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EARLY EVALUATION OF METAL EXTRACTION PROJECTS

By

Mike Dry

Arithmetek Inc., Canada

Presenter and Corresponding Author

Mike Dry
mike.dry@arithmetek.com

ABSTRACT

This paper is about computational methods that can be used to evaluate new metal extraction projects before the expenditure of substantial amounts of time, effort and money.

Simple calculations based on the broad chemistry of the envisaged process and publicly available information can be used to discern whether or not the project is potentially viable. If the project passes that gate, process modelling that generates a mass-energy balance for each option under consideration allows the variable costs to be more rigorously calculated and process options to be ranked. The mass-energy balance can then be passed electronically from the process modelling software to capital cost estimation software and preliminary capital cost and fixed operating cost estimates can be calculated and transferred to financial models to rank options and predict viability or otherwise.

The methodology is illustrated using an example based on a project in which Arithmetek Inc. has had no part, namely African Eagle's Dutwa Nickel Project in Tanzania, because African Eagle has published sufficient information on this project to enable a rational comparison between the results of the calculations advocated in this paper and those from actual work on the project.

INTRODUCTION

Mining and metallurgy are commercial ventures, governed by the principles of free enterprise and therefore by the rules of investment for profit. From that perspective, "the discovery of an economic mineral deposit is an extremely rare occurrence that involves a very difficult, costly and determined effort. For the few people or companies that do succeed, it is an extremely profitable occurrence as well. But… it is exceedingly tough to stumble across Mother Nature's buried treasures before going broke"[1.] Not all mineral discoveries make it to commercial fruition.

The discovery and definition of mineral deposits is, for the purposes of this paper, taken for granted with due deference to the substantial expertise and effort entailed in that activity. Once a deposit is discovered (or re-discovered), the challenge becomes to either prove it to be commercially viable or to show that it is not commercially viable, in both cases without going bankrupt in the process. Part of the solution to this challenge comes in the form of modern computing power. Nowadays we have powerful computers and software that can greatly facilitate the development of new mining projects.

Taking a mineral deposit from discovery to reality begins with quantification of the amount of ore present and the nature and concentrations of the valuable elements. This work includes drilling and analysis of samples, or re-compilation of results from earlier work. The results can be conveniently summarized as a total amount of ore and an elemental analysis, which, in conjunction with relevant metal prices, leads to an estimate of the potential value of the deposit. If the potential value is sufficient, attention turns to the work needed to further quantify the deposit, generate appropriate samples and evaluate extraction technology.

This paper presents a computational approach to the early-stage evaluation of new mining projects, beginning at the point at which an initial estimate of the size of the deposit has been made and the valuable elements have been identified.

MINERALOGY

In the approach advocated in this paper, the elemental analysis is first translated into a suite of valuable and gangue minerals making up the ore. Selecting the minerals in this suite would initially be guided by the geology of the deposit, and later refined *via* mineralogical analysis. Knowledge about how the various minerals respond to different chemical environments can then be used to compile one or more conceptual processes for extracting the valuable elements.

Part of the exercise of compiling conceptual processes is choosing the form of the products. For example, a nickel deposit could be processed to a nickel hydroxide intermediate or all the way to metallic nickel. As part of this exercise, selling prices need to be assigned to each product and by-product, so that the potential revenue from the deposit can be estimated. Initial estimates of selling prices can be taken from the open literature, but well before an operating plant is built, that needs to be the subject of a much more careful exercise.

Having determined the possible products and approximate selling prices to generate an estimate of the potential revenue to be had from the deposit, and having calculated that the deposit contains enough value to be mined for long enough at a rate sufficient to induce the next step in the evaluation, that next step would be an exercise to evaluate the potential cost of processing the ore. When the envisaged process is hydrometallurgical, that requires preliminary laboratory leaching tests to measure the extractions of valuable metals and the consumption of reagents such as acid. It is reasonable (but not essential) to calculate the consumption of, say, acid from the mineral suite and knowledge of the chemistry of the various minerals, before commissioning laboratory tests. It is, however, always prudent, to say the least, to obtain experimental confirmation of the leach at an early stage. One of the "tricks of the trade" is doing the "minimum but adequate"[2] amount of experimental work at each stage of a project. In the initial stages of a new hydrometallurgical project, that means limited laboratory testing.

The approach advocated in this paper is conveniently illustrated by way of example. The example chosen is based on African Eagle's Dutwa Nickel Project in Northern Tanzania, Africa. The ore is a somewhat unusual nickel laterite that is to be leached with sulphuric acid under atmospheric pressure. After neutralization and iron precipitation, the nickel is to be recovered as an intermediate product, either a mixed hydroxide precipitate or a mixed sulphide precipitate. For the exercise presented in this paper, the product was assumed to be mixed hydroxide precipitate.

This paper is not meant to accurately represent the Dutwa Nickel Project – the work presented is an example illustrating the methodology advocated. The scenario used is based on initial published information on the Dutwa Nickel project, but Arithmetek Inc. has not done any work at all for African Eagle, on this or any other project.

STOICHIOMETRIC CALCULATIONS

The analysis shown in Table 1, partial results of exploratory leach tests and a conceptual process have been published on the African Eagle web site[3]. Figure 1 illustrates the basic process concept assumed for the exercise presented in this paper. The laterite is leached with sulphuric acid and the leached slurry is neutralized with limestone. The combined leach residue and iron precipitate is removed by counter-current decantation and leaves the circuit. Magnesium oxide is used to precipitate nickel, cobalt and copper from the solution as a hydroxide intermediate that is filtered, washed and sold for further processing. The remaining solution is treated with lime to precipitate magnesium, which leaves the circuit as a hydroxide residue.

Table 1 – Initial assay

Ni grade, mass %	1.1
Co grade, mass %	0.034
Cu grade, mass %	0.007
Fe grade, mass %	8.5
Mg grade, mass %	3.5

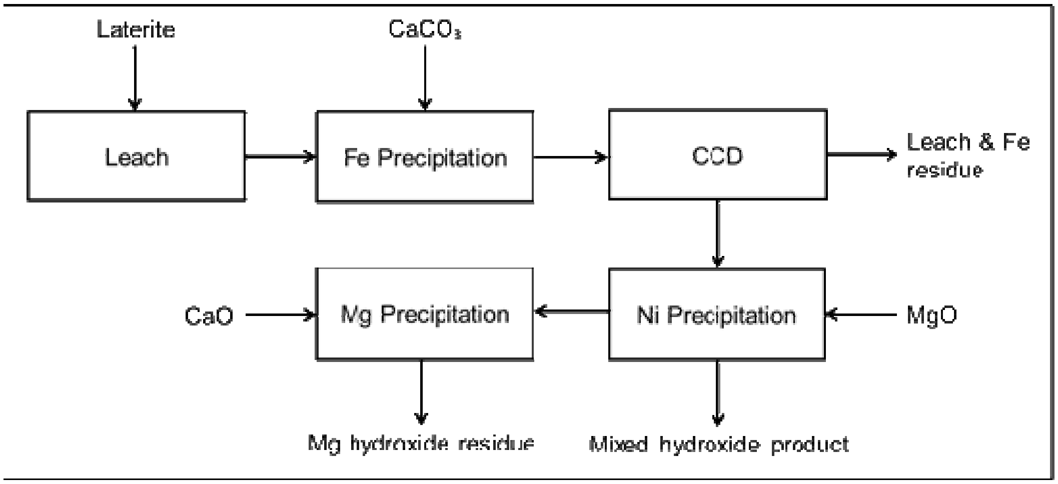


Figure 1 – Basic process

A set of preliminary tests at atmospheric pressure resulted in the dissolution of 92 percent of the nickel at a sulphuric acid addition of 210 kg/t of ore. The solid-liquid ratio and the concentration of free acid after the leach were not disclosed. Table 2 lists a mineral suite calculated from this information, assuming the leach stoichiometry shown in Table 3, that the leached slurry has 5 g/L of free sulphuric acid in the liquid phase and contains 25 percent solids by mass, and also that the iron that is dissolved does not re-precipitate at or below the atmospheric boiling point of that solution.

Manipulating the ratio of hematite (Fe_2O_3) to goethite (FeOOH) in the ore, assuming that the goethite dissolves and the hematite does not, and using the over-simplification that all the reactions listed in Table 3 have the same conversion (92%), makes this simple mineral assemblage and stoichiometry give the reported acid consumption at the assumed residual acid concentration and the assumed solids-to-liquid ratio.

There are two ways in which the leach could be done - at atmospheric pressure or at elevated pressure and temperature. Leaching at atmospheric pressure constrains the temperature to below the atmospheric boiling point of the solution, at which level the dissolved ferric iron would typically remain in solution, at the assumed level of residual acid. Figure 2 was produced from published data[4] on the solubility of ferric sulphate as a function of the concentration of sulphuric acid. The open circle represents the calculated composition of the solution after the atmospheric leach. The leach solution is well below the saturation level for ferric iron, therefore the assumption that no dissolved iron re-precipitates is plausible.

