Programme Area: Carbon Capture and Storage

Project: Mineralisation

Title: Carbon Capture and Sequestration by Mineralisation (CCSM) Stage 2b Techno-Economic Assessment (TEA) Report

Abstract:

This report describes the high level techno-economic assessment of an ammonia based process for the mineralisation of CO2. The basic process chemistry was described in a patent by Pundsack (1967, ref. 5). The overall process requirement is to remove 80% of the CO2 content in the flue gas arising from a 300 MW coal based power plant. The initial heat and mass balance was developed using the design basis as given in ref. 1, albeit with modified operating parameters for the ore leaching step. The bisulphate solution concentration and solids to liquids ratio were increased to 2.67M and 125 kg/m3 respectively. The design basis was latterly further modified with the aim of reducing the plant energy usage and equipment sizes. The revised design basis is summarized in Appendix 1. The heat and mass balance using this revised basis is described here, along with the overall energy requirements and estimated capital cost. The impact of the modified design basis is also given. The ETI is grateful for the contributions made by all participants of the Mineralisation project; Perkins Engines Company Limited, Shell Global Solutions International B.V., Natural Environment Research Council as represented by the British Geological Survey, and the University of Nottingham.

Context:

CCS by mineralisation has been identified as a promising additional method of sequestering CO2 emissions. Minerals and CO2 can react together to permanently store CO2 as a solid carbonate product, which can then be safely stored, used as an aggregate or turned into useful end products such as bricks or filler for concrete. This £1m project, launched in May 2010 carried out a detailed study of the availability and distribution of suitable minerals across the UK along with studying the technologies that could be used to economically capture and store CO2 emissions. The project consortium involved Caterpillar, BGS and the University of Nottingham. The objective was to investigate the potential for CCS Mineralisation to mitigate at least 2% of current UK CO2 emissions and 2% of worldwide emissions over a 100- year period. The project has found that there is an abundance of suitable minerals available in the UK and worldwide to meet these mitigation targets. However, challenges remain to make the capture process economically attractive and to reduce its energy use. Significant niche opportunities exist where waste materials are used as feedstock and/or the process produces value-added products, but markets would not be at the level required to meet the mitigation targets.

Carbon Capture and Sequestration by Mineralisation (CCSM) Stage 2b Techno-Economic Assessment (TEA) Report

CCSM Programme Deliverable 3.3

Stage Gate 2b Report

1st June 2012

ENERGY TECHNOLOGIES INSTITUTE

CCSM PROGRAMME

STAGE 2B TEA REPORT

Carbon Capture and Sequestration by Mineralisation (CCSM) Stage 2b Techno-Economic Assessment Report

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Loughborough Energy Technologies Institute 2012

Contents

FIGURES

TABLES

1 Introduction

This report describes the high level techno-economic assessment of an ammonia based process for the mineralisation of CO_2. The basic process chemistry was described in a patent by Pundsack (1967, ref. 5). The process chemistry is also described in the recent patent (6) granted to Wang and Maroto-Valer (University of Nottingham, 2011), and in other published papers (18). The process evaluation described in this report is based on this published material and on other research data supplied to the project by the University of Nottingham.

The overall process requirement is to remove 80% of the CO_2 content in the flue gas arising from a 300 MW coal based power plant. The initial heat and mass balance was developed using the design basis as given in ref. 1, albeit with modified operating parameters for the ore leaching step. The bisulphate solution concentration and solids to liquids ratio were increased to 2.67M and 125 kg/m^3 respectively. The design basis was latterly further modified with the aim of reducing the plant energy usage and equipment sizes. The revised design basis is summarized in Appendix 1. The heat and mass balance using this revised basis is described here, along with the overall energy requirements and estimated capital cost. The impact of the modified design basis is also given.

2 Process Overview

A schematic of the overall process is shown in Figure 1; the process flow diagrams (PFDs) are given in Appendix 3 (figures 7-9), along with the stream lists in Appendix 5.

Due to the large mass throughputs, most of the process equipment will consist of multiple plants operating in parallel. With the exception of the ore-leaching step, the PFDs show only a single main plant item, and the process description will generally refer to a single unit. Note that where aqueous solutions are concerned the composition is given in terms of equivalent molecular species rather than the actual ionic species that are present. Physical properties data was taken from a number of sources, principally from refs. 2-4. The overall mass balance model uses simplified blocks for some of the major steps, with more detailed models to look at performance of individual unit operations (ammonia absorption, flue gas scrubbing, carbonate solution recovery).

The process is based on the experimental extraction data for a specific serpentine sample, ET16, with the composition as given in ref. 1 and Appendix 1. The experimental data shows that different rocks of the same general mineral type and similar composition can exhibit significantly different leaching characteristics. ET16 is of a type with favourable characteristics.

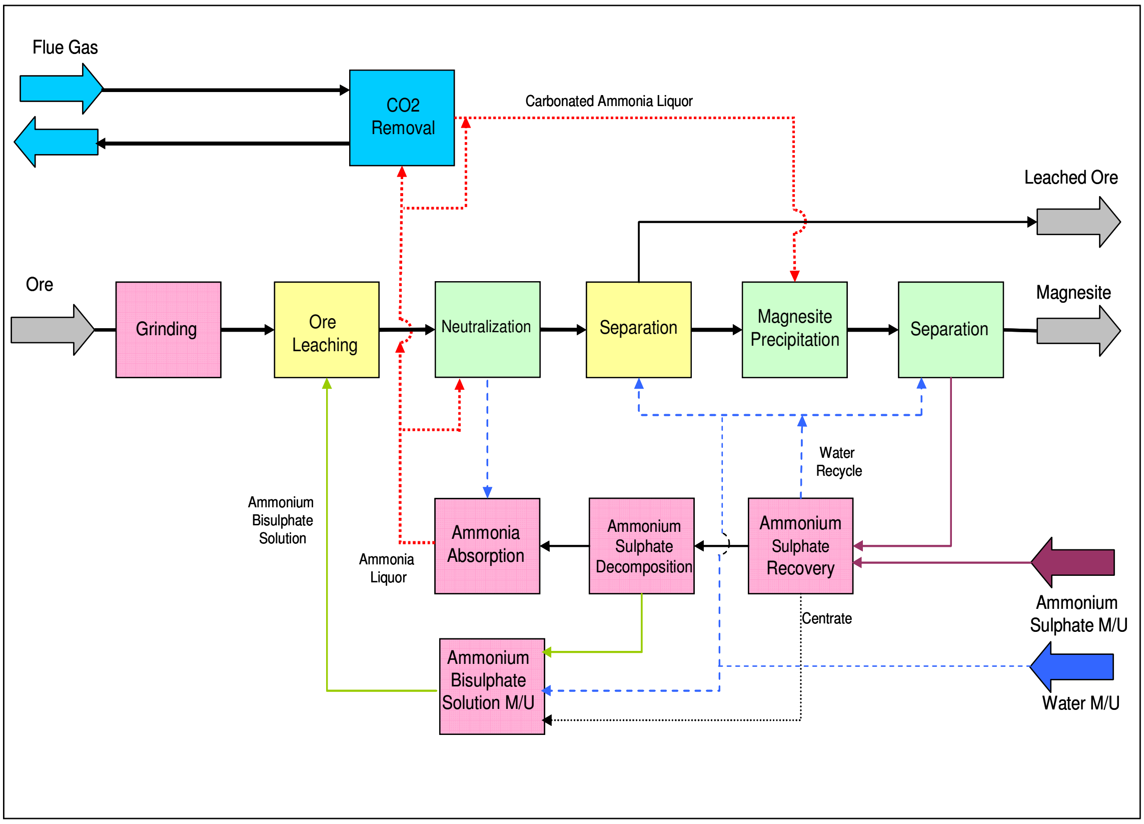

Figure 1. Process Block Diagram.

3 Process Description

Additional information on the design issues relating to a number of the process steps is given in Appendix 2.

3.1 Leaching and Separation of Leached Ore

625 Te/h of milled ore, having a nominal size range of 150-300 µm, is transferred from onsite storage to the Slurry Make-Up Vessel. The ore is taken to be dry. The make-up vessel is operated batch wise to generate a slurry of the required solids to liquids ratio by mixing the ore with 5036 Te/h of recycled ammonium bisulphate (NH_4HSO_4) solution. The bisulphate solution is acidic.

The loading of ore and solution is managed by the use of load cells on the make-up vessel. The slurry is transferred to the first of a series of extraction vessels, which operate at 127.5°C and low pressure (3 bara). The feed slurry is preheated to 122.5°C by interchange with the reactor effluent stream.

The ore composition is based on that of a lizardite sample (ET16) and is taken to be composed of the species serpentine, silica and magnetite, with c. 90% serpentine, 8% Fe_3O_4 and 2% silica (1). The feed stream has solids to liquids ratio of 152.5 kg/m^3, with an NH_4HSO_4 concentration equivalent to 2.7M (based on bisulphate and water only). The feed solution contains about 3% w/w of ammonium sulphate ($(NH_4)_2SO_4$), along with around 2% w/w of magnesium sulphate ($MgSO_4$).

