by the top impeller, and enhanced gas hold-up from the specific mixing pattern that is created by the dual-impeller configuration.

The four secondary bioleach reactors, which have considerably lower gas dispersion duties than the primary reactors, are fitted with single P4 impeller systems.

In the iron and arsenic removal circuit, shown in Figure 25, iron and arsenic are eliminated from the pregnant solution by precipitating them with limestone ($CaCO_3$) as a stable ferric arsenate in a series of reactors, and then separating the solids from the solution to produce clear liquor for nickel and cobalt recovery.
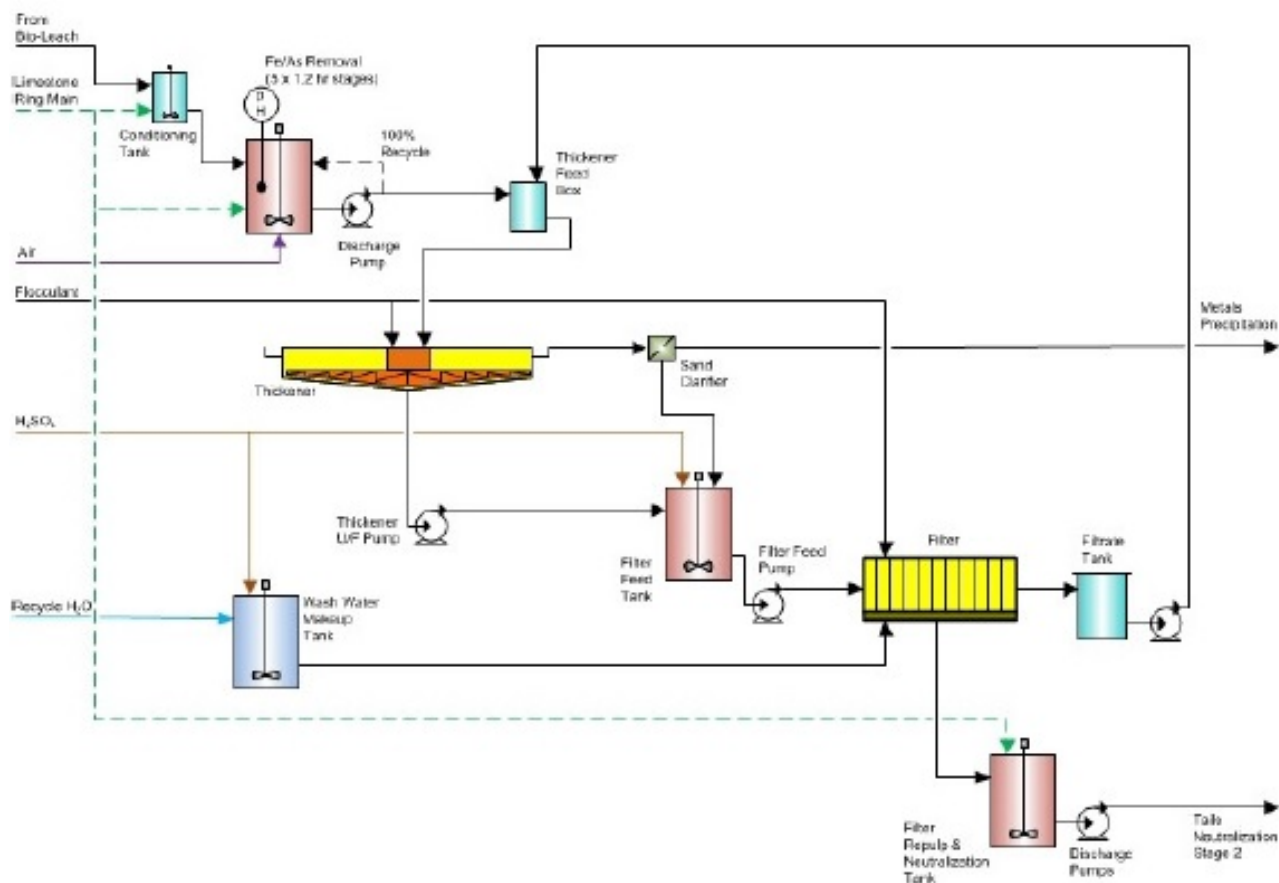

Figure 25: Mondo Minerals nickel sulphide plant process flowsheet: iron and arsenic removal[(36)]

A precipitation seed is supplied to the conditioning tank from the thickener underflow at a rate equal to the expected ferric arsenate production rate.

Slurry from the final bioleach tank is pumped to the first of five tanks arranged in series. The slurry pH is maintained at a level of 2.2 in the first tank by the addition of limestone, 2.8 in the third tank and 3.5 in the fifth tank. In this circuit, $CaCO_3$ slurry is added to the conditioning tank, and to the second and fourth of the iron-arsenic removal tanks.

Slurry from the final iron/arsenic removal tank is pumped to a thickener. The thickener overflow is pumped to a cartridge filter, and the clarified solution proceeds to the metal precipitation circuit. The thickener underflow is pumped to the filter feed tank, where recycled water is added. Slurry is pumped from the filter feed tank to a filter, where the cake is washed with acidified water at a pH level of 1.5, and then with water. The filter cake discharges to a repulp tank, from where it is pumped to the effluent neutralisation circuit. The filtrate discharges to the filtrate tank from where it is pumped to the thickener.