Table 2 – Mineral suite

NiO	1.40
CoO	0.04
CuO	0.01
MgO	5.80
$FeOOH$	3.22
Fe_2O_3	9.26
SiO_2	80.18

Table 3 – Simplified leach stoichiometry

$NiO + H_2SO_4 \rightarrow NiSO_4 + H_2O$

$CoO + H_2SO_4 \rightarrow CoSO_4 + H_2O$

$CuO + H_2SO_4 \rightarrow CuSO_4 + H_2O$

$MgO + H_2SO_4 \rightarrow MgSO_4 + H_2O$

$2FeOOH + 3\ H_2SO_4 \rightarrow Fe_2(SO_4)_3 + 4H_2O$

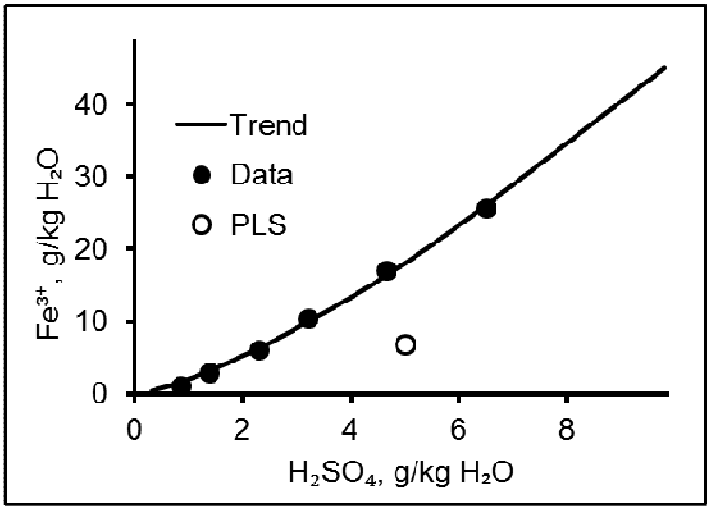


Figure 2 – Fe^{3+} solubility data at 50°C

Leaching laterite at elevated temperature and pressure (HPAL) is done when it is limonitic in nature, meaning that most of the nickel is contained in goethite-like iron minerals and those minerals make up the bulk of the ore. For such laterite ores under atmospheric leaching conditions, dissolving the iron minerals to release the nickel into solution and leaving the iron as dissolved ferric sulphate would require excessive amounts of acid because the bulk of the laterite would have to be dissolved, and then vast amounts of calcium carbonate would be required to neutralize the leach solution and precipitate out the iron. At elevated temperature (typically 240°C to 270°C in HPAL) dissolved ferric iron hydrolyses and precipitates as hematite, releasing the acid consumed in its dissolution. Leaching under HPAL conditions also greatly accelerates the leaching reactions, relative to leaching at or below the atmospheric boiling point of the solution, which may or may not be advantageous, depending on whether or not the reduced leach time results in a lower overall cost.

At the elevated temperatures found in HPAL, the second dissociation of sulphuric acid essentially does not occur[5], so instead of the two protons that are available at lower temperature, under HPAL conditions one molecule of sulphuric acid only supplies one proton. The top half of Table 4 gives stoichiometry illustrating the HPAL leach chemistry. On cooling after the high temperature leach reaction, the bisulphate salts revert to sulphuric acid and sulphate salts, as illustrated by the stoichiometry shown in the bottom half of Table 4. While the overall stoichiometry is the same as for the atmospheric leach, the behaviour of sulphuric acid under HPAL conditions is such that half of the acid added has to be neutralized after the leach, not counting the excess acid that is needed for complete leaching.

Table 4 – HPAL stoichiometry

$NiO + 2H_2SO_4 \rightarrow Ni^{2+} + 2HSO_4^- + H_2O$
$CoO + 2H_2SO_4 \rightarrow Co^{2+} + 2HSO_4^- + H_2O$
$CuO + 2H_2SO_4 \rightarrow Cu^{2+} + 2HSO_4^- + H_2O$
$MgO + 2H_2SO_4 \rightarrow Mg^{2+} + 2HSO_4^- + H_2O$
$2FeOOH + 6H_2SO_4 \rightarrow 2Fe^{3+} + 6HSO_4^- + 6H_2O$
$2Fe^{3+} + 6HSO_4^- + 3H_2O \rightarrow Fe_2O_3 + 6H_2SO_4$
$Ni^{2+} + 2HSO_4^- \rightarrow NiSO_4 + H_2SO_4$
$Co^{2+} + 2HSO_4^- \rightarrow CoSO_4 + H_2SO_4$
$Cu^{2+} + 2HSO_4^- \rightarrow CuSO_4 + H_2SO_4$
$Mg^{2+} + 2HSO_4^- \rightarrow MgSO_4 + H_2SO_4$

Figure 3 shows some data[6] relevant to the hydrolysis of ferric iron under HPAL conditions. In these measurements the slurry contained 22 percent solids and the temperature was 250°C. This can be used to calculate the acid addition required for leaching the mineralogy in Table 2 under HPAL conditions for the same dissolution of nickel. This can be done by assuming the same conversion for the first five reactions in Table 4 as for the corresponding reactions in Table 3. Adjusting the conversion of the sixth reaction in Table 4 at the same final acid concentration as for the atmospheric leach (5 g/L) in a leached slurry containing 25 percent solids, places the concentration of dissolved ferric iron on the trend shown in Figure 3. This calculation gives an acid requirement of 323 kg per tonne of ore, compared to 210 kg/tonne measured in the leach tests at atmospheric pressure. For generating the sulphuric acid from elemental sulphur, the sulphur requirements would be 69 kg/t for the atmospheric leach and 106 kg/t for the HPAL case.

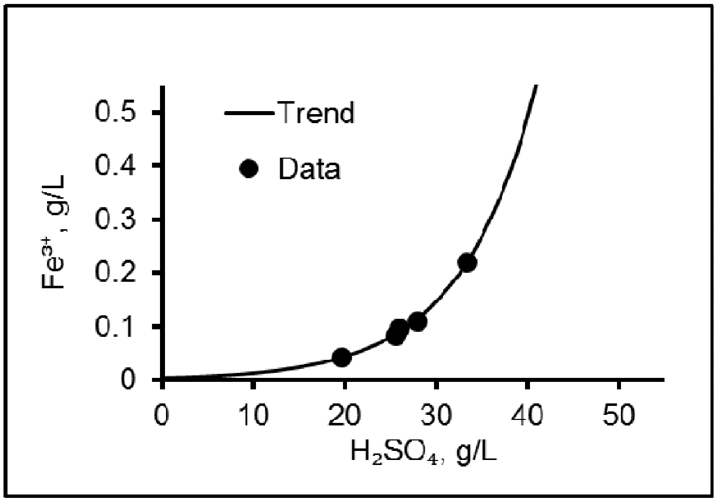


Figure 3 – Hydrolysis data

Table 5 and Table 6 show simplified chemistry representing the subsequent steps of the process illustrated in Figure 1. The calcium sulphate could be gypsum or anhydrite, depending on the exact process conditions. At this stage of the calculation, that does not matter.

Table 5 – Iron precipitation chemistry

$$H_2SO_4 + CaCO_3 \rightarrow CaSO_4 + H_2O + CO_2$$

$$Fe_2(SO_4)_3 + 3CaCO_3 + H_2O \rightarrow 3CaSO_4 + 2FeOOH + 3CO_2$$

Table 6 – Hydroxide precipitation chemistry

$$NiSO_4 + MgO + H_2O \rightarrow Ni(OH)_2 + MgSO_4$$

$$CoSO_4 + MgO + H_2O \rightarrow Co(OH)_2 + MgSO_4$$

$$CuSO_4 + MgO + H_2O \rightarrow Cu(OH)_2 + MgSO_4$$

$$CaO + H_2O \rightarrow Ca(OH)_2$$

$$MgSO_4 + Ca(OH)_2 \rightarrow Mg(OH)_2 + CaSO_4$$

Lime (CaO) is produced by heating limestone ($CaCO_3$) to between 900°C and 1000°C, driving off the carbon dioxide and leaving reactive calcium oxide. The energy consumption for this is 6 GJ per tonne of CaO produced[7]. In Africa, a convenient source of energy is coal. A typical heating value for coal is 30 MJ/kg, so a rough approximation of the amount of coal needed to produce one tonne of lime from limestone would be 200 kg.

This leads to the reagent consumption numbers listed in Table 7 for tank leaching at atmospheric pressure (ATL) and for leaching at elevated temperature and pressure (HPAL). These numbers, in conjunction with the unit costs listed, give the reagent costs for the two cases predicted by the chemistry assumed for this exercise. As might be expected, the calculated reagent cost range for the HPAL case is higher than the corresponding range for the atmospheric leach case because of the increased requirements for sulphur and limestone.

Table 7 – Reagent costs (30 kt/y Ni in product)

Reagent	Consumption, kg/t		Unit cost $/t	Variable cost, $ million/y	
	ATL	HPAL		ATL	HPAL
S	69	106	85 - 120	17 - 24	27 - 38
$CaCO_3$	214	172	100 - 250	64 - 159	96 - 239
MgO	7	7	160 - 280	3 - 6	3 - 6
Coal	17	17	85 - 100	4 - 5	4 - 5
			Sub-total	88 - 194	130 - 288

Source of unit costs: www.alibaba.com and World Bank commodity data

Table 8 lists metal price ranges and the associated potential revenue from the envisaged operation, assuming that the intermediate product can be sold for 75 percent of the nominal value of contained base metal.

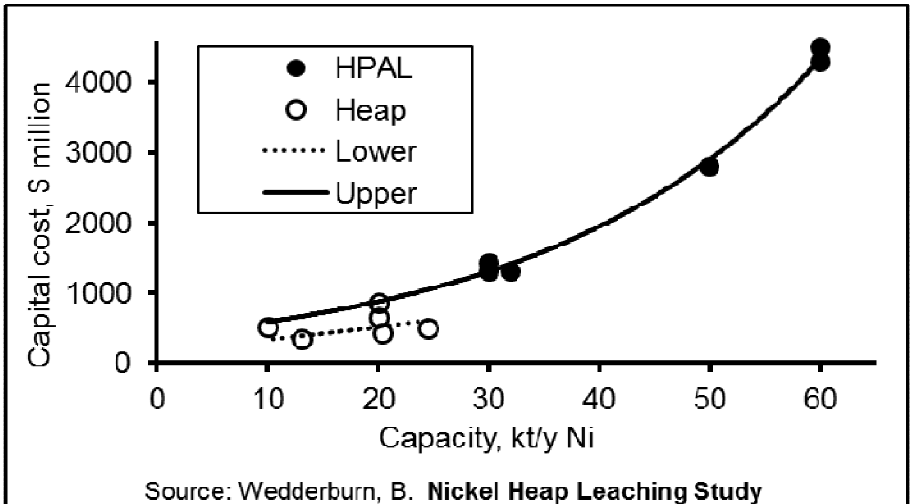
Table 8 – Revenue (75% of metal price, 30 kt/y Ni)

Metal	Metal price, $/lb	Revenue, $ million/y
Ni	8 - 10	396 - 495
Co	15 - 20	23 - 31
Cu	1 - 3	0.3 – 0.9
	Total revenue	419 - 527
Sources	MAPI, Industrial base Metal market Outlook 2013 Metal Bulletin	

The immediate conclusion is that the lower projected revenue exceeds the highest calculated reagent cost. Had the reagent costs exceeded the revenue numbers, of course, the exercise would rationally have ended and no further effort or expense would be wasted on that particular avenue of investigation, save perhaps a careful check of the assumptions leading to the results obtained.