The bisulphate reacts with the ore, solubilising both magnesium and iron. The reactions are both exothermic, and are given by:

Mg extraction:

$Mg_3Si_2O_5(OH)_4(s) + 6NH_4HSO_4(aq) =$

$3MgSO_4(aq) + 2SiO_2(s) + 5H_2O(l) + 3(NH_4)_2SO_4(aq)$

$\Delta H_r^0 = -424.6$ MJ/kmol

where :

ΔH_r^0 = heat of reaction at standard conditions (25C, 1 ata), for the components in the designated states

(s) = solid
(aq) = aqueous solution
(l) = liquid

Fe extraction :

$Fe_3O_4(s) + 8NH_4HSO_4(aq) = FeSO_4(aq) + Fe_2(SO4)_3(aq) + 4H_2O(l) + 4(NH_4)_2SO_4(aq)$

$\Delta H_r^0 = -390.3$ MJ/kmol

The heat of reaction is recovered by cooling coils within the reactor vessels to maintain the required operating temperature. The overall heat recovery from the extraction stage is 160 MW.

The total reactor liquid volume is 4350 m^3. The volume is taken to be too large for a single vessel, and consequently the reactor system is comprised of a number of agitated vessels operating in series. This approximates to the batch reaction system used for the leaching experimental work and permits direct application of the research results.

The leaching process is thought to proceed via a reaction front moving into the ore particles, leaving silica in-situ, such that the initial particle size is retained and leaving a porous matrix from which the Mg and Fe have been removed. The heat and mass balance is based on 75% removal of Mg and 47% removal of Fe, and is taken to require 60 minutes residence time. For the chosen reactor operating parameters the utilization of the bisulphate is 90%. The reaction temperature used for design is higher than the maximum temperature of 100°C in the experimental work available at the time of generating the process design. The reaction temperature was increased from the original figure of 100C (1) as part of the package of changes to the process design basis. The performance of the extraction step at the operating temperature is based on extrapolation of the research data, as described in Appendix 2 (‘Reaction Kinetics’).

The leached ore slurry (5662 Te/h) is cooled to 95°C. 120 Te/h of 28% w/w ammonia liquor, recycled from the downstream process, is mixed with the leachate to neutralize the excess bisulphate. At this point, the Fe compounds are assumed to be precipitated as the hydroxides. The leachate is cooled and neutralized upstream of the belt filters to permit operation of the filters at atmospheric pressure and to reduce possible problems associated with selection of the materials of construction.

The neutralized slurry passes to a vacuum belt filter where the leached ore is separated from the spent solution. For mass balance purposes it is assumed that the Fe precipitates are not recovered at this stage. The filter has multiple stages of counter-current washing to minimize loss of solution chemicals. The water for washing is recycled from the upstream process. The washed leached ore is cooled in a bulk flow exchanger and discharged wet, with a residual liquor content of 10-20% w/w.

3.2 Magnesite Precipitate and Recovery

The filtered leachate contains 19% w/w of $(NH_4)_2SO_4$ and 12% w/w of MgSO4. The leachate (5417 Te/h) is pumped to 20 bara and reheated to 122.5°C, passing to the Magnesite Precipitation Vessel where it is mixed with carbonated ammonia solution.

The precipitation vessel is agitated and operates adiabatically at 120°C and 20 bara, with a residence time of 30 min. The cooling from the reactor inlet temperature is due to the endothermic nature of the reaction. Under these conditions the Mg is precipitated as magnesite ($MgCO_3$). The mass balance assumes that 85% of the Mg is precipitated.

The reaction is carried out using a significant excess of the carbonate solution compared to the required stoichiometric amount. The target molar ratio of Mg to ammonium carbonate and ammonium hydroxide is 1:1.5:1, where the ammonium carbonate is taken to be an equimolar mixture of the carbonate and bicarbonate.

Assuming a nominal 50/50 mixture of ammonium carbonate and bicarbonate the overall reaction is endothermic and is given by:

$$2MgSO_4(aq) + (NH_4)_2CO_3(aq) + NH_4HCO_3(aq) + NH_4OH(aq) =$$

$$2(NH_4)_2SO_4(aq) + 2MgCO_3(s) + H_2O(l)$$

$$\Delta H_r^0 = +25.12 \text{ MJ/kmol}$$

Under the specified reaction conditions there is therefore a 50% excess of carbonate, and 100% excess of ammonium hydroxide. The carbonation index of the solution used is 0.462, where the index is defined as the ratio of total carbonate ions to total ammonium. [Ammonium bicarbonate has an index of 1, and the carbonate an index of 0.5]. In terms of speciation, the major carbonate species in solution is the bicarbonate ion [HCO_3^-].

The combined feed rate to the precipitation stage is 7596 Te/h, with an exit solids content of 5.5% w/w. The precipitated magnesite, along with the Fe hydroxide, is recovered by using rotary vacuum filters. As with the recovery of the leached ore, washing is used to limit the loss of chemicals in the solids products. The magnesite stream (515 Te/h) is cooled and discharged to storage. Neither of the solids products is dried – wet product is taken to be acceptable. Provision is required in on-site storage to recover and recycle any liquor that may drain from the material on standing.

3.3 Recovery of Excess Carbonate Solution

The filtrate liquor from the filters contains a large excess of ammoniated carbonate solution, which has to be recovered since otherwise it would neutralize part of the regenerated bisulphate solution. The filtrate is first flashed to 1 bara. The ammonia and CO_2 remaining in the flashed liquor are recovered by steam stripping at low pressure. The heat input to the reboiler is 360 MW. The stripped vapours are compressed to 1 bar and mixed with the flash vapour stream, and the combined steam used to provide the heat input to the stripper reboiler. The partially condensed stream from the reboiler is further cooled, first by interchange to preheat the liquor feed to the precipitation vessel and then against cooling water. The liquid phase stream is recycled and mixed with fresh carbonated liquor from the offgas scrubber. The mixed carbonated feed is pumped to 20 bara and heated to 122.5°C before entering the precipitation vessel.

The overall recovery process is endothermic, and can be written as:

$$(NH_4)_2CO_3(aq) + NH_4HCO_3 = NH_3(g) + CO_2(g) + H_2O(l) \qquad \Delta H_r^0 = +269.8 \text{ MJ/kmol}$$

In practice a significant amount of water is also vaporized, adding to the energy input.

The stripped solution from the bottom of the Carbonate Stripping Tower contains 26% w/w of $(NH_4)_2SO_4$ plus a residual level of $MgSO_4$ (1.6% w/w), at a total flow of 6197 Te/h. Note that for simplicity in generating the overall mass balance, the slip of ammonia and the CO_2 in the liquor are both taken as zero. The actual levels, based on more detailed modelling of the decomposition stage, are 107 and 12 ppm by wt. respectively of NH_3 and CO_2.

3.4 Recovery of Ammonium Sulphate

The ammonium sulphate is separated from the stripped solution by crystallization in a series of evaporation stages. The liquor exits the stripping tower at 81°C and is preheated before passing to the first evaporation stage. This operates at 1 bara and utilizes mechanical vapour recompression (MVR) to 1.6 bara to provide the bulk of the 1500 MW evaporative heat load. The balance of the heat input (39 MW) is provided by a separate LP steam coil. The process condensate is recovered and recycled within the process. In this stage 51% (2274 Te/h) of water in the feed is separated, to generate a concentrated solution with 41% w/w $(NH_4)_2SO_4$.

This solution passes to a 3 stage, multi-effect evaporation system, operating at successively lower pressures, to recover the sulphate as crystals. The heat into the first effect, which operates at 0.75 bara, is provided by LP steam, either raised within the process or imported. The evaporation load to the first effect is 375 MW. The vapour from each of the first and second stages is condensed in the subsequent stage to provide the evaporation load. The overheads steam from the final stage (operating at 0.2 bara) is condensed against cooling water. A vacuum system is used to eject any non-condensable gases to atmosphere. The crystallized sulphate is recovered by centrifuging. Hydrocyclones could be used to pre-thicken the centrifuge feed. The crystal product is not dried, and is assumed to contain 2% w/w liquor.

The condensed water vapour is recovered and recycled for use within the plant. The hotter condensate from the MVR system and from the first effect is directly recycled to the bisulphate solution make-up and for filter washing. The bulk of the colder condensate (from the second and third effects) is collected separately and used in the downstream ammonia absorption system and flue gas scrubber. The final centrate is also recycled within the process.

3.5 Decomposition of Ammonium Sulphate

The recovered sulphate (1460 Te/h of crystals plus 57.5 Te/h of associated liquor) is conveyed via a lock-hopper system to the sulphate melt vessel, which operates at 10 bara and 360°C. In the vessel the recovered sulphate is dissolved into a recycle stream of molten ammonium bisulphate, using a ratio of bisulphate to fresh sulphate of around 1:1.

The vessel is heated via a hot oil system to maintain the melt at 360°C and decompose the sulphate into ammonia and NH_4HSO_4.

The decomposition reaction is given by:

$$(NH_4)_2SO_4(s) = NH_3(g) + NH_4HSO_4(s) \qquad \Delta H_r^0 = +108.5 \text{ MJ/kmol}$$

where (g) = gas

The sulphate is not heated directly as it does not melt but decomposes first, partially melting in the process, and is thought to generate a range of undesired side-products, both solids and gases. The bulk of the decomposition (80%) is assumed to take place in the melt vessel, and the released ammonia, along with steam from the residual moisture in the crystal feed, is cooled and partially condensed in the vessel overheads stream. The partially decomposed melt is pumped to a tower where the remaining sulphate is further decomposed by counter-current gas stripping with nitrogen, to produce a bisulphate bottom product with low residual sulphate. The overall level of decomposition is taken as 98%. The overhead ammonia/nitrogen gas stream is cooled by interchange with the recycled nitrogen stream and then by cooling water. The stream is mixed with the recovered material from the melt vessel and passes to a gas/liquid separator. The vapour passes to the ammonia absorber, where it is absorbed into water to produce 25% w/w ammonia solution. The scrubbed nitrogen is compressed and recycled to the stripper.

