

# Development of the iron-focused laterite (ARFe) process

C.A. Biley, M. Pelsler, P. den Hoed, and M. Hove  
*Anglo American Technical Solutions - Research*

## Introduction

Global nickel production has increased near-exponentially since the late 1800s, with a large majority of the total production originating from sulphide reserves. The characteristics of base metal sulphide ores render the extraction of valuable metals simple in comparison to oxide ores of the same metals. Furthermore, sulphide ores are typically of a substantially higher grade, contain significantly less acid-consuming impurities, and are simpler in mineralogical composition. As such, there are well-developed and cost-effective processes capable of recovering valuable base metals from sulphide ores. There is significant pressure on the mining industry to explore novel techniques of recovering base metals from lateritic deposits, which are estimated to constitute approximately 70% of the total nickel reserves worldwide (Elias, 2002).

Laterite ores typically contain less than 2% nickel and consist of finely disseminated minerals, which are used to classify the ore into several key zones based on their degree of geological weathering and associated depth below ground level. Laterite resources commonly contain large quantities of iron-rich minerals, typically referred to as limonite. Limonite fractions, contrary to the magnesium-rich, saprolite fractions, are not ideally suited to pyrometallurgical smelting, where feedstock is comprised of ore with a reasonably low iron content and Si:Mg ratios where favourable degrees of slag superheat can be maintained (Voermann *et al.*, 2004). Thus, most current technologies for the treatment of limonite ores involve hydrometallurgical processes. The ammonia-based Caron process (Elias, 2002) has been applied commercially in several operations; however, the process consumes significant amounts of energy and reagents (i.e., for drying, reductive roast, and ammoniacal leaching) and achieves nickel recoveries only in the region of 80% (Elias, 2002). Another laterite technology that has been implemented on the commercial scale is high-pressure acid leaching (HPAL), which involves an aggressive, high-temperature and high-pressure leach of limonitic ores using  $H_2SO_4$ . The autoclave conditions facilitate rejection of the iron as an oxide-based residue (i.e.,  $Fe_2O_3$  [haematite] or  $H_3OFe_3(SO_4)_2(OH)_6$  [jarosite]) while regenerating the acid consumed during leaching of iron minerals from the ore.

In view of the dominant quantity of iron in typical limonite ores (i.e. 40 - 50% Fe), it has long been desired to develop a commercial-scale process that is capable of recovering iron as a saleable by-product. Such a product would benefit the process by eliminating the primary residue stream as well as providing an additional source of revenue.

## Process concept

Nickel laterites are difficult to exploit economically owing to variable mineralogical composition, stable oxide compounds, low nickel grade and limited up-grading potential. Nickel occurs as a minor metal hosted in almost all of the other minerals by substitution or inclusions, making it difficult to upgrade using conventional physical methods. The processing of laterites requires energy and/or aggressive chemical attack to recover nickel as well as near complete treatment of the bulk ore which results in high operational costs. As such, processes developed for nickel laterites need to be robust to accommodate variability in mineral composition, efficient in the utilisation of energy and reagents and highly effective in recovering nickel. The economics will also benefit from producing saleable by-product from the bulk elements in the deposit.

## Regenerative atmospheric leaching

Current trends in the development of processes for nickel laterites are moving towards atmospheric hydrometallurgical flow sheets aimed at reducing capital cost. The processes are operated under aggressive conditions resulting in almost complete digestion of the ore. The high reagent additions are compensated by regenerating the primary reagents from intermediate by-products. A number of these regenerative atmospheric leach (RAL) processes have been proposed, including systems using the sulphate, chloride, and nitrate systems (McDonald and Whittington, 2008a,b; McCarthy and Brock, 2011). Anglo American has also developed and piloted a RAL process known as the Anglo Research Nickel (ARNi) process ((Smit and Steyl (2007); Steyl *et al.* (2008)). This process utilises a mixed sulphate-chloride system and is capable of treating the full spectrum of minerals present in laterites. An additional challenge subsequently set to the

development team was to recover other saleable by-products, in particular iron from limonitic-type ore deposits, leading to the development of the iron focussed laterite (ARFe) process (Smit *et al.*, 2011).

ARFe is a sulphate based atmospheric leaching process where the primary reagents are regenerated from an intermediate by-product by thermal decomposition. The concept of the ARFe process is illustrated in Figure 1. In the lixiviant cycle, ore is digested using acid to produce a base metal sulphate solution. The acid is generated in solution by absorbing and oxidising SO<sub>2</sub> to sulphate, while reducing Fe<sup>3+</sup> to Fe<sup>2+</sup>. Ferrous sulphate is recovered as an intermediate crystal product and the solution is returned to the leach to close the cycle. The crystal product is further processed in the sulphur cycle. Ferrous sulphate is thermally decomposed to produce iron oxide and SO<sub>2</sub>. The SO<sub>2</sub> is returned to the lixiviant cycle to close the circuit. SO<sub>2</sub> is the primary reagent in this process concept and is fully regenerated.