Since the preliminary calculations indicate that revenue will exceed reagent costs, the next question is what other costs are involved and whether or not those plus the reagent costs low enough, compared to the revenue, for the project to be potentially viable. The costs for any venture are the variable operating cost, the fixed operating cost and the capital cost. The fixed operating cost comes from salaries, spares, inventory, etc. This number has to be guessed at this stage. For this exercise, the initial guess is 100 people at an average employment cost of $50 000 per year for salaries and one percent annually, of the capital cost, for spares and other consumables. The assumed employment cost is most probably not an accurate amount, but for want of a better number, it could be taken to be some sort of average between local and expatriate packages. At this stage of the calculation, over-estimation is preferable to under-estimation.

The capital cost is a number that comes from various levels of engineering design that has, at this stage of the exercise, not even been contemplated. However, an initial guess can be formulated using numbers from other similar projects. Searching the internet uncovered the capital costs shown in Figure 4, for projects processing laterite to a mixed hydroxide intermediate *via* high pressure acid leach (HPAL) and via heap leaching[8]. Three of the data points for heap leaching are very close to the regression line through the HPAL data points (Upper), while the other three lie well below this trend (Lower). Assuming that the example in this paper using atmospheric leaching can be made to fit the trend represented by the lower data points, extrapolating the lower regression line through these three points gives a capital cost of about $600 million for 30 kt/y of nickel production capacity. Assuming HPAL for the leach step, interpolating the upper regression line to 30 kt/y gives a capital cost of about $1300 million.



Source: Wedderburn, B. **Nickel Heap Leaching Study**
INSG Environmental & Economics Session, 27 April 2010

Figure 4 – Published capital costs for laterite projects

Part of the variable operating cost comes from the consumption of reagents and utilities and is thus proportional to the amount of product produced. Another part is the cost of getting the ore into the process, i.e. the variable cost of mining. A published generic cost model[9] for open pit mining, at 5000 tonnes/day of ore mined, gives a mine operating cost of about $5 per tonne of ore. Assuming a production rate of 30 000 tonnes per year of nickel in the hydroxide intermediate and an overall nickel recovery of 90 percent gives a mining rate of 6800 tonnes/day for the project in this exercise, which is of the same order of magnitude as the rate used in the published mining cost model, thus a mining cost of $5-7 per tonne of ore mined would seem to be reasonable. The mining model also gives a generic capital cost for open pit mining (which is how laterites are mined) of $1599 per daily tonne mined, or about $14 million for this exercise. However, the published costs shown in Figure 4 probably include mining, so at this stage the capital cost estimate for the mine can be ignored.

Having evolved all these very preliminary estimates for the various costs and revenues associated with the project under evaluation, simple cash flow analysis can be used to help decide whether or not the project is worth more substantial evaluation. Table 9 gives the results of this analysis for the case of atmospheric leaching. The assumptions used are a two year construction period with half of the capital expenditure in each of those years, followed by 20 years of production, with the throughput at 50 percent of design in year 3 and at 100 percent of design thereafter. Corporate tax was assumed to be 20 percent of the gross margin after recoupment of all the capital expenditure. For this calculation, the higher reagent costs and the lower metal prices were used. The mining cost was taken as $6/tonne of ore.

Table 9 – Cash flow analysis, $ million (atmospheric leach)

Year	1	2	3	4	5	6	7 to 22
Revenue			210	419	419	419	419
Costs	300	300	128	244	244	244	244
Margin	-300	-300	82	175	175	175	175
Tax	0	0	0	0	0	1	35
Profit	-300	-300	82	175	175	174	140
Before-tax IRR							22%
After-tax IRR							20%
Guesstimated capital cost, $ million							600
NPV at 10% discount rate, $ million							688

This cash flow analysis predicts an after-tax internal rate of return (IRR) of about 20 percent, for the higher calculated reagent cost and the lower calculated revenue. At the other extreme, i.e. using the lower reagent cost and the higher revenue in the calculation, the cash flow analysis gives an after tax IRR of 37 percent.

Table 10 shows the cash flow analysis for the case assuming HPAL for the leach, for the higher reagent costs and the lower metal prices. Clearly, under these assumptions the HPAL option would be a non-starter. If the lower reagent prices and higher metal prices are used in the calculation, the after tax IRR becomes 16 percent, less than half that of the atmospheric leach option under the same assumptions.

Table 10 Cash flow analysis, $ million (HPAL case)

Year	1	2	3	4	5	6	7 to 22
Revenue			210	419	419	419	419
Costs	650	650	181	345	345	345	345
Margin	-650	-650	28	74	74	74	74
Tax	0	0	0	0	0	0	0
Profit	-650	-650	28	74	74	74	74
Before-tax IRR							-3%
After-tax IRR							-3%
Guesstimated capital cost, $ million							1300
NPV at 10% discount rate, $ million							-131

The first conclusion to be drawn from these very preliminary calculations is that, for processing this particular laterite, the atmospheric leach clearly appears to be the stronger option, economically. The value of further work on developing an HPAL option for this feed would be dubious at best, unless the assumptions used can be shown to be incorrect, in which case the preliminary calculations would be repeated with better assumptions. Further work, it would seem, is merited for the atmospheric leach case.

A point to note is that the above preliminary calculations do not require sophisticated software – there is nothing that cannot be done quite easily on a spreadsheet, along with some surfing of the internet. The time needed for an exercise like this is no more than a few days of focussed effort, well worthwhile as the results can either greatly enhance confidence moving forward, or rationally discourage expensive experimental work if the numbers do not indicate a potentially viable project.

Whether or not these results lead to further effort would be a decision based on information relevant to the project in question. For the exercise presented here, the preliminary results will be taken as sufficient inducement to proceed to the next level of evaluation, which does get more sophisticated, computationally. From the preliminary numbers outlined above, it would seem logical to drop the HPAL option at this stage, but for completeness the HPAL option will also be examined in more depth.

PROCESS MODELLING

When the results of the preliminary evaluation indicate a sufficiently strong business case for the project, the next stage is to refine the calculated cost and revenue numbers so that the financial modelling can in turn be strengthened. As part of this, further experimental work becomes more than appropriate, as does a more detailed evaluation of the envisaged process. Process modelling can be used to generate and study a mass-energy balance of the envisaged circuit, optimally in parallel with the experimental work. This gives two major benefits – process modelling leads to greatly enhanced understanding of the envisaged circuit and enables the experimental work to be done much more cost effectively than would otherwise be possible; and when experimental results become available they can be incorporated into the mass-energy balance, which then becomes a very good basis for engineering design and costing work.

Circuit

For this exercise, two detailed process models were constructed using commercially available process simulation software known as Aspen Plus, and the mass-energy balances generated were used to refine the analysis of the project. Figure 5 is a diagram of the circuit modelled with the leach at atmospheric pressure and Figure 6 is a diagram of the HPAL variation of the circuit. The overall process chemistry is the same as before.

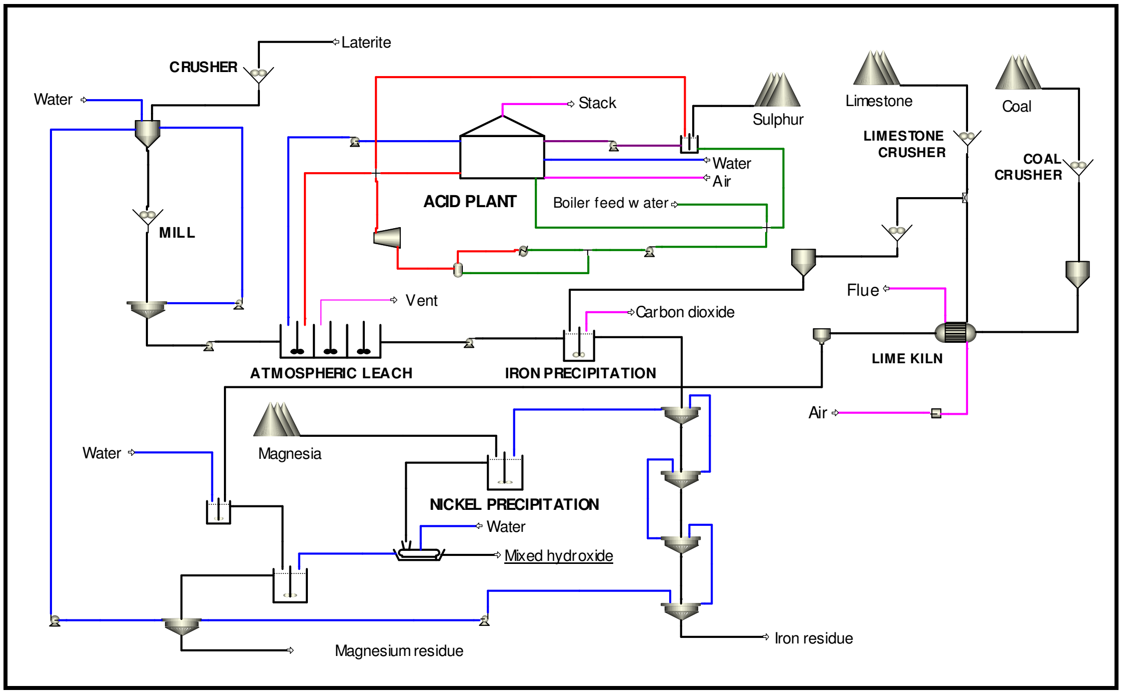


Figure 5 – Process model (atmospheric leach case)

Feed preparation

Incoming laterite is crushed and slurried with recycled and make-up water, the slurry is milled and the milled slurry is thickened, the thickener overflow returning to the slurry make-up step and the underflow being pumped to the leach.

