

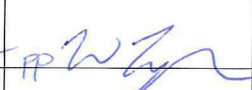


## Report

# Capacity Constraints and Suggestions for Debottlenecking to 400t/mth and 500t/mth V (contained in FeV)

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## Executive Summary

Hatch was engaged to review the Windimurra refinery (from the rotary kiln through to final product packaging) to give an opinion on whether the plant will be capable of producing 400-500t/month vanadium contained in FeV. This report is a result of a short visit of the inactive plant and a review of the plant's design basis and operating data. Hatch were also asked to study in greater depth the capacity of the rotary kilns (both Salt Roast and Reduction) using our proprietary software which models the solids flow, heat and mass transfer within kilns

The key conclusions of this study are as follows:

- The plant operating factor (a function which converts the plant's normal hourly throughput into an annual production capacity) was specified to be ~90%. This only includes the number of operating hours per year, and does not factor in lost production due to the plant ramp up/slow down etc. A long term operating factor for this type of plant is more likely to reach 85%.
- The vanadium recovery is strongly and negatively impacted by silica in the concentrate feed to the kiln. For vanadium recovery to approach the design 83%, close control of feed silica content, kiln temperature profile and soda ash addition will be required.
- The principal bottleneck of the plant is most likely to be the salt roast kiln and is the key to improved operation. We recommend the following strategy:

### Stabilise and optimize at 85t/h

- Some Capital expenditure will be required to achieve 85t/h.
- Key focus areas should be: feed only low silica concentrate (<2.5%, preferably lower) to the kiln; ensure adequate addition of sodium carbonate; and strictly control the temperature distribution in the hot zone of the kiln. It is recommended that additional instrumentation be installed on the kiln to measure and control kiln temperature profile.
- The kiln feed end should be reinstated or repaired: the paddle mixer should be eliminated and replaced with a screw feeder; the original kiln seals and double dump valves should be reinstated.

### Ramp up to 107.5t/h (400t/mth<sub>v</sub> in FeV) and higher throughputs

- It is likely that a more sophisticated burner will be required to achieve proper recovery when feeding the kiln at 107.5t/h dry feed basis.
- MVPL operating experience with the Salt Roast Kiln process appears to suggest that the dust generation rate in the kiln is much lower than Hatch's experience from other metallurgical applications. When a low dust generation rate is modeled, the results suggest that 117t/h (dry basis, fresh feed) becomes possible. It then becomes the case that the capacity for the kiln to operate at high throughputs is contingent on the salt roast kiln having a low dust generation rate, possibly attributed to sintering and stickiness of the feed as it flows through the reaction zone.

- Simulations of the kiln performance at still higher throughputs (127tph dry fresh feed equivalent to 473t/mthV) suggests limitations will be reached in terms of the off-gas handling equipment (kiln scrubber) as well as excessive freeboard gas velocities within the kiln.

**Ramp up to 135t/h (500t/mthV in FeV).**

- Our modeling indicates that the kiln is unlikely to achieve the design residence time and temperature at 135t/h (dry basis) feed at 6% moisture. The feasibility of pre-drying the concentrate with a flash or fluid bed dryer could be investigated along with the replacement of the main kiln drive motors.

Bottlenecks in other plant areas are more difficult to analyse until the plant ramps up to its design capacity. There appears to be capacity in the leaching, purification and precipitation areas of the flowsheet. It appears that the filters had not been fully commissioned during the ramp-up. It is recommended that funds be reserved to remobilize the filter vendor or filter expert to recommission and optimize these units.

Hatch can assist with future studies, optimization programs, onsite commissioning and vendor interactions.

## 1. History

Midwest Vanadium purchased Windimurra Vanadium in 2010. This plant was been operated intermittently and ramped up in 2013. Ramp up was commenced in 2013 and was completed around November 2013. On the 4<sup>th</sup> of February 2014 a fire in the ore beneficiation plant stopped operations. During the months of September to November, the refinery experienced an overall vanadium recovery of around 50% to AMV<sup>1</sup>, which is lower than the designed 82.95%<sup>2</sup>. If other data is used, the recovery can be calculated to range from 54% to 65%<sup>3</sup>. The average plant area recoveries are shown in the Table 1-1 below.

**Table 1-1: Average Vanadium Recoveries**

Plant Area	Average Vanadium Recovery (%)
Kiln Conversion	60.30
Leaching	91.42
Desilication	94.3
AMV Precipitation	89.3

The V<sub>2</sub>O<sub>3</sub> content in leach residue was calculated to be between 0.34 to 0.44%<sup>3</sup> and measured to be between 0.35 to 0.29%<sup>3</sup>. This equated to a vanadium recovery of 66% to 73%<sup>3</sup>.

The lowered recoveries through the circuit were found to be related to a vanadium loss to the tailings. The vanadium loss to the tailings was predominantly related to insoluble losses, but there are also losses of soluble vanadium. This loss needs to be decreased to improve the plants overall performance and economics.

The lowered recoveries have been attributed to the following causes:

- A high Vanadium content in the desilication solids;
- A low Vanadium recovery in AMV precipitation;
- A low Vanadium conversion in the kiln.

By all accounts, the plant was still in ramp-up and operating quite far from the design, particularly with respect to silica content of the magnetic concentrate and kiln temperature profile, and recovery would be expected to improve if these deviations were nullified.

<sup>1</sup> Windimurra Vanadium Monthly Reports

<sup>2</sup> Windimurra Vanadium Project, Process Design Criteria, 6033-G-00-F-001

<sup>3</sup> Refinery Monthly KPIs v3

## 2. Throughput Basis

Plant through-put is a product of:

- Nominal plant feed rate;
- Vanadium head grade;
- Vanadium recovery; and
- Plant Operating Factor, which itself is a product of:
  - ♦ On-line Time and
  - ♦ Capacity Utilisation.

These terms are defined and quantified for Windimurra below.

### 2.1 Nominal Plant Feed

A straightforward definition of “nominal” is that these are the numbers reported on the control system when the plant is operating as designed. These are typically the numbers which appear on the flow-sheet when the mass balance is closed. It is calculated by:

Nominal Kiln Feed (t/h) = Annual Vanadium sold (t/a) / (V head grade \* V Recovery \* 8760h/a \* Operating Factor)

### 2.2 Vanadium Head Grade

Vanadium head grade taken from the process design criteria is **0.71%V**. Silica grade is assumed to be on-spec, <2.5%.

### 2.3 Vanadium Recovery

For the purposes of the remainder of this report, the assumed vanadium recovery is (taken from the process design criteria):

- 82.95% of input vanadium recovered to AMV, of which;
- 96% is recovered to FeV, for a total recovery of;
- **79.6%** vanadium recovery from kiln feed to ferrovandium.

This is much higher than the plant has achieved so far, however in their defence, the plant was still ramping up to the design (in particular the plant was not treating on-spec material nor was the roasting temperature consistent).

In light of the discrepancy between design and actual vanadium recovery, additional site test-work was provided to Hatch to determine these sensitivities. Reference 13.06 describes tests on 4.5% silica ore (cf. on-spec ore from the PDC is listed as 2.5%). Recoveries were limited to 70% at 6% soda ash addition. Tests on 2.7% silica ore and 4.2% soda ash were 80%. Figure 4 from this report best shows the interactions, where both plant and lab test data support better recoveries (up to 85%) at 2.7% silica and 4-6% soda ash addition. It is known that vanadium recovery is a function of silica level in the concentrate, the dosing (and type) of sodium compounds and the reaction time at temperature in the kiln.

Reference 13.07, meanwhile, describes the impact of temperature on vanadium recovery. The results line up with the literature and industrial practice in South Africa, concluding that roast temperatures should be  $> 1100^{\circ}\text{C}$ .

These additional tests do not account for vanadium losses to calcine and desilication, however this is partly mitigated by recovering waste water from the tailings ponds back to the process.

Previous test-work that Hatch has performed on similar deposits indicates that:

- i. If the beneficiation plant can consistently deliver low silica concentrate (below 2.5% silica, preferably lower) to the kiln; and
- ii. if concentrate is mixed with sufficient sodium carbonate to solubilise the vanadium, after accounting for parasitic consumption from silica (expect between 4 and 6%); and
- iii. if the ore is roasted at above  $1150^{\circ}\text{C}$  (preferably higher) for a minimum of 1 hour (preferably longer);

then the vanadium recovery to AMV documented in the Process Design Criteria of 82.95% should be achievable. However any departure from these parameters risks lower recoveries.

It is recommended that additional muffle furnace tests be performed to generate more confidence in vanadium recovery, and its variation with respect to time, temperature and sodium addition. The site has muffle furnaces suitable for these tests, however CSIRO in Waterford have done similar tests in the past and have some sophisticated quantitative scanning microscopy which may assist determine the nature of the losses. Achieving the same objective using tests at full scale in the actual kiln will be extremely difficult given the lack of ability to control and measure the important process variables. Hatch can assist develop a test work plan, but in principal, we would recommend continuing the matrix of tests around:

- Producing low silica concentrate from the beneficiation plant;
- Muffle furnace tests to generate data of vanadium recovery as a function of:
  - ◆ Sodium carbonate addition (4-8%);
  - ◆ Roasting time (45 min to 2 hours);
  - ◆ Temperature (1050 to 1200);
  - ◆ Effect of other additives on silica solubilisation ( $\text{Al}_2\text{O}_3$ );
  - ◆ Effect of substituting other sodium salts (e.g. sodium sulphate). It is noted that this may require purchase of a high temperature muffle furnace, or outsourcing to CSIRO.

## 2.4 Operating Factor<sup>4</sup>

This factor is a representation of production achieved compared to design capacity. It is exactly the product of on-line time and capacity utilisation, defined below.

### 2.4.1 On-line Time

This factor represents the proportion of total time that a system, unit operation or piece of equipment is actually operating, at any feed rate greater than zero. This is defined in the PDC as 7900 hours per year, or 90.2% on-line time for the main unit operations.

### 2.4.2 Capacity Utilisation

This factor is a representation of production losses in a system due to not operating at design capacity e.g. during ramp-up, ramp-down, slowdowns or because design capacity for catch-up is greater than required for normal operation.

There is no allowance for capacity utilisation in the process design criteria (i.e. assumes 100%). This is equivalent to saying that when the kiln is running, it is running at, e.g. 107t/h, at the design temperature profile and generating on-spec product capable of 82.95% vanadium recovery.

Capacity utilisations of 95% are more typical in non-ferrous metallurgical extraction industries familiar to Hatch. Capacity utilisation has been assumed to be 100% for the purposes of this review, however there is a risk that this will underestimate the nominal throughput required to achieve annual production.

### 2.4.3 Calculated Operating Factor

The calculated operating factor in the process design criteria is 90.2%, however with a 95% capacity utilisation; the operating factor becomes 85%. For comparison, examples of other plants with 85% operating factors include nickel laterite smelting (kiln, electric furnace, refinery), copper flash smelting plants. Plants that ignore capacity utilisation in their calculations are surprised when they don't achieve annual name-plate.

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<sup>4</sup> There is unfortunately no consistent terminology or approach to converting annual production to an hourly rate, and definitions are highly client specific. The following definitions are those Hatch use in pyrometallurgical industries with kilns.

## 2.5 Nominal Kiln Feed Required for 400t/mth and 500t/mth (Contained V) in FeV

Table 2.5-1 summarises the roasting Kiln feed-rate for this debottlenecking study.

**Table 2.5-1 – Roasting Kiln Throughputs**

Parameter	Unit	Throughput A	Throughput B	Throughput C
Fresh Kiln Feed	t/h (dry basis)	107.5	134	131
% V	w/w	0.71%	0.71%	0.71
% SiO <sub>2</sub>	w/w	2.50%	2.50%	2.5%
% Recovery to AMV	% w/w	82.95%	82.95%	72%
% Recovery to FeV	% w/w	96%	96%	96%
Availability	%	90.2%	90.2%	90.2%
Capacity Utilisation	%	100.0%	100.0%	95%
Annual V in FeV	t/a	4800	6000	4825
Monthly	t/month	400	500	~402

Throughput C has been added to demonstrate what would be required if the vanadium recovery to AMV were only 72% instead of 83% - the kiln feed rate would need to be close to Throughput B while achieving only the revenue from Throughput A.

Table 2.5-2 summarises the Reduction Kilns feed-rate for this debottlenecking study.

**Table 2.5-2 – Reduction Kilns Throughputs**

Parameter	Unit	Throughput A	Throughput B
Fresh Kiln Feed	t/h (dry basis)	1.474	1.842
% V	w/w	42.94%	42.94%
Kiln Discharge	t/h (dry basis)	0.984	1.23
% V	w/w	64.31%	64.31%
% Recovery to FeV	% w/w	96%	96%
Availability	%	90.20%	90.20%
Capacity Utilisation	%	100%	100%
Annual V in FeV	t/a	4800	6000
Monthly	t/month	<b>400</b>	<b>500</b>

### 3. Capacity Review

#### 3.1 Methodology

Equipment capacities are reviewed in this section at plant throughputs of 400 and 500t/mth contained vanadium in FeV. Note: no process modelling of the overall flow-sheet has been performed; rather internal flow-rates are linearly extrapolated from the Proteus heat/mass balances. The extrapolated flow rates are summarised in Appendix A.

#### 3.2 Kiln

The salt roast kiln was the subject of a more detailed analysis because it presented the greatest single potential bottleneck in the entire process. Hatch proprietary models of the bed motion within the kiln and the thermo-chemical and combustion processes have been used to identify the key aspects in the successful operation of the kiln and to highlight those that present challenges to achieving the higher target throughput rates. Details can be found in Section 4 of this document and in the Appendix B.

##### 3.2.1 Past Studies

The MVPL salt roast kiln has been the subject of considerable past study including thermochemical modelling by Promet (2006), benchmarking and speculation by vanadium Consultant Les Ford (2008), CFD and thermochemical modelling by Drytech (2009), plant operational analysis by MVPL personnel (2013), and most recently a kiln audit by the equipment vendor, FLS (2013). Generally, the focus of these works has been on determining the ultimate capacity of the salt roast kiln and/or on the mechanical and process aspects of the operation that should be modified to increase throughput. The consensus is generally that in its current state the kiln should be able to process 100t/h of feed (nominal) and it has been suggested that 135t/h would be a feasible target. None of the aforementioned works have provided a sound technical basis nor can point to a meaningful precedent for the 135t/h feed rate. On the basis of the analysis here this target is unrealistic for the existing kiln and ancillaries without significant modification such as a pre-dryer.

##### 3.2.2 Operation at 400t/mth<sub>v</sub>

With some modification the existing kiln should have adequate capacity to support operation producing 400t/mth<sub>v</sub>. The analysis (section 6) suggests that good stable combustion, adequate secondary air supply, limited primary air addition, reduction of tramp air, and a long and stable combustion plume will allow the existing kiln to achieve the process objectives (1150°C for 60 minutes) at the required throughput. Plant experience will be essential to find the best operating point, but we expect that the primary air quantity should be restricted to less than 10% of the stoichiometric air supply and the combustion plume should extend nearly 20 metres in length. Tramp air (leakage at seals and double dump valves) should be kept to an absolute minimum and it is expected that with the installation of grease seals at the kiln feed end less than 100Nm<sup>3</sup>/hr should be possible. It will be challenging to ramp the kiln up to this production rate and careful attention to process conditions in order to find the operational "sweet-spot" will be essential. The key ingredient in successful ramp-up will be to optimize the combustion/ thermal scenario in the kiln hot-zone and most of the suggested capital spending (detailed in Sections 6 and 7) is supporting this objective.

### 3.2.3 Operation at rates in excess of 400t/mth<sub>v</sub>

At the request of MVPL Hatch have investigated the kiln process at rates in excess of 400t/mth<sub>v</sub>. To facilitate this investigation the bed residence time and fill behaviour has been modelled in detail at increasing total feed rates of 125, 135, 147, and 158, tonnes per hour.

### 3.2.4 Operation at 500 t/mth<sub>v</sub>

Our analysis suggests that achieving process objectives at production rates commensurate with a plant output of 500t/mth<sub>v</sub> is not possible with the existing kiln. The critical constraint suggested by the detailed modelling is attaining the specified time-at-temperature (60 minutes at 1150°C) which is compromised by the more rapid transit of solid material through the kiln. Simply slowing the kiln speed is not an option as it would lead to overfilling of the bed and back-spillage at the feed end. Higher heating rates are also not an option due to local overheating of the bed and excessive dust generation. The kiln main drive motors would need to be replaced due to both not being able to sustain an increase in kiln speed (the 6-pole motors are already at 164% speed, refer Section 4.2) and not having sufficient power available. The main reducers do appear to have some spare capacity available.

There may be several mitigations to allow the kilns to achieve 500t/mth contained vanadium – pre-drying the kiln feed is the solution used in cement plants and would be worth discussing with FLS. This would significantly increase the kiln off-gas temperature, requiring substantial modifications to the kiln off-gas system as well.

## 3.3 Kiln Off-gas

The kiln off-gas is cleaned in a set of dust cyclones and scrubbers. These scrubbers remove particulate by circulating dilute leach liquor, which has the additional benefit of concentrating the liquor.

The off-gas system was designed to operate with significantly greater gas flows than that achievable in the plant, at both the lower and higher throughput targets. This is shown in Table 3.3-1 below. As a result, this area of the plant is not likely to be a bottleneck.

**Table 3.3-1: Volumetric Flow rate**

Unit Operation	Volumetric flow-rate (m <sup>3</sup> /s)			
	Designed Capacity	Capacity Limit	Value at 400 t/mth <sub>v</sub>	Value at 500 t/mth <sub>v</sub>
Venturi Scrubber	33.9 <sup>5</sup>	37.8	24.62	30.78
Scrubber Tower	52.5 <sup>2</sup>	61.2	31.42	39.27
ID Fan flow-rate	52.5 <sup>2</sup>	61.2	37.24	46.55

Note: the values at the different throughputs have been scaled from the process design criteria.

<sup>5</sup> Kiln Off Gas Wet Scrubber Document No. 6033-W-19-M-023, Tender No. Q203/07, Section 1, Technical Proposal

### 3.4 Leach

In the leaching area, the sodium compounds are dissolved in vats into an aqueous solution. Based on the process design criteria, the leaching area equipment appears to be of sufficient capacity to handle an increase in calcine rate. The major limitations will be the cycle times required for the leaching process.

In the leaching process, the cooled kiln product is quenched and transferred, via a launder, into one of three leaching vats. Once that vat is filled, the quenched slurry from the launder is transferred to the next vat, and the full vat is emptied. The time required to empty the full vat will dictate the throughput of the leaching area. The bulk of the vanadium is formed as a sodium vanadates, which is readily soluble and it is likely that leaching kinetics is not rate determining. More refractory forms of vanadium are insoluble, and it is likely that no amount of time would dissolve them. Vanadium recovery could be improved by better washing of the calcine solids, and in other industries CCD thickeners are used to ensure that soluble losses are minimised. At Windimurra soluble losses are mitigated by recovering run-off water from the kiln tailings.

The leaching cycle time of 52.9 hrs per vat<sup>6</sup>, as shown in the PDC, was determined from the vat capacity and solids flow-rate (i.e. cycle time = total vat capacity / calcine feed rate). This cycle time is yet to be measured in the operating plant. In one cycle, there are three steps, filling the vat, solution draining and vat emptying. The fill time presented in the Process Design Criteria was determined from vat capacity, solids flow rate and the number of vats available (i.e. fill time = vat capacity / (calcine feed rate x number of vats). The remaining time was allocated to draining the vat and washing the calcine. As a result, it is seen that the minimum cycle time is not known. It can be assumed that the times allocated to the solution draining and calcine washing can be decreased, as well as the emptying of the calcine. This needs to be further verified with discrete event simulation. This is performed by surveying the times used in the leaching process and using this information to create the computer model. If this is possible, the throughput of the leaching process can be increased beyond the designed 131t/h<sub>ore</sub>.

If the plant throughput is increased beyond 131t/h<sub>ore</sub>, it is possible that a number of pumps in the leaching area may need upgrading. Specifically, the leach recycle pumps will need to be upgraded with a higher solids feed rate. Other pumps may also require upgrading. This will allow the flow-rates in the system to be increased, in line with the increased quenched slurry flow-rates.

Vanadium losses have been seen in the leaching area due to solution flow issues. This is directly caused by the lack of agitation present in the leaching vats and inadequate washing of the calcine. This may lead to lower vanadium recoveries than that seen in lab test-work. This issue can be remedied by using agitated tanks to perform the leaching, rather than the leaching vats.

<sup>6</sup> Windimurra Vanadium Project, Process Design Criteria, 6033-G-00-F-001

To improve the leaching area, a number of potential upgrades are possible. These include;

- Changing the leaching and dewatering of the tailings to a continuous process. This could include stirred tanks and/or CCD circuits.
- The use of clamshell reclaiming and conveyors to recover the tailings from the leaching vats.

To determine the limitations of the equipment of the leaching area it is recommended that leaching kinetics tests are performed on the calcine. It is also recommended that the cycle times for leaching are studied, with a discrete event simulation, to determine the maximum throughput of the vat.

### 3.5 Desilication

In the desilication area, the dirty pregnant liquor from leaching is purified to remove the solubilised silicates in the solution by dosing aluminium sulphate. It appears that the desilication area may be able to handle an increased ore throughput.

In the desilication process, the solution is held in stirred tanks for a number of hours. Test-work has shown that the desilication reactions are fast and that the tanks are able to handle significantly greater solution flow-rates. The reactions were shown to be close to completion in 30 minutes<sup>7</sup>, in batch reactors, while the plant is designed to operate with a tank residence time of 18 hours<sup>6</sup>.

A heat exchanger is part of the design (Desilication Feed Cooler – 30HEX501) to cool the dirty solution before Desilication. This cooler should decrease the solubility of silica in the clear pregnant solution. As a result, it is recommended that this cooler is reinstated, and possibly augmented with a stand-by unit, with further studies performed on the optimal operating temperature.

The desilication tanks operate with overflows between the tanks. This is not an issue if the solids in the tanks are sufficiently agitated and there is no short circuiting in the tanks. If there is short circuiting, the tanks would require modification to ensure that there is sufficient residence time in the tanks. If the solids are not suspended (for instance if the grind size is greater than designed), the tanks will fill up with solids, requiring significant maintenance. If this is experienced, it is recommended that the hydraulics of the system be studied in greater detail, and perhaps the agitator and motor upgraded.

The Aluminium Sulphate pump is designed to operate at a flow-rate<sup>8</sup> of 1.2 to 1.7 m<sup>3</sup>/h. This is more than the mass balance<sup>9</sup> (0.31m<sup>3</sup>/h) and process design criteria<sup>10</sup>(0.49 t/h = 0.31m<sup>3</sup>/h). The actual flow-rate required for this pump will depend on the silica loading (ultimately a function of silica in the ore). As a result, it is recommended that trials and tests are performed to optimise the Aluminium Sulphate dosage and capital set aside to upgrade this pump. A similar situation will apply to the sulphur acid pump.

<sup>7</sup> REF.13.24 – Desilication Test Work Sept 2013 Memo

<sup>8</sup> Equipment Data Sheet, Windimurra Vanadium Project, Aluminium Sulphate Feed Pump, 6033-D-30-M-030

<sup>9</sup> Windimurra Vanadium Project, Process Mass Balance, 6033-G-00-F-004

<sup>10</sup> Windimurra Vanadium Project, Process Design Criteria, 6033-G-00-F-001

All of the equipment, except the desilication filter, should be able to process the required throughputs.