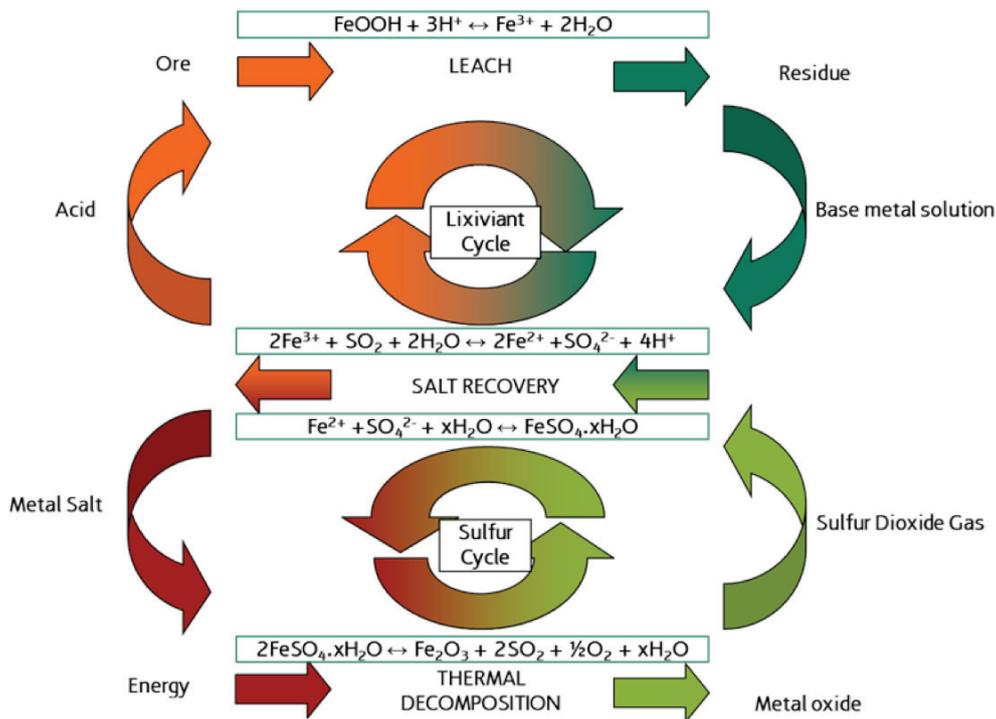


Figure 1. ARFe regenerative atmospheric leaching process: a schematic diagram of the concept (limonite ore is simplified as FeOOH)

### ARFe base-case block flow diagram

The ARFe processing concept is translated into a base-case block flow diagram in Figure 2 (adapted from Smit *et al.*, 2011). The chemistry of the key unit processes are discussed in a later section. Limonite ore is digested in the leaching step by direct addition of SO<sub>2</sub>. Reductive leaching using SO<sub>2</sub> is known to accelerate the leaching of limonitic ores (McDonald and Whittington, 2008a). The direct addition of SO<sub>2</sub> produces more protons per mole sulphur (4H<sup>+</sup> per SO<sub>4</sub><sup>2-</sup>) compared to H<sub>2</sub>SO<sub>4</sub> (2H<sup>+</sup> per SO<sub>4</sub><sup>2-</sup>). The amount of acid produced at complete reduction of Fe<sup>3+</sup> from the ore is insufficient for the leaching requirement. Additional acid can be generated by re-oxidation of Fe<sup>2+</sup> in solution or by the addition of H<sub>2</sub>SO<sub>4</sub> to the process, further described with the key unit operations. The leaching step operates close to the solubility limit of ferrous sulphate. A dissolution step is included after the leaching to dissolve any crystals that may have formed by adjusting the temperature. After solid/liquid separation, the solution is sent to the crude crystallisation step to produce FeSO<sub>4</sub>·7H<sub>2</sub>O by cooling crystallisation. The hepta-hydrate form of ferrous sulphate has a large capacity to incorporate other divalent base metals in its structure. This facilitates the recovery of the minor base metals from solution into a bulk crystalline product for further processing. It eliminates the neutralisation requirements usually associated conventional circuits and decouples the leaching and value-recovery circuits.

The FeSO<sub>4</sub>·7H<sub>2</sub>O crystals are further processed to separate the metals in a second circuit called the value-recovery section. The function of this section is not only to recover the value metals (nickel and cobalt), but also to recover a sufficiently pure ferrous sulphate product from which a saleable by-product can be produced. The crystals are dissolved and then neutralised to remove aluminium, chromium, and Fe<sup>3+</sup>. Next, nickel and cobalt are recovered in the mixed sulphide precipitation step using hydrogen sulphide. The mixed sulphide precipitate is the final nickel product in the base-case circuit. Mixed sulphide precipitation produces a purified ferrous sulphate solution which is then subjected to

the evaporation step to remove excess water and simultaneously form ferrous sulphate monohydrate ( $\text{FeSO}_4 \cdot \text{H}_2\text{O}$ ) crystals. Both the neutralisation and mixed sulphide precipitation steps require neutralising agents and it would be undesirable to introduce any foreign cations into the closed circuit. For this reason ferrous hydroxide is used as the primary neutralising agent, which is generated from neutralisation of a bleed stream with a commonly available neutralising agent. The bleed stream also aids the removal of excess water and other group (I) and (II) metals (sodium, potassium, and magnesium) from the circuit.

The  $\text{FeSO}_4 \cdot \text{H}_2\text{O}$  produced from the value-recovery section is transferred to the decomposition section. In this circuit the ferrous sulphate is decomposed thermally to iron oxide (a by-product) and  $\text{SO}_2$ . Decomposition of ferrous sulphate is an endothermic reaction occurring at temperatures approaching  $1000^\circ\text{C}$ . The process step requires a fuel source and oxygen to sustain the reaction. The product gas is dilute  $\text{SO}_2$ , which needs to be cleaned and concentrated before being transferred to the hydrometallurgical circuit. The  $\text{SO}_2$  stream is split between direct addition to the leaching step and production of  $\text{H}_2\text{SO}_4$  in an acid plant.