Acid plant

A standard sulphur burning acid plant was assumed[10]. Incoming elemental sulphur is stockpiled as a solid, reclaimed and melted in a sulphur melting pit. The molten sulphur is pumped to a sulphur burner and burned in excess air. The energy released is captured as high pressure steam that is used to melt the sulphur, heat the leach and generate electricity. The gas from the sulphur burner passes through three catalyst beds in which the sulphur dioxide is oxidized to sulphur trioxide. The oxidation being exothermic, the gas passes through heat exchangers, raising more high pressure steam, between the catalyst beds. The gas from the third catalyst bed passes through an absorber in which the sulphur trioxide is captured into circulating strong acid and water is added, making 98 percent H_2SO_4, some of which is withdrawn to leach the milled laterite. The gas from the absorber passes through more heat exchangers and a fourth catalyst bed, then through another heat exchanger to a second absorber in which the sulphur trioxide is captured into the balance of the 98 percent sulphuric acid from the first absorber. The resulting strong acid returns to the first absorber and the residual gas is vented to a stack.

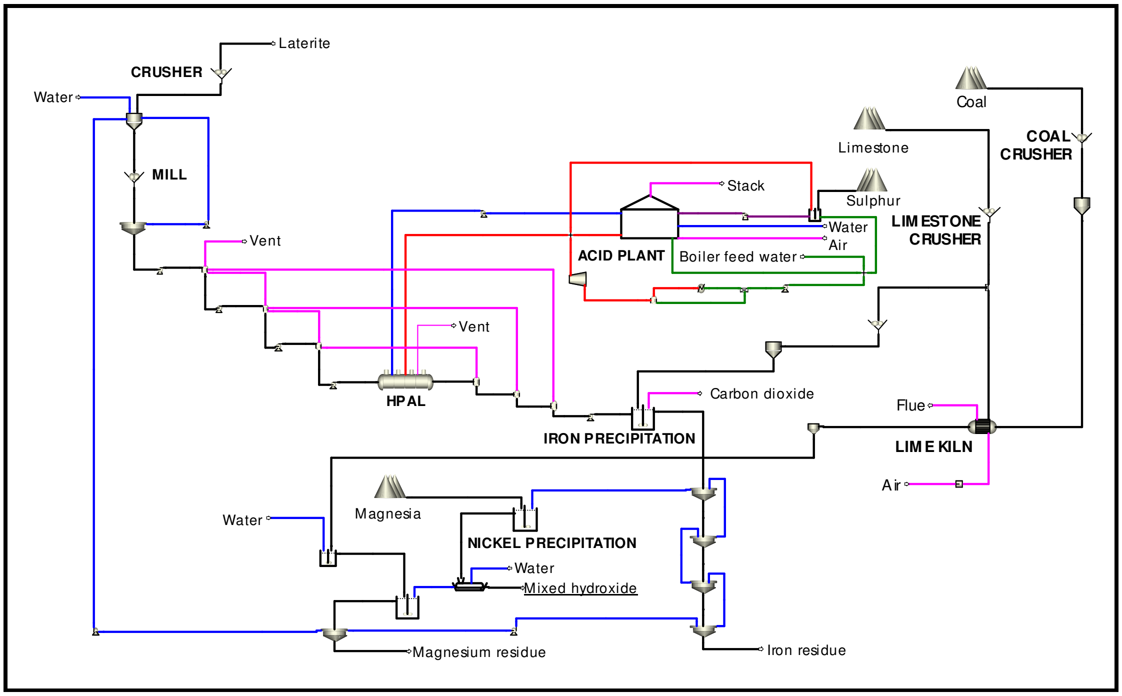


Figure 6 – Process model (HPAL case)

Leach

The leach is done at atmospheric pressure and 95°C in one case, and at 250°C and elevated pressure (HPAL) in the other case. In both cases the leach is heated by injection of live steam from the acid plant, but in addition the HPAL case assumes recycle of steam flashed off after the leach to pre-heat the incoming slurry.

The residence time for atmospheric leaching was assumed to be 20 hours, as mentioned on the African Eagle web site. For the HPAL case, the leach residence time was taken to be 75 minutes[11].

Neutralization and iron precipitation

The leached slurry is pumped to a neutralization stage, where limestone is added to neutralize the residual free acid and precipitate the dissolved ferric iron. The incoming limestone is crushed and a part of the crushed limestone is milled and used in the neutralization and iron precipitation step.

The neutralized slurry passes through a counter-current decantation train in which the solids are washed with recycled water. The washed solids leave the circuit.

Base metal precipitation

The supernatant from the counter-current decantation train is contacted with magnesium oxide to precipitate nickel, cobalt and copper as a mixed hydroxide that is filtered out and washed with water. The models assume dry MgO addition, but it could also be slaked first. The washed hydroxide leaves the circuit as the desired intermediate product.

Magnesium precipitation

The filtrate from the base metal precipitation step is contacted with slaked lime to precipitate the magnesium. The resulting slurry of magnesium hydroxide and calcium sulphate is thickened, the thickener underflow leaves the circuit and the supernatant is recycled as wash to the counter-current decantation train and to the slurry step ahead of the leach.

The slaked lime is made by calcining the crushed limestone not used in the iron precipitation step. A kiln was assumed for this duty, fired by burning coal in air. In this model, the ash from the coal accompanies the lime to the slaker.

Reagent costs

The first comparison of interest between the results from the preliminary stoichiometric calculations and the balances from the process model is between the reagent consumption numbers, as shown in Table 11. The numbers for limestone are the sum of the calcium carbonate used directly in the iron precipitation step and used to make the calcium oxide used in the magnesium removal step.

Table 11 – Reagent consumption, kg/t

Reagent	Atmospheric leach		HPAL	
	Stoichiometry	Process model	Stoichiometry	Process model
S	69	69	106	106
$CaCO_3$	214	227	323	348
MgO	7	8	7	8
Coal	17	22	17	22
Reagent cost, $M/y	88 - 194	96 - 210	130 - 288	142 - 315

Except for sulphur, the process models predict slightly higher reagent consumption numbers than the simple stoichiometric calculations. The differences are due to the solution volumes in the case of $CaCO_3$ and MgO, and the heating of solids and air in the case of coal. These effects were ignored in the earlier stoichiometric calculations. Using the reagent consumption numbers from the process models in the cash flow analyses done before reduces the IRR numbers by only 1 percent, which is not a significant difference at this level of evaluation. The process models, however, are numerically rigorous material-energy balances and the effects of solution volumes, heating or cooling and internal recycles are taken into account. It would be highly unusual to get significantly reduced reagent consumption numbers, compared to the results of stoichiometric calculations, from a process model, but it would not be unusual to get higher reagent consumption numbers. In most cases, provided the stoichiometric calculations are done correctly, there should not be a major difference between the numbers from these calculations and the process model, but the numbers from the process model would normally be a bit larger than the numbers from stoichiometric calculations.

COST MODELLING

Process engineering, equipment design and cost estimation have traditionally been undertaken by different groups of specialists who are not readily available to the people developing mining projects, particularly early in the life of a new project. Traditionally, engineering design and cost estimation might not be considered until a substantial body of experimental data has been generated, at which stage the project team would commission an engineering company for the necessary process engineering, equipment design and cost estimation. The desired and anticipated outcome of this work is that the project in question is economically viable. Appreciable levels of stress and tension are not unknown during the course of such work, because until the work is done the outcome is uncertain. No matter how deeply the proponents identify with and believe in the project, there is a risk that the time, effort and money spent up to that stage of any given project will not result in a commercially viable operation.

A second consequence of the traditional separation between the process and estimation disciplines is that when there are alternative process options, comparison of options can be based more on opinion than on dispassionate evaluation, simply because dispassionate evaluation requires capital cost estimation, traditionally unavailable for reasons of cost and time[12]. This is why approximately 80 percent of the capital cost of the process plant for any given project is sometimes determined by choices made in the conceptual design phase of the project[13], before any capital costs are known for that project. In the modern world of global competition, investment decisions bringing projects to fruition faster, at less cost and with minimum risk and uncertainty become ever more difficult.

It would be better to know sooner rather than later whether or not a project is sufficiently robust, economically, to justify its continued existence. It would also be better to know sooner rather than later that the most appropriate process option has been selected. While perfect prescience is not achievable, it is quite possible to evaluate the capital and operating costs of a project, and hence the potential economic viability of that project, significantly earlier than has previously been practical. Just as process engineers have developed their tools, so have cost estimators. Equipment sizing and costing algorithms and databases have been developed that generate equipment sizing and cost numbers from minimal process information. These databases and algorithms have been embedded into modern computer software that can be used to automate the connections between process design, equipment sizing and cost estimation. This enables the project team to iteratively evaluate project costs and compare options rapidly, without incurring major expenses.

Preliminary equipment sizing

The following illustrates the use of commercially available equipment sizing and costing software to estimate project costs. In the example used in this paper, the two cases are similar except for the leach section. The cost estimation software automatically sizes the various pumps and vessels in the circuit from the flows passed to it from the process model, assuming standard design codes and residence times, with carbon steel as the default material of construction, but the material of construction can also be specified - stainless steel or rubber lined carbon steel, for example, as can residence time and several other parameters.