The desilication filter was designed with an assumed filtration rate which was not verified with test-work, which should be performed now. If the filtration rate is able to be increased, the filter fill time could be decreased.

If the filter is not able to handle a higher solids throughput, a number of industrial solutions are available to increase the filtration capacities of the system. Flocculation/Dewatering Agent addition could be incorporated into the filtration system and this may increase the filtration rate of the filter. Otherwise the filtration area can be increased, either by upgrading the filter or adding another filter.

It is recommended that filtration tests are performed to verify the capacity of the filters. This will enable their maximum capacity to be known. Information from site was that the filters were not properly commissioned; it is therefore recommended that the filter vendor or similar expert be mobilised when recommissioning the plant to complete commissioning of the filter and to rectify any deficiencies.

### 3.6 AMV Precipitation

To produce Ammonium Meta Vanadate (AMV) from the purified solution, ammonium sulphate is added to stirred tanks. This precipitates the AMV, which is separated from the solution with the use of a thickener and filter.

The precipitation tanks are, as shown by test-work, significantly larger than required. As a result, these tanks should be able to process higher throughputs. Here again, a heat exchanger (35HEX501) is part of the design, however the heat exchanger was not in service. As designed, this cooler is used to decrease the temperature at which precipitation occurs, and this improves vanadium recovery. This may also decrease the consumption of ammonium sulphate. It is recommended that the reasons for disabling this heat exchanger be determined, and if required the heat exchanger reinstated and a standby be installed. This cooler is of sufficient capacity to handle higher flow-rates, as it is designed to operate with a solution flow-rate of 23.28 t/h<sup>11</sup>. This is significantly larger than that at the 500t/mth<sub>v</sub> case of 15.86t/h. It is recommended that if this cooler is used in the process, further studies are performed on the Heat Exchanger to resolve scaling issues.

In the precipitation area, stirred tanks are used to precipitate the solids. The first two tanks are agitated with 7.5kW motors and the last with a 3kW motors. These agitators have been observed to be insufficient for the required duty. These agitators are installed to keep the solids to remain in suspension, not to improve reaction rates. Information from site is that the AMV is settling out inside these tanks, which indicates that the tanks are under-agitated, or the design is incorrect. A site modification has added air lifters to these tanks to aid in solids transfer; however it is recommended that a study be performed on the slurry characteristics in the tanks, and potentially a larger motor installed. This will reduce the potential issue of short circuiting in the tank, which is present when there is insufficient mixing in the tank.

<sup>11</sup> Windimurra Vanadium Project – 06033-P54, Quotation for Spiral Heat Exchanger, 13/02/2008, from: Paul Tuckwell to: Annette Fong.

The precipitated AMV is separated from the barren solution with the use of a thickener and a belt filter. This dewatering equipment was designed with minimal test-work. As a result, the maximum capacity of the equipment is not able to be determined. It appears that the thickener and filter have the capacity to handle a higher throughput if they have the dewatering characteristics outlined in the Process Design Basis.

If this equipment is able to handle a higher throughput, the pumps and other equipment that feed them may require upgrading to suit.

To increase the throughput of the AMV dewatering equipment, a number of upgrade options are industrially proven. Flocculant can be added to both the thickener and belt filter. This may increase the capacity of this process equipment. If this is not practical, the thickener and belt filter needs can be increased. This can be done with the addition of another high rate thickener and belt filter.

Recycling thickener underflow back to the first precipitation tank may provide seed crystals to the precipitation process, and will possibly increase the size of the AMV crystals. This will in turn improve the solids filtration and settling characteristics.

### 3.7 AMV Dryer

The Ammonium Meta Vanadate filter-cake produced in Area 35 contains approximately 20% moisture. This is removed by the flash dryer, before being calcined in the reduction kilns. The flash dryer works by heating the AMV precipitate with hot air. The dryer was supplied by Drytech, and was designed to dry 2t/h(dry basis) AMV<sup>12</sup>. This is higher than the 1.91t/h (dry) solids that are to be processed in the 500t/mth<sub>v</sub> case. As a result, this system is able to process at the higher throughput, but will lose its catch-up capacity. If this is deemed to be vital, the dryer's capacity can be expanded with a new burner.

Any constraints on the dryer are more likely to be in the solids handling equipment. If the solids rates are different from those the equipment is designed for, further studies may recommend some equipment replacement or modification. The capacity of the dryer is also impacted by the moisture content of the feed, with a higher moisture content of the feed decreasing the dryer capacity.

If the thermal capacity of the dryer is to be increased in the future, a number of options are available. These include:

- Increase the dewatering occurring in the AMV filter (mechanical dewatering is “always” cheaper than thermal drying);
- Installing a larger dryer burner; and
- Install a new dryer.

### 3.8 Reduction Kilns

The Ammonium Meta-Vanadate (AMV) from the dryer is fed into a buffer silo before then being fed into two, indirectly fired reduction kilns. The reduction kilns are heated by 6 gas burners. The ammonia liberates and cracks in-situ to provide a reducing environment.

<sup>12</sup> Windimurra Vanadium, Flash Dryer Operating & Maintenance Manual

Drytech<sup>13</sup> specify kilns capable of processing 1.515t/h of AMV (listed as 1.166t/h of V<sub>2</sub>O<sub>5</sub> equivalent). The 400t/mth<sub>V</sub> and 500t/mth<sub>V</sub> cases would require the kilns to process 1.474t/h and 1.842t/h of AMV respectively. The 400t/mth<sub>V</sub> and 500t/mth<sub>V</sub> cases are therefore 97% and 122% of the equipment design. Drytech have stated in email correspondence that the kilns can handle 120% of their current design. Refer to Section 5 for a more detailed kiln analysis of the reduction kilns.

### 3.9 Smelter Reagent Handling

One of the operational bottlenecks described on site was materials handling of furnace charge, i.e. V<sub>2</sub>O<sub>3</sub>, lime, aluminium and scrap iron.

The lime and aluminium conveying systems were reportedly acceptable. The V<sub>2</sub>O<sub>3</sub> and scrap iron addition systems were criticised.

As the plant was not operating when we visited it is difficult to see what issues were caused by pneumatic conveying V<sub>2</sub>O<sub>3</sub> to the top of the building. V<sub>2</sub>O<sub>3</sub> from the reduction kilns should flow and convey easily because it is warm and dry. If there is difficulty discharging from the bin, consideration should be given to having the material tested for flow properties to check that the hopper geometry is optimal. Hatch has a strong history in optimising bin geometry and installing wear plates to bins that discharge poorly.

The scrap iron addition system does not look suited to the form of scrap iron received; it appeared to be more suited to conveying steel punchings. The scrap iron that we saw appeared of dubious metallurgical quality, and not the type of material that one would expect would be added to a final product. Consideration should be made to purchasing better quality scrap (e.g. steel punching) at least until the plant ramps up. Once ramped up, the business case to replace the existing handling system to handle lower quality steel can be assessed.

### 3.10 FeV Smelting

In the FerroVanadium Smelting area, the product from the reduction kilns is mixed with aluminum, iron and lime. The aluminium reduces the vanadium oxides and the DC furnace melts the different components to form a slag and ferrovanadium.

The capacity of the FeV Smelting area is determined by the various steps in the smelting cycle. The furnace is designed to smelt 637t/mth<sub>V</sub> at its maximum capacity<sup>14</sup>. If the equipment is able to achieve its designed capacity, the 400t/mth<sub>V</sub> and 500t/mth<sub>V</sub> targets will be able to be met. If it is not able to be met, it is recommended a discrete event simulation is performed. This will allow the cycle time to be analysed and methods to decrease the cycle times can be determined.

<sup>13</sup> Windimurra Vanadium, Reduction Kilns Operating & Maintenance Manual

<sup>14</sup> Calculated from data in; Windimurra Vanadium Project, Process Design Criteria, 6033-G-00-F-001

### 3.11 FeV Crushing and Packaging

The FeV crushing plant reportedly had low availability due to the hardness of FeV in the jaw crushers. Again, without viewing the plant operating we cannot comment on this other than to say that anecdotally, within Hatch there is experience in high carbon ferrochrome, which is similarly difficult to crush. In one plant the practice was to drop a wrecking ball from height to perform the initial break, and then use a (very) oversized jaw crusher for sizing. There was also the possibility to decant off the slag, and cast the button of metal into sand moulds on the ground. Moulded metal is easier to break up and crush.

It is recommended that Windimurra operations visit as many FeV plants as possible to determine their experience, both operational and equipment based. In this type of situation, a successful plant should be found and copied as faithfully as possible, even sole-sourcing equipment from the same supplier. Hatch have design offices in South Africa, and our engineers could accompany Windimurra's teams. This would permit a South African design and procurement team to install a proven system.

For now, this report includes in the capital estimate Hatch a factored feasibility estimate from the Primero group.

## 4. Kiln Modelling

### 4.1 Kiln

Hatch proprietary models of the bed motion within the kiln and the thermo-chemical and combustion processes have been used to investigate the salt roast kiln in greater detail. The approach has been to benchmark the model initially against plant data during a period of relatively stable kiln operation (Dec 15, 2013). During this time the kiln feed rate was nominally 85t/hr and the kiln recoveries were reported to be 85%. The second step was to use the model to predict the kiln operating behaviour at the 400 and 500t/mth<sub>V</sub> production rates.

Of course, when models are used to predict real plant behaviour, questions are invariably asked regarding reliability and accuracy. Unfortunately, there are no easy answers. While the Hatch model has been used successfully in the past for similar processes (laterite calcination, spodumene roasting, nickel carbonate drying and reduction), and has reasonably well predicted key operating parameters (gas temperature and refractory wall temperatures within +/-100 C, off-gas temperature within +/-50C, product temperature within +/- 50C) there has been limited time or available data to fully benchmark/calibrate or validate the existing model for the salt-roast process across a range of operating conditions. Nevertheless, because the salt-roast process itself is relatively benign (i.e. consists mostly of moderate drying and sensible heating) and because these are process features of kilns that we have modelled successfully in the past, we have reasonable confidence in the model results. It should be stressed however, the lack of rigorous benchmarking and calibration creates greater uncertainty in the ability to make absolute predictions (i.e. the exact value of the bed peak temperature, or the exact point along the kiln axis when the target temperature is reached) and makes the model most useful in a "delta analysis" for comparing between two operational modes, for example

#### 4.1.1 General Process Considerations

The significant process steps occurring within the kiln are best illustrated by considering typical gas and bed axial temperature distributions along the length of the kiln. Shown in Figure 4.1.1-1 are plots for operation at nominally 85t/h (wb) with 6% feed moisture. The axial temperature distributions of bed, gas and shell are shown in the figure as well as the site measurements of the kiln shell temperature measured during May 2013. The prevailing process conditions at that time may not be exactly the same as used in the simulations so the data are shown for reference only. The reaction progress of the reactions liberating free moisture and consuming V<sub>2</sub>O<sub>5</sub> (i.e. producing sodium metavanadate) are displayed in terms of the percent of the parent material. The significant behaviours include:

1. A slow period of preheating and drying associated with the reduced heat transfer rates that are available at the feed end of the kiln. It can be seen that the concentrate is predicted to heat slowly to near the boiling temperature of water (~100°C) and, due to the limitations on the rate at which energy can be transferred to the bed, maintains a near constant temperature until the bed is free of moisture (~40 metres from the feed point).
2. A more rapid period of sensible heating to a point 65 metres from the feed inlet.

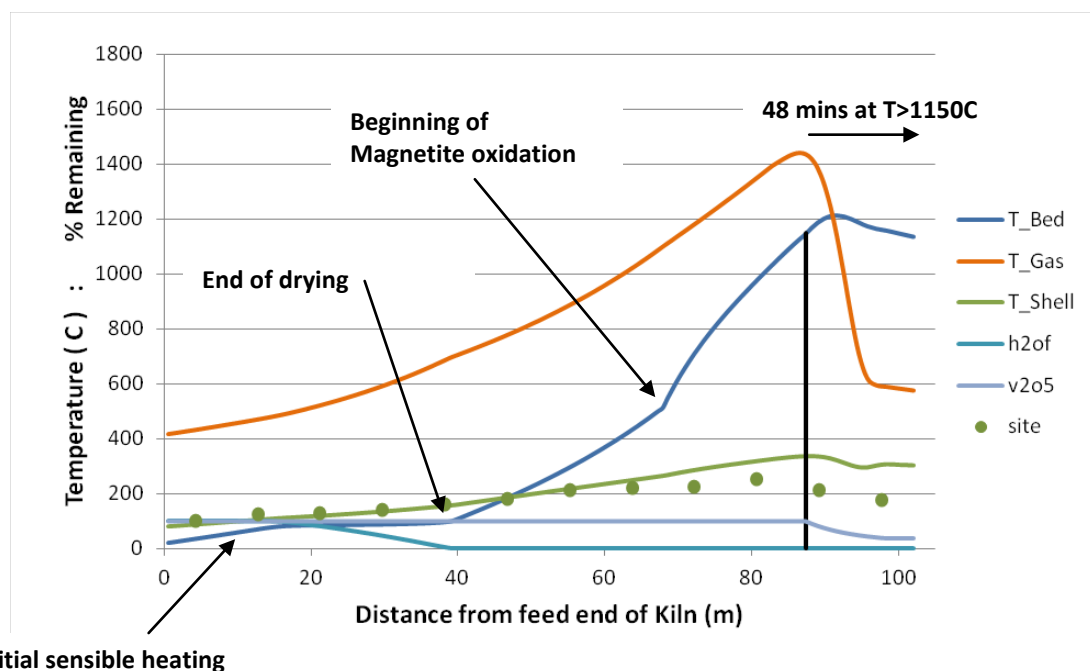


Figure 4.1.1-1 – Typical Roasting Kiln temperature and reaction profiles

1. A rapid increase in the bed temperature once it rises above 550°C, the temperature at which the magnetite oxidation reaction is assumed to commence.
2. Attainment of the key process temperature (1150°C) at approximately 88 metres from the feed inlet. At this temperature the production of sodium metavanadate is assumed to occur. Under these conditions, and for the predicted feed rate and kiln rotational speed, the predicted residence time at temperature is approximately 48 minutes.

3. For these process conditions it is noted that the peak gas temperature is predicted to be slightly in excess of 1400°C. The peak gas temperature is important because higher peak gas temperatures can be expected to promote the formation of ring accretions in the hot zone of the kiln.
4. For these process conditions the model predicts an off-gas temperature of 414°C, dust losses of 3.2% of the kiln feed, and heat losses from the kiln of nearly 5MW. The heat loss represents 25% of the total thermal energy transferred to the bed and approximately 16% of the total energy supplied by the kiln main burner and preheated secondary air.

#### 4.1.2 **Scope of model investigation**

The kiln model investigation included analysis of the effect of primary and secondary air quantity, flame length, and feed moisture on the kiln operation at nominal throughput (85 tonnes/hr), and throughputs corresponding to the 400t/mth<sub>v</sub> and 500t/mth<sub>v</sub> targets. A list of the model inputs has been provided in the Model Design Basis document presented in Appendix A.

A summary of the cases investigated is presented in Table 4.1.2-1:

**Table 4.1.2-1 – List of cases examined using the Kiln model**

Case	Throughput	Secondary Air Flow-rate*	Primary Air Flow-rate**	Flame length
1	Basecase (85 t/h feed)	normal	normal	Normal (15 m)
2	Basecase (85 t/h feed)	75% normal	normal	Normal (15 m)
3	Basecase (85 t/h feed)	50% normal	normal	Normal (15 m)
4	400 t/mth <sub>v</sub>	normal	normal	Normal (15 m)
5	400 t/mth <sub>v</sub>	normal	Low	Normal (15 m)
6	400 t/mth <sub>v</sub>	normal	Low	Long (20 m)
7	500 t/mth <sub>v</sub>	normal	Low	Long (20 m)
8	500 t/mth <sub>v</sub> – dry feed	normal	Low	Long (20 m)
<ul style="list-style-type: none"> <li>• * Normal secondary air flow rate = 48,000 Nm<sup>3</sup>/hr @ 85 t/h feed</li> <li>• ** Normal primary air flow rate = 9,600 Nm<sup>3</sup>/hr @ 85 t/h feed</li> <li>• ** Low primary air flow rate = 10% stoichiometric air</li> </ul>				

#### 4.1.3 **Importance of Secondary Air**

To illustrate the importance of providing adequate secondary air Cases 2 and 3 repeated the basecase simulation with reduced quantities of secondary air. With decreasing quantities of secondary air the peak gas temperature increases and the residence time at temperature is reduced. As the secondary air is reduced to 75% (Figure 4.1.3-1) and then to 50% (Figure 4.1.3-2) of the base case value, the residence time at temperature decreases to 45 and then 37 minutes respectively. The corresponding peak gas temperatures are 1560 and 1690°C. While there is a degree of uncertainty with regards to the absolute value of these numbers, the trend is clear. Operation of the kiln with reduced volumes of secondary air will reduce the quality of the product and promote the formation of accretion rings in the hot zone of the kiln.

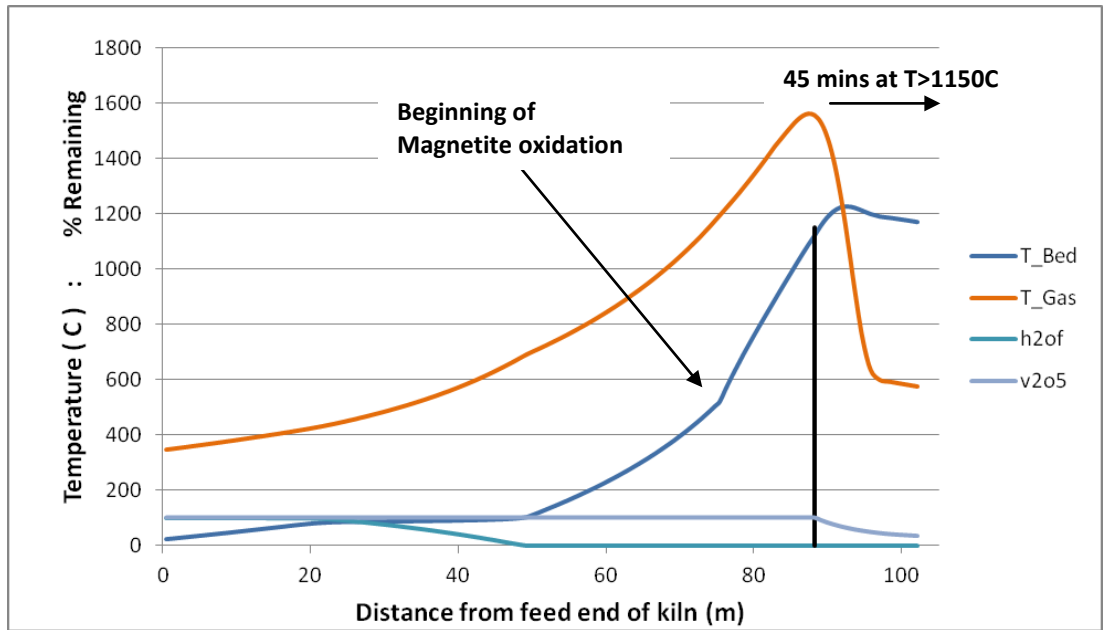


Figure 4.1.3-1 – Case 2: Nominal 85 t/h feed, 75% Secondary air

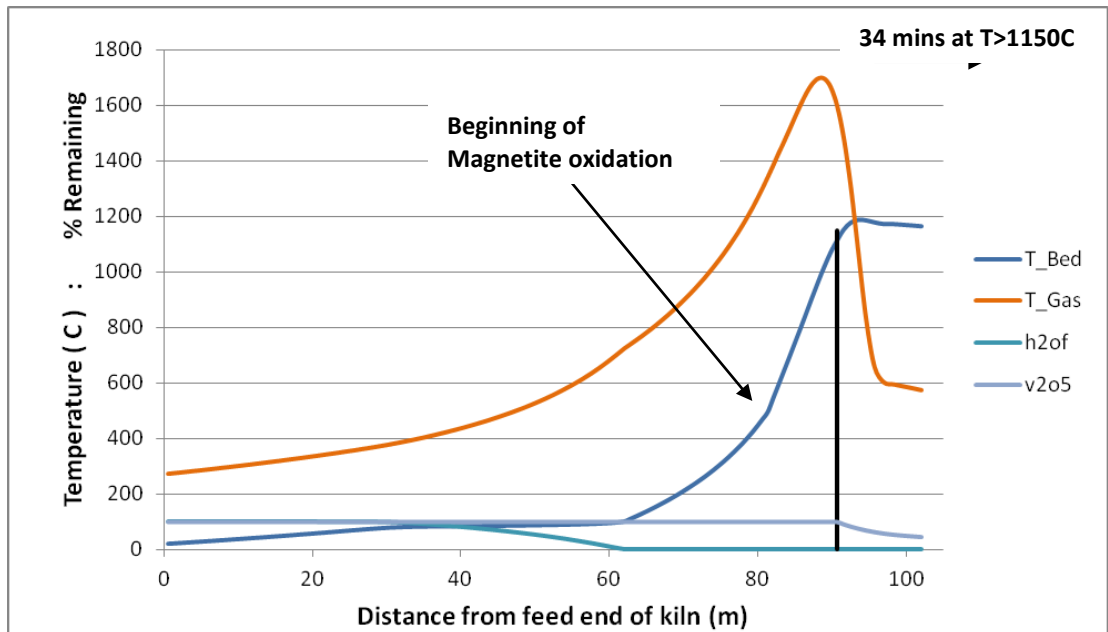


Figure 4.1.3-2 – Case 3: Nominal 85 t/h feed, 50% Secondary air

#### 4.1.4 Operation at higher throughputs (400t/mth<sub>v</sub>)

To achieve the 400t/mthV production target will require an optimized combustion system and careful operation. The formation of ring accretions in the hot zone of the kiln are expected to dictate the ultimate throughput. The model predicts that simply using the same operational parameters for the higher throughput (i.e. extrapolated from the Basecase) will only give 41 minutes time at temperature. This is largely a function of the more rapid motion of the bed through the kiln that is required at the higher throughput.