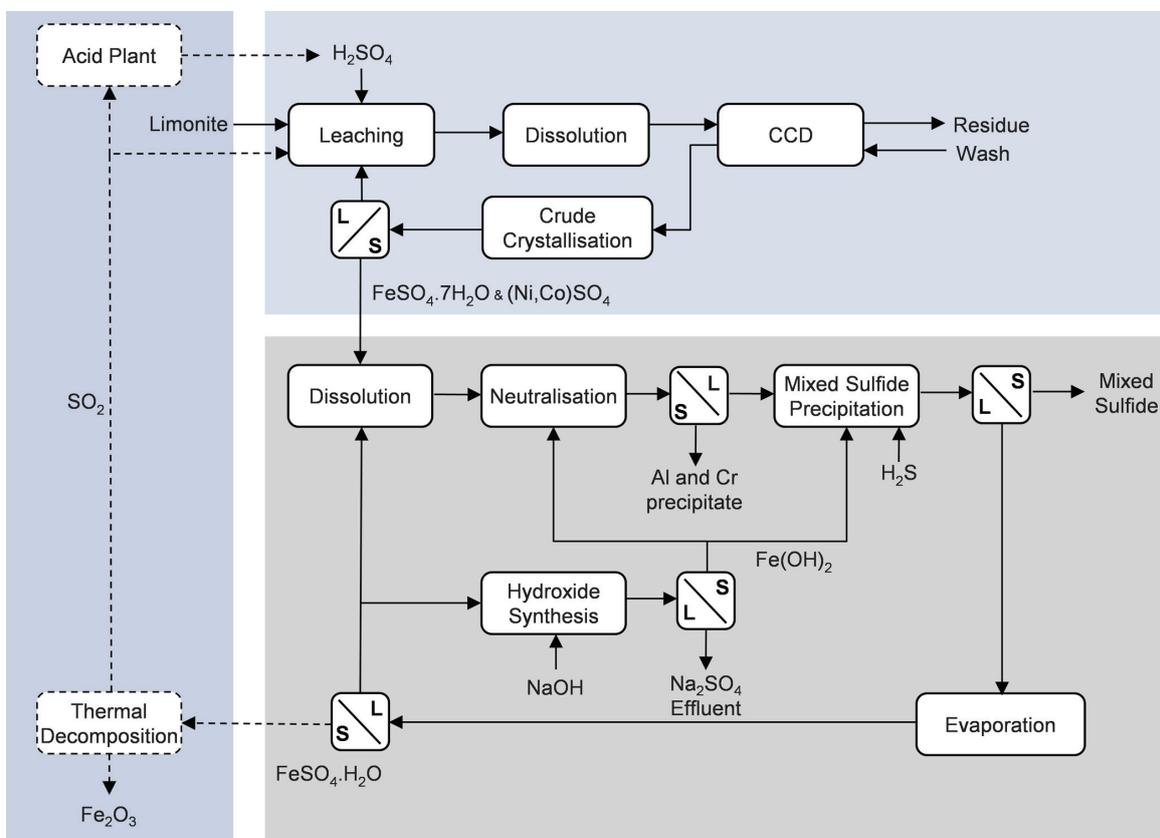


Figure 2. Block flow diagram: the ARFe base case (adapted from Smit et al., 2011)

### Iron deportment control

Iron is the predominant metallic species in limonitic ores and it is crucial to manage the solubility of ferrous sulphate in the circuit to control the deportment of iron. In the leach section the ferrous sulphate needs to be retained in solution during the solids-liquids separation step to minimise losses to the residue, and then allowed to crystallise in the crude crystallisation step to recover the intermediate product. Illustrated in Figure 3 is the solubility behaviour of pure ferrous sulphate as a function of temperature between  $25^\circ\text{C}$  and  $100^\circ\text{C}$ . The leach is operated close to the boiling point of the solution to achieve high leaching kinetics. The operating conditions and the amount of iron extracted would result in supersaturation at point 1. It exceeds the solubility of  $\text{FeSO}_4 \cdot \text{H}_2\text{O}$  causing ferrous sulfate to crystallise to the solution composition at point 2. To avoid ferrous sulfate losses during solid/liquid separation, the temperature is lowered causing the  $\text{FeSO}_4 \cdot \text{H}_2\text{O}$  to dissolve (point 3). The solution is then further cooled in the crude crystallisation step causing the formation of  $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}$  with a saturated solution (point 4) which is returned to the leaching step.

The circuit is a closed loop and the solution can only accommodate as much iron from the leaching step as can be removed during the cooling crystallisation step. The concentration of point 1 is determined by the amount of iron extracted and is related to the leach pulp density. Point 4 lies on the solubility curve and the extent of crystallisation is

therefore determined by the operating temperature of the crystalliser. As a result, the leach pulp density is limited to approximately 10% (solids/(solids+water)) for a typical limonite ore and a crystalliser temperature of 20°C. The value recovery section is similarly constrained between the solubility in the dissolution step and the evaporation step.