The software used for the exercise presented in this paper is known as Aspen Process Economic Analyzer (APEA). This software can be activated directly from the process simulation software, thus the entire mass-energy balance is electronically transferred from the process model to the APEA software. The various unit operations in the process model are computationally mapped to process equipment that is automatically sized, based either on default parameters or parameters that are specified for individual items of equipment. The mapping can be left to the defaults in the software, or specified as input. For non-standard items such as leach tanks and autoclaves, the appropriate parameters were set manually. The sizing calculations give the amounts of material and labour required to fabricate and install the process equipment, as well as the materials and quantities required for foundations, supporting structures and process piping, instrumentation, electrical wiring, etc. In the exercise presented in this paper, the software was set to give costs for a project in Africa. The equipment sizing was set up as follows.

Ore preparation

In the ore preparation section, incoming ore is crushed, slurried in water and milled. The milled slurry is thickened and the thickened slurry proceeds to leaching. The surplus water from the milled slurry is recycled.

The minimum process information required for the crusher by the sizing/costing algorithm is the power input, which can be approximated from the work index of the ore, the crushed product size and the throughput. Since the work index has not been measured (it probably has, but the number is not available to this exercise), a plausible number from the published literature has to be used instead. One published number for an Indonesian laterite is 9.7 kWh/t[14] and the ore throughput for this exercise is 360 t/h. Crushing the ore to 3 mm particles requires a crushing power of 638 kW (or 1.8 kWh per tonne of ore). The crusher was assumed to be fed from a belt conveyor 100 m long and 1 m wide.

The information required for costing the mill is the internal diameter and the length. Outokumpu (now part of Outotec) have published useful information on preliminary mill sizing[15], specifically a correlation linking the mill power to the mill type and dimensions. Assuming the same throughput and work index as for the crusher, and that the ore is milled from 3 mm to 100 microns, gives a milling power requirement of 2854 kW (7.9 kWh per tonne of ore). Using that power and the factors tabulated in the Outukumpu pamphlet gives a mill diameter of 3.7 m and a length of 17 m.

The slurry tank between the crusher and the mill was assumed to be a carbon steel agitated tank with a residence time of an hour, to allow the crusher to be run intermittently if necessary.

The mill thickener was assumed to be a standard thickener made from carbon steel. The thickening rate of the milled slurry not having been measured yet, 0.1 $t/h/m^2$ was assumed. This is the same number as that used for the counter-current decantation train after leaching, which was inferred from published data from a pilot plant (details in the section on counter-current decantation below). The resulting thickener diameter is 68 m.

Acid plant

The sulphur melting pit was assumed to be an agitated tank holding two hours of molten feed to the sulphur burner and to be heated by a steam coil using steam from the acid plant. The material of construction was set to stainless steel.

The cost of the sulphur burner was approximated by the cost of a process furnace, the material of construction being stainless steel. The steam raising part of the sulphur burner was approximated by a standard heat exchanger, also of stainless steel.

The converters were sized from a standard design[16], assuming the same superficial gas velocity entering the first catalyst bed and the same bed depth, which gave the vessel size and the volume of the catalyst beds. The catalyst bulk density and price were taken from a catalogue of sulphuric acid catalysts[17]. The various pumps and heat exchangers were sized automatically by the software. The material of construction was stainless steel throughout.

The two absorber columns and the various pumps and heat exchangers in the acid plant were sized from the mass/energy balance, using correlations embedded in the software. The material of construction was set to stainless steel for all the pumps and heat exchangers.

Atmospheric leach

In the circuit using atmospheric leaching the underflow from the mill thickener is pumped into the leach train, where it is leached for 20 hours. The reactors were specified as ten agitated tanks, each having a residence time of two hours, with the tanks and agitators made of rubber lined steel. The material for the slurry pump after the leach train was specified as stainless steel.

HPAL leach

In the circuit using HPAL for the leach section, the milled and thickened ore slurry is passed through three direct contact heat exchange vessels in which it is progressively heated by condensing steam flashed from the hot leached slurry. These vessels were sized for a slurry residence time of ten minutes each, and specified as titanium clad steel lined with acid resistant bricks to protect the vessel against erosion.

The pre-heated ore passes to the leaching step. The autoclave design was based on data from the Ambatovy pilot plant[11]. Anaconda's Murrin Murrin project had four autoclaves, each 35 m long and just less than 5 m in internal diameter[18], made from titanium clad steel. The residence time in HPAL given in the work on the Ambatovy pilot plant[11] is 75 minutes and the slurry into the preheating step contained 30 percent solids. That solids content and residence time translates to a required active volume of 1500 m^3 for the 360 t/h solids throughput of the exercise in this paper. That requires four autoclaves like the ones at Murrin Murrin, assuming the active volume to be 60 percent of the total internal volume of the autoclave. For costing purposes, each autoclave was assumed to be a six-compartment agitated horizontal tank reactor with the wall thickness designed for a working pressure of 40 Bar, made from titanium clad steel, (8 mm titanium). The agitators were assumed to be six per autoclave, fitted with mechanical seals and made from titanium.

The hot pressure leached slurry is flashed to atmospheric pressure in three let-down vessels. The residence time was left as the default for flash vessels and the material of construction was set to titanium clad steel lined with acid resistant bricks.

Specifying titanium cladding and acid resistant bricks for the heating and let-down vessels may well be overkill, but this was done for preliminary cost estimation purposes. The cost database used does not have costs on the actual high-pressure slurry pumps found in HPAL, so the slurry pump costs were based on data for less sophisticated pumps, and therefore probably under-estimated. Specifying double linings for the heaters and flash vessel would hopefully compensate for this. In a real-life exercise, this gap in the cost modelling would be exposed and more work would be done if the HPAL option were to be pursued seriously.

Iron precipitation

The leached (and depressurized, in the case of HPAL) slurry is neutralized with limestone. The process equipment was assumed to be three agitated tanks in series, at a residence time of 20 minutes per tank. The material of construction for the tanks and the agitators was set to rubber-lined steel.

The incoming limestone was assumed to be crushed to 3 mm and then milled to 100 microns. A typical work index for limestone is 103 kWh/t. The limestone throughput is 82 t/h for the crusher and 24 t/h for the mill, in the atmospheric leach case. That gives a crusher power of 153 kW and a mill 3.7 m in diameter and 1.2 m in length. For the HPAL case the throughput of limestone through the crusher is 125 t/h and through the mill is 68 t/h. In this case the crusher power is 408 kW and the mill is 2.8 m long, with the same diameter as for the atmospheric leach case.

Counter-current decantation

The solids are separated from the neutralized leach solution and washed in a seven-stage counter-current decantation train. (The process model diagrams show four thickeners in the train, but the costing was done for seven.) The Ambatovy pilot plant data was used to derive a thickening rate, as follows:

- The working volume of the pilot autoclave was 30 L, the feed slurry contained 30 percent solids and the residence time in the autoclave was 75 minutes. That gives a solids feed rate of 7.2 kg/h.
- The thickeners in the counter-current decantation train were 30 cm in diameter. Assuming the rate of solids into the autoclave to be approximately the same as the rate of solids out of the first neutralisation step, each thickener settled about 7.2 kg/h of solids, making the solids settling rate about 0.1 t/h/m^2.
- In the exercise presented in this paper the solids flows through the counter-current decantation train are 375 t/h for the atmospheric leach, and 447 kg/h for the HPAL case. At the assumed settling rate of 0.1 t/h/m^2, that gives thickener diameters of 69 m for the atmospheric leach case and 75 m for the HPAL case, assuming single-train counter-current decantation.

The required parameter for thickener costing is the total volume of the thickener. Using the settling rate to calculate a diameter and assuming a sidewall depth of 3 m and a cone angle of 10° leads to the total volume in the thickener. Optional specifications selected were that the bridge is from the centre to the edge of the thickener, the drive and rake are heavy duty with overload alarm, a flocculation system is included and that the whole thickener is made from rubber-lined steel. Each thickener was specified to include an underflow pump, made from stainless steel.

Nickel precipitation

The supernatant from the counter-current decantation train is contacted with powdered magnesium oxide to precipitate the nickel, cobalt and copper, in three agitated tanks in series. The residence time in each tank was set to 20 minutes. The material of construction was left at the default, carbon steel. The MgO was assumed to be purchased, stored in a conical bottomed hopper and metered into the precipitation tanks via a screw feeder. It could also be slurried in batches and the slurry fed into the precipitation reactor. The dry feed choice is arbitrary.

The resulting slurry is filtered. A rotary drum filter was specified for costing purposes. Since no filtration test results were available, the filtration rate was set to a default representing slow filtration and the software was left to size the filter from the data transferred from the process model and its default numbers for slow filtration. The material of construction was set to epoxy lined steel.

Magnesium precipitation

The filtrate from the nickel precipitation step is contacted with slaked lime to precipitate the dissolved magnesium. For this, crushed limestone is calcined in a coal-fired kiln and the burnt lime is held in a surge vessel before being slaked with water in an agitated tank for 10 minutes. The slaked slurry overflows into a second tank where it is diluted with more water and grit is settled out and removed by a screw conveyor. The dry lime is fed into the first tank from one of two hoppers by a rotary feeder (24 hours storage capacity, CaO bulk density 0.5 t/m^3). The material of construction was left as carbon steel throughout, for this equipment.

The input power to the coal crusher was calculated from a generic work index for coal (11.3 kWh/t) and the assumption that it is crushed to 100 microns. The coal throughput is 8 t/h and the crusher power is 90 kW.

The lime kiln was approximated as a horizontal drum lined with fire bricks and heated by direct contact with the coal-air combustion gas. The coal was assumed to be fed from the crusher to the kiln by a volumetric belt feeder. Air was assumed to be blown in using a fan.

The filtrate from base metal precipitation and the slaked lime are contacted in two agitated tanks in series (carbon steel, residence time 30 minutes each). Calcined lime is held in an intermediate storage hopper (200 m^3) and metered into the slaking system by a volumetric belt feeder.