Generation of hotter off-gasses and elutriation of greater quantities of dust are also predicted, resulting from the need to fire the burner harder to meet the increased thermal input requirements. Results are shown in Figure 4.1.4-1

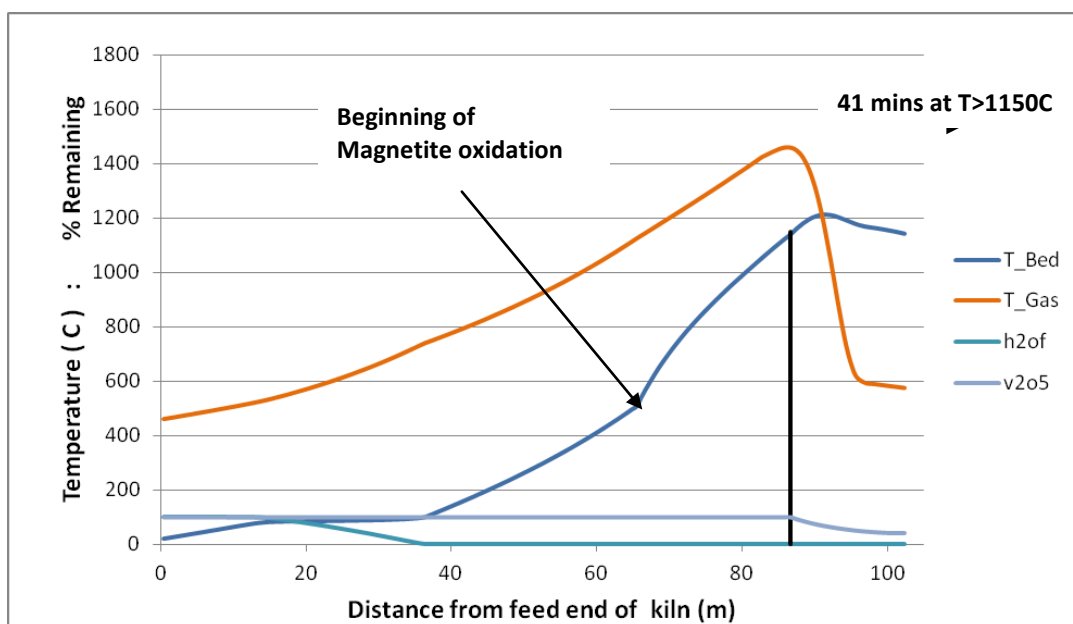


Figure 4.1.4-1 – Case 4: 400t/mthV production

By optimizing the burner performance through reduced primary air and a longer flame length it is possible to achieve 54 minutes at the target temperature (Figure 4.1.4-2 ). To reiterate, successful operation will require improved equipment and a focus on the kiln combustion behaviour.

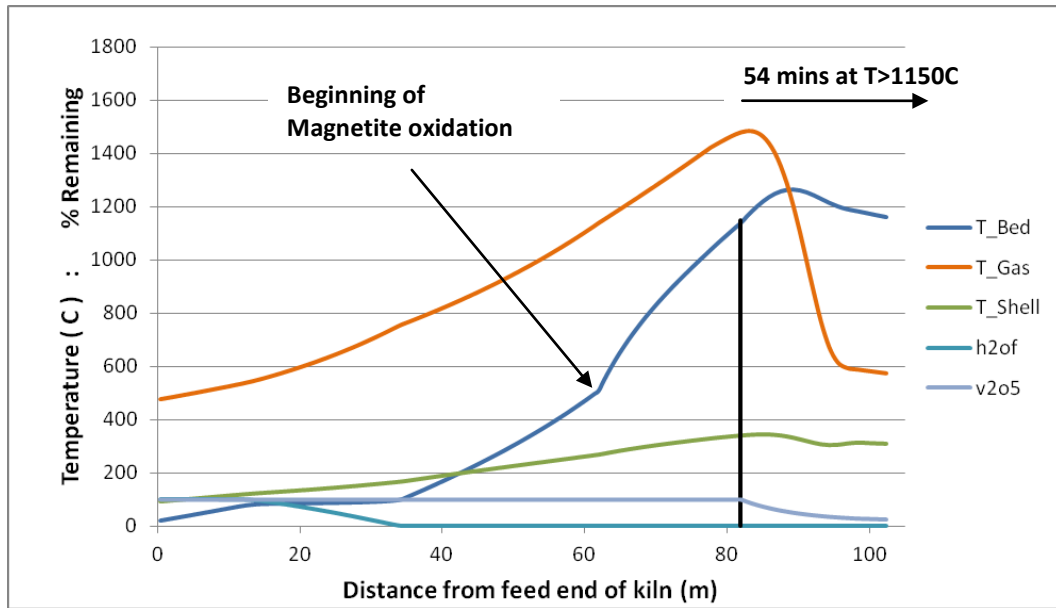


Figure 4.1.4-2 – Case 6: 400t/mthV production, reduced primary air, 20m flame length

**4.1.5 Operation at higher throughputs (500t/mth<sub>v</sub>)**

To achieve the 500t/mthV production target will require both an optimized combustion system and pre-drying of the feed. Predictions of the kiln performance under these conditions are shown in Figure 4.1.5-1.

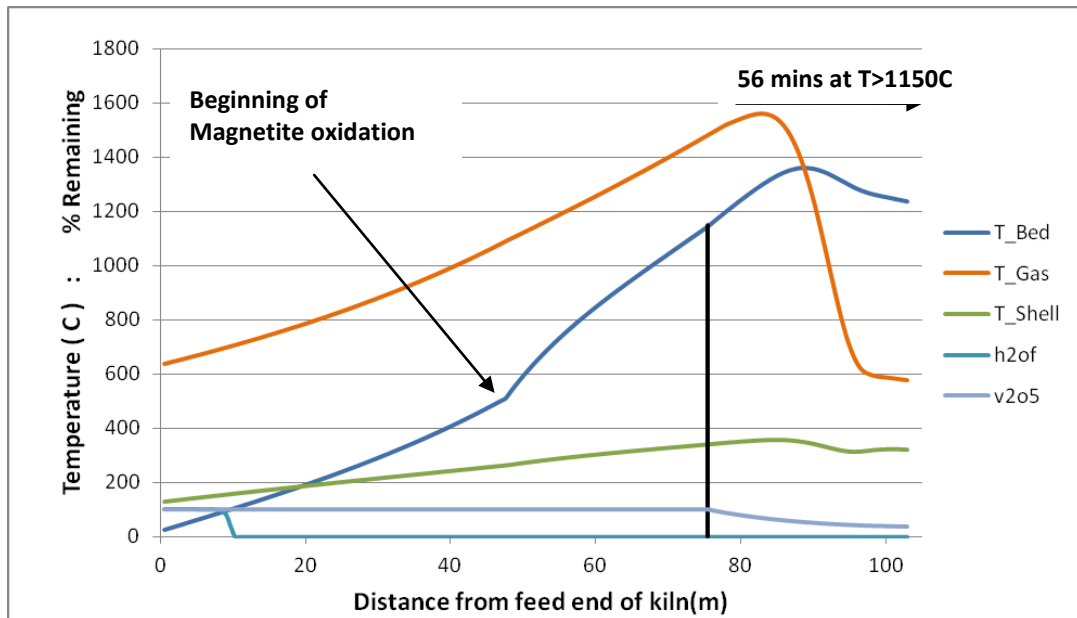
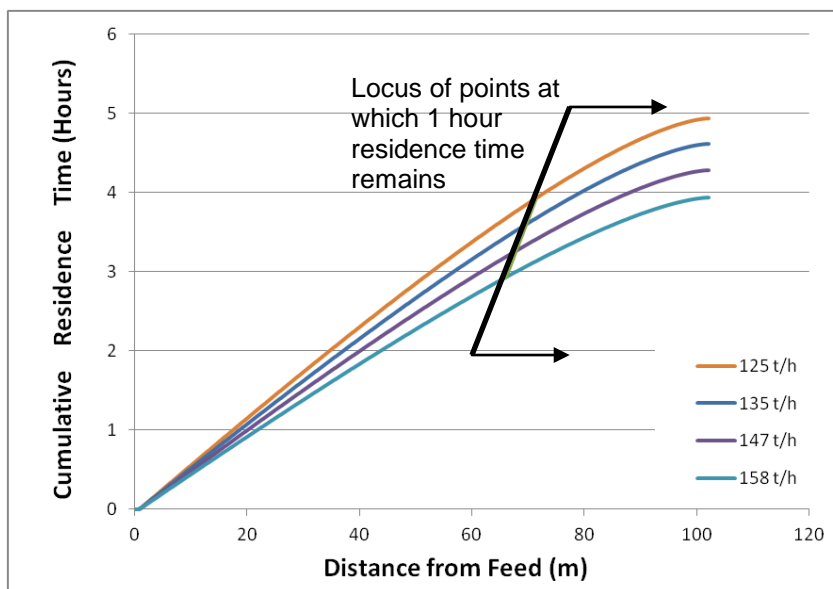


Figure 4.1.5-1 – Case 8: 500t/mth<sub>v</sub> production, reduced primary air, 20m flame length, pre-dried feed

**4.1.6 Operation at throughputs greater than 400t/mth<sub>v</sub>**

At the request of MVPL Hatch have investigated the kiln process at rates in excess of 400t/mth<sub>v</sub>. To facilitate this investigation the bed residence time and fill behaviour has been modelled in detail at increasing total feed rates of 125, 135, 147, and 158, tonnes per hour (corresponding to fresh dry feed rates of 107.5, 117, 127, and 135tph). A close examination of the bed behaviour demonstrates the difficulty of operating at increasing throughputs. As shown in Figure 4.1.6-1 with increasing kiln production rates the bed moves more quickly through the kiln. Hence, to achieve the process requirements of 1 hour residence time at temperature requires the bed to heat more rapidly, reaching the target temperature closer and closer to the feed end of the kiln. This in turn necessitates increased firing rates generating higher peak gas- and bed-temperatures, higher freeboard gas velocities and greater dust generation.



**Figure 4.1.6-1 – Residence time behaviour at higher feed rates**

The results of the complete simulations, key results presented in Table 4.1.6-1, suggest that operation at the 400 t/mth<sub>v</sub> production rate (this corresponds to dry solids feed rate of 107.5tph) is realistically the maximum production rate that can be achieved. Further increases lead to excess bed temperatures, or excess dust losses. Although the dust is assumed to be dry on leaving the kiln with the off-gasses this puts additional thermal load on the kiln as the dust leaves the kiln hot and is re-introduced cold. Shown in the table are the results for the 107.7, 117, and 127tph feed rates with flame lengths of 15 and 20 metres whilst limiting the bed maximum temperature to either 1400 or 1600°C. All cases have achieved the minimum required residence time at temperature (RT@T) of 1150 °C for 60 minutes.

**Table 4.1.6-1 Cases studied, throughput  $\geq$  400t/mth<sub>v</sub>. All achieved required RT@T**

Case	Solids Feed (tph dry)	Gas Feed kg/hr	Secondary Air (Nm <sup>3</sup> /hr)	Flame Length (m)	Dust (tph)
Peak Bed Temperature limited to 1600°C					
A1	107.7	3600	85100	15	8.3
A2	107.7	3420	81000	20	9.6
A3	117	4920	95255	20	22.4
A4	127	5900	114228	20	36.3
Peak Bed Temperature limited to 1400°C					
B1	107.7	4800	124364	15	40.82
B2	107.7	5000	129545	20	47.23

#### 4.1.7 Operation at throughputs greater than 400t/mth<sub>v</sub>. - Low dusting behaviour

Because of the limitations imposed by the predicted high dust generation MVPL requested that the previous simulations be re-run assuming dust is not elutriated from the bed under the following conditions:

- i. When the bed material contains free moisture;
- ii. When the bed material is at a mean temperature greater than 550°C (the point at which magnetite oxidation begins and the vanadium compounds become sticky).

As with the simulations presented in the previous section the kiln behaviour has been modelled in detail at increasing total feed rates of 125, 135, and 147, tonnes per hour. Results are given in the table below. In all cases the required RT@T was achieved whilst limiting the peak bed temperature to below 1500°C.

**Table 4.1.7-1 Cases studied, throughput  $\geq$  400t/mth<sub>v</sub>. Low dusting behaviour. All achieved required RT@T**

Case	Solids Feed (tph dry)	V production (t/mth <sub>v</sub> )	Gas Feed kg/hr	Secondary Air (Nm <sup>3</sup> /hr)	Flame Length (m)	Dust (tph)
Peak Bed Temperature limited to 1500°C						
C1	107.7	400	3420	81000	20	2.8
C2	117	435	3870	91500	20	4.8
C3	127	473	4500	106375	20	6.8

By suppressing the dust generation the kiln model predicts that the higher throughputs scenarios are (at least theoretically) feasible. Typical results for the C1, C2, and C3 cases (107 – 127 tph dry solids feed – equivalent to 400, 435, and 473 t/mth<sub>v</sub> production respectively) are given in Figure 4.1.7-1, Figure 4.1.7-2, and Figure 4.1.7-3

It is worth noting that while the low dusting rate scenarios that have been investigated here no longer suggest that the higher throughput rates will be impossible due to excessive elutriation or excessive bed temperatures, the model predicts kiln freeboard velocities that are higher than what would normally be expected in rotary kilns. (See Figure 4.1.7-4). The C3 case simulation shows a peak gas velocity of 18 m/s. It is difficult to imagine a total dusting rate of only 6.8 tonnes per hour (~ 5% of total feed) at such a high gas velocity.

Assuming the prescribed dust generation behaviour is realistic there are other limitations to the kiln throughput that may become critical at the higher throughput scenarios. In particular the off-gas quantity and temperature are very near to the specifications of the scrubber system in the C3 scenario.

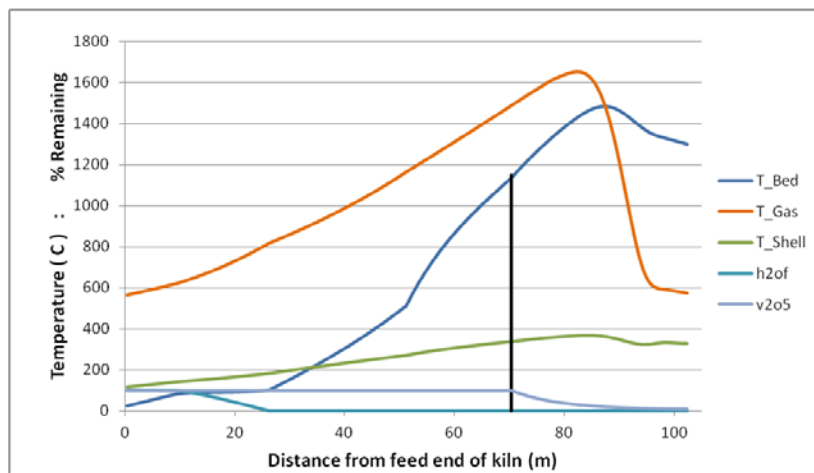


Figure 4.1.7-1 Predicted Kiln behaviour at 400 t/mth<sub>v</sub> production rate – low dusting behaviour

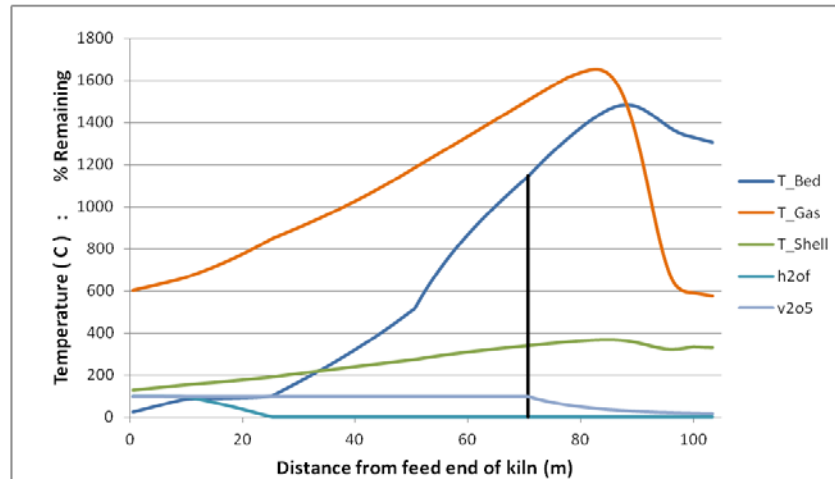


Figure 4.1.7-2 Predicted Kiln behaviour at 435 t/mth<sub>v</sub> production rate – low dusting behaviour

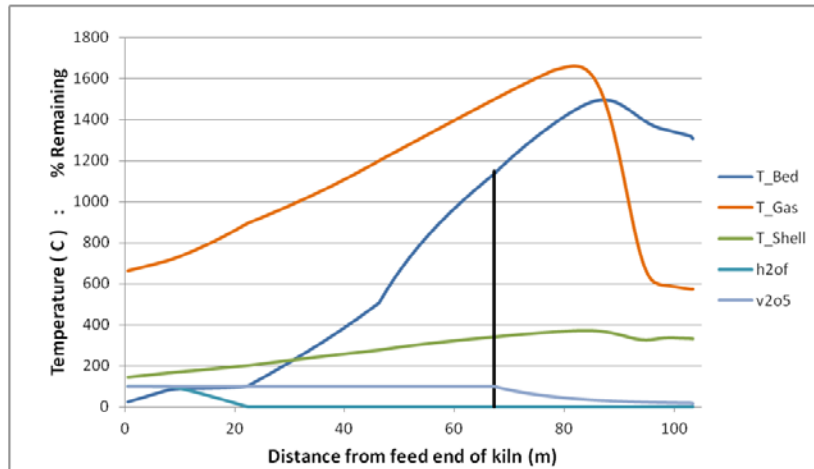


Figure 4.1.7-3 Predicted Kiln behaviour at 473 t/mth<sub>v</sub> production rate – low dusting behaviour

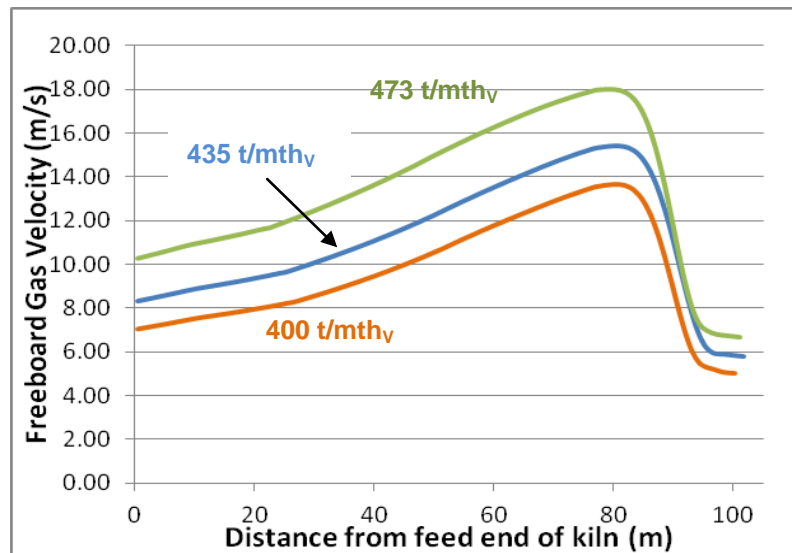


Figure 4.1.7-4 Freeboard Gas Velocities at Production Rates of 400, 435, and 473 t/mth<sub>v</sub>

#### 4.1.8 **Bed Profile Modelling**

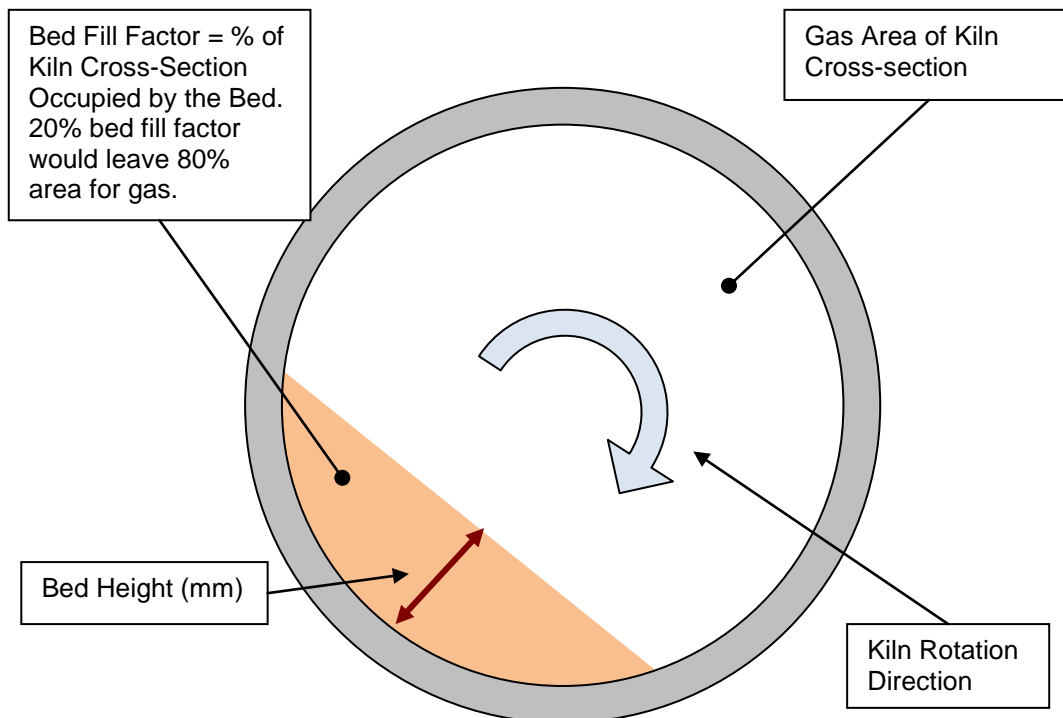
The following wet feed rates were used for the first pass of the kiln modelling:

- Current stable operation – 85wt/h;
- 400t/mthv – 135wt/h;
- 500t/mthv – 170wt/h.

During the course of the project the actual wet t/h values for the 400 and 500t/mthv cases has reduced slightly. The kiln modelling has not been recalculated to account for this at this stage. The current inputs were used for the models:

- Wet bulk density of the feed = 2600kg/m<sup>3</sup>;
- Dynamic angle of repose of the material = 37 deg;
- Kiln slope = 2%.

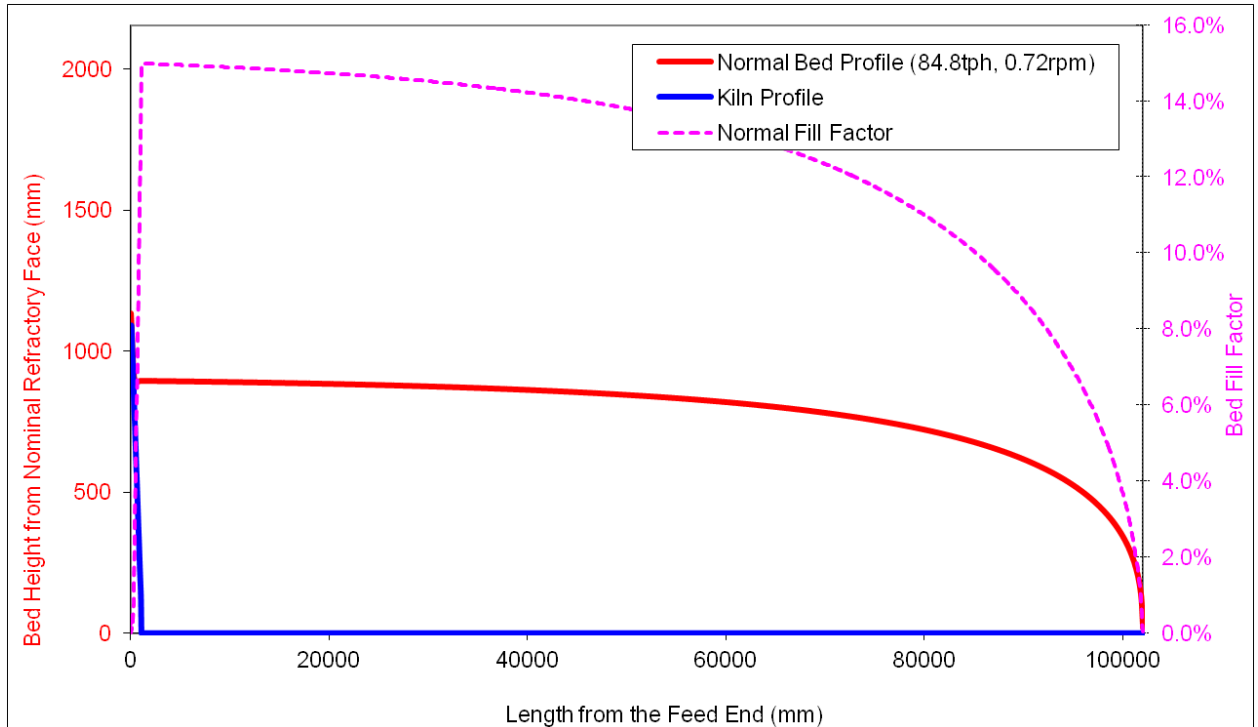
The Hatch bed fill model has been benchmarked against testwork, Discrete Element Modelling and operating commercial kilns. The model calculates the bed height, fill factor, residence time, bed mass, power required to rotate the bed and more. The bed height is the height of the bed, measured perpendicular to the inside of the shell / refractory, at the deepest part of the bed for any axial location in the kiln. The bed fill factor is the % of the kiln internal cross-sectional area occupied by the bed for any axial location. These two concepts are directly related and are shown in Figure 4.1.8-1.