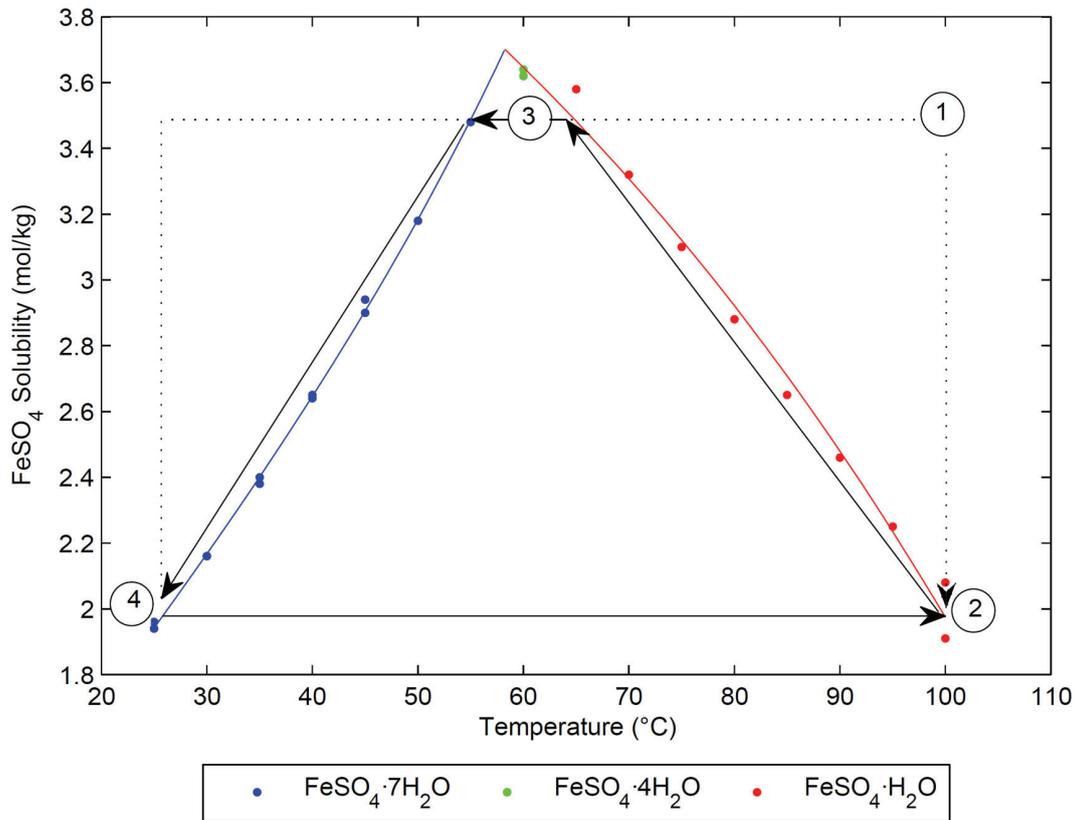


Figure 3. Solubility behaviour of ferrous sulphate at selected points in the leach section. Data points from Linke and Seidell (1965), solid lines represent a minimum-parameter Pitzer model of the system

## Key process chemistry and laboratory tests

### Atmospheric Leaching

The dissolution of FeOOH (i.e., the main component of the ore) in acidic sulfate media produces Fe<sup>3+</sup> ion in solution and consumes 3 mol H<sup>+</sup>/mol Fe, presented in Equation [1]. Central to the ARFe process concept is the exploitation of the oxidized form of leached iron to oxidize SO<sub>2</sub> in solution and thereby partially regenerating the acid consumed during leaching, shown in Equation [2].



If sufficient ferric ion is available in solution, the reduction to Fe<sup>2+</sup> using SO<sub>2</sub> is capable of producing two thirds of the acid demand of the idealised FeOOH phase. In order to maintain the acid balance of the circuit, additional Fe<sup>3+</sup> ion can be generated by oxidizing process solution with oxygen or air, according to Equation [3].



A complete reduction-oxidation cycle of Equations [2] and [3] will produce the balance of the acid demand of the FeOOH phase and additional cycling will be required to meet the complete acid demand of the ore. Alternatively, H<sub>2</sub>SO<sub>4</sub> can be added directly to the leaching circuit to fulfil acid requirements.

Critical to the operation of the ARFe process is a high utilisation efficiency of SO<sub>2</sub> in the leaching section. It has been found that the rate of SO<sub>2</sub> absorption depends, to a large extent, on the concentration of Fe<sup>3+</sup> in solution. The results of

batch tests indicate that the extraction of iron increases with increase in the concentration of  $\text{Fe}^{3+}$  in solution (Figure 4). The mechanism of laterite leaching with  $\text{SO}_2$  in sulfate media has been reported to proceed via either a combination of acid attack of the mineral surface and/or surface electrochemical processes (Nicol (1983), Senanayake (2003, 2011)). The strong dependence on  $\text{Fe}^{3+}$  and  $\text{H}_2\text{SO}_4$  concentrations (see Figure 4) suggests that the solution phase reduction of  $\text{Fe}^{3+}$  by  $\text{SO}_2$  and the subsequent acid dissolution of the mineral is likely the dominant mechanism of laterite leaching in this system.