The product bisulphate is taken as a purge from the melt recycle at 333°C, and contains around 3% of residual sulphate. As ammonium bisulphate melts at 145°C the product is cooled only to 160°C and then mixed with water recovered from the evaporation area and with the centrate to re-constitute the bisulphate solution for recycle to the ore leaching stage. Water is also added here to make up for losses in the process, principally in the product mineral streams.

Losses from the process of sulphate and ammonia are made-up by addition of imported $(NH_4)_2SO_4$ into the melt vessel. Depending on the exact ratio of ammonia to sulphate lost from the process it may be necessary to also add either sulphuric acid or ammonia. Alternatively, the losses could be made up solely by addition of these chemicals.

3.6 Ammonia Absorption

The ammonia absorber is a two stage packed tower, operating at 9.5 bara. In the upper stage, 451 Te/h of ammonia free water, recycled from the sulphate crystallization area and cooled to 40°C, is used to reduce the ammonia content of the nitrogen to 100 ppm. The water rate has been set to produce the maximum practical concentration of ammonia in the bottoms liquor (without resorting to refrigeration). The bottom stage uses a cooled recycle of ammonia liquor to scrub out the bulk of the ammonia and remove the substantial heat of absorption (62.5 MW), with the bottoms product at 60°C. The recovered ammonia liquor is mixed with the ammonia liquor stream from the separator to give a stream of 28% w/w ammonia. The ammonia rate is 189 Te/h, in 477 Te/h of water.

The absorption process is exothermic and is given by:

$$NH_3(g) + H_2O(l) = NH_4OH(aq) \qquad \Delta H_r^0 = + -30.6 \text{ MJ/kmol}$$

The ammonia solution from the absorber area is split three ways. One part of the stream (120 Te/h) is sent to neutralize the leachate upstream of the belt filter. Of the remaining ammonia liquor, 342 Te/h passes to the Flue Gas Scrubber, where it absorbs CO_2 from the gas stream to produce a bottoms product of carbonated ammonia liquor, and the balance (204 Te/h) is by-passed around the scrubber to mix with the scrubber bottoms stream, this mixed stream being fed forward to the magnesite precipitation vessel. Only part of the liquor to the scrubber area is passed through the scrubber in order to control the extent of CO_2 absorption whilst limiting the problem of ammonia slip into the flue gas stream.

3.7 CO_2 Absorption / Liquor Carbonation

The flue gas, which contains 251.5 Te/h of CO_2 (14.4% molar), is compressed at the inlet to the scrubber to a nominal 1.2 bara to offset the additional pressure drop imposed by the scrubber system. Due to the high volatility of ammonia and the low operating pressure of the scrubber, it is necessary to operate the scrubber at low temperature to limit the loss of ammonia into the flue gas stream. It is also necessary to minimize the amount of water used in the scrubber to avoid excessive water recirculation within the overall process. As a result of the low operating temperature and minimum water rates, much of the CO_2 in the scrubber liquor (48%) is as solid ammonium bicarbonate. The scrubber process is similar to other processes proposed for recovery of CO_2 from flue gases using chilled ammonia, for example the Alstom process (19,20), which also operates under conditions where ammonium bicarbonate is precipitated.

The carbonation reactions are exothermic, and for example the reaction to solid bicarbonate is given by:

$$CO_2(g) + NH_4OH(aq) = NH_4HCO_3(s) \qquad \Delta H_r^0 = -93.4\ \text{MJ/kmol}$$

The heat release arising from carbonation is approximately 102 MW.

The scrubber is a three-stage system. This is comprised of a two-stage scrubbing column where the majority of the CO_2 is absorbed, and a wash column for which the major function is the removal of ammonia from the flue gas stream.

In the bottom stage of the scrubbing column, the compressed flue gas is cooled to 25°C against cooling water and contacted with a recycle stream of chilled liquor at 12.5°C. The column bottoms (at 25°C) contain around 8% by wt of solid ammonium carbonate. A purge is taken off this stream and the balance is chilled to 12.5°C, causing further bicarbonate to be crystallized out. The chilled stream is centrifuged to recover the bicarbonate solids, which are mixed with the purge liquor.

The chilled centrate stream is recycled to the top of the bottom stage of the flue gas scrubber, where it mixes with liquor from the upper section. The liquor to the top of the scrubbing column is chilled ammonia liquor at 12.5°C.

The gas from the scrubber is cooled in an overheads condenser against cooling water and passes to the wash column. 400 Te/h water from the evaporation area is cooled, initially to 25°C against cooling water, then chilled to 12.5°C. This is fed to the top of the wash column, and is used to reduce the ammonia content of the flue gases to ppm levels. The scrubbed flue gas is reheated by interchange with the blower exhaust and discharged. The wash column bottoms stream is mixed with 342 Te/h of ammonia liquor, and the mixed stream is cooled and then chilled (to 12.5°C). The chilled ammonia liquor is fed to the top of the scrubber.

The overall removal of CO_2 is 80%. Of this, some 90% is removed in the scrubbing stage, with the remaining 10% in the wash stage.

The solid ammonium carbonate in the scrubber bottoms is readily dissolved as the stream is heated upstream of the magnesite precipitation stage.

4 Overall Mass Balance

The overall mass balance is summarized in Table 1 below. The table does not include the water make-up to the associated cooling water system (estimated at 1576 Te/h) or the fired heater for the hot oil system.

Table 1. Overall Mass Balance.

		Feed Streams						Product Streams		
Stream Description		Ore	FG	Water MU	Sulphate Feed	N2 MU		Export Mineral	Magnesite Export	Return FG
H2O	Te/hr		13.1	111.5				65.7	91.6	14.8
N2	Te/hr		882.0			0.1			0.0	882.0
O2	Te/hr		53.5						0.0	53.5
CO2	Te/hr		251.5						0.0	50.7
(NH4)2SO4_AQ	Te/hr							3.4	7.8	
NH4OH	Te/hr								0.7	
MGSO4_AQ	Te/hr							2.2	0.5	
(NH4)2CO3_AQ	Te/hr								0.1	
NH4HCO3_AQ	Te/hr								0.4	
SERPENTINE	Te/hr	563.0						140.8		
QUARTZ	Te/hr	12.2						195.3		
MAGNESITE	Te/hr								383.6	
FE(OH)3_PPT	Te/hr							0.1	21.7	
FE(OH)2_PPT	Te/hr								9.1	
FE3O4	Te/hr	50.2						26.6		
(NH4)2SO4_SOLID	Te/hr				13.9					
TOTAL	Te/hr	625.4	1200.1	111.5	13.9	0.1		434.0	515.5	1001.0

5 Overall Energy and CO_2 Balance

5.1 Energy Balance

A nominal heat integration scheme has been developed. A detailed assessment and optimization of the process operating conditions and heat integration scheme, for example, the evaluation of alternative integration schemes, optimization of evaporator pressures and heat exchanger duties, has not been carried out. The basic approach adopted is to use the heat released from the extraction reactor system to provide preheating for the feeds to the magnesite precipitation stage. The extraction temperature has been raised marginally from 125°C to 127.5°C to facilitate this.

The energy input into the sulphate decomposition stage is taken to be provided by a hot oil system heated by a stand-alone natural gas fired heater. The only source of relatively high grade heat within the process is from the cooling of the bisulphate melt (126 MW) and the overheads vapour stream from the melt vessel (30 MW to 120°C). In principle it is possible to use these two heat sources to raise 20 bara steam, generating approximately 13 MW of power via a pass-out steam turbine exhausting to 2 bara, this low pressure steam being used within the process. This scheme has not been adopted, although it would provide a marginal improvement to the overall process energy requirements. For simplicity, the heat is assumed to be used to directly generate LP steam for use within the process. The preheat duty for the feed to the first sulphate evaporation stage (81-92°C) is met by heat recovery from the extraction area. Preheating of various recycle streams is by heat recovery from the carbonate solution recovery area. There is a deficit of heat available at sufficient temperature to provide the heat input required for the sulphate crystallization stage and the net requirement of 241 MW is taken to be provided by import of saturated 2 bara steam.

The LP steam import and associated export of steam condensate is not included in Table 1.

The overall energy balance is summarized in Table 2. A breakdown of the electrical power usage is given in Table 3. Note that all grinding or milling operations are assumed to take place off-site and energy requirements for these operations are not included.

Table 2. Process Energy Summary.

Electrical Power Import	143.1	MW
Natural Gas Firing	722.8	GCV MW(1)
LP Steam Import	240.5	MW(2)
Load to Cooling Water	999.1	MW

Notes : 1. assumes 85% GCV efficiency
2. condensing load

Table 3. Electrical Power Usage (MW).