The objective of the metal precipitation circuit, illustrated in Figure 26, is to precipitate a mixed nickel and cobalt hydroxide product with a nickel content of typically 42 %, bagged for sale. The metal recovery section recovers any remaining nickel or cobalt for recycle before tails treatment.
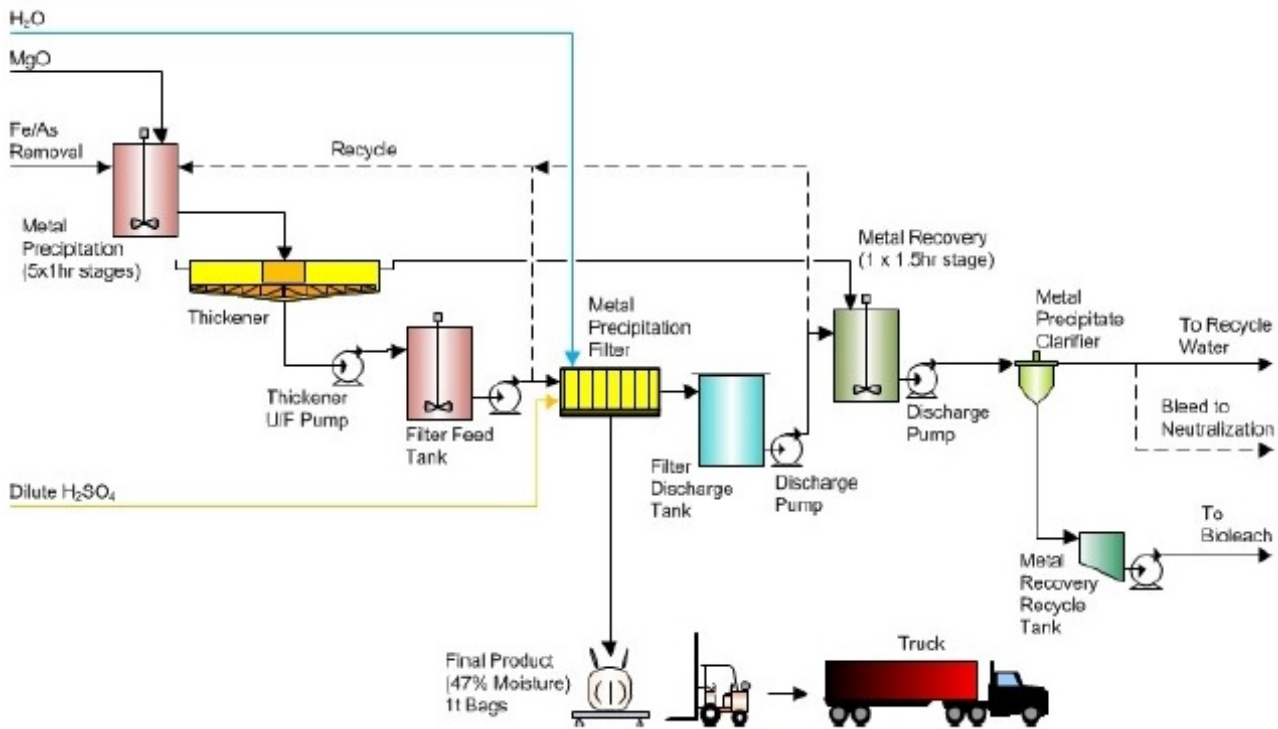
Figure 26: Mondo Minerals nickel sulphide plant process flowsheet: metal precipitation (MHP production)[36]

Clarified solution from the iron and arsenic removal circuit is pumped to the first of five metal precipitation tanks, operating in series. Each tank has a residence time of one hour, and the target operating temperature is 30 to 34 °C. The pH is raised to between 7.0 and 7.2 by the addition of magnesia (MgO). An important aspect of this process is to ensure that the MgO slurry is fresh. To achieve this, the MgO is fed dry into a small mixing tank with a residence time of 5 minutes, where water is added to give a 20 % MgO slurry.

The nickel product is a basic nickel sulphate.

Slurry from the last tank is pumped to a thickener. The thickener overflow discharges to the metal recovery tank. The thickener underflow is pumped to the filter feed tank, with a bleed back to the first metal precipitation tank to facilitate seeding of the precipitate. The slurry is pumped to the filter where the cake is subjected to a water wash, followed by an acid wash. The filtrate discharges to the filter discharge tank from where it is pumped to the metal recovery tank.

The metal recovery tank has a residence time of 1½ hours. The pH level in this tank is raised by the addition of CaO to between 7.5 and 8.0. Slurry from the tank is pumped to a filter. The filter underflow is repulped and recycled to the conditioning tank in the iron-arsenic removal circuit to recover any residual nickel. The mixed hydroxide filter cake is sampled, bagged and weighed prior to being loaded onto trucks for sale.

Neutralised and clarified water from the metal recovery circuit is used as a process water source for the plant. However, it has too high a magnesium sulphate level for reuse in the bioleach process or for safe environmental discharge, and so it is processed in the recycle water treatment circuit, as shown in Figure 27. The water is fed to a neutralisation tank, where CaO is used to raise the pH to a level of around 9.5. The resulting precipitated hydroxides and gypsum are then removed from the solution via a thickener. The thickener underflow is pumped to tailings, while the water, now containing low levels of sulphates and magnesium, is reused in the process.