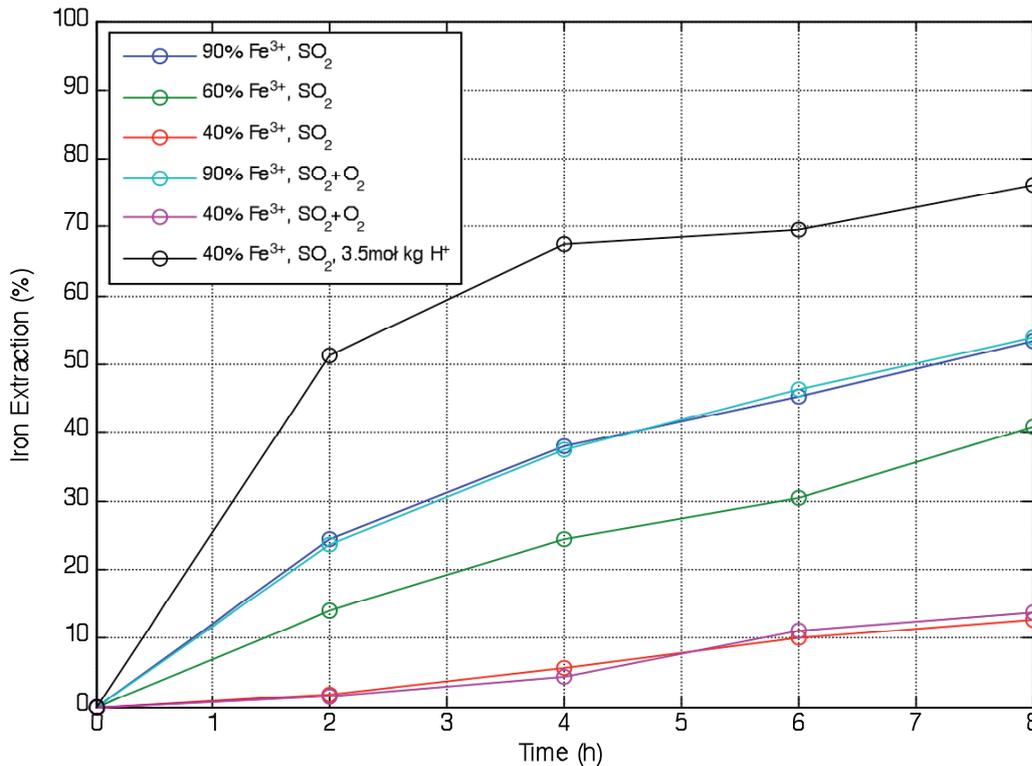


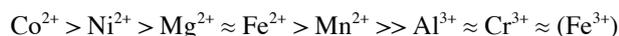
Figure 4. Typical batch extraction curves for iron from a limonitic laterite. Conditions:  $95^\circ\text{C}$ ,  $[\text{Fe}]_{\text{initial}} = 2.6 \text{ mol/kg}$ ,  $[\text{H}^+]_{\text{initial}} = 0.6 \text{ mol/kg}$ , 20% solids/(solids+water), 1 standard litre per minute (SLPM) gas flow (100% or 80%  $\text{SO}_2$ )

The feasibility of performing *in-situ* oxidation of  $\text{Fe}^{2+}$  in the leaching process was also investigated (see results in Figure 4), and it was found that similar leaching profiles were obtained with and without added oxygen. This suggests that oxygen mass transfer into the leach solution was limiting the oxidation of  $\text{Fe}^{2+}$  and a separate pressure oxidation step may be required if  $\text{SO}_2$  is to be utilised as a sole lixiviant source. For the first integrated continuous circuit of the ARFe technology, therefore, both  $\text{SO}_2$  and  $\text{H}_2\text{SO}_4$  were fed in order to simplify the flow sheet as well as (partially) decouple the acid supply from any  $\text{SO}_2/\text{O}_2$  reactions.

It was also noted that the extractions of nickel and iron were highly correlated and a simple linear model sufficiently related the extraction of these metals ( $R^2 = 93\%$ ,  $n = 20$ ), confirming the fine dissemination of nickel in the bulk iron phase of ferruginous laterites (Senanayake and Das, 2004).

### Metal sulphate co-crystallisation

The transport of metals from the aggressive leaching circuit to the more benign purification and recovery circuit is performed by co-crystallisation with a bulk  $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}_{(s)}$  phase. These crystals have been observed to have a large capacity for incorporating other divalent metal cations into their crystal structure, making it possible to co-crystallise minor metals at concentrations far below their solubility limits in pure systems. The underlying mechanism of incorporation is not well understood; however, initial test work appears to indicate that the incorporation is a result of crystal growth rates exceeding rates of diffusion of metal ions in solution, i.e. a kinetic rather than a thermodynamic phenomenon. Synthetic continuous crystallisation tests in the laboratory and an integrated mini-plant campaign have confirmed the levels of incorporation into  $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}_{(s)}$  crystals occur according to the series:



While tests have indicated that the co-crystallisation process is feasible, there are several important implications that require consideration. Firstly, the distribution coefficient of divalent metals is sufficient to maintain the solution concentration of metals at reasonably low concentrations in the leaching circuit. However, the low incorporation of trivalent cations into the crystals, primarily as a result of entrained solution, leads to considerable build-up of  $\text{Al}^{3+}$  and  $\text{Cr}^{3+}$  in the leaching circuit. Since these metals have no explicit outlet, they will result in inoperable conditions. Additionally, since the  $\text{Fe}:\text{H}_2\text{O}$  ratio of the ferrous sulphate crystals is fixed, water addition into the leach circuit needs to be carefully managed to avoid the need for additional water removal steps.

### Impurity removal

Typically neutralisation is used to remove tri-valent metals from solutions containing base metals due to their hydroxides having significantly lower solubilities (Smith and Martell, 1976). Limestone is typically used for this purpose in sulfate media due to it being relatively inexpensive and the reasonably low solubility of  $\text{CaSO}_4$ . However, due to quality requirements of the iron oxide by product, it would be detrimental to add foreign metal cations (i.e. Ca, Na, Mg) into the process by utilising common neutralising agents. The ARFe process aims to fulfil all neutralisation requirements with  $\text{Fe}(\text{OH})_2$  which is generated within the ARFe circuit (see Figure 2).

Laboratory synthesis testwork has shown that both NaOH and MgO can effectively precipitate  $\text{Fe}^{2+}$  and other metals from a concentrated  $\text{FeSO}_4$  solution and that the resulting  $\text{Fe}(\text{OH})_2$  solids remain relatively stable against oxidation if re-slurried in water or process solution. Furthermore, pH controlled titrations using 0.1M HCl in deionised water, on a small mass of synthesized  $\text{Fe}(\text{OH})_2$  slurry have shown that 95% of the solids reacted in under 15 minutes at pH 2 and 3, presented in Figure 5. However, some laboratory tests have indicated that the dissolution of  $\text{Fe}(\text{OH})_2$  in concentrated  $\text{FeSO}_4$  solutions is considerably slower than in deionised water (Hove, 2012).