**Figure 4.1.8-1 – Bed Height and Fill Factor Definition**

4.1.8.1 *Current Stable Operation*

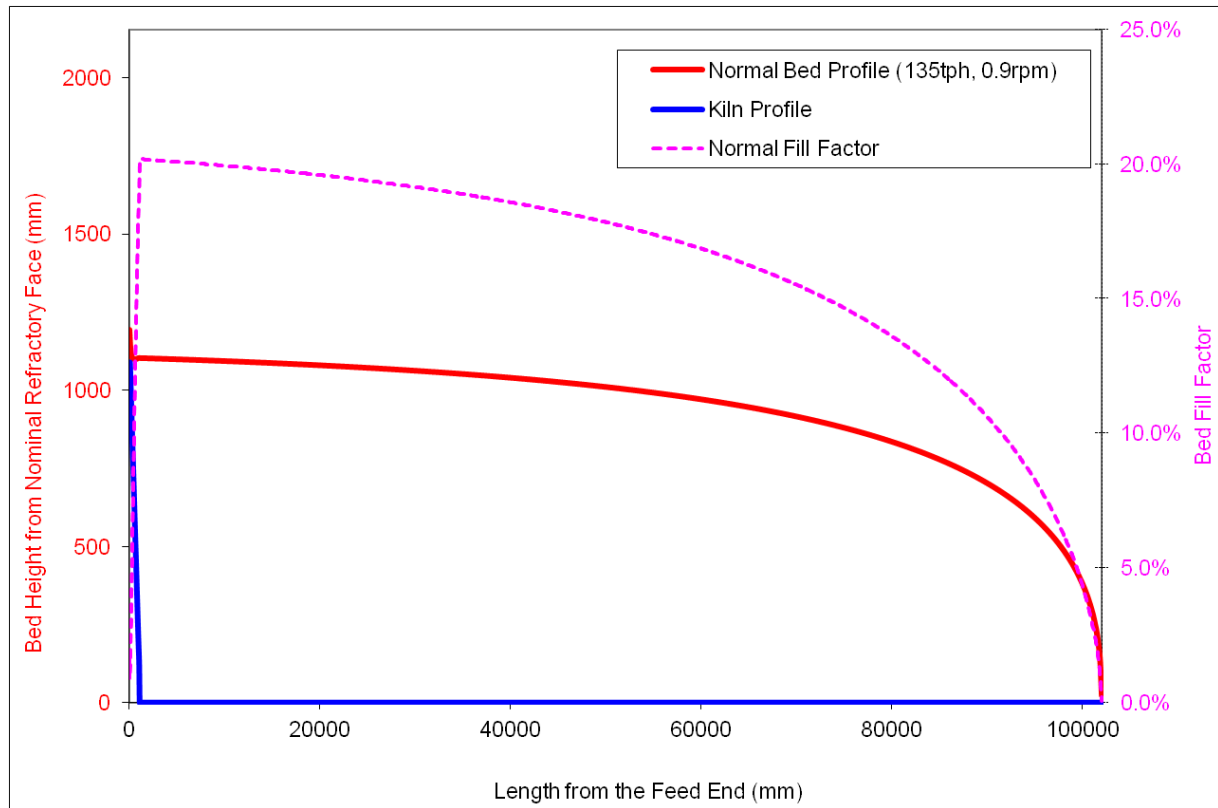
Current stable operation appears to be ~85wt/h of feed. The initial bed profile was run at 0.72rpm. This was because the %speed in the control system was 80% and it was assumed the kiln had 4-pole motors (which would give a 100% motor speed equal to 0.9rpm of the kiln). The bed profile is shown in Figure 4.1.8.1-1. The residence time was calculated to be 5.68hrs. If the 80% speed equates to 0.8rpm, then the residence time would reduce to 5.08hrs. The bed height at the feed end would reduce from 893mm to 818mm.



**Figure 4.1.8.1-1 – 85wt/h Bed Profile @ 0.72rpm**

A preliminary estimate of the power required to run the kiln was also made. The estimate used only the mass of the kiln bed, as the mass of the kiln and all its components cannot be determined with the currently available drawings. The power drawing was estimated to be 487kW (495kW @ 0.8rpm). This is almost 50% of the installed 1000kW of drive power. Most kilns are designed to run at 50-60% power draw for normal operation so that there is sufficient capacity available for upset conditions and start-up. The calculated motor torque was compared to the motor torque observed during operation for a period of representative kiln operation from the site data (site motor torque was calculated from current draw measurements) and agreed to within 10%.

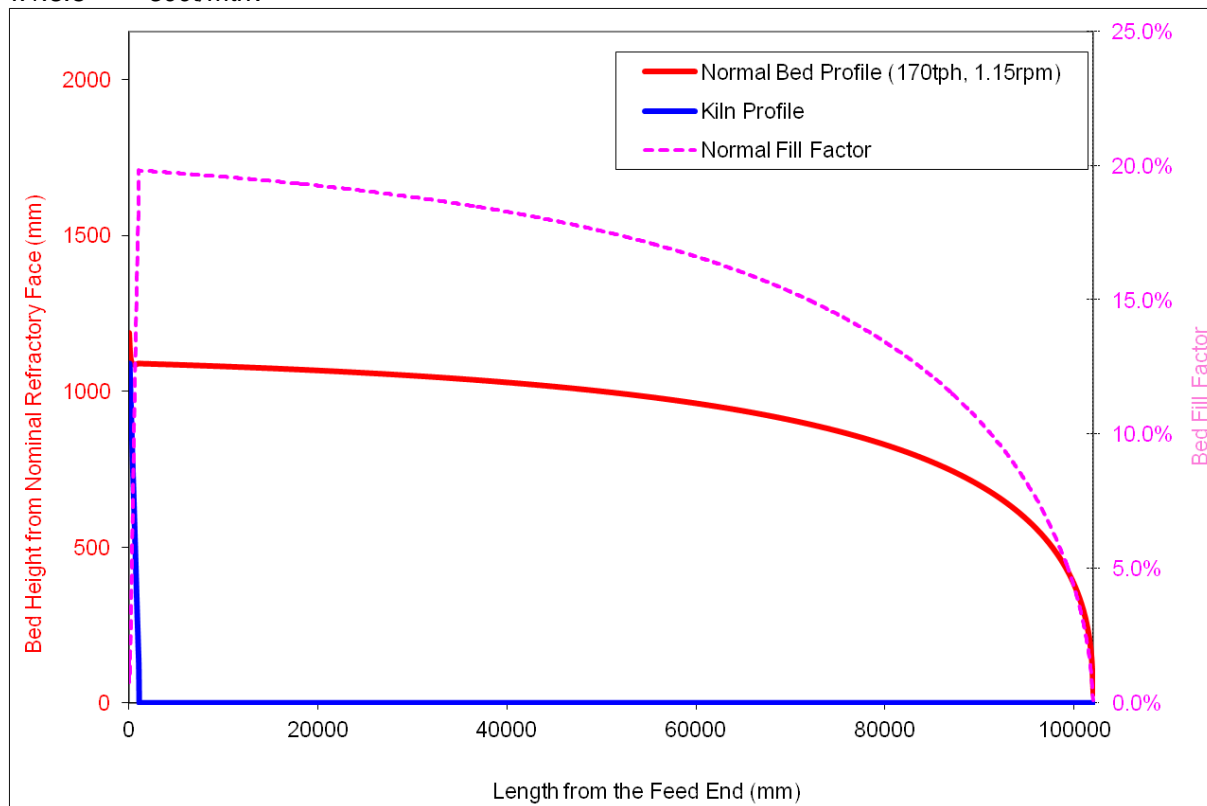
4.1.8.2 400t/mthv



**Figure 4.1.8.2-1 – 135wt/h Bed Profile @ 0.9rpm**

A bed profile was created for the 400t/mthv case which is shown in Figure 4.1.8.2-1. It was decided to limit the kiln fill factor at the feed end to 20% (this is a typical maximum feed end fill rate used in other industries, it normally correlates to a level just below the inside diameter of the feed end opening). Feed spirals will likely be required to help prevent back spillage into the feed hood at this flow rate (refer to Section 4.3). In order to limit the bed fill the kiln speed was run at 0.9rpm. The residence time in this scenario was 4.61hrs. The calculated power draw is 741kW, ~74% of the installed power or a 52% increase on the current stable operation power requirements. The drive can likely handle this. The reducer data sheet specifies it is designed for a normal operating power draw of 375kW (x 2 = 750kW).

4.1.8.3 500t/mthv



**Figure 4.1.8.3-1 – 170wt/h Bed Profile @ 1.15rpm**

A bed profile was created for the 500t/mthv case which is shown in Figure 4.1.8.3-1. As with the 400t/mthv case the feed end bed fill was limited to 20% and feed spirals will be required. In order to limit the bed fill the kiln speed was run at 1.15rpm. The residence time in this scenario was 3.60hrs. The calculated power draw is 936kW, ~94% of the installed power or a 92% increase on the current stable operation power requirements. It is very unlikely the drives will handle this power requirement. The larger issue is that the kiln speed of 1.15rpm relates to a 189% speed for the 6-pole drive motors. It is expected the motor will not be able to operate at constant power up to this speed.

For these 2 reasons it would be expected new drives (and VSDs) would be required, likely two 4-pole 750kW drives (running up to 128% of synchronous speed). The existing reducers are designed for up to 750kW drives, the additional speed should be checked with the vendor.

The upgrade to 750kW will be a difficult task. It would involve the replacement of the motors, cables, switchgear and VSDs. The existing motor is a high current, low voltage 415V motor; this requires a large amount of current and therefore large cables. A 750kW 415V motor would have to be custom ordered as low voltage motors are not normally produced at that size. Large rotary kilns with such large drives normally use high voltage motors because of this. Converting to high voltage would be an even larger task and would probably require new switchrooms to be installed. The estimate has been based on sourcing a pair of special order 750kW 415V motors.

## 4.2 Kiln Drive Motors

The reducer datasheet shows that the input speed is 161 to 1613rpm (relating to a kiln speed of 0.1 - 1rpm). This can be achieved with either:

- A 4-pole motor (~1485rpm synchronous speed) running from 11 - 109% speed, or
- A 6-pole motor (~985rpm synchronous speed) running from 16 - 164% speed.

The roasting kiln has been fitted with two 500kW 6-pole motors. The advantage of the 6-pole motors is that they will run at the lower rpm range better as the cooling fan will be spinning faster and they will have more torque available at this speed. The available torque vs. kiln speed graph for a 500kW 4 and 6 pole motor is shown in Figure 4.2-1. The figure also shows that the motors are constant torque up to their 100% synchronous speeds and then constant power after that (constant power means that the torque must drop as the speed is increased).

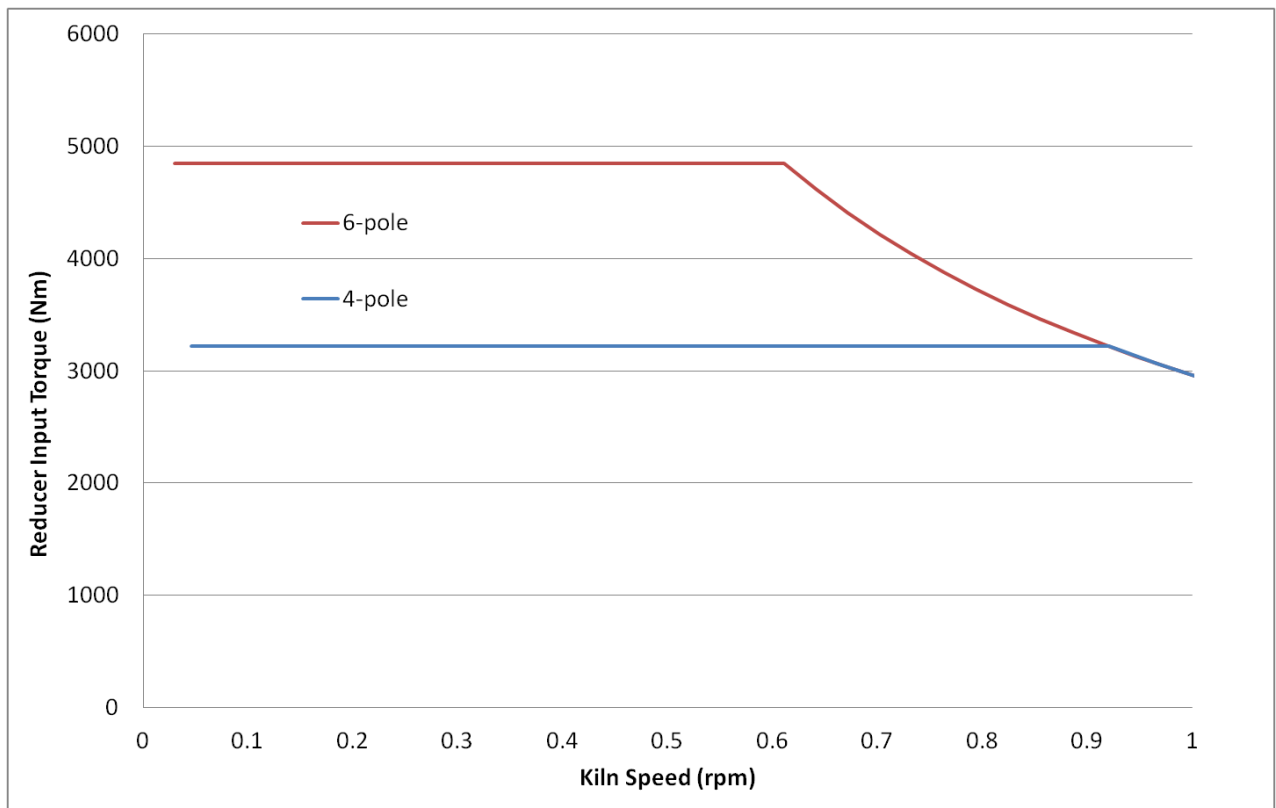
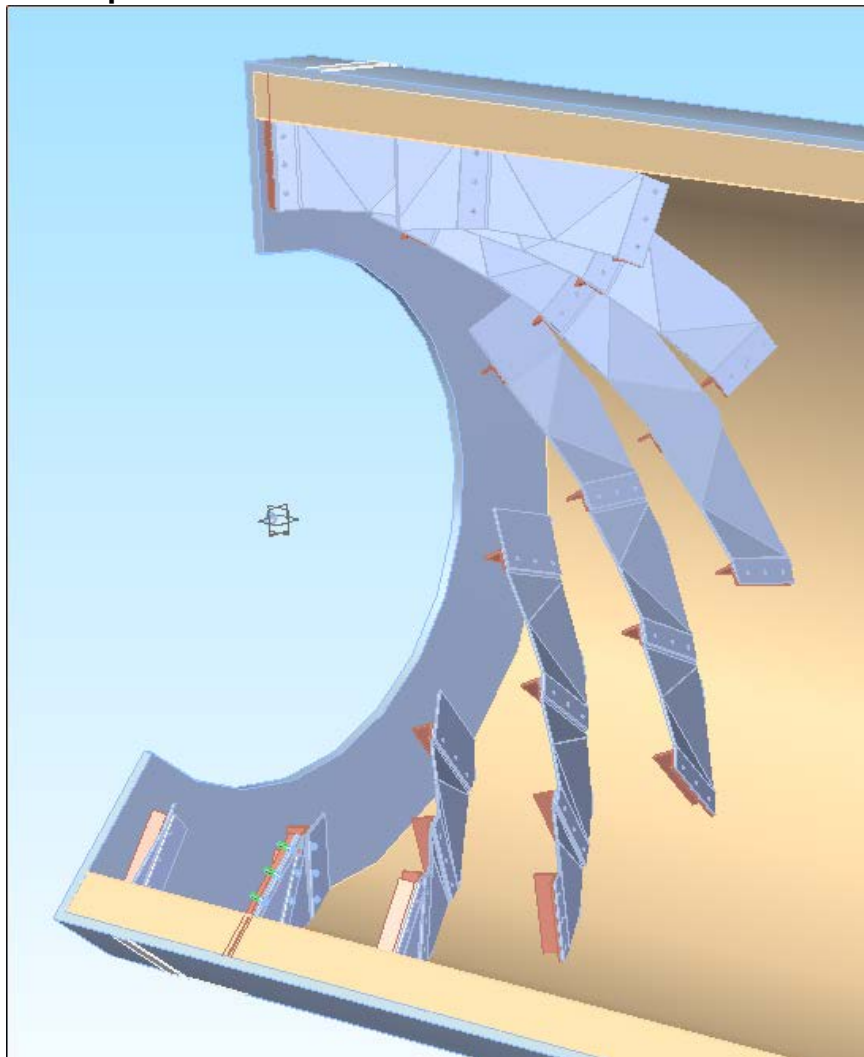


Figure 4.2-1 – Comparison of Available Torque vs Speed of the 4 and 6 Pole Motors

Below 0.9rpm (100% speed of the 4 pole motor), the 6-pole motor will have more torque available for rotating the kiln. Above that speed they will both have the same torque available. There is a limit to whether the 6-pole motor can continue to produce constant power up to 164% speed, this is dependent on the motor characteristic and would need to be checked with the motor manufacturer. The manufacturer would also need to verify the motor can safely operate at 164% speed.

It is unlikely the 6-pole motor can be used much higher than 164% speed. This would therefore mean it would need to be replaced if the kiln is going to exceed 1rpm.

### 4.3 Feed Spirals Discussions



**Figure 4.3-1 – Bolt in Feed Spirals**

Feed spirals are often used to push the material away from where it lands after it is discharged from the feed screw. This can help eliminate backpillage. Figure 4.3-1 is a feed spiral design used by Hatch in the past. The installation requires some of the original castable refractory lining to be removed in order for the mounting posts to be welded to the shell. The spiral plates are then bolted to the posts.

For difficult to handle material the feed spirals are sometimes fitted with chains either between or along the spirals to help dislodge material from between them. This is shown in Figure 4.3-2 which is taken from the 1941 US Patent # US2261403.

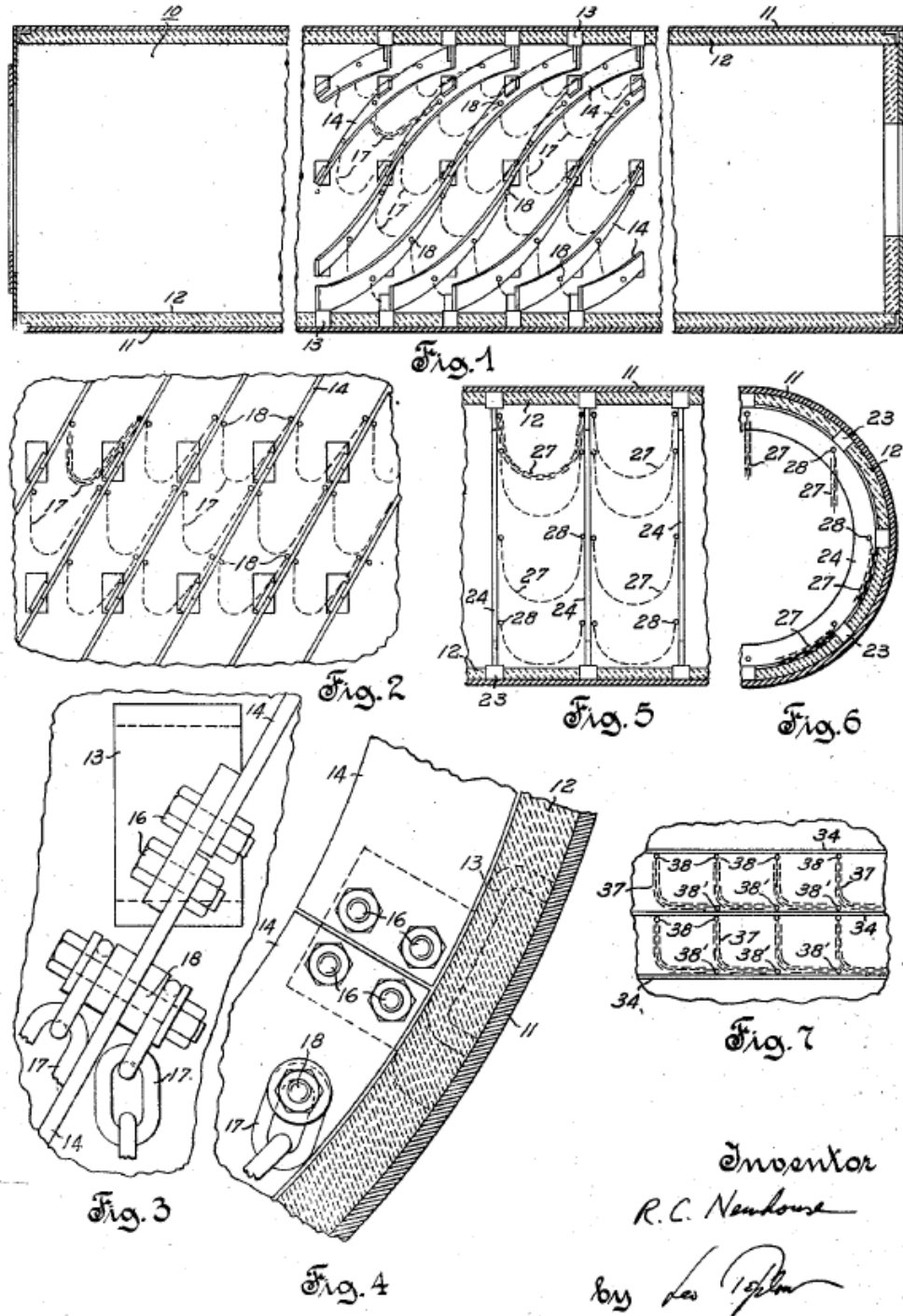
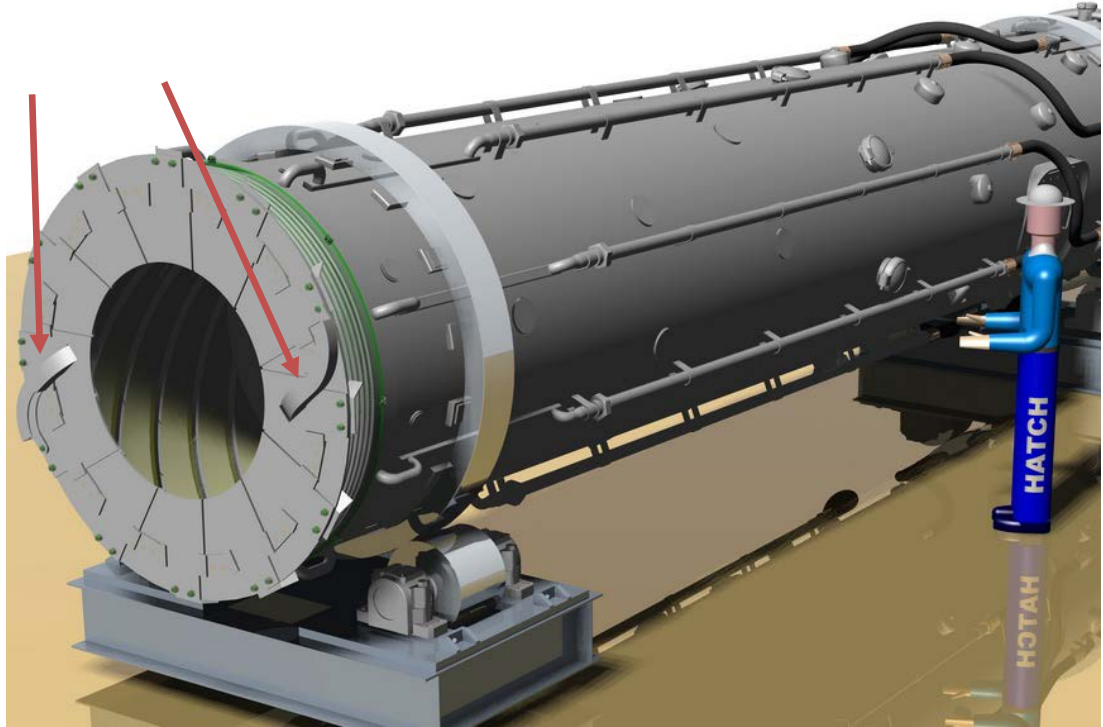


Figure 4.3-2 – Feed Spirals with Cleaning Chains



**Figure 4.3-3 – Hatch Feed Reclaim Scoops**

Feed spirals would normally be sufficient to prevent back-spillage into the feed hood. Occasionally the off-gas velocity may carry feed over the feed end and into the feed hood. For an application with this problem Hatch added reclaim scoops which rotate with the kiln (cooler in this case) and scoop the material from the feed hood and drop it back into the kiln. A dry feed is necessary for this device to work. Some grease seal designs use a similar concept of using the kiln rotation to help reclaim material from the feed hood area back into the kiln.