In the tailings neutralisation circuit, which is shown in Figure 28, any residual metals are precipitated prior to discharge of the tailings to the tailings dam.

The filter cake from the iron-arsenic removal circuit is repulped. The slurry is then pumped to the neutralisation tank, which has a residence time of 1 hour. The pH is raised to a level of between 10.0 and 10.3 by the addition of CaO.

The neutralised tailings from the sulphide treatment plant are discharged to the main plant's existing tailings collection tank. Here, tailings are mixed with the magnesite tailings from the talc production plant, and then pumped to the existing tailings dam[36,37].
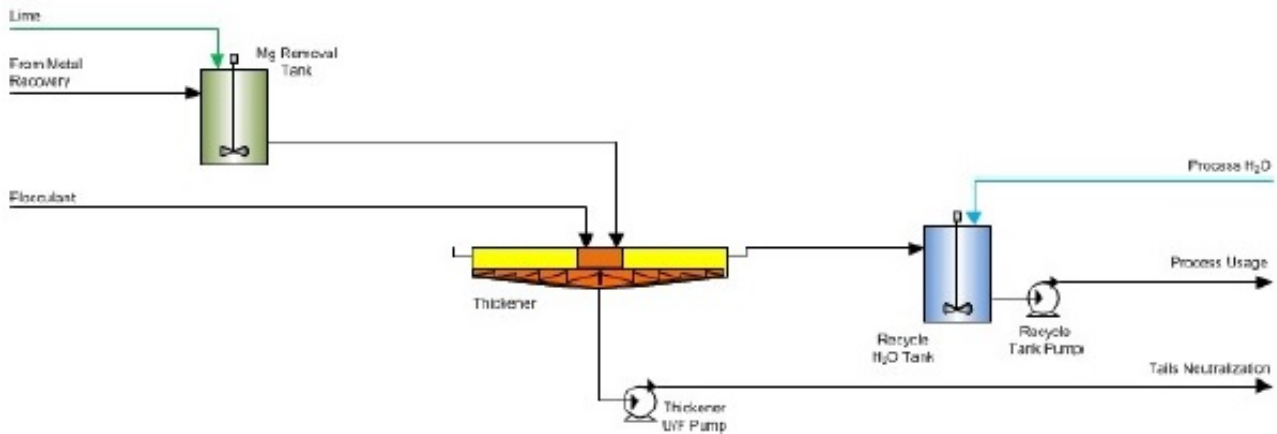
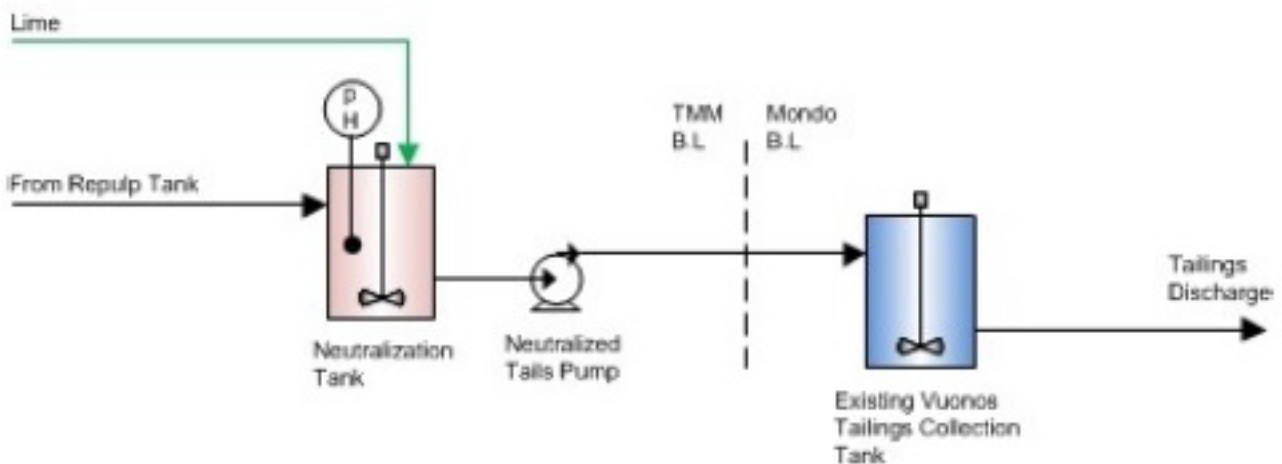**Figure 27: Mondo Minerals nickel sulphide plant process flowsheet: recycle water treatment[(36)]**

Figure 28: Mondo Minerals nickel sulphide plant process flowsheet: tailings neutralisation[(36)]

Commissioning and Troubleshooting

The inoculum for the bioleach plant was built up on site in a succession of reactors, starting with a small-scale continuous mini-plant, and progressing through batch operation in 1, 3 and 6 m^3 reactors.

The inoculation and start-up of the production plant commenced in late September 2015. The first two attempts failed, and this was attributed to the highly reactive pyrrhotite in the Sotkamo concentrate (which was used in the commissioning phase) reacting with acid and producing hydrogen sulphide (H_2S) gas, which poisoned the inocula as soon as they were introduced into the vessels. This behaviour was not encountered during the metallurgical test work programme, for several reasons: the scale of operation in the laboratory- and pilot-scale test work prevented reducing conditions from forming, and pre-acidification was conducted under aerated (and therefore oxidising) conditions. The inoculation method was adjusted, and on the third attempt, successful inoculation of one of the primary bioleach reactors was achieved. From there, the other bioleach reactors were filled, and feeding of the bioleach plant could begin.