In the absence of measured information, the thickening rate of the limed slurry was assumed to be the same as the rate in the counter-current decantation train, i.e. 0.1 t/h/m^2. That gives a thickener diameter of 40 m for that thickener in both variations of the process. The optional parameters for this thickener were set to the same as the thickeners in the counter-current decantation train, except that in this thickener the material of construction was left as carbon steel.

Infrastructure

For ranking alternatives, the infrastructural cost components can be ignored because these items would be common to the competing alternatives. In this exercise the infrastructural items were included because the capital cost for a new project does not consist only of the cost of the process equipment. The software also has cost models for various items of infrastructure.

Tailings

The process model for the atmospheric leach case predicts 155 m^3/h solids and 615 m^3/h water in the leach residue plus the magnesium precipitate (45% solids by mass, combined). Assuming the final consolidated tailings to contain 80 percent solids, the volume would be 379 m^3/h. At a rise rate of 4 m/year[19] gives a tailings dam area of 793 thousand square meters, with a perimeter of 3567 m.

The process model for the HPAL case predicts 411 m^3/h of consolidated tailings. At the assumed rise rate of 4 m/y, that gives a tailings dam of just over 890 thousand square meters with a perimeter of 3775 m. The higher volume of consolidated tailings in the HPAL case is due the higher production of gypsum, from higher consumption of sulphuric acid and limestone.

A tailings slurry pump station and a pipeline 2000 m long and 0.4 m in diameter were assumed for both cases. The pipe material was assumed to be FRP (more because there is a variety of material in the cost modelling database than for any real reason).

For both cases the tailings dam was assumed to require a starter dyke 5 metres high. The base of the tailings dam was assumed to require brush clearing and excavation to a depth of half a meter. The bottom of the tailings dam was assumed to be a clay liner and two geosynthetic membranes, plus a grid of piping to assist drainage. Each year the dyke would need to be raised by at least 4 meters. The cost of the starter dyke ($0.7 million) was thus added to the fixed operating cost, to cover the annual raising of the dyke around the tailings dam.

Process water

The process water was assumed to circulate through a fenced pond holding 48 hours of the water going to the counter-current decantation stage as wash and to the feed preparation section to slurry the incoming crushed laterite.

Make-up water was assumed to come from 25 km away, via a pipeline and pumping station.

Cooling water

Cooling water, 4767 t/h in the atmospheric leach case and 5500 t/h in the HPAL case, mostly for condensing the residual steam from the acid plant after electricity generation, was assumed to circulate through an evaporative cooling tower. The HPAL case uses more sulphuric acid and thus generates more steam, hence the higher cooling load there.

Other

The project was assumed to require a heavy duty rail link 80 km long and a paved road 20 km long. The required buildings were assumed to be a plant office and a laboratory, each 200 square meters. A parking lot (250 m^2) and a security fence (3 m high, 2000 m long) were assumed. A power line (11 kV, 20 km) and a standby generator (1500 kVA) were also assumed.

Cost estimates

The cost modelling software generated the capital cost estimates shown in Table 12 and Table 13 from the two mass-energy balances and the equipment sizing described above. These costs exclude the capital cost for the mine, which would be the same in both cases. The direct field cost is the cost of the process equipment and its installation, including items like a motor control centre that the software adds automatically. The equipment costs are the material and manpower costs for manufacturing and installing the various items of process equipment. Process piping, instrumentation, electrical wiring, foundations and structural steel are included in this number. The software has default piping allocations, electrical allowances, foundation standards, etc. When more detailed engineering work is done, these defaults are replaced by the results of the relevant work. The indirect field costs cover home office costs, field supervision, start-up and commissioning. The non-field costs are for freight, taxes and permits, basic and detailed engineering, procurement, overheads and contract fees. These are all default numbers at this stage, but defaults based on extensive real data. The contingency of 15 percent was set to allow for the process being new.

Table 12 – Capital cost estimate for the atmospheric leach case, US$

Account	Man-hours	Wage Rate, $/h	Labor Cost, $	Material Cost, $	Total Cost, $
(2) Equipment	23,358	29.88	697,969	108,990,984	109,688,953
(3) Piping	291,832	29.61	8,641,373	37,808,814	46,450,187
(4) Civil	2,230,368	24.19	53,958,053	94,724,889	148,682,943
(5) Steel	7,397	27.79	205,591	1,157,698	1,363,289
(6) Instruments	22,558	30.23	681,875	6,402,698	7,084,573
(7) Electrical	24,973	29.17	728,510	3,293,603	4,022,113
(8) Insulation	70,540	22.57	1,592,398	1,653,707	3,246,105
(9) Paint	36,343	22.29	810,201	668,398	1,478,599
Direct Field Costs	**2,707,367**		**67,315,970**	**254,700,791**	**322,016,761**
Indirect Field Costs	**297,593**				**94,443,005**
Total Field Costs	**3,004,960**				**416,459,766**
Freight					30,564,100
Taxes and Permits					10,188,001
Engineering and HO	245,221				26,112,402
Other Project Costs					30,296,888
Contingency					92,451,808
Non-Field Costs	**245,221**				**189,613,199**
Project Total Costs					**606,072,965**

Table 13 – Capital cost estimate for the HPAL case, US$

Account	Man-hours	Wage Rate, $/h	Labor Cost, $	Material Cost, $	Total Cost, $
(2) Equipment	39,704	29.93	1,188,378	182,940,004	184,128,382
(3) Piping	603,245	28.41	17,139,774	70,867,200	88,006,975
(4) Civil	2,163,936	23.96	51,846,017	97,183,546	149,029,563
(5) Steel	9,743	27.77	270,587	1,533,526	1,804,113
(6) Instruments	26,456	30.23	799,837	11,472,847	12,272,683
(7) Electrical	31,708	29.20	926,038	5,067,739	5,993,777
(8) Insulation	114,558	22.57	2,585,976	2,666,900	5,252,876
(9) Paint	37,886	22.30	844,694	686,587	1,531,281
Direct Field Costs	**3,027,236**		**75,601,302**	**372,418,349**	**448,019,651**
Indirect Field Costs	**352,708**				**106,535,406**
Total Field Costs	**3,379,944**				**554,555,057**
Freight					44,690,000
Taxes and Permits					14,896,701
Engineering and HO	261,872				27,910,702
Other Project Costs					39,101,788
Contingency					122,607,776
Non-Field Costs	**261,872**				**249,206,967**
Project Total Costs					**803,762,024**

The total installed equipment costs calculated are $283 million for the atmospheric leach case and $394 million for the HPAL case, or about 88 percent of the total direct field cost. The remaining 12 percent of the direct field cost covers items like electrical sub-stations and other hardware added by the software.

Interestingly, while the extrapolations from published capital cost information predict that the HPAL case would cost about twice the cost of the atmospheric leach case, the numbers found in the exercise presented here indicate that the total capital expenditure for the HPAL case will be about one third higher than the capital cost of the atmospheric leach case.

The algorithms inside the cost modelling software predict the fixed operating costs listed in Table 14. Inserting the capital and operating cost estimates generated by the cost modelling software into the cash flow analyses done before gives the results shown in Table 15 for the atmospheric leach case and in Table 16 for the HPAL case. In the cash flow analyses the capital cost is the total capital cost calculated for the process plant plus the capital cost for the mine ($14 million added to the cost of the process plant).

Table 14 – Fixed operating costs, $million/year

Cost item	ATL	HPAL
25 operators/shift (S20/h, 4 shifts, 2000h/shift)	4.0	4.0
2 supervisors/shift (S35/h, 4 shifts, 2000 h/shift)	0.6	0.6
Maintenance cost	4.1	6.0
Utility cost	9.0	16.1
Operating charges, plant overhead, G&A	7.3	9.1
Tailings dam dike	0.7	0.7
Total operating cost	25.7	36.4

Table 15 – Cash flow analysis, atmospheric leach, base case, $ million/year

Year	1	2	3	4	5	6	7	8 to 22
Revenue			210	419	419	419	419	419
Costs	310	310	135	244	244	244	244	244
Margin	-310	-310	75	175	175	175	175	175
Tax	0	0	0	0	0	0	31	35
Profit	-310	-310	75	175	175	175	144	140
Before-tax IRR								21%
After-tax IRR								19%
Capital cost estimate, $ million								620
NPV at 10% discount rate, $ million								334

Table 16 – Cash flow analysis, HPAL, base case, $ million/year

Year	1	2	3	4	5	6	7	8 to 22
Revenue			210	419	419	419	419	419
Costs	409	409	201	366	366	366	366	366
Margin	-409	-409	9	54	54	54	54	54
Tax	0	0	0	0	0	0	0	0
Profit	-409	-409	9	54	54	54	54	54
Before-tax IRR								-2%
After-tax IRR								-2%
Capital cost estimate, $ million								818
NPV at 10% discount rate, $ million								-417

Even though the capital cost of the HPAL option decreased from the $1300 million assumed initially to the $804 million calculated from the process and cost modelling exercise, the HPAL case remains a non-starter under the assumptions used for these cash flow analyses, i.e. the lower metal prices and the higher reagent unit costs from Table 7, a two year construction period, production at half of design in year three and at design thereafter and tax at 20 percent of the gross margin after recoupment of the capital costs. At the higher metal prices, the after tax IRR for the HPAL case becomes 11 percent, still considerably lower than the after tax IRR for the atmospheric leach case, using the lower metal prices. At the higher metal prices the after tax IRR for the atmospheric leach case is 29 percent.

The point of this is that the calculations based purely on the overall stoichiometry and information from the literature gave results showing the HPAL case to be weak, economically. The results of the substantially more detailed process and cost modelling done on the HPAL option confirm that HPAL is not a good candidate for the feed of this exercise. A reasonable generalisation would be that if preliminary calculations indicate that a project is economically weak, it most probably is.