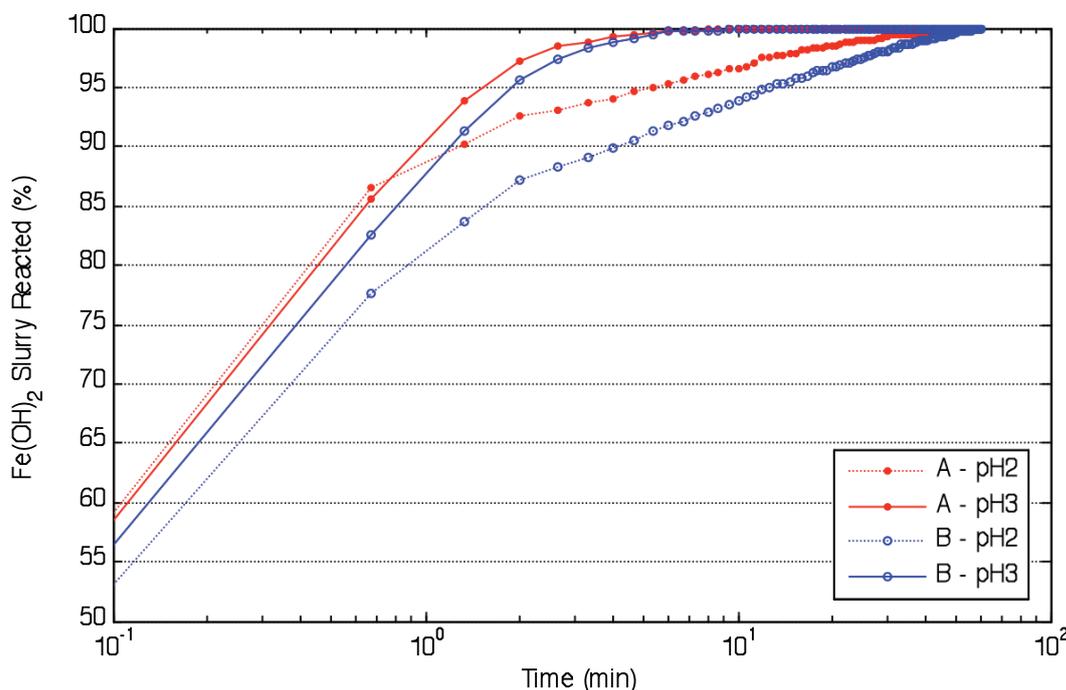
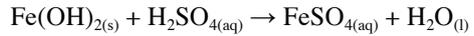


Figure 5.  $\text{Fe}(\text{OH})_2$  reactivity tests by pH-controlled titration

### Value recovery

Base metals are recovered from the ARFe process by precipitation with  $\text{H}_2\text{S}$  gas, shown in Equation [4] for nickel. The primary difficulty with this process is the sparing solubility of  $\text{H}_2\text{S}$  in even slightly acidic solutions, and typically the precipitation is performed under pressure to increase  $\text{H}_2\text{S}$  solubility. In order to avoid the use of  $\text{H}_2\text{S}$  under pressure, *in-situ* acid neutralisation with  $\text{Fe}(\text{OH})_{2(s)}$  was incorporated (Equation [5])





[5]

The results of laboratory test work in synthetic solutions are presented in Figure 6, and emphasize the importance of pH on the recovery and selectivity of base metals and iron using  $\text{H}_2\text{S}$ . Obviously, the large background of ferrous sulphate in solution results in even minor recoveries of iron as  $\text{FeS}_{(s)}$  constituting a considerable fraction of the final precipitate and will likely increase refining costs of the mixed sulphide product.

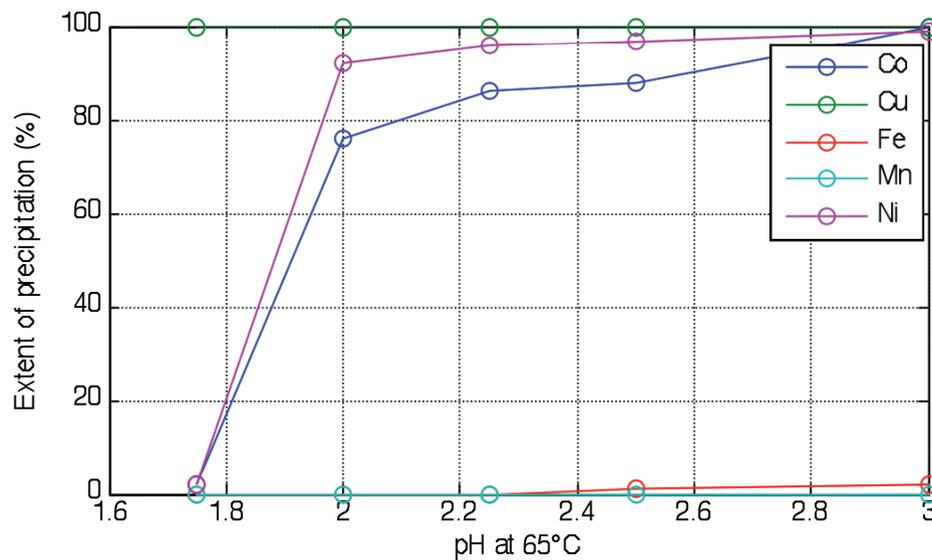


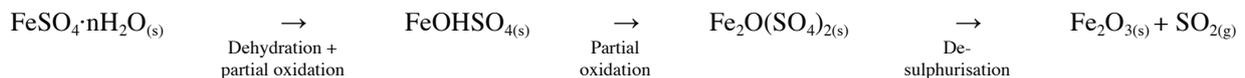
Figure 6. Recovery of base metals as a function of pH at 65°C

Data from the continuous campaign showed high base metals recoveries (i.e. >97% Ni) using a dilute (5% m/m)  $\text{Fe}(\text{OH})_2$  slurry to control the pH at 2.8. The iron content of the precipitate at these conditions averaged approximately 40% m/m and particle size analysis showed a  $d_{50}$  of 20  $\mu\text{m}$ . With further optimisation of this unit operation in the context of the value-recovery circuit, it may be possible to attain more desirable properties.