During commissioning, several practical problems were encountered that needed to be addressed. These included a build-up of agglomerated lumps and small pebbles in the conically-bottomed concentrate repulp tank, which then passed into the suction of the peristaltic pump that feeds the regrind mill feed tank, causing regular blockages of the pump. This was fixed by the installation of a makeshift filter basket in the repulp tank, and lifting the withdrawal point (which was close to the bottom of the tank) above the conical section of the tank.

Initially, the upgrading circuit – comprising magnetic separation and flotation – was not commissioned, which resulted in the appearance of a very persistent foam in the bioleach reactors. This foam was undoubtedly caused by the talc that occurs in the non-upgraded concentrates, as evidenced by the very light colour of the foam seen in Figure 29. The flotation section of the upgrading circuit was designed to remove almost all of the talc in the concentrates. The foam was combated to some extent by the use of an antifoaming agent (Figure 30)[(37)].

Figure 29: Foaming in a primary bioleach reactor at the Mondo Minerals nickel sulphide plant caused by the presence of talc in the feed concentrate[37]

Figure 30: Surface of a primary bioleach reactor at the Mondo Minerals nickel sulphide plant, showing the beneficial impact of antifoam addition to reduce foaming[38]

One of the main features of the design of this bioleach plant is that it needs to withstand the harsh Finnish winter. In January 2016, this aspect of the design was fully tested. The temperatures in January were particularly low for a sustained period, with lows of -25 °C and lower being experienced for almost three weeks,

including a period of a week when lows of -30 °C were experienced. The impact on the plant was quite severe, as illustrated in Figure 31, with all of the pipework and instrumentation above the bioleach reactors becoming enveloped in ice. More seriously, in the coldest week, regrind mill feed tank pump failed, which in turn resulted in the bioleach feed pump being stopped. Despite the extensive lagging that is a feature of all the external pipework, the bioleach feed line froze. This caused the exothermic sulphide oxidation reactions in the bioleach reactors to slow down, and the pulp temperatures to drop. This in turn caused the cooling water flow rates to decrease, and so the cooling water supply lines also froze. Steam generators were rushed to site, the lines were defrosted, and normal operation resumed. The bioleach pulp temperatures began rising once feeding was re-instituted, and the cooling water flow rates returning to the expected levels[37].

Figure 31: The big freeze in January 2016 at the Mondo Minerals nickel sulphide plant

In April and May 2016, a mass balance sampling campaign was instituted over the bioleach plant, which showed that the average nickel and cobalt extractions attained in the bioleach plant over this period were 97.4 % and 98.4 %, respectively. Although the full design throughput was yet to be achieved, these results provided confidence in the robustness, stability and efficiency of the bioleaching section of the plant[37].

Commissioning of the downstream sections of the plant proceeded well. The residue from the iron-arsenic precipitation circuit was subjected to environmental stability testing, which showed that the neutralised product met the requirements for classification as a regular waste.

The metal precipitation plant commissioning also proceeded well, and a product containing around 42-43 % nickel hydroxide and about 1 % cobalt hydroxide was produced[37].

Over the following two years, the complete circuit was commissioned, and production was ramped up, although a few bottlenecks remained which prevented the attainment of the design throughput. The quantity of the mixed hydroxide precipitate (MHP) being produced increased steadily, and it was of excellent quality, typically containing 47 % nickel and 2 % cobalt. The MHP produced in this plant was sold and used for the production of battery-grade nickel and cobalt sulphates. By 2018, the nickel sulphide treatment plant was able to achieve 80 % nickel recovery to the MHP product from the bioleach feed. It was thought that further improvement to a desired range of 85 to 90 % would be possible, particularly with optimisation of pH control in the precipitation circuits and of the iron/arsenic filter wash sequences[38].

In the last quarter of 2018, Mondo Minerals was acquired by the British speciality chemicals company Elementis plc, from U.S. private equity firm Advent International. In view of the decline in the nickel price during the second half of 2018, the new owners took the decision to suspend operations, and the plant is currently being kept in "care and maintenance" mode. The stop in production is considered to be temporary. Prior to the cessation of the operation, good progress had been made in ramping up the production, and the throughput was just below the design value[49].

And so the story ends…but perhaps, as was the case with Talvivaara/Terrafame, there is another chapter yet to be written…

Projects in the Pipeline

Two projects that have contemplated the application of bioleaching for nickel or cobalt are worthy of mention: they are the Cosmic Boy nickel project in Western Australia and the Walford Creek copper-cobalt project in north-west Queensland.

Cosmic Boy Mill Recovery Enhancement Project

The BioHeap™ technology that was used in the Mt Sholl/Radio Hill heap leach project was acquired by Western Areas in 2009[39], and has been adapted for treating concentrates arising from the company's mining operations at the Forrestania Nickel Project, located 400 km east of Perth in Western Australia. The Spotted Quoll and Flying Fox mines produce nickel-bearing ores that are treated at the Cosmic Boy nickel concentrator. A feature of the site is that the process water contains a very high concentration of chloride (51 g/L) and has a total dissolved solids (TDS) concentration of around 100 g/L[41].