Figure 7 shows how the installed equipment costs are distributed between the various parts of the process. The leach section makes up a much greater proportion of the total in the HPAL case, with the estimated cost of the HPAL autoclaves and ancillaries being about ten times the estimated cost of the atmospheric leach reactors. The acid plant is also bigger and therefore more expensive in the HPAL case because of the higher overall acid requirement arising from the nature of sulphuric acid at the HPAL leach temperature, i.e. having only one proton per molecule of sulphuric acid instead of two at the lower temperature of the atmospheric leach.

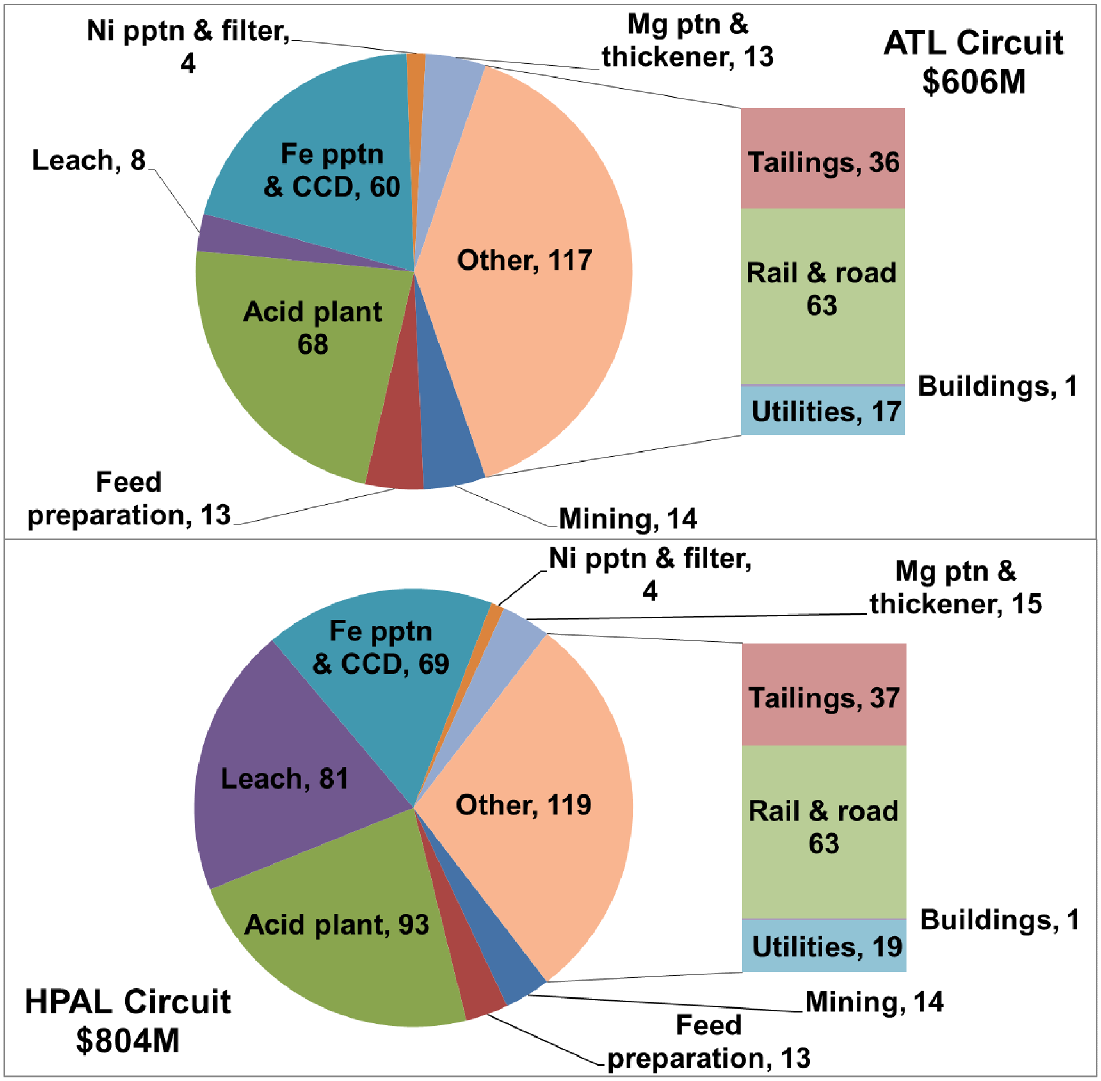


Figure 7 – Breakdown of the installed equipment costs, $ million

Design changes

Once the process models and the costing models have been set up, it is quick and easy to evaluate different scenarios. For example, if the settling rates in the various thickeners are actually half or double the value assumed (i.e. 0.05 or 0.2 instead of 0.1 t/h/m²) the thickener diameters would change accordingly, leading to a corresponding change in the estimated capital cost. Evaluating the impact of the thickening rate on the projected economics of the project is a simple matter of changing a few numbers in the thickener sizing calculations and the input to the cost model, re-running the cost model and saving the output to another spreadsheet.

To illustrate this, Figure 8 shows the impact of changing the settling rate on the installed equipment cost, for the atmospheric leach case. The settling rate in the base case is 0.1 t/h/m², in the slow settling variation it is 0.05 t/h/m² and in the fast settling variation it is 0.2 t/h/m². The projected total capital cost changes from $606 million in the base case to $747 million for the slow settling rate of 0.05 t/h/m² and to $538 million for the fast settling rate of 0.2 t/h/m². The cost of the equipment for the iron precipitation and counter-current decantation section changes from $60 million in the base case, to $104 million for the slow-settling variation and to $37 million in the fast settling variation.

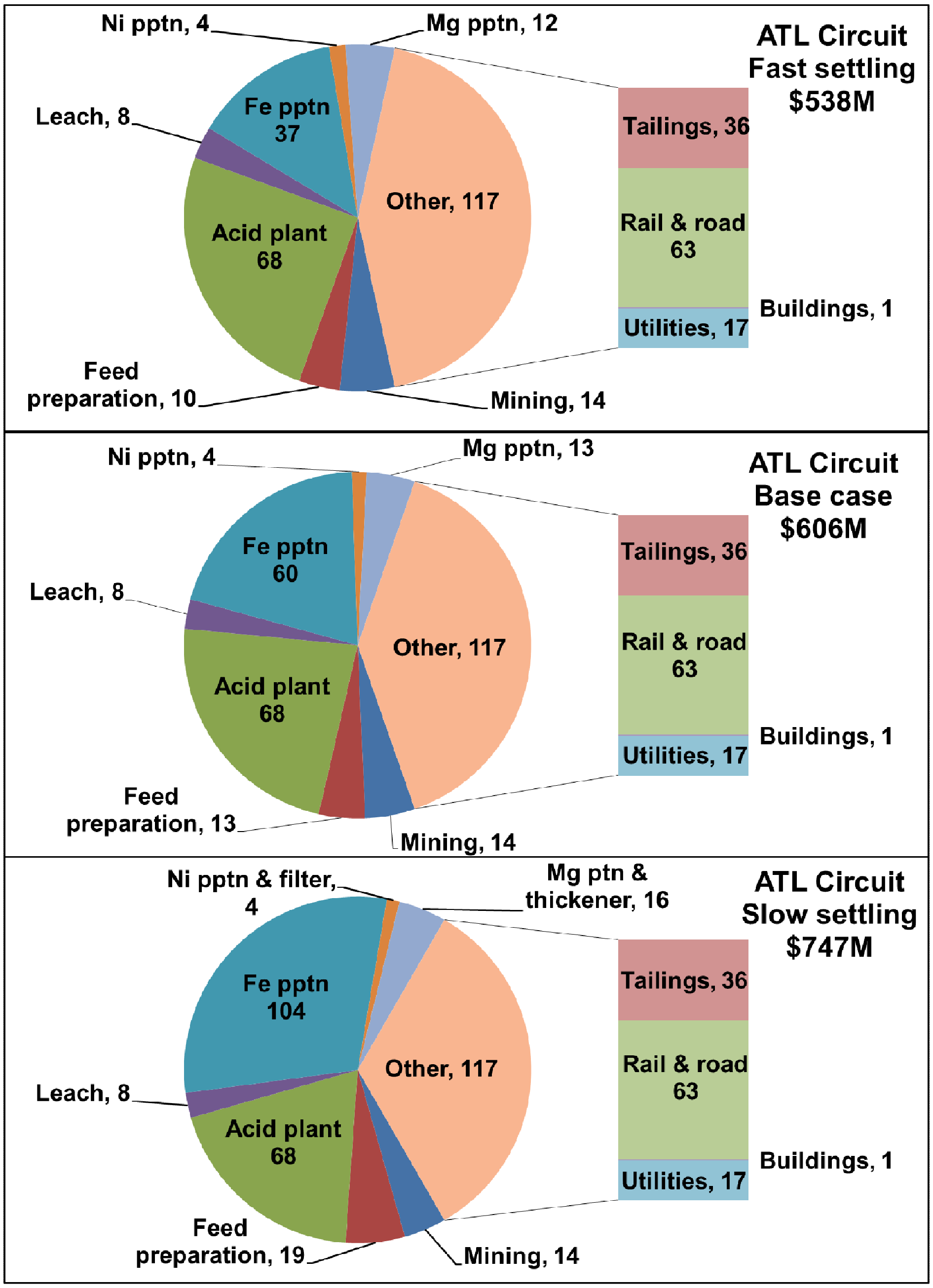


Figure 8 – Equipment cost distributions, atmospheric leach case, varying settling rate, US$ million

Settling rates in thickening steps are not always measured in preliminary experimental work, but given their potential impact on the capital costs, they should be.