### Iron oxide by-product

The intermediate product from the hydrometallurgical circuit in the ARFe process is ferrous sulphate monohydrate ( $\text{FeSO}_4 \cdot \text{H}_2\text{O}$ ). These solids are dried and thermally decomposed to produce a saleable iron oxide by-product and regenerate all  $\text{SO}_2$  required for leaching. It is a primary objective that the by-product is sufficiently pure to be blended into an iron ore concentrate and thereby derive additional revenue for the process.

The decomposition of  $\text{FeSO}_4 \cdot \text{H}_2\text{O}_{(s)}$  has been reported (Pelovski *et al.*, 1996) to occur in a number of parallel reaction steps that are separated into three distinct phases: (1) dehydration with partial oxidation (290-330°C), (2) intermediate partial oxidation (410-525°C), and (3) desulphurisation (550-800°C):



Blast furnaces can tolerate only trace amounts (0.02%) of base metals, sulphur, and chromium, and minor amounts (0.5%) of manganese in the feed. Such impurities in the  $\text{FeSO}_4 \cdot \text{H}_2\text{O}_{(s)}$  filter cake therefore need controlling. Fortunately, in contrast to the large capacity of the  $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}_{(s)}$  crystal lattice to include metal cations (in solid solution), the monohydrate salt accommodates significantly less foreign cations. However, magnesium and manganese, which are not directly removed from the process can accumulate and may be incorporated to a significant degree in the  $\text{FeSO}_4 \cdot \text{H}_2\text{O}$  solids.

The decomposition of  $\text{FeSO}_4 \cdot \text{H}_2\text{O}_{(s)}$  produces a mixture of iron oxides (i.e. primarily  $\text{Fe}_2\text{O}_3$  and  $\text{Fe}_3\text{O}_4$ ) depending on the oxygen partial pressure during decomposition. Thermogravimetry tests of the material produced in the continuous campaign emphasize several important reaction regimes, presented in Figure 7. Points 1 and 2 in Figure 7 correspond to the endothermic removal of crystal waters from  $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}$  and  $\text{FeSO}_4 \cdot \text{H}_2\text{O}$  at 121°C and 321°C respectively. Further analysis is required to determine the presence and extent of partial oxidation according to the mechanism of Pelovski *et*

al. (1996) in this region. Preceding the endothermic desulfurization (Point 4, 736°C) there is an exothermic region (Point 3, 542°C) that corresponds to a slight but repeatable mass gain; presenting evidence for the (partial) oxidation of the material. Point 5 shows a further endothermic mass loss region at 918°C. A number of tests on synthetic material have shown that the magnitude of this mass loss correlates with the content of impurity metals in the solids and corresponds to the desulfurisation of these metal salts. The achievable degree of desulfurisation of the material analysed in Figure 7 was determined in separate laboratory tests conducted in a muffle furnace with a stationary solid bed which produced an iron oxide with a sulfur content of less than 0.03%.