#### **4.4 New Burner**

Initial recommendations and cost estimates are for the supply of a Duoflex burner from FLS. There are, however, a number of very capable burner manufacturers that could custom design a burner to suit Windimurra's kiln. Because of the unique burner attributes that would best achieve the salt roast kiln process objectives, consideration could be given to tendering this item competitively.

## 5. AMV Reduction Kiln Results / Discussion

### 5.1 Bed Profiles and Thermal Performance

The reduction kiln was analysed using a basic sensible heating only kiln model. The model inputs are shown in Table 5.1-1:

**Table 5.1-1 – Reduction Kiln Model Inputs**

Parameter	Value	Comment / Reference
AMV bulk density	1035 kg/m <sup>3</sup>	Drytech Manual – Plant Specification
AMV bulk density abnormal case	730 kg/m <sup>3</sup>	Email Flormirza Cabalteja 28/05/14
AMV Specific Heat	1105 J/kg.K	Drytech Mass Balance
Product Feed Temperature	20 deg C	Drytech Mass Balance
Dynamic angle of repose	37 deg	Assumed (no data found)
Kiln slope	0%	Drytech Drawings
Feed moisture	0%	Upstream Flash Dryer
Kiln speed	2.78rpm	Drytech Manual (19.8rpm x 27/192 teeth)
Shell Temperature	950 – 900 deg C	Drytech 22/08/2012 Memo
Dams?	No	Midwest drawing 10-336-ME-DET-0006
Feed Spirals	50 tall x 500 pitch x 1500 long	Midwest drawing 10-336-ME-DET-0006
Discharge Spirals	50 tall x 500 pitch x 500 long	Midwest drawing 10-336-ME-DET-0006
Lifters	50 tall straight lifters	Midwest drawing 10-336-ME-DET-0006

The indirect kiln / cooler model used is based on sensible heating only. Any heats of reaction need to be converted to an artificially high Cp value at the expected reaction temperature. The expected reactions are:

- $2\text{NH}_4\text{VO}_3(\text{s}) \rightarrow \text{V}_2\text{O}_5(\text{s}) + 2\text{NH}_3(\text{g}) + \text{H}_2\text{O}(\text{aq})$ , occurring at ~400 deg C;
- $2\text{NH}_3(\text{g}) \rightarrow \text{N}_2(\text{g}) + 3\text{H}_2(\text{g})$ , occurring at ~600 deg C;
- $\text{V}_2\text{O}_5(\text{s}) + 2\text{H}_2(\text{g}) \rightarrow \text{V}_2\text{O}_3(\text{s}) + 2\text{H}_2\text{O}(\text{g})$ , occurring at ~800 deg C and requiring 30minutes residence time above that temperature.

The heat of reactions for the first two reactions are required, they are calculated to be 961kJ/kg and 462kJ/kg respectively. The third reaction is exothermic and therefore will occur as long as the temperature is reached and maintained for the desired residence time.

Note this is a total reaction heat of for the first 2 reactions of 1423kJ/kg. This is significantly more than the 585kJ/kg reaction heats in the Drytech mass balance. The Drytech heat balance assumed all the reactions happened at once, which is not the case. Their simplified mass balance also does not appear to account for the fact the various reactions happen at different temperatures and that the last reaction is exothermic. The Drytech and Hatch sensible heats are the same as the same specific heat value was used.

**Table 5.1-2 – Specific Heat Values Used**

Temp deg C	Specific Heat J/kg.K
0	1105
399.9	1105
400	33138.33
430	33138.33
430.1	1105
599.9	1105
600	16505
630	16505
630.1	1105
10000	1105

The heat of reactions are incorporated into the simple sensible heating model by artificially raising the specific heat at the reaction temperature as shown in Table 5.1-2. A temperature spread of 30 deg C has been used for the reactions to improve model convergence compared to a narrower temperature range. The equivalent heats for each reaction is expressed in the calculations below:

$$\text{Reaction equivalent Cp} = \text{Sensible Cp} + (\text{Reaction Energy} / \text{Temperature Range})$$

$$400\text{-}430^{\circ}\text{C reaction Cp} = 1105 \text{ J/kg.K} + (961,000\text{J/kg} / 30 \text{ deg C}) = 33,138.33 \text{ J/kg.K}$$

$$600\text{-}630^{\circ}\text{C reaction Cp} = 1105 \text{ J/kg.K} + (462,000\text{J/kg} / 30 \text{ deg C}) = 16,505 \text{ J/kg.K}$$

This can be cross checked by calculating the total energy per mass of feed using the simplified modified Cp method (1) and a method incorporating the heats of reactions (2) as follows:

1. Total energy / kg mass  
= Sum of Specific Heats x Temperature Ranges  
=  $1105 \times (400 - 20 + 600 - 430 + 950 - 630) + 33138.33 \times (430 - 400) + 16505 \times (630 - 600)$   
= 2450.65kJ/kg
2. Total energy / kg mass  
= Specific Heat x Total Temperature Range + Heat of Reactions  
=  $1105 \text{ J/kg.K} \times (950^{\circ}\text{C} - 20^{\circ}\text{C}) + 961,000 \text{ J/kg} + 462,000 \text{ J/kg}$   
= 2450.65kJ/kg

The first model run is the kiln processing 830kg/hr of AMV as per the Proteus PDC<sup>2</sup>. The predicted bed profile produces a 1.02hr residence time and is shown in Figure 5.1-1. The slope of the bed causes the material to move along the kiln during rotation, rather than the slope of the kiln for a traditional inclined kiln. This causes a ramped up bed height at the feed end which gives more residence time at the start of the kiln. The spirals are assumed to work perfectly, moving the product at the rate of the spiral pitch / rotational speed (23.2mm/s).

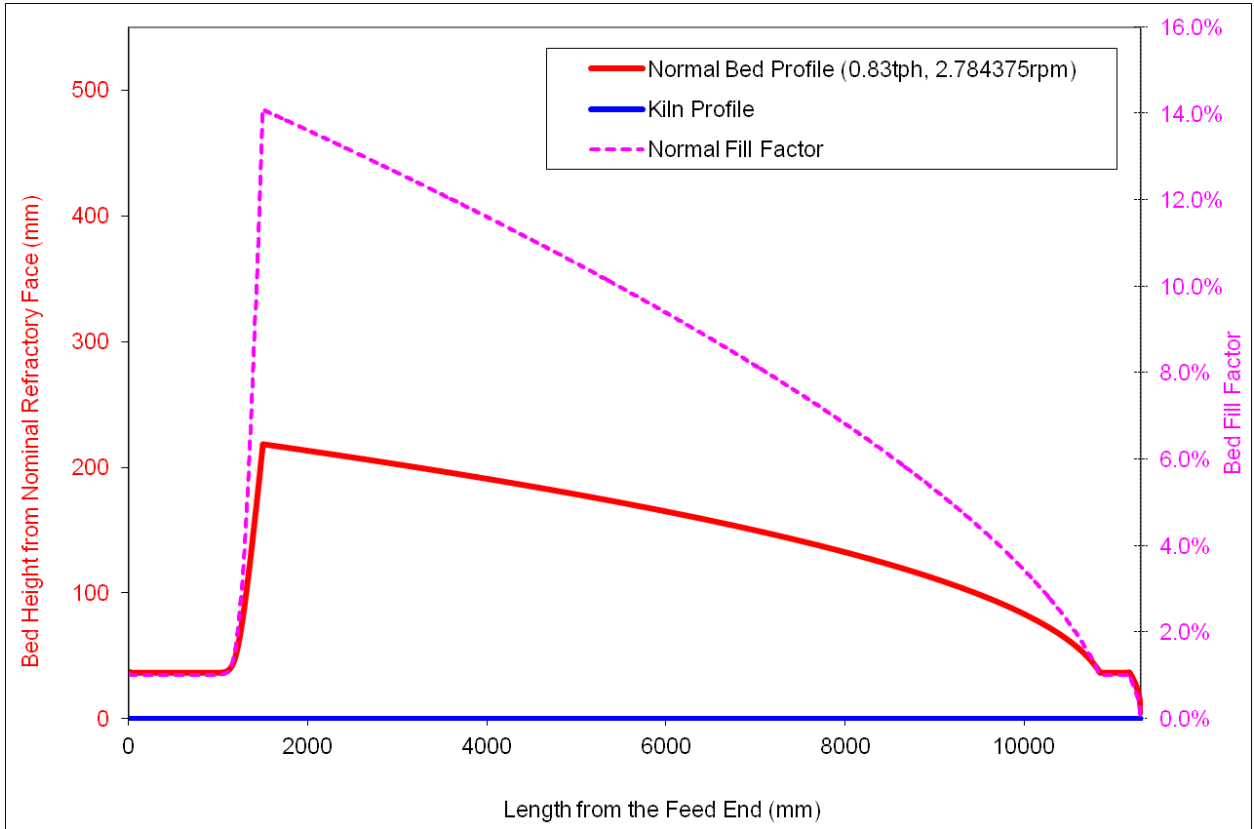


Figure 5.1-1 – 830kg/hr AMV Bed Profile, 1.02hr Residence Time

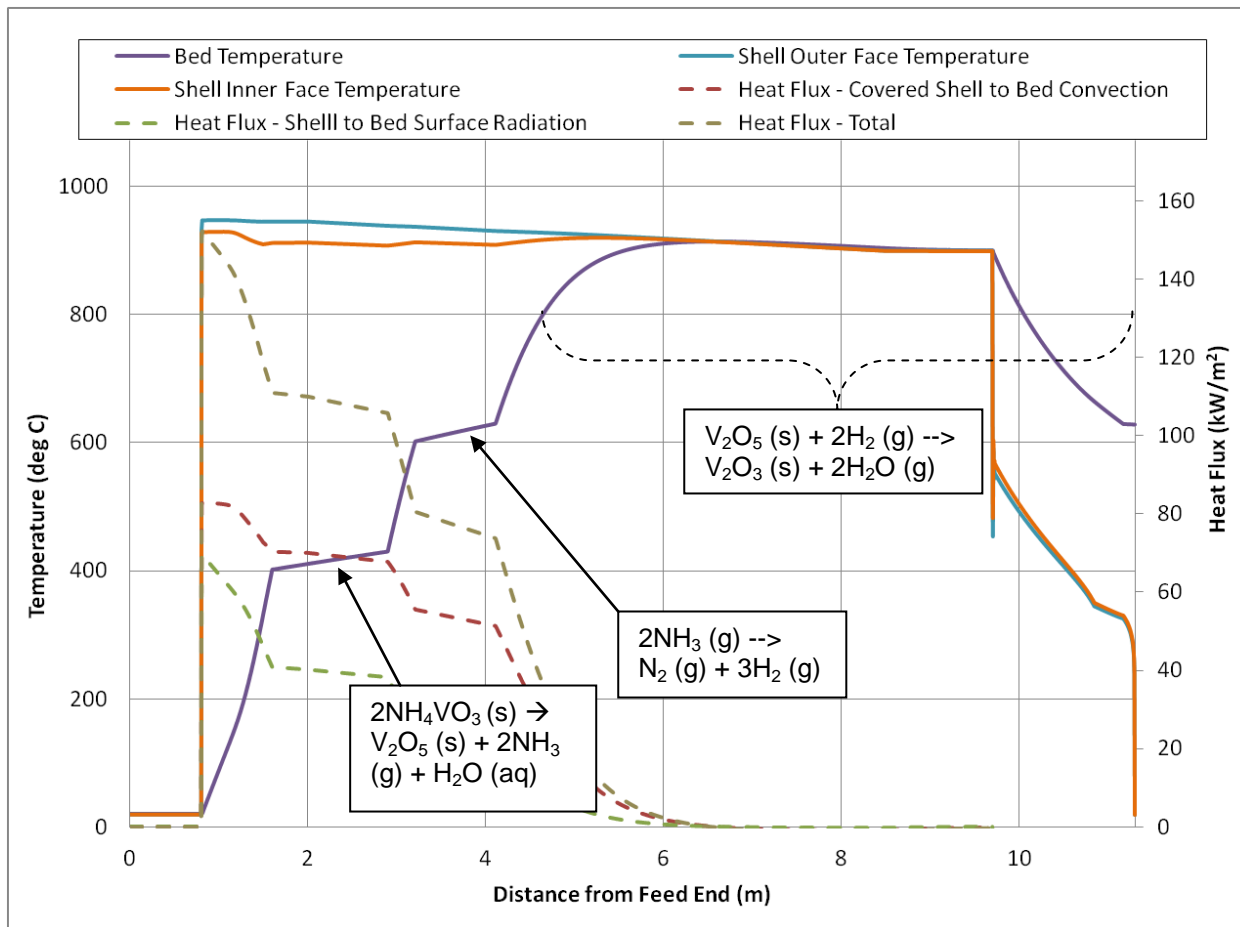


Figure 5.1-2 – 830kg/hr Thermal Profile, 30.6min > 800 deg C

The calculated thermal profile is given in Figure 5.1-2. Features to note are:

- The blue line is the shell temperature targeted by the burners, it is ramped from 950 to 900 deg C.
- The purple line is the bed temperature; the plateau's at 400 and 600 deg C are for the first 2 reactions.
- The exothermic heat of reaction of the third reaction  $\text{V}_2\text{O}_5 + \text{H}_{2(\text{g})} = \text{V}_2\text{O}_3 + \text{H}_2\text{O}_{(\text{g})}$  was excluded from the heat balance. If it were included it would cause the bed temperature to start to increase rapidly above 800 deg C (and approach the kiln shell temperature quicker). This is a conservative approach, as only the residence time remaining after it reaches 800 deg C is important (target is 30 minutes to allow the reactions to occur).
- The heating hood is only in the middle of the kiln; hence the cooler temperatures either side.
- The model assumes no axial heat conduction along the shell outside of the heating hoods, and no radiation in the axial direction either. In reality the material will start heating at the very start of the kiln due to these effects.

- Likewise the model assumes the product will start cooling as soon as it exits the heating zone (because the effects listed above are excluded). If the product in this final exposed section of the shell actually did cool to a point that conversion to  $V_2O_3$  was affected, it may be necessary to either insulate or refractory line the shell (the latter would have a dam effect).

Hatch calculates the 400t/mth case would relate to 737kg/hr AMV to the reduction kiln. This is less than the Proteus PDC rate and therefore does not need to be examined.

The 500t/mth case is calculated to be 921kg/hr of AMV feed to the reduction kiln. The bed height is shown in Figure 5.1-3 (0.98hr residence time).

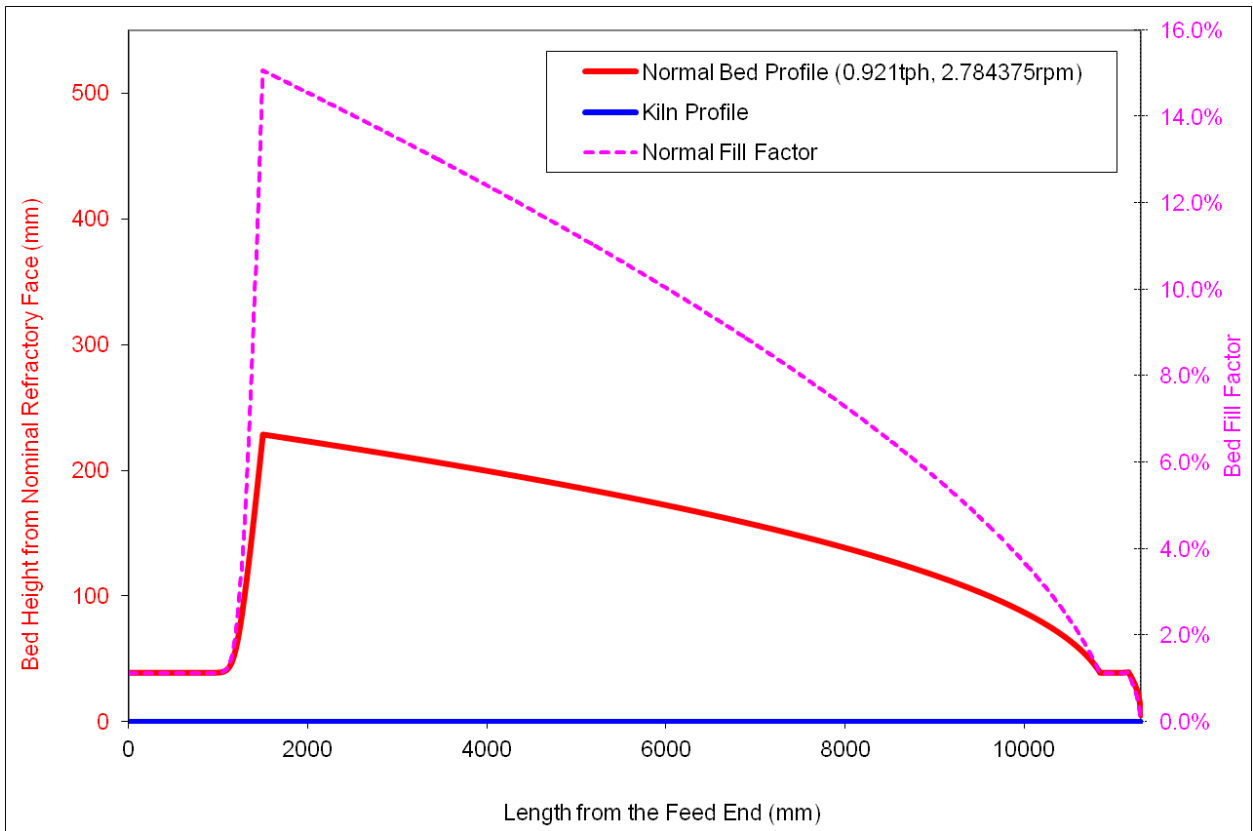


Figure 5.1-3 – 921kg/hr AMV Bed Profile, 0.98hr Residence Time

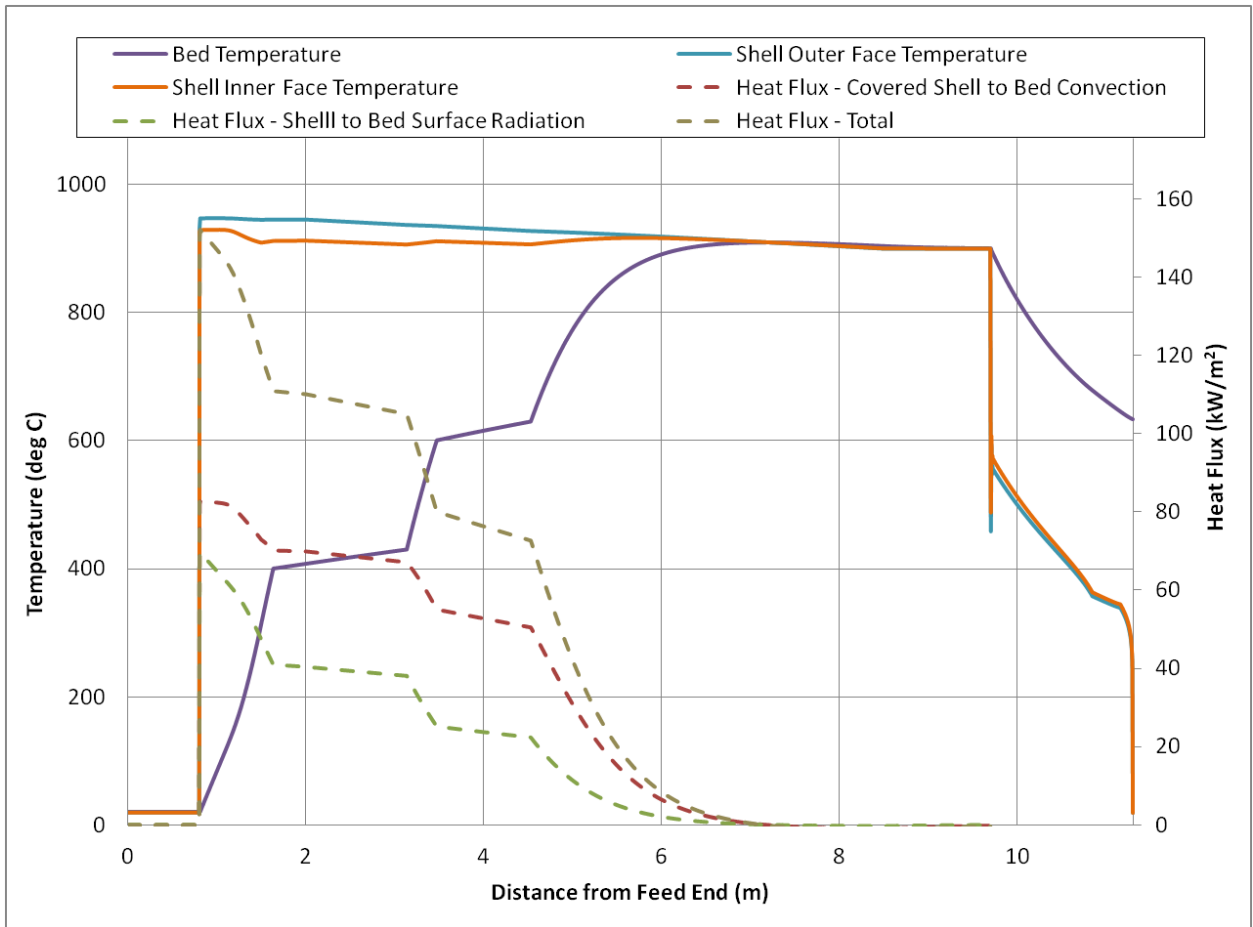


Figure 5.1-4 – 921kg/hr Thermal Profile, 26min > 800 deg C

Figure 5.1-4 shows the thermal profile for the 921kg/hr case. In this scenario the 30 minutes residence time cannot be maintained. At this point it should be noted that the discharge spirals serve no purpose, they only reduce the amount of material in the kiln. The feed spirals are also excessively long (along the kiln axis), refer Figure 5.1-5. Because of the indirect kiln design and co-current gas flow there is no risk of feed back-spillage and therefore the spirals are not necessary.