Compressors		
FG Blower	4.8	
Sulphate Vapour Compressor	64.9	
Stripper Gas Blower	0.3	
Carbonate Stripper Vapour Compressor	29.0	
Refrigeration Compressor Set	6.9	
Sub Total		105.8
Pumps		
Slurry Feed Pump	0.5	
Reactor Transfer Pumps	0.7	
Leachate Transfer Pump	3.6	
Condensate Feed Pump	1.6	
CW Circulation Pumps	12.4	
AS Feed Pump	0.4	
Hot Oil Pumps	0.8	
Other pump duties	2.7	

Sub Total		22.7
Agitators		
Slurry M/U vessels	1.2	
Extraction reactors	2.2	
Magnesite Pptn	1.9	
Sulphate melt vessel	0.9	
Bisulphate solution vessel	0.2	
Sub Total		6.4
Other		
Sulphate centrifuges	2.8	
Scrubber Carbonate Centrifuge	0.9	
Conveyors	2.5	
Filters	1.9	
Sub Total		8.2
Total		143.1

5.2 CO_2 Balance

The overall CO_2 balance is given in Table 4 and Table 5 below. As can be seen, there is a net reduction in CO_2 emissions of 11%. The CO_2 associated with the energy import streams are taken to be as follows:

- Natural gas
    - 0.18 Te/h CO_2 per GCV MW, based on a typical NG composition
- LP steam import
    - 0.18 Te/h CO_2 per MW of condensing load, based on pass-out 2 bara steam from a NG fired power station

Table 4. CO_2 release.

Source	**Te/h CO_2**
Flue gas CO_2 for coal fired power plant	251.5
CO_2 from NG firing	130.1
CO_2 for LP steam import	43.3
CO_2 sequestered	200.8
Nett release	**224.1**

In the Shell report, the power usage is accounted for by assuming power generation from a NG fired power plant at 60% efficiency (basis undefined but taken as NCV). On the same basis, the following figures are obtained. The comparable (scaled) figure obtained for the Shell process is (+54.9 Te/h). This is slightly different from the figure quoted by Shell (the report fails to define the assumed NG properties).

Table 5. CO_2 avoided.

Source	**Te/h CO_2**
Flue gas CO_2 for coal fired power plant	251.5
CO_2 from NG firing	130.1
CO_2 for LP steam import	43.3
CO_2 for power import	47.6
CO_2 sequestered	200.8
Nett release	**271.7**
Avoided CO2	**(-20.2)**

6 Review of Process Energy Efficiency

Ignoring the presence of iron compounds, the overall reaction is given by:

$$Mg_3Si_2O_5(OH)_4(s) + 3CO_2(g) = 2SiO_2(s) + 3MgCO_3 + H2O(l)$$

$$\Delta H_r^0 = -187.2 \text{ MJ/kmol}$$

The reaction is exothermic, and at flowsheet rates the heat release is 106 MW.

Despite the overall exothermic nature of the process, the energy input is substantial due to four main factors, namely:

1. Large material flows:
    - for 1 Te CO_2 removed,
        - 3.1 Te of ore feed
        - 3.9 Te of dry solids generated
        - 7.4 Te ammonium sulphate turn over
    - aqueous chemistry: large water flows are required to move reagents around, for example the water flow through the carbonation step is 25.2 Te/Te CO_2 removed
    - significant power inputs (pumps, agitators, compressors) associated with large material flows and large reaction volumes
2. Material inefficiencies:
    - Extraction: reaction rate limits the bisulphate conversion per pass
    - Carbonation: uses 150-200% of the stoichiometric amount of reagents, with limited conversion of Mg per pass (85%)
3. Energy intensive steps

    - Sulphate crystallization - total evaporation load is 2792 MW
    - Sulphate decomposition - 614 MW heat input at 360°C
    - Carbonate solution recovery - 360 MW reboiler duty
4. Exothermic steps release heat at low temperature
    - Extraction - 160 MW at 127.5°C
    - Ammonia absorption - 63 MW at 60°C
    - CO_2 scrubbing - 102 MW at ambient or below

The energy input for the sulphate decomposition stage is due to the reaction being very endothermic, but in addition the required operating conditions, coupled with practical limitations on the system design, mean that the heat input (at 360°C) is almost twice that of the standard heat of reaction (332 MW).

7 Capital Cost Estimate and Comparison of Initial and Current Designs

A simple 'Process Step Scoring' method due to Taylor (21,22) was used to generate an order of magnitude cost for the initial process flowsheet and was also applied to the revised heat and mass balance to give an indication of the likely level of cost saving arising from the changes. A detailed cost estimate was also generated for the revised design basis and is described in section 7.2 below.

7.1 Initial High Level Capital Cost Estimates

The Taylor method is one of several broadly similar approaches that can be used at the early stage of project definition to generate indicative capital costs, since they require a minimum of information.

The method is based on the concept of a 'significant process step', which refers to the operation performed on the process stream. Generally the number of significant steps will be much lower than the number of main plant items (MPI's). A step may require several MPI's in series (e.g. multiple reactors), although it can be that one item may be capable of performing multiple steps. The definition of the steps will be, to some degree, arbitrary, in that it may differ depending on the individual interpretation of Taylor's guidelines and possibly also the degree of understanding or the detail available of the process under consideration. The degree of solids handling involved in this process places it outside the strict range of applicability of the method (indeed the sheer scale of the process is probably outside the range of applicability of any of the generalized estimating methods) and so the method has been adapted slightly. The factor for parallelization has been excluded since it results in unreasonably high figures. The figure given by Shell is (23) for their process is \$1153M for c. 100 Te/h CO_2 removal from flue gas, taken as equivalent to £740M. Scaling on a 0.6 power law to 201 Te/h CO_2 removal gives an estimate of £1125M for the scale of process used here. The Shell cost figure includes unspecified allowances for contingency, new technology and OSBL costs.

As applied to the original flowsheet, the capital cost figure of £860M generated using the Taylor approach is broadly in line with that estimated by Shell for their process. The comparable cost figure for the modified design basis is £700M, a reduction of 19%. These figures include an allowance of 30% to cover OSBL costs. The cost breakdown by area is shown in Figure 2 below.

The main differences between the initial and current design basis are summarized in Table 6.

Table 6. Design basis comparison for the ammonia based CCSM processes evaluated.

	Initial Basis	**Current Basis**
Extraction		
Temperature	100 C	127.5 C
Bisulphate utilization	80%	90%
Solid/liquid ratio	125 gm/l	152.5 g/l
Mg precipitation :		
Temperature	80 C	120 C
Pressure	2 bara	20 bara
Mg carbonate form	hydromagnesite	magnesite
Reagent excess	150%	50-100%
Mg recovery per pass	94%	85%
Leachate neutralization	downstream of belt filters	upstream of belt filters

The two major changes are the increased utilization of sulphate, and the shift from hydromagnesite to magnesite. Minimizing the turnover of ammonium sulphate is the key factor in improving the process energy efficiency.

Hydromagnesite has the formula $Mg_5(CO_3)_4(OH)_2.4H_2O$, and therefore a Mg to carbonate ratio of 5:4, rather than the 1:1 in magnesite. For a given removal rate of CO_2, the required rate of Mg is therefore 20% lower for magnesite than for hydromagnesite. These gains are offset to some extent in the Mg precipitation stage despite the reduction in the amount of reagents used, due to the increase in operating pressure (required to maintain the CO_2 partial pressure in the solution) and the lower reaction of Mg per pass.

Since any unreacted bisulphate has to be neutralized to the sulphate, increasing the bisulphate utilization also reduces the turnover of sulphate in the process, in this case a reduction of 11%.

The bisulphate solution strength is virtually the same in the two cases. It would be beneficial to increase the solution strength since this reduces the amount of water circulated through the process, and in particular reduce the amount of water to be removed from the sulphate solution. However, it is not possible to independently increase the concentration without, for example, reducing the degree of utilization of the bisulphate. As described in ref. 1, for a given particle size, the basic extraction process can be characterized by six parameters:

1. temperature
2. initial bisulphate concentration
3. % conversion of bisulphate
4. % extraction of Mg
5. solid/liquids ratio
6. residence time

If the operating temperature is fixed, then there are three degrees of freedom - three of the other five parameters can be chosen although there are also constraints, for example, the solubility of $MgSO_4$ in the system, and any upper limit on the solids/liquid ratio. The relationship between residence time, operating temperature and bisulphate utilization is based on analysis of experimental results as briefly described in Appendix 2.

Referring to Figure 2 below, in most areas the comparable capital cost as derived by application of Taylor's method is lower due to the reduced mass throughput. The largest reductions are in the ore filtration area where, as well as the reduction in throughput, less expensive materials are required due to upstream neutralization of the feed, the carbonate solution recovery area and in the magnesium carbonate recovery. The only area where the cost has increased is for the magnesium precipitation, due to operating at higher pressure.