The BioHeap™ microbes are claimed to be able to operate at temperatures of between 15 and 95 ºC, in solutions containing up to 220 g/L of TDS, at pH levels of up to 3.5, and not to require any nutrients[39,40,41].

Amenability tests were conducted over a period of several years to evaluate the use of the BioHeap™ technology to treat nickel-bearing concentrates from the Cosmic Boy concentrator, some of which contained arsenic. The tests were reportedly carried out at a chloride concentration of 51 g/L, a TDS concentration of 100 g/L, and a pH level of 3.5, using materials with a grind size of 80 % passing 150 μm. The operating temperature was not reported. The results of three batch tests indicated that nickel recoveries of between 80 and 90 % could be attained, but in periods of between 40 and 80 days[39,40].

Continuous test work was conducted in a small pilot plant, but was beset with problems, including persistent foam formation and the failure of the equipment (agitators, heating coils) in the harsh environment. As a result of these failures, no detailed metallurgical results could be obtained, but it was reported that a "high" nickel recovery was obtained at a 7-day residence time, a pH level of between 2.8 and 3.5, an operating temperature of 55 ºC, on a material with a grind size of 80 % passing 110 μm[40].

Downstream test work included pressure filtration to separate the leached products, and sulphide precipitation using sodium sulphide to produce a nickel sulphide product containing about 50 % nickel[40].

In 2016, it was reported that a favourable feasibility study had been concluded, and that much of the equipment had been ordered and delivered to the site[41]. However, because of the prevailing low nickel price, the project had been suspended, and no technical updates have been published since then.

Walford Creek Copper-Cobalt Project

The Walford Creek copper-cobalt project in north-west Queensland is owned by Aeon Metals Limited. In July 2019, the company announced that bioleaching had been selected to treat a cobalt-bearing pyrite concentrate arising from the processing flowsheet. It was selected ahead of several other options, including atmospheric leaching, atmospheric oxidative leaching, low-pressure pressure oxidation (POX), roasting, and roasting and atmospheric leaching. Bioleach test work showed that a cobalt recovery of around 97 % could be obtained, and cost comparisons with POX and the roast-leach options indicated favourable economics, both in terms of capital and operating costs, for the bioleach option[42].

In September 2019, the company released the results of a scoping study for the project, which included the following metallurgical treatment process[43]:

- Bioleach-assisted heap leaching of low-grade ore at a nominal treatment rate of 1.5 million t/a.

- A 2 million t/a flotation circuit, consisting of crushing, milling, flotation, thickening and filtration, producing copper, lead, zinc and cobalt-pyrite concentrates.

- A 900,000 t/a bioleach circuit for treatment of the cobalt-pyrite concentrate.

- Precipitation circuits for production of copper sulphide precipitate with silver by-product credits, zinc sulphide precipitate and cobalt sulphide precipitate with nickel by-product credits.

- A sulphur burner and acid plant for generation of sulphuric acid.

In the bioleaching world, this is an ambitious project: a plant treating 900,000 t/a of concentrate equates to around 2,700 t/d, which is over ten times the treatment capacity at the Kasese cobalt project, and larger than the largest BIOX® plant in the world, at Kokpatas in Uzbekistan.

On the basis of the favourable scoping study, the company embarked on a prefeasibility study for the project. Very recently, in September 2020, the company announced that the delivery of the study had been delayed because of excessive costs arising from the substantial cooling requirements in the bioleach process, as well as a higher-than-expected quantity of zinc precipitation from the bioleach liquor incurring extra precipitation costs. The company is now exploring alternative processing routes that include heap leaching and vat leaching. The prefeasibility study is now expected to be completed in the first quarter of 2021[44].

THE POTENTIAL – A FUTURE LINKED TO AUTOMOBILES AND ENERGY STORAGE

The suitability and versatility of bioleaching as a primary extraction process for nickel and cobalt sulphides has clearly been demonstrated in a wide variety of applications. Commercially-proven technologies exist to treat both ores and concentrates, and they have been tested in a wide variety of locations and under widely differing climatic conditions. Nevertheless, application of these technologies remains rare, and some unique feature is usually required for bioleaching to be selected. In the case of the Kasese project, the drivers were both commercial and environmental; at Talvivaara/Terrafame, the complexity of the orebody and consequent poor metallurgical response was the overriding factor that led to the selection of a heap-leaching process; at Mondo Minerals, the presence of arsenic in the material was the reason for the selection of a hydrometallurgical treatment process.

These inherent factors will probably continue to feature when processing technology choices are made. However, demand-side factors which determine whether new deposits are discovered and exploited will also play a role. Opportunities for the application of technologies that are able to deliver benefits down the value chain will increase if the markets for these metals widen. Nickel and cobalt are both battery metals, and future demand for these commodities is therefore likely to be closely linked to the emergence globally of a low-carbon economy, and the impact this is expected to have on the energy sector. The emergence of light electric vehicles (LEVs) is undoubtedly a big part of this picture, but the development of battery-based technologies for large-scale energy storage may also emerge as a significant factor.