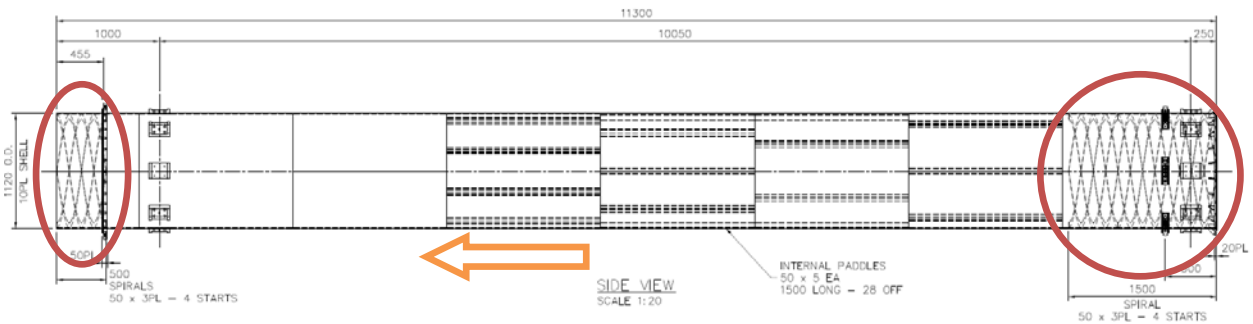
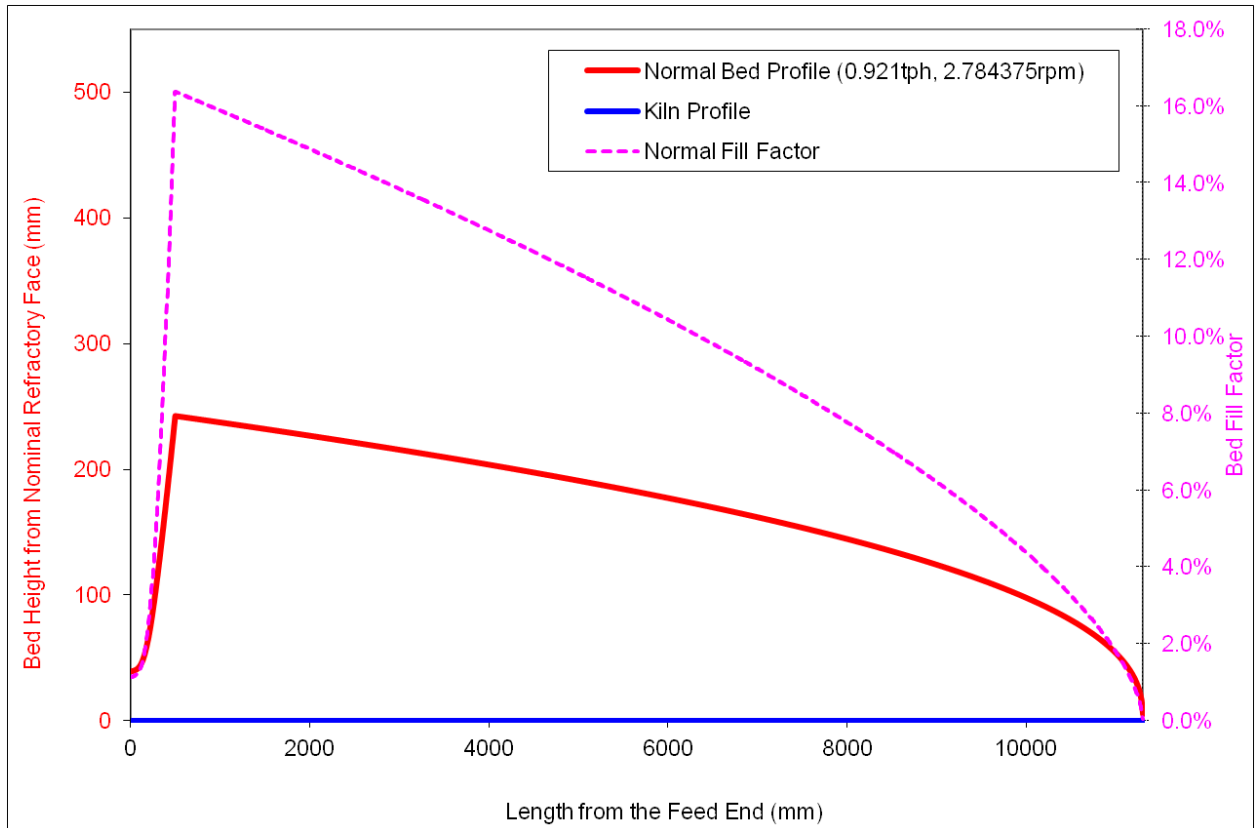
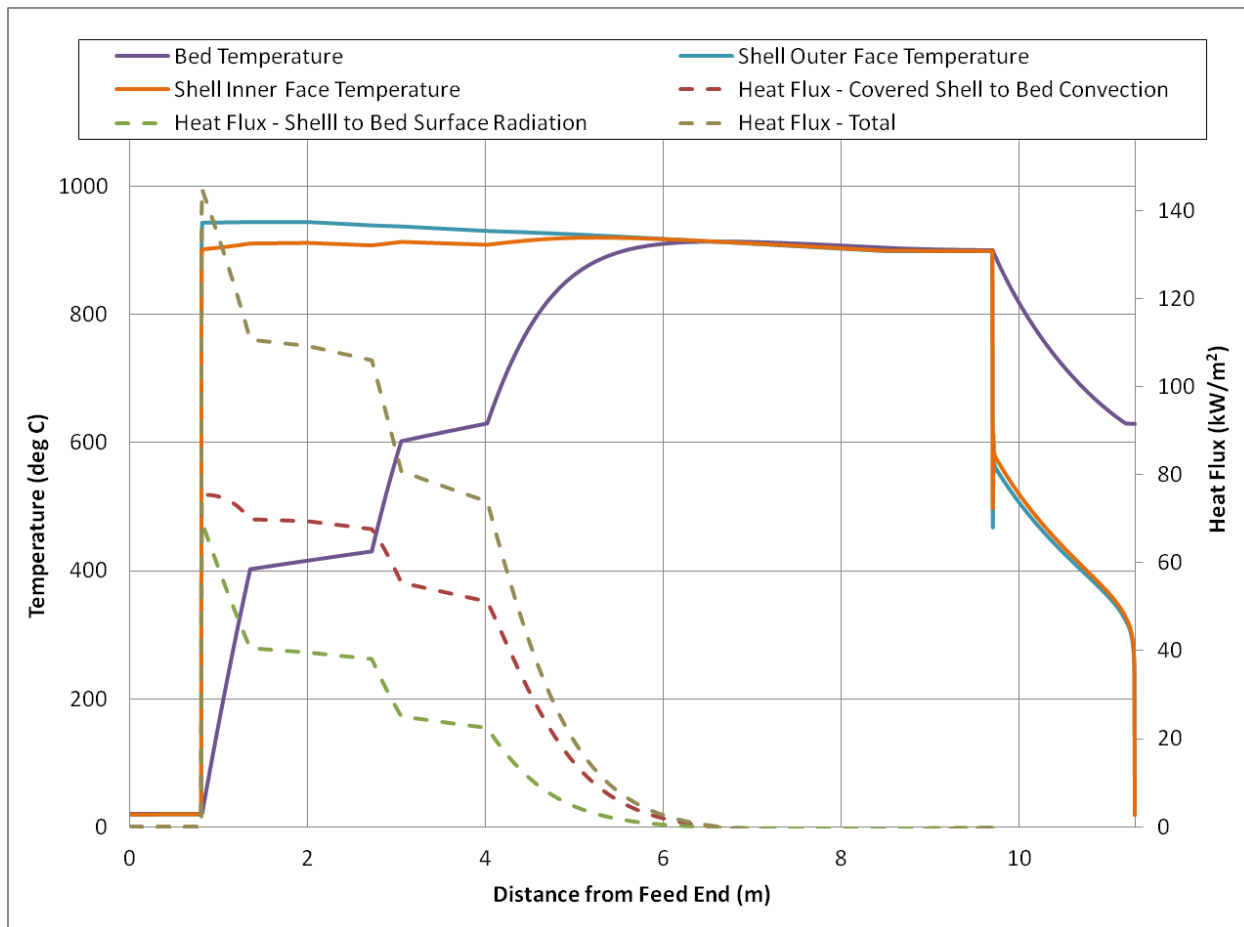


Figure 5.1-5 – Kiln Spirals



**Figure 5.1-6 – 921kg/hr AMV Bed Profile with Modified Spirals, 1.19hr Residence Time**

Figure 5.1-6 and Figure 5.1-7 show the bed and thermal profiles for the 921kg/hr case with the feed spirals shortened to 500mm long and the discharge spirals removed. The kiln is then capable of achieving the 30mins residence time at temperature.



**Figure 5.1-7 – 921kg/hr Thermal Profile with Modified Spirals, 32.2min > 800 deg C**

A sensitivity to feed material bulk density was also performed by changing the density from 1035kg/m<sup>3</sup> to 730kg/m<sup>3</sup>. This reduced the kiln total residence time to 1.05hrs and the time at reaction temperature to 30 minutes (for the 921kg/hr case).

These reduction kilns still have a lot of flexibility to accommodate even higher material flow rates. The kiln speed could be slowed or small discharge dams fitted to increase the heat transfer area and time at temperature even further.

The impact on the shell strength and support tyre life due to heavier bed loads has not been analysed at this stage. This work could be completed in the future if required. A preliminary calculation suggested the kiln shell bending stress is very low, approximately 5MPa.

Site data for the reduction kiln was reviewed. In general there were not many period of continuous operation of the kilns. The highest daily average flow was produced on the 12<sup>th</sup> of November 2013, refer Figure 5.1-8. Operation in excess of 921kg/hr (500t/mthv) was continued for most of the day. If the product produced during this period was acceptable then the operation of the kiln at 921kg/hr could be considered proven in the existing configuration.

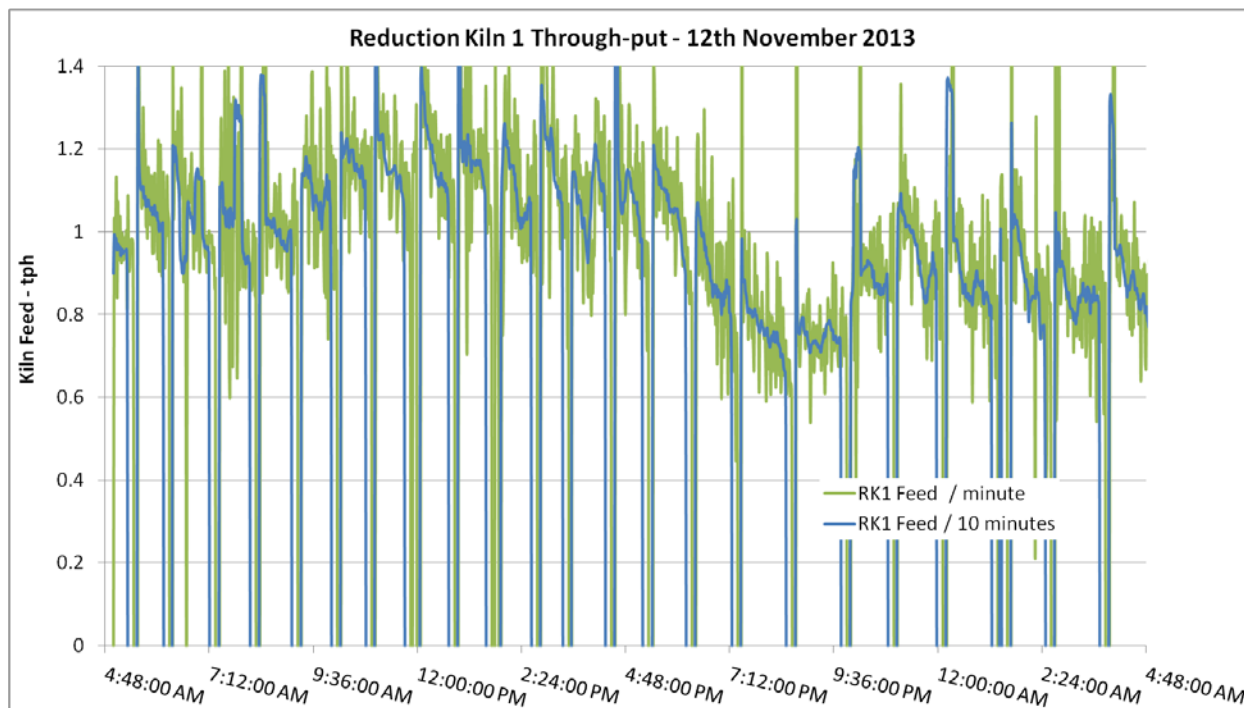


Figure 5.1-8 – Site Data Showing Highest Reduction Kiln Throughput

## 5.2 Burner Capacity

Discussion with plant personnel during the site visit 15/5/2014 suggested that the reduction kiln burner system was undersized for the required duty. Calculations based on our current understanding of the kinetics of the de-ammoniation and reduction reactions indicate that the burner capacity is marginal for the 400t/mth<sub>V</sub> duty and is undersized to sustain production at the 500t/mth<sub>V</sub> level. The installed burner capacity is understood to be 6 x 318kW burners for a total thermal input of 1.91MW.

The high gas supply temperature and high heat losses that result in order to maintain a 900 – 950°C shell temperature substantially increase the burner duty beyond simply the heat required to drive the process reactions. Table 5.2-1 below shows the estimated burner requirements for the 400 and 500t/mth<sub>V</sub> production scenarios. The required burner nominal capacity is calculated as:

$$\text{Required Burner Capacity} = \frac{\text{Total Heat required}}{\text{De rating factor} \times 0.80}$$

Table 5.2-1 – Required Burner Capacity for Reduction Kiln

Case	Total Heat Required (Including losses) (MW)	Burner De-rating with 950 Exhaust Temperature	Required Burner Nominal Capacity based on 80% duty (MW)
400 t/mth <sub>V</sub>	0.75	0.51	1.83
500 t/mth <sub>V</sub>	0.87	0.51	2.13

The simplest approach to supplying the additional heat would be to provide two additional packaged burners and adjust the burner spacing near the feed end of the reduction kiln to accommodate the new burners.

### 5.3 Feed and Discharge Screw Capacity

The capacity of the feed and discharge screws have not been analysed in detail at this stage. It is expected that both screws will be easily capable of handling the small proposed capacity increase, whether a speed increase is required or not has not been determined. Any such speed increase should be a low cost change due to the small motor and reduce sizes (or free if there is speed capacity left in the VSD).

Data from site regarding changes to the kiln screw speeds is shown in Figure 5.3-1. The reason for these changes and the flow rate of material through the kiln at the time that the changes were required is not understood at this stage.

#### Changes made to the kiln feed screw:

1. RK1 feed screw 41FDS501 increased from 50Hz to 60Hz on 04/11/13
2. RK2 feed screw 41FDS503 increased from 50Hz to 60Hz on 16/10/13, then from 60Hz to 70Hz on 17/12/13
3. RK1 discharge screw 41FDS502 increased from 50Hz to 60 Hz on 13/11/13 (CM)
4. RK2 discharge screw 41FDS504 increased from 50Hz to 60Hz on 14/12/13 (MG)

Figure 5.3-1 – Kiln Screw Motor Frequency Changes

The cooling capacity of the discharge screw has not been investigated in detail. If it is not sufficient an alternative cooling device or equipment arrangement may need to be investigated. Preliminary information calculations suggest the existing watercooled discharge screw may be capable of cooling the material below 250 deg C.

It should be noted that changing the speed of the screw (50-60Hz shown above) will directly affect its cooling capacity. As the screw is choke fed (refer Section 5.4) the area of contact between the material and the cooled jacket is fixed. The feeder flights will be 100% full and the following coarse flights will be ~50% full (depending on the relative spacing of the flights). Running the screw faster will increase the mass flow rate through the screw, which will reduce the cooling as the cooling area is fixed. When the screw is stopped (while level is building in the discharge hood) the material in the screw will get cooled the entire time and approach the water temperature, but the material in the hood will stay hot. When the screw starts the hot material is pushed through the screw at a flow rate proportional to the screw speed.

Therefore cooling would expect to be better if the screw was run slowly (leading to a long period of time between screw stops and starts as the material in the hood fluctuates between the high and low levels). This leads to the slowest mass flow rate through the screw.

The current design has no allowance for the differential thermal expansion between the screw trough and its cooling jacket. Cracking issues and water leaks are expected at this location. This design flaw could be corrected by fitting an expansion joint into the outer cooling jacket.

### 5.4 Kiln Gas Seal Design and Philosophy

The current gas seal philosophy at the discharge end of the kiln assumes a level of material is maintained in the discharge hood of the kiln. This material choke feeds the discharge screw creating a gas seal. The design will prevent air entering the kiln, but does not stop the hot material in the screw from being exposed to air from the product bin. It is expected that the material will need to be maintained in a reducing environment until it is cooled sufficiently to prevent reoxidation type reactions.

The screw is shown in Figure 5.4-1. The first section of flights are close pitched feeder flights designed to run completely full, creating a good gas seal. These cannot be used the full length of the kiln as it often leads to the material gripping onto the screw and rotating in the trough (therefore not moving along the trough and blocking). As the pitch changes to the coarser pitch the level of material in the screw drops. The rest of the screw may only be 50% full. This gives a greater potential for gas to backflow along the screw.

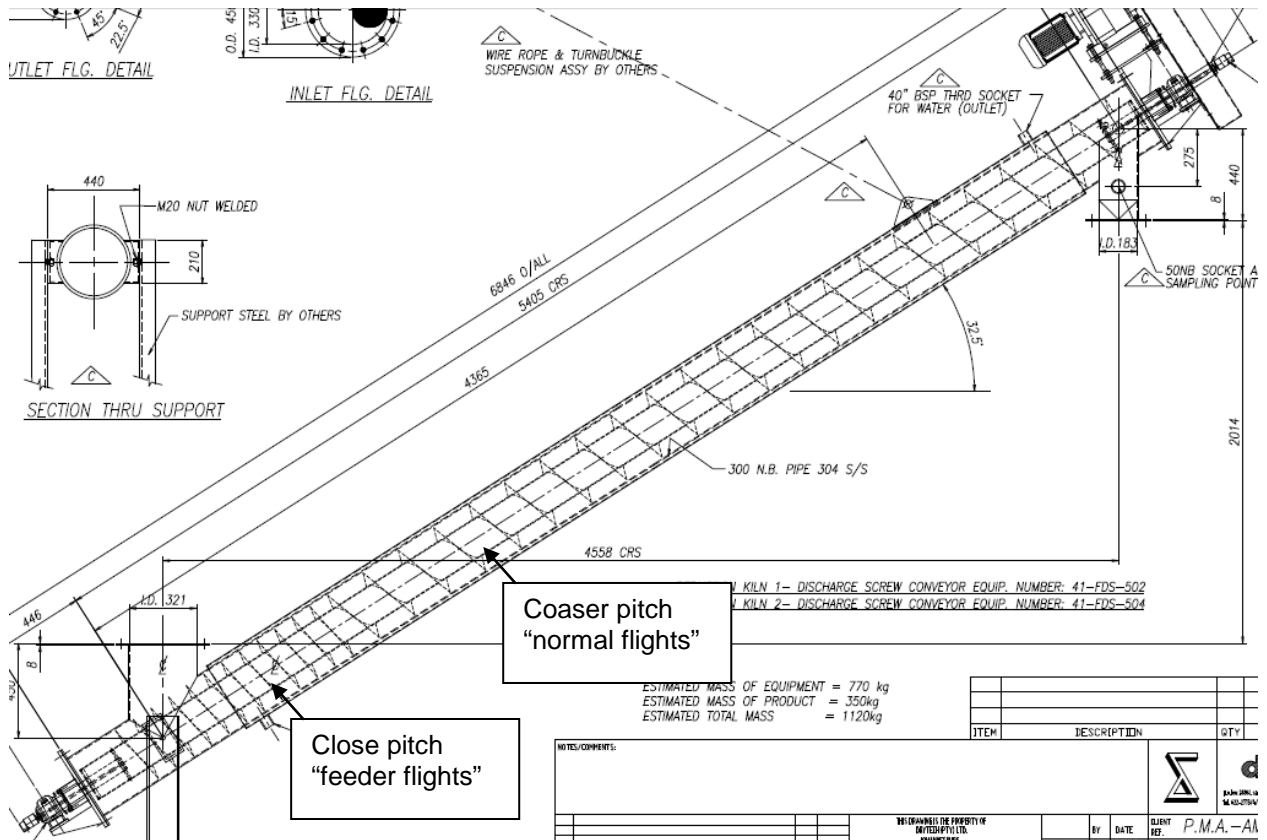


Figure 5.4-1 – Discharge Screw

A better solution would be to install a better gas seal either between the discharge screw and product bin, or after the product bin. This would also solve the issues of trying to accurately determine the material level in the kiln discharge hood. Running the discharge hood empty would also greatly reduce the chance of blockages in the hood.

An alternative would be to look at alternative pneumatic conveying equipment, either a vacuum based system or a dense phase blow-pot system to prevent ingress air returning to the kiln.

## 5.5 Kiln Removal / Replacement

Site advised it is difficult to remove the kilns due to the congested nature of the plant. Hatch has experience in designing custom monorails and lifting equipment to assist in these situations. Hatch has designed a curved monorail for a rotary cooler which allowed the cooler to be removed from a building while avoid existing equipment between the cooler and the possible crane pick-up point.

The cooler also had headroom constraints so conventional rigging equipment could not be used to connect the cooler to the monorail. Hatch designed custom lifting equipment to connect the coolers tyres as directly as possible to the monorail trolleys. Similar approaches may be of interest to assist with the removal and replacement of these reduction kilns.

## 6. De-bottlenecking priorities

The priorities for debottlenecking the plant are:

- Re-commission and ramp-up the plant to 85t/h;
- Stabilize operation at 85t/h and 80% vanadium recovery;
- Increase production to 400t/m FeV;
- Increase production to 500t/mth FeV.