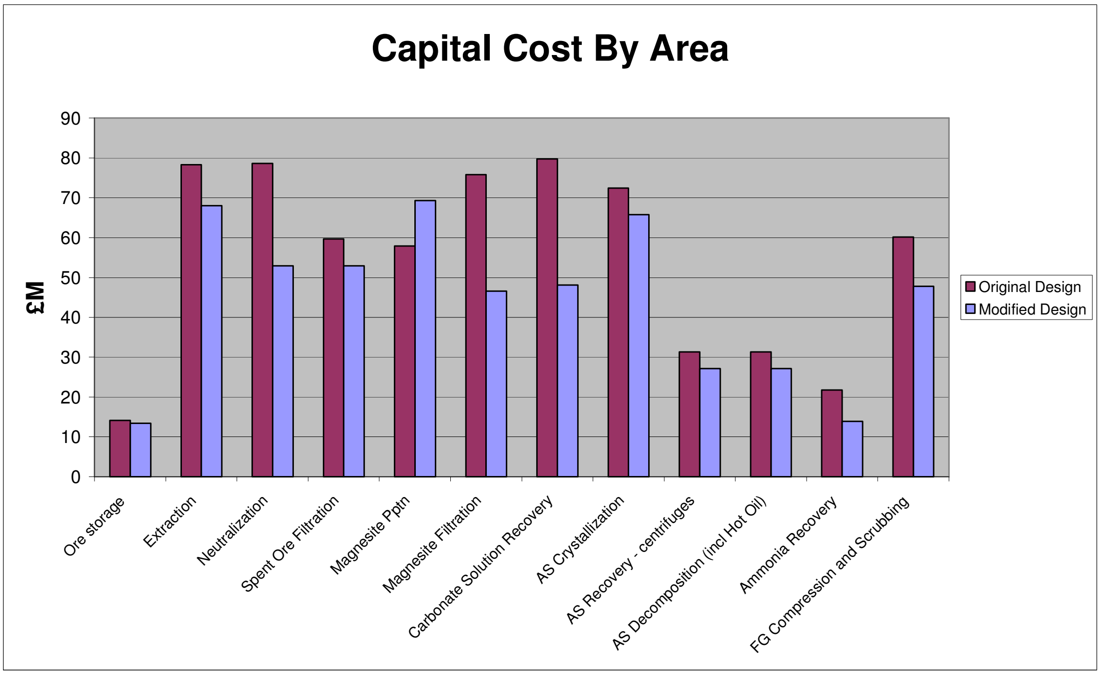

Figure 2. Capital cost comparison using Taylor's method for the ammonia-based CCSM processes evaluated.

7.2 Detailed Cost Estimate

7.2.1 Overview

This section summarizes the detailed capital cost estimate for the revised design basis, set appropriate to the level of process definition. The cost is given in 2009 £.

The Main Plant Item (MPI) costs use parametric correlations based on key sizing factors, as available from the literature (24,25,26,27). The net installed cost is obtained by using typical installation factors appropriate to the MPI, allowing for the MPI size and materials of construction. The gross installed cost is obtained by use of an assumed typical net to gross factor. In line with the approach used for the evaluation of the Shell process (23), additional percentage allowances are made for a project contingency (15%) and as a new technology allowance (5%). No breakdown of the Shell capital cost figure is given, and it is therefore not known if these figures are consistent with the assumptions made by Shell.

Costing of the process presents a number of difficulties, arising both from uncertainties in the process design and the sheer scale of the process. For many of the process steps the mass throughput is well above the proven capacity range of the process equipment concerned, requiring the use of multiple parallel streams. Assumptions about the maximum feasible capacity of items, in particular the filters, centrifuges and crystallizers, whilst based on published information, are probably on the generous side. If further development of the process is undertaken, then a thorough assessment of maximum equipment capacity will be required.

For the filters, the absence of information on the filtering characteristics of both the leached ore and the magnesite precipitate means that the derived filter areas are based only on notional filtration rates, thought to be representative of the nature of the solids being recovered. The sizing and cost of the filters is therefore subject to a high degree of uncertainty.

In general, the maximum diameter for vessels is taken as c. 6 m. For columns, the diameters are defined by consideration of the packing hydraulics, using available vendor software (28), and subject to the assumed nominal upper limit.

Area requirements for heat transfer equipment are determined from typical shell or tube side heat transfer coefficients, with allowance for dirt and wall coefficients, appropriate to the nature and state of the fluid being handled. For the purposes of defining the number of parallel items required, the maximum heat transfer surface for an individual unit has been taken as a nominal 18,000 m^2.

Pump duties are defined by the flowsheet rates and an estimate of the head required. Costs for pump system generally assume 2 off 50% pumps, with an installed 50% spare. Where multiple pump sets are assumed to be required due to multi-streaming of associated vessels (for example in the sulphate recovery area), the number of spare pumps is halved.

The details of some elements of the heat integration scheme are not fully defined, particularly in respect of the arrangement for the heat recovery from the extraction reactors. The costs for the two evaporators include the associated heat transfer surface. The two crystallizers are costed on the basis of separate vessels with external heat exchangers.

The capital estimate includes the costs for the refrigeration, cooling water and hot oil systems. The process imports electrical power and LP steam; no capital cost element is included for the provision of these services.

Materials of construction (MOC) are generally taken as stainless steel, with some carbon or low alloy steel where deemed appropriate. In the extraction area, the selection of the MOC is problematic given the highly corrosive nature of the leaching solution. For costing purposes only, the MOC for exchanger surfaces exposed to the leaching solution is taken as Monel, as representative of a more expensive alloy. The extraction reactors are costed on the basis of being in stainless steel.

The total volume required for the extraction reaction is 4343 m^3. This is assumed to include the volume in the slurry make-up vessels.

The receiver vessels and catch-pots are sized on liquid hold-up requirements and on the diameter required for effective demister operation (where appropriate).

7.2.2 Cost Estimate Figures

The MPI and installed costs are given in Tables 16-20 in Appendix 5 on an item by item basis, arranged by MPI type.

The basic sizing parameters of the key items of equipment are given in Tables 21-23 in Appendix 5.

The physical dimensions of the two evaporator stages are not given in Table 22 as these are costed on the required heat transfer surface area (given in Table 21). Similarly, the dimensions of the crystallizers are not given in Table 22 as these are costed on the basis of the crystallization rate. However, the likely size can be estimated from information from Whiting/Svenson (29). This shows a crystallizer built for Shell producing 600 t/day of ammonium sulphate, at 22ft (6.7m) diameter. The required sulphate recovery rate here is 35,040 Te/day. For costing purposes the maximum production rate per crystallizer is taken as c. 900 Te/day. The likely maximum practical diameter for a crystallizer appears to of the order of 40-42ft (c. 12.5 m), which might indicate a possible maximum capacity of up to c. 2000 Te/day.

The total cost is summarized in Table 7 below.

Table 7. Capital Cost Summary.

		2009 £M
Nett Installed Cost		
Table	**Equipment Type**	
2	Pumps	15.9
3	Exchangers	60.4
4	Vessels	297.2
5	Machines (incl. filters)	307.6
6	Other (Fired heater, CW towers)	35.6
Total Nett Installed		**717**
Gross Installed		968
Contingency		145
New Technology Allowance		48
Total Capital		**1161**

7.2.3 Review of Cost Breakdown

The cost breakdown by area is shown in Figure 3 below. The capital cost is dominated by two main areas, the precipitation and recovery of magnesite, and the recovery of ammonia sulphate, which together constitute 76% of the cost.

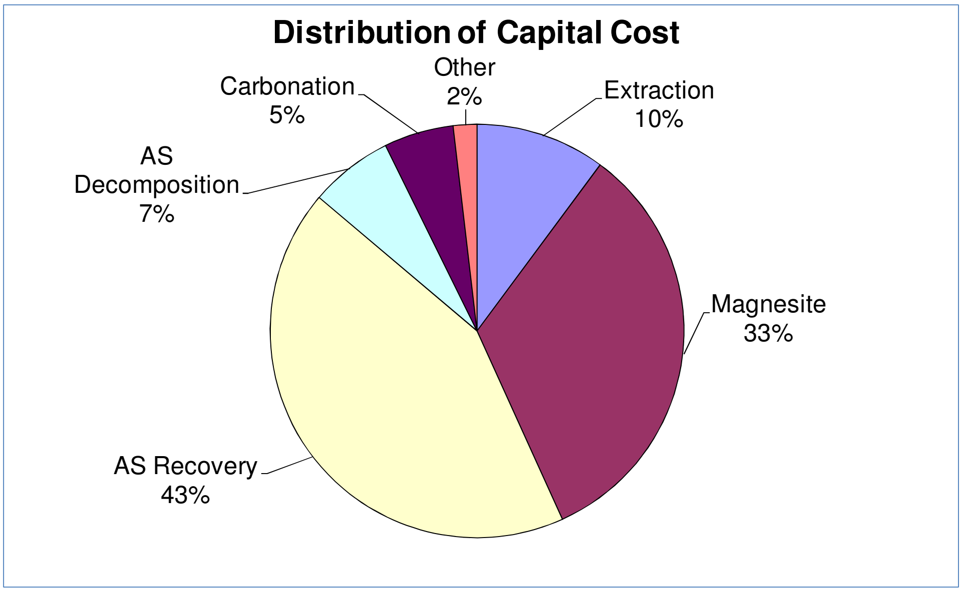

Figure 3. Cost Breakdown by Area

The breakdown of costs in sulphate recovery is shown in Figure 4. Some 48% of the cost is in the initial evaporation stage, split roughly 40/60 between the vapour compressors and the evaporator vessels, with a further 34% of the cost attributable to the two crystallization stages. The evaporative load on the initial evaporation stage is 1499 MW; the power requirement of the compressors is estimated at 69 MW. These figures reflect the large throughput of ammonium sulphate solution (6197 Te/hr).