The point of this illustration is not that the settling rate is or should be any particular value - that has yet to be determined. (In the case of the actual project, it probably has been determined, but the result is not available to this exercise.) The point of this illustration is that the settling rates in the various thickeners can have an appreciable impact on the overall capital cost, therefore settling rates should be measured at as early a stage of the experimental work as can be arranged, even knowing that later work will generate improved or more accurate settling rates. In this illustration, doubling the settling rate has a substantially smaller impact on the calculated capital cost than halving this number, which means that early-stage experimental work would be more focussed on finding out whether or not the various slurries settle “normally”. If early work uncovers especially slow settling, work can be planned to improve the settling rate, confirm that it is slow enough to make the circuit of which it is a part unviable, or to find a better solid-liquid separation technique for the project in question. If reasonably fast settling rates are found early on, there would be little rational incentive for extensive experimental work aimed at further improving the settling rate, as the impact of further improvements on the capital cost of that project would be small.

Capacity changes

A question that commonly arises when processes are under evaluation is what effect the throughput has on the economics. If the changes in throughput are not excessively large, the fixed operating cost is, at least to a first approximation, unaffected and the variable cost is directly proportional to the throughput. That leaves the capital cost, which is traditionally scaled using the “0.6 rule”, i.e. the baseline capital cost is multiplied by the ratio of the new throughput to the baseline, to the power 0.6. The cost modelling software used for the exercise presented in this paper contains a feature that enables the effect of throughput on the capital cost to be evaluated more rigorously. As the capital cost is based on equipment sizing results that are based on input from a numerically rigorous mass/energy balance, and that balance can be scaled linearly, the cost modelling software can easily recalculate the capital cost for throughputs other than the baseline by scaling the balance and re-sizing the various items of process equipment, support structures, foundations, etc., and recalculating the capital cost from the various quantities of material and labour so recalculated. If any item of process equipment, when re-sized, falls outside the range covered by the capital cost database, the software warns the user and then the user would specify duplicate items or make whatever other design change is required. Figure 9 shows the results obtained from the “0.6 rule” and using the cost modelling software to calculate the capital cost of the atmospheric leach case, from half to double the baseline capacity of 30 kt/y of nickel in the mixed hydroxide product. In this exercise the mine cost was excluded.

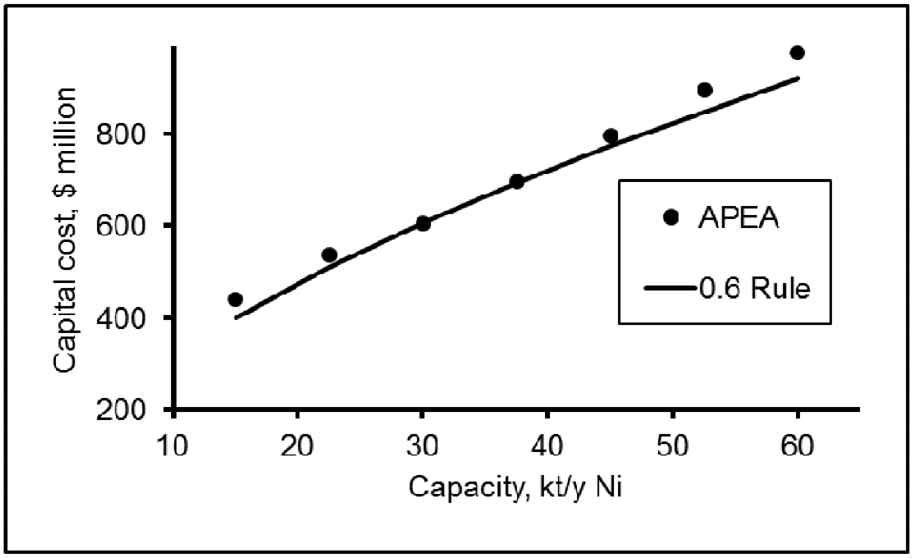


Figure 9 – Effect of throughput on capital cost, atmospheric leaching

The “0.6 rule” actually gave reasonably good results for this exercise, compared to the more rigorous calculation from the cost modelling software (APEA). The “0.6 rule” does seem to under-estimate slightly as the extrapolation gets larger (in the case of this example, at least), but the difference is unlikely to be significant at the preliminary stage of evaluation assumed for the exercise presented here. This actually supports the initial calculations based on the process stoichiometry and capital cost information from the literature. The “0.6 rule”, it would appear from this exercise, is completely adequate for producing initial guesses of capital cost from numbers published for something similar to the process in question. Naturally, if no numbers can be found in the literature for any given process, the "0.6 rule" cannot be used. The cost modelling software makes no use whatsoever of published project costs, as it is based on the material and labour costs for manufacturing and installing process equipment.

CONCLUSION

This paper presents and advocates a computational approach to the early-stage evaluation of new metal extraction projects. By way of calibration of the approach advocated in this paper, a press release in 2009 reports a capital cost estimate of $435 million for African Eagle's Dutwa Nickel Project, at an annual throughput of two million tonnes per year of ore[20]. Another press release from African Eagle, released in December 2010, gives a capital cost estimate of $600 million for a production rate of 23 kt/y of nickel[21]. A third release gives capital costs of $600 million and $840 million for facilities producing 30 kt/y and 50 kt/y, respectively, of nickel in a mixed hydroxide precipitate[22]. In Figure 10 the symbols plot the capital costs published by African Eagle for the Dutwa Nickel Project and the lines are the capital costs calculated for the exercise presented in this paper, using the process and cost modelling software.

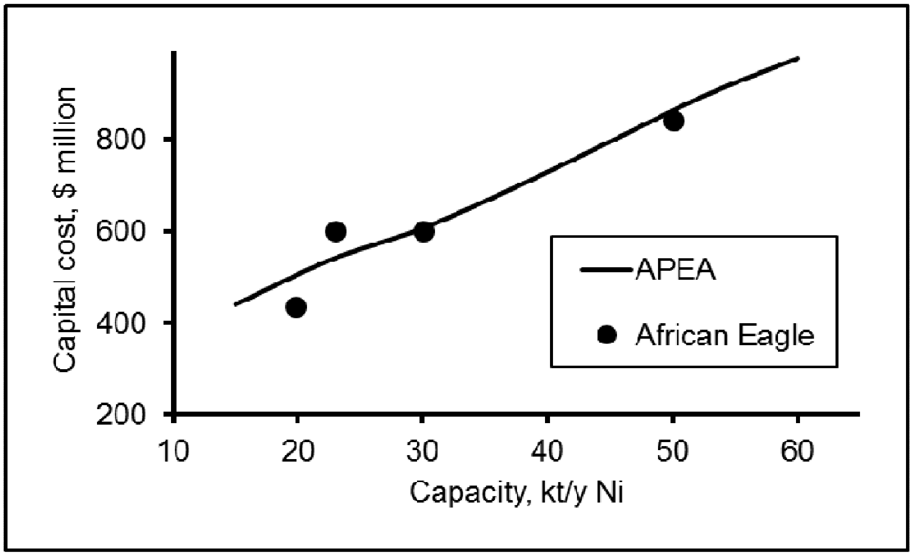


Figure 10 – calculated and published capital costs

The capital costs calculated in the exercise presented in this paper pass through the numbers published by African Eagle, which, having been generated by process engineering companies, are most probably based on much more engineering design work than was used in the cost modelling exercise presented in this paper.

The capital cost numbers calculated using the capital cost modelling software differ from the corresponding published numbers by less than ten percent. That is well within the uncertainty generally associated with initial capital cost estimates. This does not, of course, mean that the capital cost estimates developed in this work are accurate to plus or minus ten percent – getting to that would take the appropriate experimental and engineering design work. What it does mean is that it is possible to generate capital cost numbers that are entirely plausible, early in the life of a project, from general knowledge of the technology and a minimum of specific information.

The objective of the exercise presented in this paper is not to determine whether or not the Dutwa Nickel Project has business muscle. This project was chosen because there is enough published about it to enable sensible comparison between the calculated capital costs and corresponding published numbers.

A particular point of note is that the capital costs calculated in the work done for this paper were based on the mass-energy balances from the process models, default design codes and very basic equipment sizing calculations. No plant layout drawings, P&ID diagrams or piping isometric drawings are required in this approach, yet still the cost modelling software generated realistic capital cost estimates. The bulk of the effort in this work was building the process models. After that, transferring the balances electronically to the costing software, setting up the mapping of process equipment and generating the capital cost estimates took only a fraction of the effort that went into the process models.

The message of this paper is that new projects can be sensibly and very cost effectively evaluated early in the development cycle, essentially as soon as enough is known about the ore to generate estimates of its overall quantity and composition, and how it behaves in a leach. Preliminary calculations based on the overall stoichiometry of the envisaged process and information in the open literature can be used to generate sensible preliminary evaluations of whatever process options are under consideration, quickly and inexpensively. When the preliminary calculations indicate a process option to be potentially viable, process and cost modelling can be used to predict rational capital and operating costs from limited data, and these numbers can be used to confidently rank process options and justify work or the termination of work on any given project much earlier in the game than would be possible without these tools.

No skilled person would knowingly choose to spend time, effort and money on a doomed project, but all too often, even though the experimental work is completed without uncovering any technical flaws in a circuit, the economics may not stand up. Realization that the project is doomed then comes after the expenditure of considerable amounts of effort, time and money that could have been deployed elsewhere. Since process modelling and cost estimation are inevitable steps in the development of a project, it would be wise to not commit to extensive experimental and other work before starting the mass/energy balancing and cost estimation that will be needed anyway to evaluate its economics. Far better to use preliminary experimental results for setting up the process and cost models, then use the models to evaluate the project and, assuming that the models predict sufficiently strong economics, use them to guide the experimental work, feed the data back into the models, etc.

In the case of viable projects, judicious use of these tools can make the whole project development cycle much more cost and time effective, especially for dispassionately ranking process options, including novel technology, and determining the best program of experimental work. Identifying and testing the parameters most strongly affecting the process economics is also possible using these tools, again leading to better results and less wasted effort.

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