It has been shown that the capital and operating costs of producing battery-grade nickel sulphate from nickel concentrate are competitive with those utilising other potential feed sources (Figure 32)[38]. The decision by Terrafame to build a battery chemicals plant to produce nickel and cobalt products for use in the electric vehicle industry illustrates that bioleaching produces products that are ideally suited for this. In addition, during its productive life, the MHP product from the Mondo Minerals nickel sulphide plant was sold to a refinery that produces battery-grade nickel and cobalt sulphates[38]. The decision by Terrafame to invest in a downstream plant is clearly driven by the scale of their operation: their current nickel production target is 30,000 t/a, as against just 1,000 t/a at the Mondo Minerals nickel sulphide plant. For context, a nickel plant with a production rate of 50,000 t/a of nickel is considered "large"[49].

The links between battery metals, including nickel and cobalt, and the emerging LEV industry have been discussed in this forum for the past two years – a dozen or more papers have addressed the topic at the ALTA Nickel-Cobalt-Copper Forums held in 2019 and 2020. Last year's keynote paper in this forum provided an in-depth evaluation of the future of nickel, cobalt and copper processing in the age of lithium-ion batteries[35]. It is clear that, if the "LEV revolution" rolls on, there will at some point be a deficit in the supply of these metals. It is equally clear that the flowsheets and plants required for the production of battery-grade chemicals from these commodities are being developed, tested, evaluated and implemented. The extraction and processing technologies exist. Two things are less certain, though: the projected rate of uptake of LEVs in the medium term, and even more crucially, how rapidly the battery technologies of today, which utilise nickel and cobalt, will evolve into ones that utilise other raw materials, driven by factors such as availability, the need for improved energy storage capacity, and cost reduction.
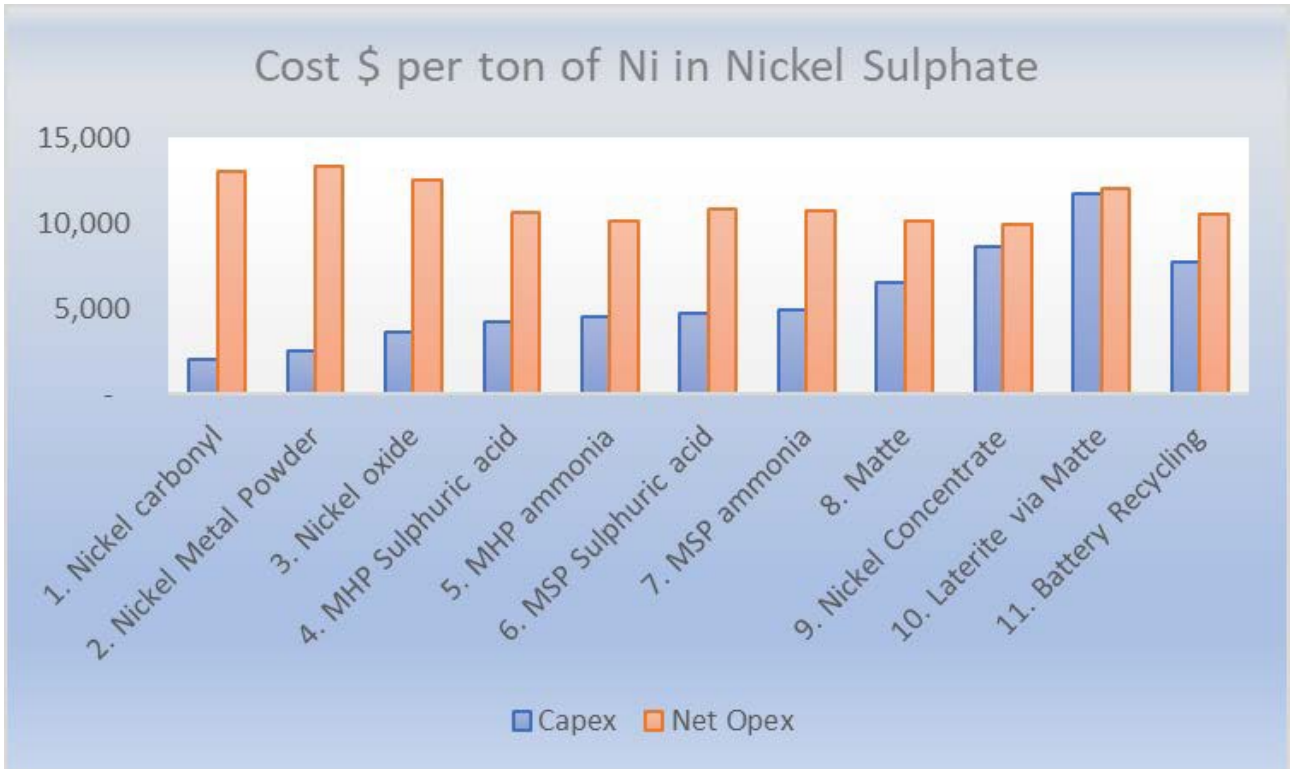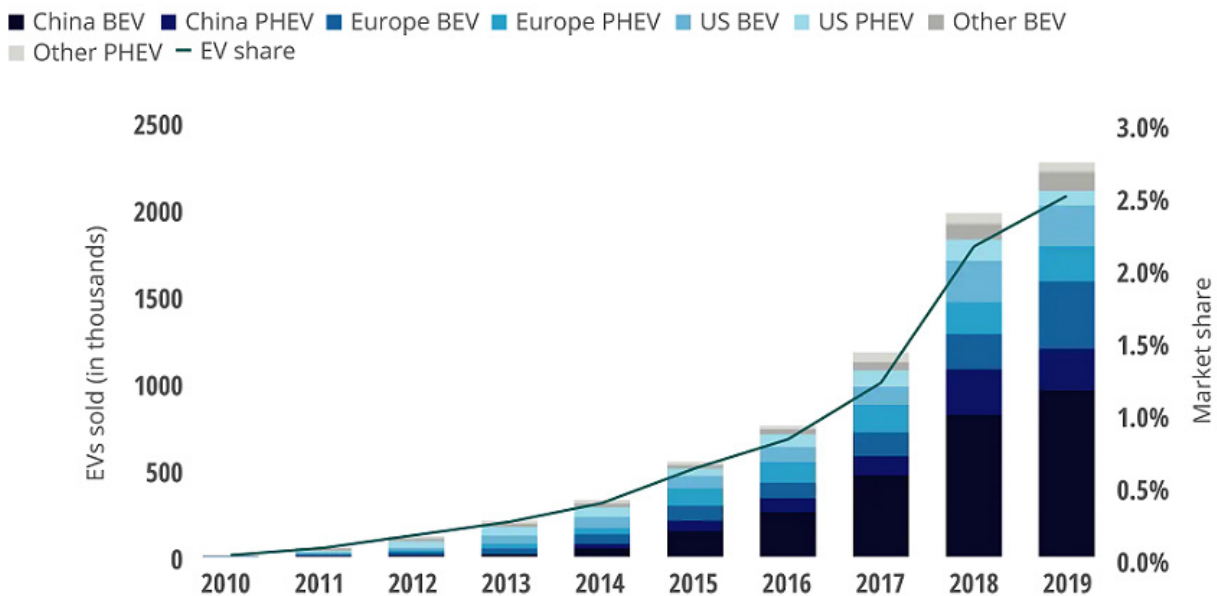
Figure 32: Cost of producing LEV battery-grade nickel sulphate from various feed sources[38]