Figure 5 shows the cost breakdown in the magnesite precipitation area. There the cost is dominated by the filters, with the cost of solution recovery and recycling, and the precipitation vessels also being major contributors. The filters are expensive because such a large number of units are required. The maximum filtration area for a rotary vacuum filter has been taken as 150 m^2 per unit. Given an assumed filtration rate of 500 $kg/hr/m^2$, the area for the total filtration rate of 6894 Te/hr is 13788 m^2, requiring 100 units. The vessels are relatively expensive because of the volume required (4270 m^3 total) and their operating pressure (20 bar).

Figure 6 shows the top ten cost items, where the figures are the percentage of the total installed cost. The most costly unit operation is the filtration of the precipitated magnesite, which accounts for some 17% of the capital cost. The first evaporation stage, consisting of the evaporator itself and the overheads vapour compressor, are the next two mostly costly items, and together account for over 20% of the capital cost.

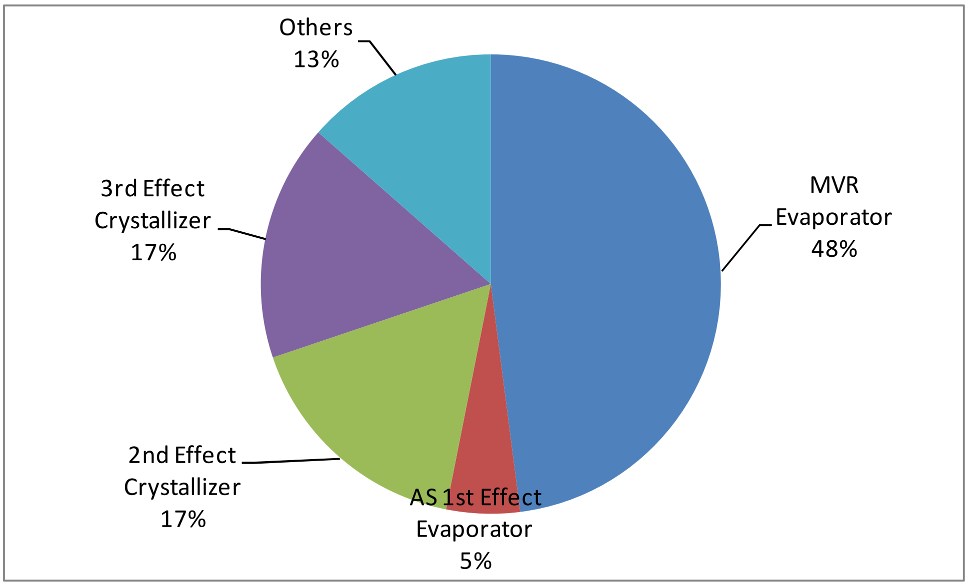

Figure 4. Costs in the Ammonium Sulphate Recovery Area.

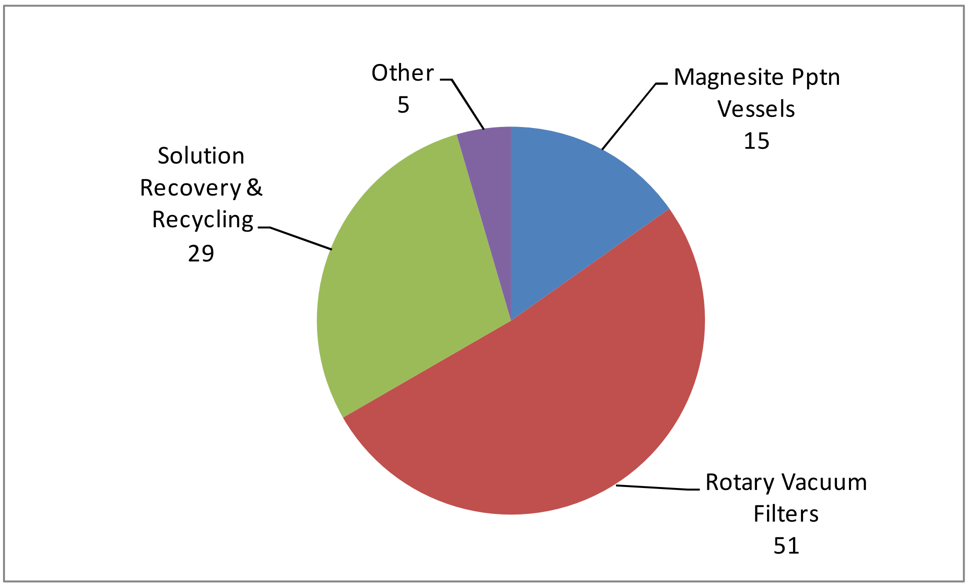

Figure 5. Costs in the Magnesite Production Area.

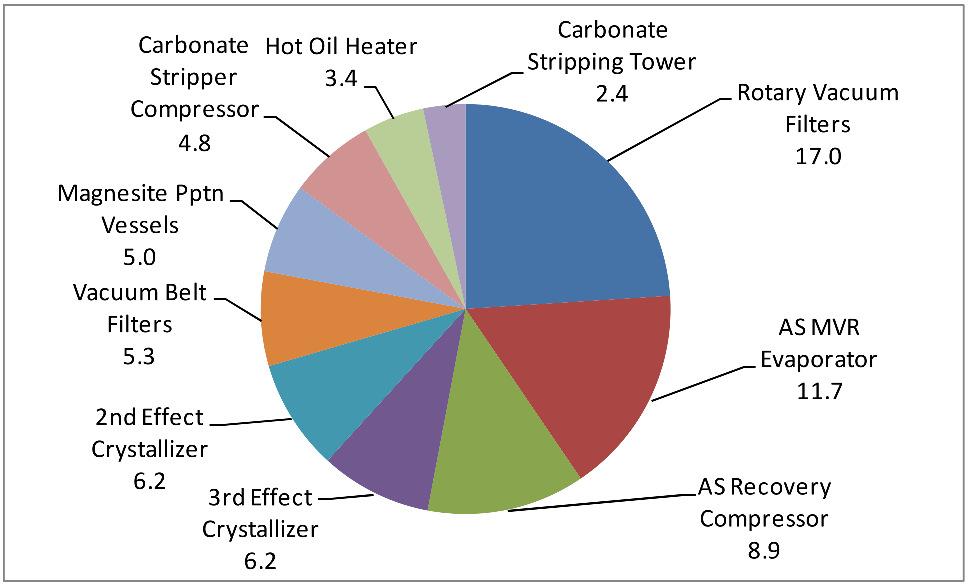

Figure 6. Top Ten Most Costly Unit Operations (% of total capital cost).

7.2.4 Capital Cost Summary

At £1161M, the estimated capital cost is very similar to that of the Shell process (scaled to the same flue gas CO_2 removal capacity).

A major area of uncertainty in the cost estimate is the size of the filtration equipment, and any further evaluation should fully assess the filtration properties of both solids streams and the selection of filtration equipment. The possibility of using a single filtration stage should also be considered.

The other major cost area is the ammonium sulphate recovery system. Although increasing solution strengths would reduce costs, the scope for improvement (in both energy usage and costs) by this means may be fairly limited.

8 Estimated Variable Costs

8.1 Raw Materials and Energy Costs

The variable costs for feedstock and energy usages are given in Table 8 below. The on-site delivered cost of the feed ore is not included, nor are costs associated with disposal of solids product streams. In the original process concept the solids products were intended to be recovered as products with a significant sale value (i.e. as silica, iron hydroxide and magnesite). However, it was considered by the project that the probable additional costs associated with meeting the required product specifications was unlikely to be justified. There are no specifications for the process product streams, and they are taken to have no net value.

Table 8. Variable costs and energy consumption for the modified ammonia based CCSM process.

Feed	Rate	Units	Unit cost £	Cost	Units
Process Water	111.5	Te/h	0.5 / Te	56	£/h
CW System make-up	1740.6	Te/h	0.5 / Te	788	£/h
Ammonium sulphate	13.9	Te/h	70 / Te	970	£/h
Natural gas	722.8	MW GCV	19.11 / MWhr	13813	£/h
Electricity	143.1	MW	50 / MWhr	7156	£/h
LP Steam Import	393.3	Te/h	12 / Te	4720	£/h
			Total	27503	£/h
			Annualized	218	£M/yr

8.2 Other Manufacturing and Fixed Costs

Other annual costs are summarized in Table 9, based on a capital cost of £700M and using typical factors based on the plant capital cost, throughput and complexity.

Table 9. Manufacturing and fixed annual costs for the modified ammonia based CCSM process.

Item	**£M/yr**
Operator salaries	12.4
Supervisor salaries	2.5
Laboratory charges	1.2
Plant overheads	6.2
Maintenance	35
Capital charges	70
Insurance	7
Local taxes	7
Total	**141.3**

8.3 Total Variable Cost

The total variable cost is therefore estimated at £359 M/yr. This equates to a cost of £226/Te of CO_2 removed from the flue gas. This cost does not include any costs for grinding of the ore, nor for transport of the feed and product material.

9 Process Comparison

Table 10 compares the Shell process (as documented in ref. 23) with the ammonia based process, for both the initial and revised design basis. For the Shell process the figures are scaled to an equivalent rate of flue gas CO_2 removal, as in the design basis used here; the capital cost figure is scaled as described in section 7. Of the significant energy users in the Shell process, the power given for the Flue gas blower is 30.2 MW. This is surprisingly high, and may not be accurate.