### 6.1 Engineering Changes Required During Re-commissioning and Ramp-up

Working capital is needed during ramp-up. As with a contingency fund, it is not possible to predict what it will be spent on, but experience is that between 5-10% of the original capital cost of the plant will be spent re-designing equipment to suit site specific circumstances. The more customized the plant, the higher the cost: while a cement plant would have a lower ramp-up cost requirement, a first of a kind plant will have a higher capital expenditure in the first year. Note this is to cover Capital Changes and is in addition to the costs of initial commissioning, re-commissioning and ramp-up.

While it might be expected that untested process innovations would be responsible for ramp-up problems, underestimating material handling difficulties (such as flow of solids through kilns; kiln ringing) is arguably responsible for more plant failures during ramp-up.

Two recent examples illustrate the point:

- BHP's HBI plant in Port Hedland, which cost at completion \$2.5B AUS, had a budget in excess of \$100M to cover capital projects in the first year. This was customized new technology of which there was only one other plant in the world.
- The Ravensthorpe Nickel Refinery, which cost \$2.2B on completion, was eventually purchased and ramped-up by First Quantum. The new owners reportedly spent an additional \$300M on capital fixes during ramp-up, much in the comminution and beneficiation circuits.

Windimurra is now using a well documented soda ash roast/aqueous leach design with few innovations (e.g. only satellite coolers and scrubbing kiln off-gas with pregnant liquor are significant innovations), therefore a ramp-up budget at the lower end of the scale is assumed to allow the plant to reach the minimum design capacity. For this purpose it is assumed that \$10M in capital changes (approximately \$5M in new equipment and \$5M for installation) is allowed for in the estimate.

## 6.2 Expenditure to stabilize at 85t/h

In order to stabilize operations at 85 t/h kiln feed, and to maximize vanadium recovery in the kiln, it is recommended that capital be allocated to:

- Re-instate the original feed end (hydraulic) seals and double dump system. This will improve on-line time, and improve in plant hygiene in the kiln feed area.
- Replace the pug mixer with a screw feeder to improve on-line time.
- Update kiln instrumentation to allow the kiln to maximize vanadium recovery (Unlike a cement kiln, Windimurra's kiln is a chemical reactor similar to SR kilns, SR kilns measure bed temperature along the kiln length). Useful instrumentation can include shell mounted thermocouples and a burner end pyrometer camera. The pyrometer is very useful for monitoring the process visually and allowing the operator to measure temperatures at any location on the image, refer Figure 6.2-1. Many industries find the ability to visually monitor the bed behaviour very useful in correctly operating the kiln.
- Re-commissioning the de-silication and AMV filters. These units were not fully commissioned. The costs included cover remobilizing the vendor to site to complete commissioning his scope of supply.
- An allowance to build a crushing plant has been made since the existing crushing plant is underperforming. Since FeV is a hard material very difficult to crush, it is recommended that extensive study be undertaken, including visiting FeV plants in South Africa.



Figure 6.2-1 – Screenshot from a Mirion Pyrometer Camera (M555/M215)

### 6.3 Recommended modifications to reach 400t/mth V in FeV

Further modifications to reach 400t/mth V in FeV (equivalent to 107.5t/h fresh magnetite) are again mostly in the kiln area:

- New burner. The cost included here is for supply of a Duoflex burner from FLS, however there are a number of very capable burner manufacturers that can custom design a burner to suit Windimurra's kiln and consideration could be given to tendering this new item competitively.
- Feed-end spirals, to stop feed spilling out the back of the kiln.

### 6.4 Recommended modifications to reach 500 t/mth V in FeV

Modifications to go from 400t/mth to 500t/mth V in FeV are likely to be substantial:

- The kiln will not likely be able to maintain high vanadium recoveries while fed at this rate, so a pre-dryer should be investigated. As can be seen, 3 m – 45m of the kiln length is spent drying the feed.
- As an option to further de-bottleneck the kiln, the feed can be pre-heated with the kiln off-gas. This is established technology in cement plants now, and is also used in a variety of industries (lime, potash, laterite smelting, etc). This has not been included yet, as pre-drying should be sufficient.

- An alternative to flash drying the feed would be to pelletise the magnetic concentrate with the soda ash and a suitable binder, such as barren liquor. The pellets could then be indurated on a travelling grate kiln such as a Lepol. Note that this style of preheater has been supplanted by suspension preheat systems, however it might be an opportunity to pre-dry and pre-heat the feed while minimising the dust generation rate in the kiln. There is precedent in this, and Hatch has looked at this previously, however we had assumed that a ball mill was required in the leach circuit. Test-work would be needed.
- While not necessarily a bottleneck, the current scheme for handling vat leach residue should be compared to a more automated flow-sheet.

## 7. Cost Estimates

### 7.1 Capital Cost

The capital cost estimate has been developed using the factored method, and has been divided as follows:

- Minimum capital expenditure to ramp up to 85t/h fresh feed (per section 6.1);
- Additional cost to go to 400t/mth contained vanadium;
- Additional cost to go to 500t/mth contained vanadium.

Factors used are typical of non-ferrous extraction plants. Where available, in-house costs have been relied on (either from quotes to Midwest Vanadium or from Hatch databases) to estimate mechanical equipment costs; if exact items are not available, costs have been scaled from similar units using the “6/10” rule:

$$\text{Capital cost 2/ Capital cost 1} = (\text{Capacity 2/Capacity 1})^{0.6}$$

Industry standard factors have been applied for installation to develop the total direct cost as follows:

- Mechanical installation: 50% for complex brownfield equipment;
- Civils and foundations: 10%;
- Structural supports: 20%;
- Associated electrical and instrumentation: 30%.

These costs are summed to give an estimated direct cost.

Project Indirects (Studies, EPCM, test-work, vendor support etc) are factored and summed to generate the total project cost direct and indirect.

A contingency of 30% has been added below the line. Owner’s cost is at the discretion of Midwest Vanadium, however 10% has been assumed for now.

		Australian Dollars X 1,000			
Plant Areas		Kiln Throughput t/h fresh feed (dry basis)	85	107.5	134
		FeV (t/m Contained V)	320	400	500
	<b>General</b>				
		Equipment changes required to ramp-up to stable operation	5,000	-	-
	<b>Kiln</b>				
		Testwork (vanadium recovery sighter tests)	100	-	-
		Reinstate Feed-end Seals	100	-	-
		Reinstate double-dump valves	20	-	-
		Replace pug mill with screw feeder	100	-	-
		Burner end Spyrometer	200	-	-
		Gas Supply Instrumentation	40	-	-
		Shell mounted thermocouples		90	-
		2° air thermocouples		50	-
		New Burner		500	-
		Feed End Spirals		200	-
		New Main Drive		-	1,700
		Flash/fluid bed dryer		-	3,200
		Single stage pre-heat (not required yet)		-	-
	<b>Leach</b>				
		Vat recycle pumps leach		-	30
		New spent calcine handling system		-	-
	<b>Desilication</b>				
		Agitators	-	-	-
		Standby cooler	40	-	-
		Filter Re-commissioning	120	-	-
	<b>AMV</b>				
		AMV Precipitation Agitators	-	-	-
		Filter recommissioning	120	-	-
		Re-instate cooler	10	-	-
		Duty/Standby cooler	40	-	-
		Thickener testwork		40	-
	<b>Reduction</b>				
		Repair existing cooler	20	-	-
		Engineering for New Cooler (Scoping study and new equipment)		50	200
		Discharge Seal Arrangement	60	-	-
		Blowpot under bin		100	-
		Cladding of Discharge Chute	50	-	-
		New Burners for reduction kilns		-	150
	<b>Furnace Feed</b>				
		V203 bin	100	-	-
		Purchase better scrap		-	-
		New scrap metal handling		-	1,000
	<b>Smelting</b>				
		Arena modelling	70	-	-
	<b>Crushing to packaging</b>				
		New Crushing Plant (Allowance)		-	-
	<b>Mechanical Equipment Installation</b>		8,140	940	6,080
	<b>Mechanical</b>		2,440	280	1,820
	<b>E&amp;I</b>		2,690	310	2,010
	<b>Civils</b>		250	-	610
	<b>Structural</b>		490	-	1,220
<b>Subtotal - Directs</b>			<b>14,010</b>	<b>1,520</b>	<b>11,730</b>
	<b>EPCM</b>	25% of (Capital + Installation)	3,500	380	2,930
	<b>Vendor Support</b>	Recommissioning Filters	240	-	-
	<b>Engineering studies</b>		170	90	200
	<b>Testwork</b>		100	-	-
<b>Subtotal - Indirects</b>			<b>4,020</b>	<b>470</b>	<b>3,130</b>
<b>Total - Directs and Indirects</b>			<b>18,030</b>	<b>2,000</b>	<b>14,870</b>
	<b>Owners costs @ 10%</b>		1,800	200	1,490
	<b>Contingency @ 30%</b>		5,410	600	4,460
<b>TOTAL</b>			<b>25,240</b>	<b>2,790</b>	<b>20,820</b>

Figure 7-1 – Cost Estimates



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## 7.2 Operating Cost

The operating cost has been developed from process design criteria, with a number of assumptions. These are:

- All cost data are provided by Midwest, and are assumed accurate for June 2014.
- With variation in plant feedrate, only the power, natural gas and reagent consumption rates change.
- The Process Design Criteria is accurate for scaling the reagent usage from.
- Mining and beneficiation costs are not included in the operating cost of the refinery.

From these assumptions, the operating cost for the plant can be estimated to  $\sim \pm 30\%$  for the two cases. This is presented in the table below.

	Lower Throughput	Higher Throughput
Vanadium Production (t/mth)	400	500
Total (\$/y)	46,927,395	54,314,515
Cost per unit of Vanadium (\$/t)	9,808	9,052

It is apparent in this table that unit vanadium cost is relatively independent of the plant throughput. As such, the reagents and power will have the largest contribution to the operating cost of the plant, as seen in the figures below.

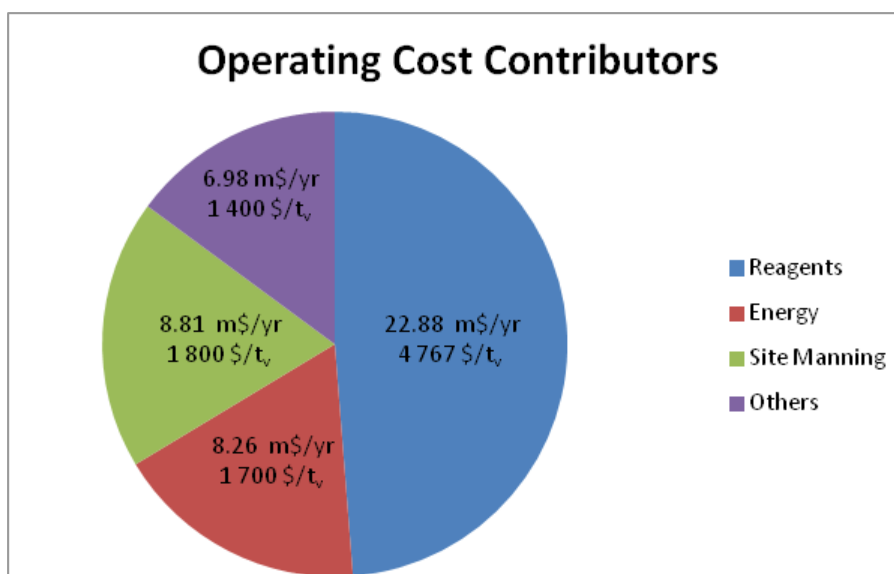


Figure 2: Contributors to the Operating Cost of the Refinery (400 t/mth<sub>v</sub>)

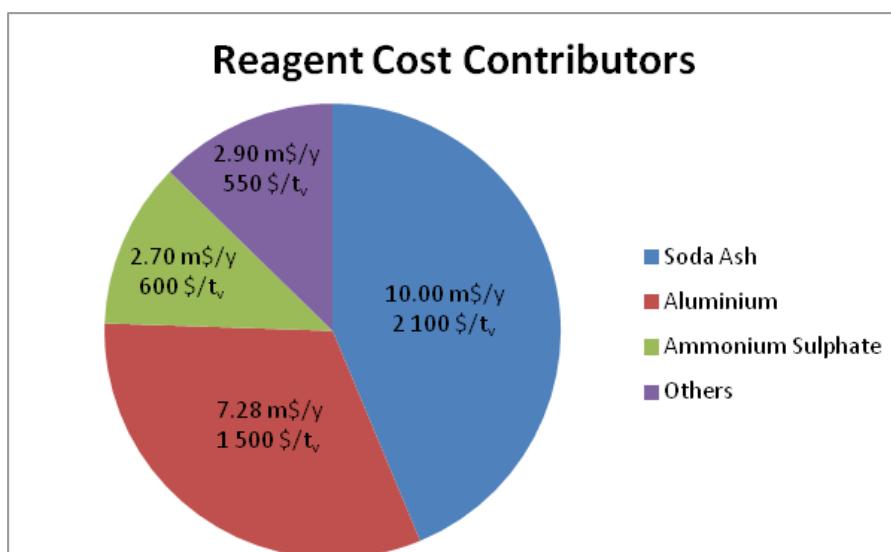


Figure 3: Contributors to the Reagent Costs of the Refinery (400 t/mth<sub>v</sub>)

As a result, to reduce the operating cost, it is recommended that the costs associated on the reagents and natural gas is focussed on. Of the reagents, the largest contributor is the soda ash cost, and this can be reduced in a number of ways. This includes substituting it with another sodium containing compound such as crystallising and recycling sodium sulphate from various process streams. This is practiced at some South African refineries; however there is a vanadium recovery penalty. The sensitivity of the operating cost to soda ash is shown in the figure below. Note that during ramp-up the plant was known to have over-dosed with sodium carbonate (in response to high silica content in the concentrate): lowering the soda rate to the PDC target of 3.9% significantly lowers the operating cost.

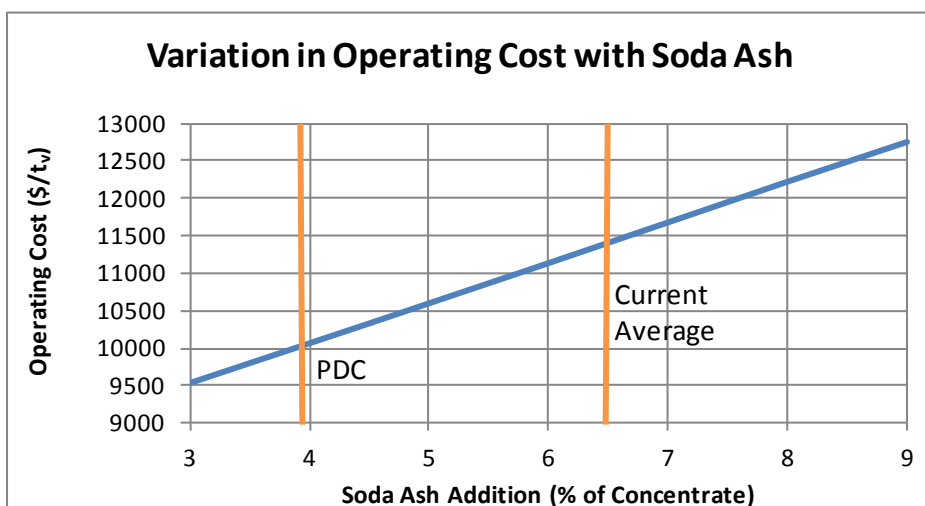


Figure 4: Sensitivity of the operating cost to the soda ash (400 t/mth<sub>v</sub>)

The other significant contributors to the operating cost include natural gas, aluminium sulphate, ammonium sulphate and aluminium. Reducing the costs associated with these reagents will also have a significant impact upon the operating cost.

# Appendix A

## Process Calculation

## PROCESS CALCULATION

This document contains information which is proprietary to Hatch and must be kept strictly confidential. Disclosure of this information to a third party or other use of such information is strictly prohibited, unless the prior written consent of Hatch is obtained.

Parameter	Units	Flowsheet Requirement	Installed Capacity	Value at 400 t/m V	Value at 500 t/m V
Magnetite Concentrate	Feed Rate	131	131	107	134
	Vanadium (% w/w)	0.71	0.71	0.71	0.71
	Silica (% w/w)	2.5	2.5	2.5	2.5
Recovery to AMV	% w/w	0.830	0.830	0.830	0.830
Recovery to FeV	% w/w	0.96	0.96	0.96	0.96
Availability	h/h	0.902	0.902	0.902	0.902
Capacity Utilisation	t/h per t/h	1	1	1	1
Annual V in VeV	t/a	5851	5851	4800	6000
Monthly	t/month	488	488	400	500
Feed Moisture	t/h water	11.4	11.4	9.3	11.7
Soda Ash	t/t fresh feed	0.039	0.039	0.039	0.039
Dust Recycle	t/t fresh feed	0.028	0.028	0.05	0.05
Total Kiln Feed	t/h (dry basis)	140	-	117	146
Total Kiln Feed	t/h (wet basis)	151	-	126	158
Kiln Discharge	t/h (dry basis)	136	272	112	140

Equipment	Units	Flowsheet Requirement	Installed Capacity	Value at 400 t/m V	Value at 500 t/m V	References
Leaching Vats	cycle time (hrs)	52.9	26	65	52	Memorandum, Calcine Removal, From: Blair Duncan, To: Ian Scott, 14/09/2007, and 6033-G-00-F-001, Process Design Criteria
Scrubber Discharge Pump Duty/Standby	Flowrate (m3/h)	70	91	6.38	7.97	Equipment Data Sheet - Scrubber Discharge Pump, 6033-D-25-M-013
Scrubber Discharge Tank	Residence Time (hrs)	0.86	-	1.05	0.84	Scrubber Discharge Tank Agitator, MC223/04/00 A
Quench Scrubber	Flowrate (m3/h)	23000	50000	18979	23723	Equipment Data Sheet - Quench Scrubber, 6033-D-25-M-069
Quench Scrubber Fan	Flowrate (m3/h)	22232	48330	18979	23723	Equipment Data Sheet - Quench Scrubber, 6033-D-25-M-069
Leach Vat Discharge Pumps	Flowrate (m3/h)	100	130	-	-	Equipment Data Sheet - Leach Vat Discharge Pumps, 6033-D-25-M-008
Leach Vat Recycle Pumps	Flowrate (m3/h)	300	300	246	307	Equipment Data Sheet - Leach Vat Recycle Pump, 6033-D-25-M-020
Slug Dose Tank Pumps	Flowrate (m3/h)	300	340	-	-	Equipment Data Sheet - Slug Dose Tank Pumps, 6033-D-25-M-003
Slug Dose Tanks	Volume (m3)	1200	1200	-	-	6033-G-00-F-001, Process Design Criteria
	Vat Fill Time (h)	4	4	-	-	-
Pregnant Liquor Storage Tank	Residence Time (hrs)	>25	82	100	80	Pregnant Liquor Storage Tank 25TNK505, Arrangement and Details, 6033-25-M-1085, and 6033-G-00-F-001, Process Design Criteria
Pregnant Liquor Transfer Pumps	Flowrate (m3/h)	28	35	23	29	Equipment Data Sheet - Pregnant Liquor Transfer Pumps, 6033-D-M-006
Leach Wash Pump	Flowrate (m3/h)	150	-	-	-	Equipment Data Sheet - Leach Wash Pump, 6033-D-25-M-002
Leach Wash Tank	Volume (m3)	1354	-	-	-	Leach Wash Tank 25TNK501, Arrangement and Details, 6033-25-M-1065
Leachate Tank Return Pump	Flowrate (m3/h)	100	-	-	-	General Arrangement Drawings, Blakers Unistream 65/20 Pumpset, L999.75, <a href="http://blakerspumps.com.au/downloads/BPE%20Unistream%20Brochure.pdf">http://blakerspumps.com.au/downloads/BPE%20Unistream%20Brochure.pdf</a>
Leachate Return Tank	Volume (m3)	-	-	-	-	-
Desilication Tanks	Residence Time (hrs)	18.12	0.5	22	18	REF13.24 - Desilication Test Work Results - Sept 2013
Dirty Preg Tank	Volume (m3)	320	-	-	-	6033-G-00-F-001, Process Design Criteria
Dirty Preg Tank Discharge Pump	Flowrate (m3/h)	14	22.4	11	14	Equipment Data Sheet - Dirty Preg Tank Discharge Pump
Desilication Feed Cooler	Slurry Flowrate (t/h)	15.5	23.28	13	16	Equipment Data Sheet - Desilication Heat Exchanger
Desilication Filter Feed Pump	Flowrate (m3/h)	33	35	-	-	Equipment Data Sheet - Filter Feed Pump, 6033-D-30-M-025
Desilication Wash Water Tank	Residence Time (hrs)	12	-	-	-	6033-G-00-F-001, Process Design Criteria
Desilication Wash Water Pump	Flowrate (m3/h)	-	-	-	-	-