In the short term, the COVID-19 pandemic has had a significant impact on the global EV market. Nevertheless, a recent analysis of the market by Deloitte[50] shows that there has been a pattern of continued growth in the EV market (Figure 33), which is expected to be sustained throughout the 2020s (Figure 34), despite the short-term impact of COVID-19. Battery electric vehicle (BEV) and plug-in hybrid electric vehicle (PHEV) sales surpassed two million vehicles in 2019, with EVs claiming a 2.5 % share of all new car sales. Deloitte forecasts a global compound annual growth rate of 29 % in EV sales over the next ten years. In this scenario, total EV sales will grow from 2.5 million in 2020 to 11.2 million in 2025, reaching 31.1 million by 2030, thereby securing approximately 32 % of the total new car sales market. The impact of COVID-19 on this market is clearly seen in Deloitte's forecast, and annual car sales are projected to reach pre-COVID-19 levels only in 2024. Interestingly, the pace of recovery is projected to be as a result of a slowdown in internal combustion engine (ICE) sales, while EVs are projected to continue to have a positive trajectory during the COVID-19 recovery period, which may in turn result in them capturing a disproportionate share of the market in the short term.

Figure 33: Annual passenger-car and light-duty vehicle sales in major regions[50]

Legend: Global ICE ■ Global BEV ■ Global PHEV — EV share
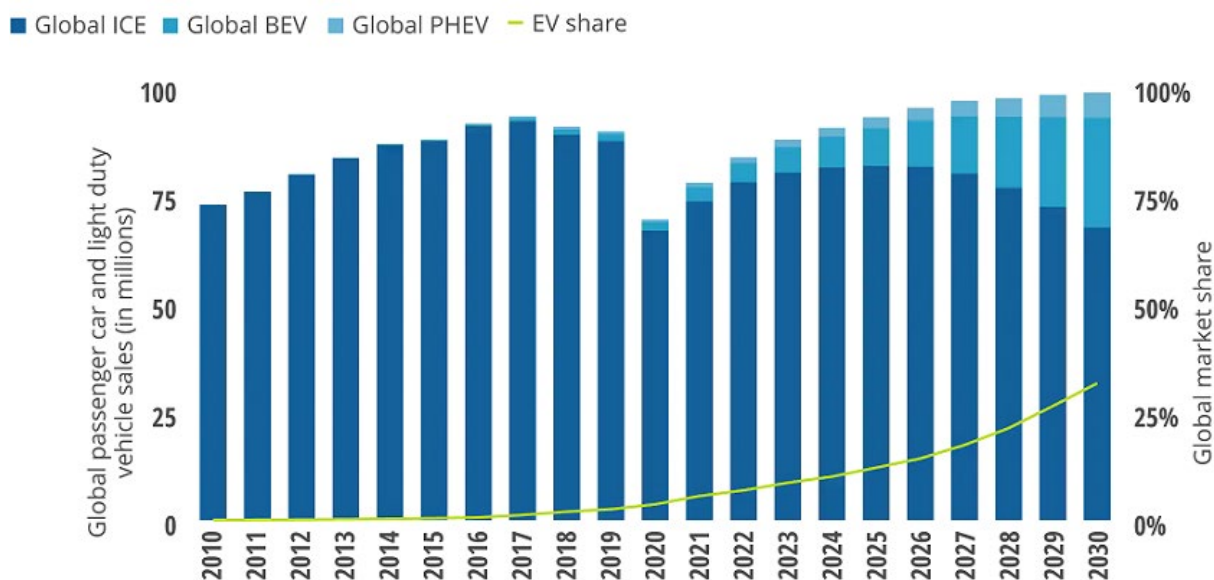
Source: Deloitte analysis, IHS Markit, EV-Volumes.com[16]