The Shell report gives the rate of NG as 1.91 kg/s, but not the composition of GCV value. The figure given in Table 8 for the Shell NG GCV rate assumes a nominal NG GCV value of 55.25 MJ/kg. The Shell report assumes that the power requirement for their mineralisation process is generated by a NG fired power plant at 60% efficiency. The basis of the efficiency figure is not defined, but is assumed to be on an NCV basis (i.e. c. 54% on a GCV basis). The CO_2 data for the current design is taken from Table 5.

Table 10. Process comparison between Shell and the two ammonia based CCSM processes evaluated.

	Shell	**Ammonia**	
		Initial Design Basis	**Current Design Basis**
Capital (£M)	1125	860	700
CO_2 captured	80%	80%	80%
CO_2 parasitic load	30%	n/a	88%
CO_2 avoided	50%	n/a	(-8%)
£/Te CO_2 avoided	306	n/a	n/a
NG firing (MW GCV)	212	n/a	773
Power usage (MW)	156	n/a	147
Significant energy users	• Thermal pretreatment • Flue gas blower • Absorption column pump	• sulphate recovery & decomposition • recovery of carbonate solution	

10 Life Cycle Assessment

Life cycle assessment (LCA) is an ISO standardised process that attributes life cycle material consumption, energy use, and solid, liquid and gaseous emissions to environmental impacts. In this assessment, we have focused on quantification of life cycle carbon footprint (per tonne CO_2 sequestrated) of CCSM technology and geological CCS technology to address the impact of these technologies on future climate change risk mitigation. Energy, material and emission data have been collected through the whole life cycle steps of both technologies with consideration of UK geological sites, including mineral mining, transportation, mineral crush and grinding, CO_2 capture and sequestration, etc. Carbon footprint of unit electricity, natural gas, steam and materials are obtained from EcoInvent [ref. 30] or European Life Cycle Database (ELCD) [ref. 31] with UK specific dataset or European average dataset. The preliminary analysis has demonstrated net carbon balance benefits from both technologies. We have also identified that steam and electricity demand of the mineralisation process contributes the most to the overall carbon footprint of CCSM technology. Benefits of CCSM technology can be significantly improved when waste steam is available for the heating requirement in the CCSM process. In next steps, the preliminary analysis will be validated and a few scenarios will be further explored to finalize the analysis, including electricity grid options for each life cycle steps, waste heat energy availability, uncertainty and sensitivity analysis, etc.

10.1 Carbon footprint of major energy inputs

Major energy inputs to the CCS technology life cycles are electricity, natural gas, diesel, etc. Due to variability of energy sources and production process, the carbon footprint of each energy type can be quite different as shown in Table 11. In this analysis, electricity of coal-fired power plant is used in the CO_2 capture and CCSM process and UK average electricity grid is applied for all other electricity requirements.

Table 11. Carbon footprint of energy sources.

Energy type	Carbon footprint	Unit	Reference
Electricity, UK average	599	$kgCO_{2e}$/MWh	EcoInvent [30]
Electricity, from hard coal	1050	$kgCO_{2e}$/MWh	EcoInvent [30]
Natural gas combustion, HHV	214.1	$kgCO_{2e}$/MWh	EcoInvent [30]
Diesel	2.7	$kgCO_{2e}$/litre	CarbonTrust [32]

10.2 Serpentine Mining

Material and cost data of Mount Keith nickel mining operation from 1994 to 2007 is used to estimate the serpentine mining activities because serpentine is the base rock for the nickel metal in this mine. This analysis focuses on energy consumptions of the operation process, including drilling, excavating, loading, hauling and crushing steps. The overall waste-to-product ratio is assumed to be 1:3. Capital equipment, explosives and maintenance are not included in the calculation. The overall carbon footprint of serpentine mining is shown in Table 12.

Table 12. Carbon footprint of serpentine mining process.

Energy input	Unit	Range (1994-2007)	Average
Diesel	liter/tonne serpentine	0.79-1.02	0.85
Electricity	kWh/tonne serpentine	3.62-7.20	5.08
Carbon footprint	CO_{2e}/tonne serpentine	2.39-4.57	3.27

Without the milling and treatment steps in Nickel production, serpentine mining is not an energy intensive process and has a small carbon footprint. Table 13 shows that serpentine mining carbon footprint is similar to a few other mineral mining processes.

Table 13. Carbon footprint of various mineral mining process. [after ref. 30]

Minerals	Carbon footprint, $kgCO_{2e}$/tonne
Clay	2.92
Gravel	2.39
Gypsum	2.03
Limestone	1.91
Sand	2.39
Average	2.33

10.3 CCSM process

The CO_2 capture and sequestration process includes dissolution of the mineral (chemical pretreatment) with ammonium bisulphate to extract aqueous magnesium from the rock, separation of insoluble impurities such as silica, separation of iron impurities by pH adjustment with ammonium hydroxide, CO_2 capture with ammonium hydroxide, carbonation reaction with flue gas and ammonium hydroxide, precipitation and separation of magnesium carbonate product, recovery of ammonium sulphate, decomposition of ammonium sulphate to ammonium bisulphate, ammonia, recycle and reuse ammonium bisulphate, ammonia and water.

The evaporation of water from the ammonium sulphate solution uses waste steam from the power plant as the heat source, and the decomposition of ammonium sulphate will use separate natural gas combustion as additional heat energy. These two steps are the most energy intensive steps in the CCSM progress. The Centre for Process Innovation (CPI) has provided the CCSM process mass balance and energy balance data. Serpentine mineral grinding energy required before chemical pretreatment is estimated as 33 kWh/tonne mineral as it was described in Chapter 4.3.

Table 14. Mineralisation process inventory.

Mineralisation Process Inventory					
Functional unit: per tonne CO_2 sequestrated					
Input					
Material	Amount	Unit	Carbon footprint	Unit	Comments
Ore	3.1	tonne		kg CO_{2e}	
Flue gas	6.0	tonne		kg CO_{2e}	
Water	0.6	tonne		kg CO_{2e}	
Ammonium Sulphate	0.069	tonne	38.1	kg CO_{2e}	EcoInvent
N_2	4.85E-04	tonne	0.2	kg CO_{2e}	EcoInvent
Product					
Magnesite	2.6	tonne		kg CO_{2e}	
Other minerals	2.2	tonne		kg CO_{2e}	
Waste					
Flue gas	5.0	tonne		kg CO_{2e}	
Energy consumption					
Electricity for CCSM process	0.7	MWh	779.3	kg CO_{2e}	
Natural gas (HHV)	3.6	MWh	770.9	kg CO_{2e}	
Low pressure steam	1.3	MWh	0.0	kg CO_{2e}	Power plant waste heat
Electricity for grinding	0.10	MWh	107.9	kg CO_{2e}	

10.4 Transportation

The transportation mode and distance for CCSM technology and geological CCS technology is quite different. In geological CCS technology analysis, CO_2 is captured from Fellside power plant and transported through pipeline to an offshore drilling location. The pipeline transportation distance is 1450 km [ref. 33]. In CCSM technology analysis, CO_2 is captured and sequestrated in the Fellside power plant and the mineral carbonate waste is transported using ocean barge to Ballantrae Ultramafic Rocks mine site. The ocean transportation distance is 183 km. The carbon footprint of the transportation is shown in Table 15.

Table 15. Transportation carbon footprint.

Transportation mode	Transportation distance, km	Carbon footprint, $kgCO_{2e}$/tonne CO_2 sequestrated	Comment
Pipeline transportation	1450	1.2E-03	1.4 tonne fugitive emission per year per km [Ref 34]
Ocean shipping	183	1.91	2845.9 tonne-miles/gal diesel [Ref 35]

10.5 Geological sequestration

As described in the Zero Emission Platform (ZEP) report [ref 36], there are three primary stages for geological carbon capture and storage (Geo CCS): capture, transport, and storage. The carbon footprint associated with capture stage is heavily dependent on both the type of power plant being captured from and the capture technology. Based on chapter 4.3, the energy required in CO_2 capture process is 257.3 kWh/tonne CO_2 captured and the compression energy required is 91.2 kWh at 35.5 outlet pressure. The corresponding carbon footprint of these two steps is 325 kg CO_{2e}/tonne CO_2 captured. Although leakage is a serious concern in geological CCS, no direct scenario for accounting leakage risk is provided as part of the ZEP cost study. Sensitivity analysis of the potential leakage needs to be investigated in the next step.

10.6 Result & Conclusion

In this analysis, energy, material and emission data have been collected through the whole life cycle steps of both technologies with consideration of UK geological sites, including mineral mining, transportation, mineral crush and grinding, CO_2 capture and sequestration, etc.

- The preliminary analysis has demonstrated Geological CCS technology has lower carbon footprint than CCSM technology.
- The high heating and electricity demand of the CCSM process also results in negative carbon footprint benefits of the technology. This needs to be further validated in next steps with more detailed analysis.

A few scenarios will also be further considered in next steps, including electricity grid options for each life cycle steps, uncertainty and sensitivity analysis, etc.