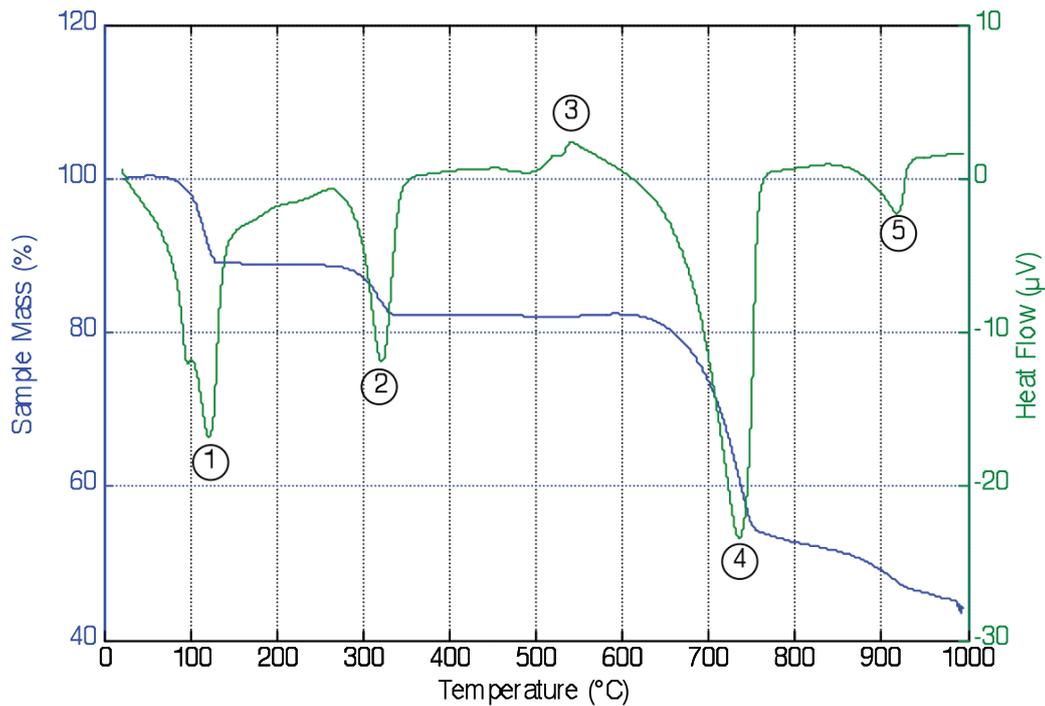


Figure 7. Thermogravimetric (TG) and differential thermal analysis (DTA) curves of a sample of  $FeSO_4 \cdot H_2O_{(s)}$  produced in the continuous campaign. Heating rate:  $10^\circ C/min$ , atmosphere: air

### Integrated continuous circuit

The results of a 4-week continuous mini-plant campaign showed that the ARFe process concept is technically feasible and the integration of the major unit operations operates as expected on the basis of individual laboratory batch tests. A total of 134kg of dry ore was fed to the leaching circuit over a period of stable operation selected for mass balance reconciliation. The metal extractions achieved during this period were similar to those achieved in the laboratory batch tests and are presented in Table I. The total consumption of  $H_2SO_4$  and  $SO_2$  in the leaching circuit during the campaign was ca. 850 kg/t ore as  $H_2SO_4$  equivalent.

Table I. Metal extractions achieved in the continuous campaign

Al	Cr	Co	Fe	Mg	Mn	Ni
51%	9%	96%	80%	90%	96%	97%

Total nickel and cobalt recovery to the mixed sulphide product over the campaign was 89% and 87% respectively. Cobalt recoveries were assumed equal to those of manganese as leach residue samples were typically below the lower limit of detection for cobalt. Losses were experienced to the value-recovery bleed stream in the initial stages of the campaign due to operational difficulties in approaching steady state. Recoveries of the base metals were in excess of 97% during stable operation.

A bulk sample of approximately 70 kg of  $FeSO_4 \cdot H_2O$  filter cake was produced during the campaign and is to be used for further development of the decomposition circuit. Using the experimental mass losses noted during decomposition

of the material by thermogravimetry (see previous section), the expected metal concentrations of the material could be calculated. Nickel and manganese contents of 0.05% and 1.14% were calculated for the decomposed material, which are typically in excess of blast furnace specifications. The improved removal of these metals in the circuit to decrease their deportment to the  $\text{FeSO}_4 \cdot \text{H}_2\text{O}$  filter cake needs to be further addressed.

The continuous circuit also confirmed the low incorporation of trivalent cations into the  $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}(\text{s})$  crystals and emphasised that a bleed stream from the leaching circuit is required to maintain the concentrations of  $\text{Al}^{3+}$  and  $\text{Cr}^{3+}$  at reasonable levels (Smit *et al.*, 2011).

### Summary of development

ARFe is a unique and novel process that aims to maximise value from low-grade limonitic laterite ores by producing a saleable iron oxide by-product from the iron-rich ore feed. Not only does this derive additional revenue from the iron contained in the laterite, but simultaneously dramatically reduces the total solid residue from the process and facilitates regeneration of the primary reagents. It is based on the processing concept underpinning the ARNi technology in which the primary reagents are regenerated from an intermediate by-product. The ARFe concept and circuit cater for variability by utilising aggressive leaching conditions that result in almost complete digestion of the stable oxide compounds in limonitic ore. The conditions are generated by the direct addition of  $\text{SO}_2$ , which results in reductive leaching conditions at high acid concentrations. The direct utilisation of  $\text{SO}_2$  also proves to be more efficient in generating acid compared to the  $\text{H}_2\text{SO}_4$ -only approach. The leaching conditions are maintained independently by decoupling the leaching section from the value-recovery section through crystallisation. This not only saves the neutralisation reagent cost for the free acid after leaching, but also retains the acid for further leaching. The intermediate ferrous sulphate by-product is further refined in a value-recovery circuit to recover the value metals (nickel and cobalt) and produce a purified ferrous sulphate intermediate by-product. Decomposition of this intermediate by-product will yield a potentially saleable iron oxide by-product and enough  $\text{SO}_2$  to meet the demand from the leach section.

Development of the ARFe technology builds on the experience gained from developing the ARNi process. The processing concept was established at the start of development and the key process chemistry involved was evaluated through a series of batch laboratory tests. A block flow diagram for the base-case circuit was derived from the test work and the hydrometallurgical component of the circuit was evaluated in an integrated mini-plant, which operated continuously over a four-week period at the end of 2012. The campaign confirmed the chemistry and integration of the key unit operations and showed the deportment of the major elements through the circuit. A sufficiently large batch of ferrous sulphate was also produced for future decomposition test work. The results from the campaign will now be used to update the mass-and-energy balance model. A conceptual level cost estimate for the circuit will then be conducted with the help from an EPCM company. This will bring the development of the technology to its first major milestone in which the technical and economic feasibility of the flow sheet will be evaluated at a conceptual level.

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#### The Author



**Max Pelser**, Senior Research Scientist, Anglo American Technical Solutions - Research

Obtained BEng and MScEng degrees in chemical engineering from the University of Stellenbosch. Started employment at Anglovaal Mining in 2003 and was seconded to Chambishi Metals Plc. in Zambia and then Marula Platinum in Limpopo. Joined Anglo American Technical Solutions – Research in 2004, working as a research scientist in the field of hydrometallurgy. Been involved in the development of the ARNi and ARFe technologies for the treatment of lateritic ores.