## PROCESS CALCULATION

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Equipment	Units	Flowsheet Requirement	Installed Capacity	Value at 400 t/m V	Value at 500 t/m V	References
Desilication Filter Press	Flowrate (m3/h)	-	-	-	-	-
Clear Preg Storage Tank	Residence Time (hrs)	6	21.38	7	6	6033-G-00-F-001, Process Design Criteria
Precipitation Feed Pump	Flowrate (m3/h)	15	24	12	15	Equipment Data Sheet - Precipitation Feed Pump, 6033-D-30-M-028
Aluminium Sulphate Feed Pump	Flowrate (m3/h)	0.31	1.2	0.25	0.32	Equipment Data Sheet - Aluminium Sulphate Feed Pump, 6033-D-30-M-030
Sulphuric Acid Feed Pump	Flowrate (m3/h)	0.48	2	0.39	0.49	Equipment Data Sheet - Sulphuric Acid Feed Pump, 6033-D-30-M-029
Sulphuric Acid Storage Tank	Residence Time (hrs)	211	-	-	-	Sulphuric Acid Storage Tank 30TNK507, Arrangement and Details, 6033-30-M-1059
Aluminium Sulphate Storage Tank	Residence Time (hrs)	192	-	-	-	Aluminium Sulphate Storage Tank 30TNK508, Arrangement and Details, 6033-30-M-1065
Vanadium Precipitation	Residence Time (hrs)	12	2	15	12	REF13.22 - AMV Precipitation Investigation - SWL
Pregnant Liquor Cooler	Flowrate (m3/h)	16.25	19.52	13	17	Equipment Data Sheet- Pregnant Liquor Cooler, 6033-D-35-M-067
AMV Product Hopper	-	-	-	-	-	-
AMV Product Pump	Flowrate (m3/h)	20.2	30.2	17	21	Equipment Data Sheet- AMV Product Pump, 6033-D-35-M-182
AMV Thickener Feed Box	-	-	-	-	-	-
AMV Thickener	Feed Capacity (t/h)	1.79	3.46	1.47	1.83	6033-G-00-F-004, Mass Balance
AMV Thickener Underflow Pump	Flowrate (m3/h)	3.71	-	3.04	3.80	6033-G-00-F-004, Mass Balance
Barren Solution OF Hopper	-	-	-	-	-	-
Barren Solution OF Pump	Flowrate (m3/h)	23.26	-	19	24	6033-G-00-F-004, Mass Balance
AMV Wash Water Tank	-	-	-	-	-	-
Seal and Gland Water Pump	Flowrate (m3/h)	4.5	-	-	-	Equipment Data Sheet - Seal and Gland Water Pump, 6033-D-35-M-195
Seal and Gland Water Return Pump	Flowrate (m3/h)	5.5	-	-	-	Equipment Data Sheet - Seal and Gland Water Return Pump, 6033-D-35-M-196
Seal and Gland Water Return Hopper	Residence Time (hrs)	3	-	-	-	6033-G-00-F-001, Process Design Criteria
Belt Wash Return Sump	-	-	-	-	-	-
AMV Belt Wash Return Pump	Flowrate (m3/h)	-	-	-	-	-
Barren Storage Tank	Volume (m3)	150	-	-	-	6033-G-00-F-001, Process Design Criteria
Barren Solution Purge Pump	Flowrate (m3/h)	0.59	8	0.48	0.60	Equipment Data Sheet - Barren Solution Purge Pump, 6033-D-35-M-038
AMV Filter Cloth/Belt Wash Pump	Flowrate (m3/h)	-	-	-	-	-
Amsul Dilution Pump	Flowrate (m3/h)	4.28	15	3.5	4.4	Equipment Data Sheet - Amsul Dilution Pump, 6033-D-35-M-119
Amsul Mixing Tank 1	Volume (m3)	104	-	-	-	Amsul Mixing #1 Tank 35TNK511, Arrangement and Details, 6033-35-M-1017
Amsul Mixing Tank 2	Volume (m3)	104	-	-	-	Amsul Mixing #1 Tank 35TNK511, Arrangement and Details, 6033-35-M-1017
Amsul Liquor Pump	Flowrate (m3/h)	6.8	9.8	5.6	7.0	Equipment Data Sheet - Amsul Liquor Pump, 6033-D-35-M-042
Amsul Screw Feeder	Flowrate (t/h)	25	-	-	-	Equipment Data Sheet - Amsul Solids Feed System, 6033-D-35-M-092
Scrubber Circulation Pump	Flowrate (m3/h)	480	-	-	-	Equipment Data Sheet - Scrubber Circulation Pump, 6033-D-19-M-178
Scrubber Circulation Tank	-	-	-	-	-	-
Scrubber Discharge Pump	Flowrate (m3/h)	14	21	-	-	Equipment Data Sheet - Scrubber Discharge Pump, 6033-D-19-M-177
Scrubber Tower	Volumetric flowrate (m3/h)	189144	220342	131237	164047	Kiln Off Gas Wet Scrubber, 6033-W-19-M-023
ID Fan flowrate	Volumetric flowrate (m3/s)	54.2	-	37	47	C24-086 Rev 1 Performance Curve
Flash Dryer	Solids (t/h)	1.87	2	1.53	1.91	Windimurra Vanadium, Flash Dryer Operating & Maintenance Manual

# Appendix B

## Model Design

## Midwest Vanadium Windimurra Debottlenecking Study

### Model Design Criteria

for

### Salt Roast Kiln Analysis

C	25-Jun-2014	Final	R. Haywood	W Taylor	R. Haywood	C Crowe	Not Required
B	06-Jun-2014	Internal Review	R. Haywood	W Taylor	R. Haywood	C Crowe	Not Required
A	21-May-2014	Draft	R. Haywood	W Taylor	R. Haywood	C Crowe	Not Required
REV	Date	Description of Issue	Originator	CHKD	Discipline Lead	Project Manager	Client

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PROJECT NUMBER: H346368  
PROJECT: Windimurra Debottlenecking Study  
AREA: -  
AREA DESCRIPTION: -  
CALCULATION: -  
DOCUMENT NUMBER: H346958-0000-00-122-0001  
BY: R. Haywood  
DATE: 25-Jun-2014

#### COVER SHEET

**Equipment name** Salt Roast Kiln Analysis  
Equipment number -

#### Revision History

Revision number :	A
Revision date :	21-May-2014
Reviewed by :	Ross Haywood
Review comments :	<a href="#">Wes Taylor</a>
Review comments incorporated :	<a href="#">Ross Haywood</a>
Revision number :	B
Revision date :	05-Jun-2014
Reviewed by :	C. Crowe
Review comments :	Nil
Review comments incorporated :	
Revision number :	C
Revision date :	25-Jun-2014
Reviewed by :	Client
Review comments :	<a href="#">Wes Taylor</a>
Review comments incorporated :	<a href="#">Wes Taylor</a>



**CLIENT :** Midwest Vanadium  
**PROJECT :** Salt Roast Kiln Analysis  
**PLANT AREA :**  
**DOCUMENT NUMBER :** H346958-0000-00-122-0001

**MODEL DESIGN CRITERIA**

PROJECT NUMBER: H346368  
 BY: Ross Haywood  
 CHECKED: Wesley Taylor

	UNITS	VALUE	REF/CONF.	SOURCE	NOTES	REV
<p> <b>The following reference codes are used in this document :</b>            Assumed / Indicative / Estimated            Consultants Data            Process Design Basis &amp; Capacity Basis            Engineering calculation            Hatch Recommendation            Literature            Mass Balance            Regulatory Standards and Codes            Standard Industry Practice            Testwork            Vendor Originated Data            Data from MVPL         </p> <p> <b>The following codes are used next to the reference code to specify the confidence level of the data :</b>            A Execution / Implementation Phase            C Feasibility stage            D Definition Phase / Pre-feasibility stage            E Preliminary data            H Scoping level / Sighter tests            L            M            R            S            T            V            Z         </p>						

**Model Input Data**

<b>Physical Constants</b>						
Temperature @ Normal Conditions	K	273.15	S			A
Pressure @ Normal Conditions	kPa(a)	101.325	S			A
<b>Geometry</b>						
<b>Kiln</b>						
Shell Inside Diameter	m	4.750	V	Drawings 3516-20-VM-001 and -V-002.	Listed in drawing title only	A
Length	m	102.030	V	Drawings 3516-20-VM-001 and -V-002.	Feed end to centerline of cooler inlets	A
Inclination	%	2.0	V	Drawings 3516-20-VM-001 and -V-002.		A
Refractory Thickness	mm	220	Z	"Windimurra Project Data and info, as 23-5-06" document	Original lining 180mm feed end, then 220.	A
<b>Burner</b>						
Operating Insertion Length	m	4.5	V	FLSmith kiln audit report, 26/10/2013 - 07/11/2013	Shown in dam diagram.	A
Full Insertion length	m	5.7	V	Drawing 3516-20-VM-067, length scaled as dimension not listed.	Insertion past centerline of cooler inlets	A
Flame Length	m	15.0	Z			A
<b>Coolers</b>						
Number of satellite coolers	#	9.0	Z / V	Drawing 6033-20-M-1110 Rev 1 / FFE Equip. Spec. 3516/SC-01		A
Cooler Inside Diameter	m	2.1	Z / V	Drawing 6033-20-M-1110 Rev 1 / FFE Equip. Spec. 3516/SC-01		A
Cooler Length	m	15.0	Z / V	Drawing 6033-20-M-1110 Rev 1 / FFE Equip. Spec. 3516/SC-01	From centerline of inlets to discharge end	A



**MODEL DESIGN CRITERIA**

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**PLANT AREA :**  
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PROJECT NUMBER:  
H346368

BY: Ross Haywood

CHECKED: Wesley Taylor

	UNITS	VALUE	REF/CONF.	SOURCE	NOTES	REV
<b>Mechanical Specifications</b>						
<b>Kiln Drives</b>						
Number of Main Drives	#	2.0	V	FFE Equip. Spec. 3516/SC-01		A
Drive Power (per Drive)	kW	500	V	FFE Equip. Spec. 3516/SC-01		A
Girth Gear to Pinion Ratio	:1	5.642	V	FFE Drawing 1.151917	Calculated from pitch diameters	A
Reducer ratio	:1	285.91	V	Santasalo Technical Specification B071830 - 22.1.1999		A
Motor Synchronous Speed (100% speed)	rpm	985	Z	Photo of nameplate from site - 6 pole motor		A
Kiln Speed at Motor 100% Speed	rpm	0.611	E			A
Design Maximum Speed of Reducer (in terms of Kiln Speed)	rpm	1.0	V	Santasalo Technical Specification B071830 - 22.1.1999		A
Motor Speed at Maximum Kiln Speed	%	164%	E		Motor and VSD needs to handle this	
Reducer Rated Power, Normal / Maximum	kW	375 / 750	V	Santasalo Technical Specification B071830 - 22.1.1999		A
<b>Kiln / Cooler Internals</b>						
Kiln		Refractory Only	Z	"Windimurra Project Data and info, as 23-5-06" document	HY11 / Harchem 1550 in hot zone?	A
Kiln Refractory Thermal Conductivity	W/m.K	?				
Kiln Dams		None	Z	"Windimurra Project Data and info, as 23-5-06" document		A
Cooler		Cast refractory elbows, 1/3 Bricks, 2/3 steel with lifters	Z	"Windimurra Project Data and info, as 23-5-06" document		A
<b>Material / Stream Properties</b>						
<b>Kiln Feed</b>						
Bulk Density (wet)	kg/m <sup>3</sup>	2,600	D	320900-00-F-22-0001 Hatch Design Criteria and Basis for Design	Unclear whether this is wet or dry	A
Moisture at which Bulk Density was Tested	%w/w	6.6%	E		Calculated from ore and	A
Dynamic Angle of Repose	degrees	37.0	A		Only static repose values were found	A
Bed Target Temperature	deg C	1,200	D	320900-00-G-24-0001 Hatch Process Design Criteria from 2006		A
Bed Time at Temperature	hr	1.0	D	320900-00-G-24-0001 Hatch Process Design Criteria from 2006		A
Kiln Dust Loss - For bed profile model	%feed	0.0	A		Assumed for bed profile model only, effect is small. Bed profile past end of dust loss location is not affected.	B
Kiln Dust Loss - thermochemical model	%feed	Variable	C		For the thermochemical model the dust loss is calculated based on bed properties and freeboard gas velocity	B
<b>Kiln Discharge / Cooler Inlet</b>						
Bulk Density (dry)	kg/m <sup>3</sup>	2,430	E		Calculated from wet density and moisture	A
Dynamic Angle of Repose	degrees	37.0	A		Only static repose values were found	A



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<b>Cooler Discharge</b>						
Calcine temperature	deg C	425.0	D	320900-00-G-24-0001 Hatch Process Design Criteria from 2006		A
<b>Analysis input - Mechanical</b>						
<b>85 tph nominal throughput (base case)</b>						
Wet Feed Rate Including Dust Recycle	tph	84.800	E	Preliminary Process Estimate		A
Kiln Speed	rpm	0.720	E	All Based on Hatch Proprietary Kiln model.		A
Kiln Fill Factor - average	%	12.5	E			A
Kiln Residence Time	hr	5.7	E			A
Average Fill height (note fill level varies from feed to discharge)	mm	780.00	E			A
<b>Throughput to achieve nominal 400 tpm V</b>						
Wet Feed Rate Including Dust Recycle	tph	135.000	E	Preliminary Process Estimate		A
Kiln Speed	rpm	0.90	E	All Based on Hatch Proprietary Kiln model.		A
Kiln Fill Factor - average	%	16.10	E			A
Kiln Residence Time	hr	4.61	E			A
Average Fill height (note fill level varies from feed to discharge)	mm	930.00	E			
<b>Throughput to achieve nominal 500 tpm V</b>						
Wet Feed Rate Including Dust Recycle	tph	170.000	E	Preliminary Process Estimate		A
Kiln Speed	rpm	1.15	E	All Based on Hatch Proprietary Kiln model.		A
Kiln Fill Factor - average	%	15.90	E			A
Kiln Residence Time	hr	3.600	E			A
Average Fill height (note fill level varies from feed to discharge)	mm	930	E			A



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<b>Concentrate Composition (Dry)</b>						
Fe3O4	(%w/w)	47.0	Z	Report by Promet Engineers, June 2006, C5241-RP-002 Rev 0 also generally in agreement with original Proteus design criteria (6033-G-00-F-001) and original Signet data sheet D-20-F-01. Note there is some discrepancy in the relative quantities of Hematite and Magnetite, but the overall quantity of Fe-oxides is relatively consistent.		A
Fe2O3	(%w/w)	23.0	Z			A
SiO2	(%w/w)	2.0	Z			A
TiO2	(%w/w)	18.0	Z			A
V2O5	(%w/w)	1.0	Z			A
Al2O3	(%w/w)	Balance	E			A
<b>Feed to Kiln</b>						
Free Moisture	(%w/w)	6.3	Z	Plant Data for December 2013. Provided by Flor Cabalteja, 16 May 2014		A
Soda Ash Addition Ratio based on Dry concentrate feed	(%w/w)	3.9	Z	Throughput estimate (this study). Also in original PDC		A
<b>Fuel Gas Composition</b>						
CH4	(%v/v)	0.904	R	Australian Standard for Natural Gas composition.		A
CO2	(%v/v)	.031	R			A
N2	(%v/v)	Balance	R			A
Density	kg/Nm3	0.79	R			A
<b>Air Supply</b>						
Primary Air Temperature	deg C	25.0	A			A
Vendor Specified Secondary Air Temperature from Coolers	deg C	625.0	D	320900-00-G-24-0001 Hatch Process Design Criteria from 2006		A
Secondary Air Temperature (used in model)	deg C	575.0	H	Actual Value used in model to best fit existing plant data.		A



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	UNITS	VALUE	REF/CONF.	SOURCE	NOTES	REV
<b>Modelling Assumptions</b>						B
<p>1D model: Both the bed and gas phase are assumed to be single valued (Temperature and composition) at each axial slice through the kiln</p> <p>1D model: Refractory and shell temperatures also assumed to be single valued at each axial slice. External heat loss to atmosphere at 25C</p> <p>Heat transfer occurs by radiation, convection and conduction from Gas to Wall, Gas to Bed and Wall to bed. A grey gas enclosure model is used for radiation heat transfer, standard correlations are used for convection.</p> <p>The gas phase is assumed to be in equilibrium, calculated by Gibbs free energy minimization.</p> <p>The gas-phase combustion has been specified with a finite length (ie the flame length) , the actual heat release over the flame length specified using a Beta function distribution.</p> <p>Solids specific heats calculated according to correlations/ testwork of Barin and Knacke (for the most part), some taken from HSC chemistry.</p> <p>Free moisture vapourisation assumed to occur according to an Arrhenius rate expression.</p> <p>Magnetite oxidation assumed to occur at rate proportional to the oxygen partial pressure in the freeboard gas but only at bed temperatures greater than 525 C</p> <p>Formation of Sodium metavanadate assumed to occur at a fixed overall rate at bed temperatures greater than 1150 C. The overall rate is reduced according to the law of mass action to account for the reduced contact as solid-phase reactants are depleted from the bed.</p>						

# Appendix C

## Operating Cost Calculation



Safety • Quality • Sustainability • Innovation

## PROCESS CALCULATION

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PARAMETER	VALUE	UNITS	SOURCE	REMARKS	REV
<b>Consumable Costs</b>					
<b>Reagents</b>					
Soda Ash	302	\$/tonne			A
Sulphuric Acid	214	\$/tonne			A
Ammonium Sulphate	198	\$/tonne			A
Aluminum Sulphate	286	\$/tonne			A
Scrap Steel	700	\$/tonne			A
Aluminium	2445	\$/tonne			A
Lime	205	\$/tonne			A
<b>Site Manning</b>					
Production Operator	107000	\$/annum.unit			A
Production Supervisor	130000	\$/annum.unit			A
Maintenance - Fitter	120000	\$/annum.unit			A
Maintenance - Planner	140000	\$/annum.unit			A
Electrician	122000	\$/annum.unit			A
Technical (Metallurgist)	150000	\$/annum.unit			A
Engineering Team (1 elec, 1 mech, 1 reliability)	450000	\$/annum.unit			A
Engineering Manager	220000	\$/annum.unit			A
Refinery Manager	220000	\$/annum.unit			A
Assistant Manager	150000	\$/annum.unit			A
Maintenance Manager	220000	\$/annum.unit			A
Technical Services Manager	220000	\$/annum.unit			A
HSE Manager	150000	\$/annum.unit			A
Envtl Health and Safety	105000	\$/annum.unit			A
Commercial - Manager	110000	\$/annum.unit			A
Commercial - Stores	95000	\$/annum.unit			A
Human Resources Manager	110000	\$/annum.unit			A
Human Resources	60000	\$/annum.unit			A
Admin Assistance	45000	\$/annum.unit			A
<b>Metallurgical Analysis</b>					
Lab samples	15.5	\$/sample			A
Number of samples/day	92	samples/day			A
Additional samples for troubleshooting	5	samples/day			A
<b>Power</b>					
Electricity Cost Estimate	13	cents/kWh			A
Natural Gas Cost	5	\$/GJ			A
Natural Gas Transport Cost for 6TJ/day	3	\$/day			A
<b>Research and Development</b>					
Consultant Fees	250000	\$/annum			A
External testwork	250000	\$/annum.consult			A
<b>IT Costs</b>					
Communications Link Infrastructure	168000	\$/annum			A
Printers (assume 4 large, 4 small)	3360	\$/annum			A
ERP software	16000	\$/annum			A

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PARAMETER	VALUE	UNITS	SOURCE	REMARKS	REV
Historian software	16000	\$/annum			A
Production Accounting Software	10000	\$/annum			A
Lab Management Software	6000	\$/annum			A
Site Management Software	4000	\$/annum			A
Document Management Software	4000	\$/annum			A
Business Reporting Software	6000	\$/annum			A
Phone Hardware (assume 100)	4800	\$/annum			A
Office Software (assume 250 users)	60000	\$/annum			A
IT administration / support labour	120000	\$/annum			A
<b>Health, Safety and Environment</b>					
Training Allowance	7	% of manning cost			A
Pest Control	50000	\$/annum			A
Emergency Services	50000	\$/annum			A
Medical Centre Costs	50000	\$/annum			A
<b>Miscellaneous</b>					
Camp Cost	65	\$/person/day			A
Travel Cost	300	\$/person/flight			A
General Admin Cost	2400000	\$/annum			A
<b>Reagent Addition Rates</b>					
Soda Ash	0.039	t/tore	Process Design Criteria		A
Sulphuric Acid	0.0054	t/tore	Process Design Criteria		A
Ammonium Sulphate	0.0160	t/tore	Process Design Criteria		A
Aluminum Sulphate	0.0037	t/tore	Process Design Criteria		A
Scrap Steel	0.2240	t/TV	Process Design Criteria		A
Aluminium	0.6202	t/TV	Process Design Criteria		A
Lime	0.2713	t/TV	Process Design Criteria		A
<b>Operating Costs</b>					
	Annual Cost (\$/yr)				
<b>Reagents</b>	400 t/mthv	500 t/mthv			
Soda Ash	10,002,467	12,468,191	\$/yr		A
Sulphuric Acid	975,498	1,215,969	\$/yr		A
Ammonium Sulphate	2,695,558	3,360,045	\$/yr		A
Aluminum Sulphate	908,503	1,132,460	\$/yr		A
Scrap Steel	752,574	940,718	\$/yr		A
Aluminium	7,279,208	9,099,010	\$/yr		A
Lime	266,992	333,741	\$/yr		A
<b>Total</b>	<b>22,880,800</b>	<b>28,550,133</b>	<b>\$/yr</b>		<b>A</b>
<b>Site Manning</b>					
Production Operator	2,996,000	2,996,000	\$/yr		A
Production Supervisor	260,000	260,000	\$/yr		A
Maintenance - Fitter	2,640,000	2,640,000	\$/yr		A
Maintenance - Planner	140,000	140,000	\$/yr		A
Electrician	488,000	488,000	\$/yr		A
Technical (Metallurgist)	300,000	300,000	\$/yr		A
Engineering Team (1 elec, 1 mech, 1 reliability)	450,000	450,000	\$/yr		A

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PARAMETER	VALUE	UNITS	SOURCE	REMARKS	REV
	Annual Cost (\$/yr)				
	400 t/mthv	500 t/mthv			
Engineering Manager	110,000	110,000	\$/yr		A
Refinery Manager	220,000	220,000	\$/yr		A
Assistant Manager	150,000	150,000	\$/yr		A
Maintenance Manager	220,000	220,000	\$/yr		A
Technical Services Manager	110,000	110,000	\$/yr		A
HSE Manager	75,000	75,000	\$/yr		A
Envtl Health and Safety	52,500	52,500	\$/yr		A
Commercial - Manager	55,000	55,000	\$/yr		A
Commercial - Stores	380,000	380,000	\$/yr		A
Human Resources Manager	55,000	55,000	\$/yr		A
Human Resources	60,000	60,000	\$/yr		A
Admin Assistance	45,000	45,000	\$/yr		A
<b>Total</b>	<b>8,806,500</b>	<b>8,806,500</b>	<b>\$/yr</b>		<b>A</b>
<b>Metallurgical Analysis</b>					
Lab samples	548,778	685,972	\$/yr		A
<b>Power</b>					
Electricity Cost Estimate	1,844,856	1,844,856	\$/yr		A
Natural Gas Cost	6,411,838	7,992,430	\$/yr		A
Natural Gas Transport Cost for 6TJ/day	1,095	1,095	\$/yr		A
<b>Total</b>	<b>8,257,789</b>	<b>9,838,381</b>	<b>\$/yr</b>		<b>A</b>
<b>Research and Development</b>					
Consultant Fees	250,000	250,000	\$/yr		A
External testwork	250,000	250,000	\$/yr		A
<b>Total</b>	<b>500,000</b>	<b>500,000</b>	<b>\$/yr</b>		<b>A</b>
<b>IT Costs</b>					
Communications Link Infrastructure	168,000	168,000	\$/yr		A
Printers (assume 4 large, 4 small)	3,360	3,360	\$/yr		A
ERP software	16,000	16,000	\$/yr		A
Historian software	16,000	16,000	\$/yr		A
Production Accounting Software	10,000	10,000	\$/yr		A
Lab Management Software	6,000	6,000	\$/yr		A
Site Management Software	4,000	4,000	\$/yr		A
Document Management Software	4,000	4,000	\$/yr		A
Business Reporting Software	6,000	6,000	\$/yr		A
Phone Hardware (assume 100)	4,800	4,800	\$/yr		A
Office Software (assume 250 users)	60,000	60,000	\$/yr		A
IT administration / support labour	120,000	120,000	\$/yr		A
<b>Total</b>	<b>418,160</b>	<b>418,160</b>	<b>\$/yr</b>		<b>A</b>
<b>Health, Safety and Environment</b>					
Training Allowance	616,455	616,455	\$/yr		A
Pest Control	50,000	50,000	\$/yr		A
Emergency Services	50,000	50,000	\$/yr		A
Medical Centre Costs	50,000	50,000	\$/yr		A
<b>Total</b>	<b>766,455</b>	<b>766,455</b>	<b>\$/yr</b>		<b>A</b>

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PARAMETER	VALUE	UNITS	SOURCE	REMARKS	REV
	Annual Cost (\$/yr)				
	400 t/mthv	500 t/mthv			
<b>Miscellaneous</b>					
Camp Cost	976,114	976,114	\$/yr		A
Travel Cost	1,372,800	1,372,800	\$/yr		A
General Admin Cost	2,400,000	2,400,000	\$/yr		A
<b>Total</b>	<b>4,748,914</b>	<b>4,748,914</b>	<b>\$/yr</b>		<b>A</b>
<b>Total</b>					
Total Operating Cost	46,927,395	54,314,515	\$/yr		A