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Figure 34: Outlook for annual global passenger-car and light-duty vehicle sales, to 2030[50]

In recent months, reports have emerged of motor manufacturers being in discussions with miners to secure direct lines of supply. For example, Brazilian miner Vale SA is in talks with Tesla and others in the EV supply chain about securing nickel from its Canadian operations[51], Tesla is reportedly also in talks with BHP on a nickel deal[52], and Glencore, which is already BMW's main cobalt supplier, has indicated that it is close to securing nickel supply deals with both automakers and battery producers[53]. This indicates that, for the foreseeable future, motor manufacturers expect that nickel and cobalt are going to feature in the batteries that power their EVs.

Currently, up to 20 % of the cost of an EV is the cost of the battery, and there is a drive to reduce EV costs to improve the chances of widespread adoption of the technology. One way to do this is to eliminate the more costly cobalt component of the lithium-ion batteries used in EVs with less costly nickel. Nickel also holds more energy, allowing manufacturers to reduce the volume needed. On the other hand, the advantage of cobalt is that it doesn't overheat or catch fire easily, which means manufacturers will need to make safety adjustments if they use a substitute for cobalt[54].

It has also recently been reported that the demand for cobalt may be significantly boosted by the need for larger rechargeable batteries and more energy storage for 5G technology. This could place the sector in direct competition with EV manufacturers for cobalt. 5G mobile phones require larger batteries, using lithium cobalt oxide chemistry, because the antennae need more power than those in 4G phones. Base station antennae for 5G technology also need significantly more power, placing pressure on power grids, and necessitating the use of energy storage systems, which are now being built with cobalt-containing lithium-ion batteries. Stationary storage demand is expected to grow by 35 % per annum through the 2020s[55].

There's no doubt that, for now, nickel and cobalt will feature as the switch to a low-carbon economy gathers momentum. On the supply side, sulphide resources are diminishing, and although it is expected that laterites will feature strongly in meeting demand, laterite processing is complex. The development of recycling technologies to recover copper, nickel and cobalt from spent EV batteries is currently receiving a lot of attention, and it has even been speculated that the exploitation of undersea resources, such as deep-sea manganese nodules and cobalt crusts, may eventually be the only way to meet the demand for some of these battery metals[49].

Concerns around the responsible production of cobalt is another factor that could impact both the supply and demand of both cobalt and nickel. The fact remains that roughly 60 % of the world's cobalt is produced in the Democratic Republic of Congo (DRC)[56], and it holds more than 50 % of global cobalt reserves[57]. Within the DRC, it is estimated that 20 % of the production is by artisanal mining, involving up to two million people, many of them underage children, driven by poverty to work in unsafe conditions, and causing significant environmental damage[57,58,59,60,61,62]. Various "responsible cobalt mining initiatives" have emerged recently, including the Responsible Cobalt Initiative of the Chinese Chamber of Commerce for Metals, Minerals & Chemicals Importers and Exporters and the Organisation for Economic Co-operation and Development[63], the Eurasian Resources Group Cobalt Initiative[59], the Impact Facility's Fair Cobalt Alliance[60] and various programmes within the Responsible Minerals Initiative[57]. Miners, motor manufacturers and battery makers

have joined these initiatives[61,62,64,65], although their effectiveness and impact has been questioned in some quarters[58]. Tesla recently announced that it is aiming to manufacture cobalt-free EV batteries[66], but (as alluded to previously) that will present a number of challenges regarding battery safety, and will also – if successful – likely result in an increased requirement for nickel in EV batteries[54,65].

As much as necessity is the mother of invention, in the mining industry demand is the mother of exploration. If demand for these metals grows as forecast, and the motor manufacturers and EV producers continue to cosy up to each other, it is inevitable that exploration will increase, and new sulphide deposits may emerge from these activities. That is where the potential lies, and this review of notable past successes in the application of bioleaching for the processing of nickel and cobalt sulphides shows that, if the need arises, the technology is ready and waiting.

CONCLUSIONS

This paper has attempted to document the successes achieved in the development of bioleaching technologies as the primary extraction process for nickel and cobalt sulphides. In mechanically-agitated bioleaching of nickel and cobalt, the successes included the development of the BioNIC® technology for nickel, and culminated in the commercial application of technologies at the Kasese cobalt project in Uganda, and the Mondo Minerals nickel-cobalt project in Finland. Heap bioleaching of nickel and cobalt was developed at the Radio Hill project in Australia, and has been implemented in the large Talvivaara/Terrafame operation, also located in Finland.

A feature of the various projects has been the wide variety of downstream processing options that have been employed, from the production of a variety of intermediate products, including MHPs and individual metal sulphide precipitates, to the production of high-quality cathode metal through well-established SX and EW processes.

Successes there have been, but they are far and few between. The future of nickel and cobalt is inexorably linked to the emerging LEV and bulk energy storage markets, where they will be required as key raw materials used to construct the batteries that are needed to drive the switch to a low-carbon economy. It has been postulated here that, if demand grows to the extent forecast, exploration may lead to the discovery of new nickel- and cobalt-containing sulphide deposits, which may increase the potential for future applications of these technologies, particularly since they are ideally placed to enable integration with downstream value addition.

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