Life cycle assessment (LCA) is an ISO standardized process that attributes life cycle material consumption, energy use, and solid, liquid and gaseous emissions to environmental impacts. In this assessment, we have focused on quantification of life cycle carbon footprint (per tonne CO_2 sequestrated) of CCSM technology and geological CCS technology to address the impact of these technologies on future climate change risk mitigation. Energy, material and emission data have been collected through the whole life cycle steps of both technologies with consideration of UK geological sites, including mineral mining, transportation, mineral crush and grinding, CO_2 capture and sequestration, etc. Carbon footprint of unit electricity, natural gas, steam and materials are obtained from EcoInvent [ref. 30] or European Life Cycle Database (ELCD) [ref. 37] with UK specific dataset or European average dataset. The preliminary analysis has demonstrated net carbon balance benefits from both technologies. We have also identified that steam and electricity demand of the mineralisation process contributes the most to the overall carbon footprint of CCSM technology. Benefits of CCSM technology can be significantly improved when waste steam is available for the heating requirement in the CCSM process. In next steps, the preliminary analysis will be validated and a few scenarios will be further explored to finalize the analysis, including electricity grid options for each life cycle steps, waste heat energy availability, uncertainty and sensitivity analysis, etc.

11 Possible Process Development Options

The process is inherently inefficient, the major problem areas being the ammonia sulphate decomposition stage, the ammonia sulphate recovery, and the recovery and recycling of excess reagents around the precipitation stage.

Development of the process within the existing design concept is likely to yield some reduction in energy usage and capital cost, although the gains are unlikely to fundamentally alter the underlying picture. Possible areas for incremental development include:

- optimization of Mg extraction operation. e.g.
    - reactor type and configuration – a different reactor type, for example a as a filter bed system, could enable a higher feed solution concentration and/or higher bisulphate utilization
    - particle size, solution concentration and degree of utilization
    - operating temperature and materials of construction
- optimization of precipitation stage, e.g.
    - reagent use, residence time and per pass efficiency
- direct precipitation of Mg by flue gas
- refinement of the heat and power integration scheme
- cost/benefit analysis of potential energy reduction schemes, e.g.:
    - power recovery from letdown of precipitation reaction liquors
    - power generation from sulphate decomposition vapours
    - pre-heating of ore feed
    - options for combined power/heat generation associated with NG firing, and integration of heat recovery from sulphate decomposition stage

There are in addition certain design issues that require further consideration, in particular:

- materials of construction for extraction reactors
- the operation of the sulphate decomposition stage and the impact of pressure on its performance
- the filtration characteristics of the leached ore and magnesite solids, the washing efficiency, and the selection and sizing of the filtration equipment
- solids formation in the flue gas scrubber

- fate of minor components

A more radical approach is required to significantly reduce or even eliminate some of the inherent inefficiencies. Possible concepts include:

- separation of $MgSO_4$ from the leachate solution:
    - avoids neutralization of unreacted bisulphate
    - separates the carbonation and extraction solution cycles
    - possible direct reaction of ammoniated $MgSO_4$ solution with flue gas.
- application of bipolar electrodialysis as an alternative bisulphate regeneration technology:
    - simultaneous regeneration of bisulphate and alkali solutions
    - avoids evaporative crystallization of ammonia sulphate
    - low temperature
- freeze crystallization for sulphate recovery

12 Conclusions

The ammonia based mineralisation process that this group has developed is uneconomic due to its energy requirements and high capital cost Much of the process equipment is large and multi-streamed and the process will be complex to operate. The capital cost is equivalent to that of the Shell designed process, but as a result of the ammonium sulphate decomposition stage, ammonium sulphate recovery, and the recovery and recycling of excess reagents from the precipitation stage, the plant has high variable costs of operation and is unable to capture the carbon dioxide generated by the power station without significant additional energy usage. This energy usage equates to a carbon dioxide production rate which exceeds the amount that the recovery plant is designed to capture. At this stage of development of the process, therefore, it is less favourable than the Shell designed process that it has been compared to. There are a number of further technology development options that can be explored. These are listed in Section 11. It is believed that these are likely to raise the plant capital cost, but have the potential to significantly reduce its operating costs and increase its carbon capture potential.

The project team will investigate these further improvement opportunities as required to meet the needs of the stakeholder group

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Appendix 1 Summary of Revised Design Basis

Flue gas:

- 39600 kmol/h at 50°C, 1.1 bara, saturated with water
- gas composition (molar basis, dry, as per Table 1)
  - CO_2 : 14.7%,
  - N_2 : 81.0%
  - O_2 : 4.3%
  - CO_2 target removal : 80%
  - CO_2 mass rate : 251.5 Te/h, equivalent to 1.99 MTe/yr at 330 days/yr operation

Mineral Feed:

- by wt, dry (as per Table 1)
  - $Mg_3Si_2O_5(OH)_4$: 90.035%
  - Fe_3O_4 : 8.02%
  - SiO_2 : 1.95%
- size range 150-300 µm

Extraction Conditions:

- 127.5°C
- feed solution 27M NH_4HSO_4
- solid/liquid feed ratio 152.5 g/l
- target Mg extraction 75%
- Fe extraction 47%
- bisulphate utilization 90%
- residence time 60 min

Mg Precipitation

- 120°C, 20 bara
- target feed ratio [Mg]:[ammonium salts]:[free ammonia] 1:1.5:1 (molar basis)
- particle size range 30-130 µm
- Mg precipitation per pass : 85 %
- Mg precipitate is magnesite [$Mg(CO_3)$]
- residence time 30 min

Silica Product Specification

- none

Mg Product Specification

- none

Fe Precipitation Step

- not included

Utilities and Services

- steam & water as may be required
- Cooling Water:
  - supply 20°C
  - return 32°C
- N_2 as may be required

Appendix 2 Additional Process Design Information

RECOVERY OF AMMONIUM SULPHATE

In the crystallization step the energy requirement is heavily and adversely impacted by the large elevation of boiling point. For a saturated solution of ammonium sulphate at 1 ata, the boiling point is 108.5°C, an elevation of 8.5°C. For MVR, the compressor delivery pressure has to be set such that the vapour saturation temperature is sufficiently above the solution boiling point to provide an adequate temperature differential for heat transfer. Taking a minimum temperature difference of 5°C, then for an evaporator operating at 1 bara, the required saturation temperature is 113°C, with a compressor delivery pressure of 1.6 bara.

The major energy input is the power required for the MVR system. The power requirement is 64.9 MW, equating to 29 kWhr/Te of steam evaporated. As might be expected, this is at the high end of the expected range for MVR (typically 10-30 kWhr/Te). There is a shortfall in heat load provided by the compressed vapour, which is made up by heat input from LP steam. This can be added to the compressor exhaust but here it is taken to be put to a separate heating coil so that the steam condensate can be recovered.

For the multi-effect evaporation system, the elevation of boiling point also limits the number of effects that are possible. The operating pressure in the first effect is set at 0.75 bara. The number of effects used is three, with the second and third evaporators operating at 0.43 and 0.2 bara respectively. The heat input into the first effect is 375 MW. The heatload on the second and third stages increases progressively due to additional steam flashing off as the pressure is reduced (399 and 454 MW on stages 2 and 3 respectively).

AMMONIUM SULPHATE DECOMPOSITION

The design of the ammonium sulphate decomposition step is based on information in the open literature, in particular ref. 6. There are a number of other patents which also deal with decomposition of ammonium sulphate, for example ref.s 8-17. Wang (18) also reports data.

No reference has been found to any one practising the decomposition of ammonium sulphate, so there is no obvious practical experience to guide the choice of technology.

Ammonium sulphate cannot be melted at atmospheric pressure without decomposition, releasing ammonia and leaving bisulphate. The ammonia vapour pressure of pure, anhydrous ammonium sulphate is effectively zero up to 80°C, with decomposition starting at 80-100°C. At temperatures of perhaps 300-350°C, decomposition is reasonably rapid and more or less complete decomposition to ammonia and bisulphate can be attained, but at higher temperatures, particularly above around 400°C, decomposition also gives N_2, SO_2 and H_2O, as well as ammonium pyrosulphate.

The design approach described in Brennan (US3929977) uses a thin film reactor with a sweep gas, and is backed up by their reported experimental data. Amongst other possibilities, Brennan suggests using equipment such as a falling film exchanger, a wiped film exchanger and a rotary kiln. The concern with using this type of device is that given the required residence time quoted by Brennan (> 20 min) it will be very expensive to provide the required surface area. Brennan also mentions the possible use of a packed bed, which would be a far more cost-effective approach, but again it would be difficult to provide the required liquid residence time. Brennan states that the rates of decomposition using a simple heated pot are 'unsatisfactory'. Using a thin film system requires separate preparation of the melt, and is made by dissolving the sulphate in recycled molten bisulphate product. This step is essential since the melt relies on the relatively low melting point of the bisulphate (c. 145°C) to get the sulphate into the liquid phase.

The Koppers patent (6) suggests the use of a packed bed with an external recirculation of molten salt, with a closed loop nitrogen stripping gas, and recovery of ammonia by absorption in water. The heat input is by heating the stripping gas to 450°C, with a target outlet temperature of 300°C. The patent suggests using a 50% sulphate feed (taken to be by weight), with a 5 ft packed bed. Data in the patent indicates up to 100% decomposition, varying from 80-100% depending on the level of ammonia in the stripping gas.

The data in the Koppers patent clearly implies that increasing the ammonia partial pressure will decrease the degree of stripping, but that levels of up to (say) 10% in the recycled stripping gas will have only a minor effect.

Evaluation of the Koppers design shows that it is impractical for full decomposition of the bisulphate as the gas rates required to achieve the required heat input are very high, but it does nevertheless offer a basis for achieving a low residual level of sulphate.

The approach adopted here is to assume the use of a simple agitated vessel to effect the bulk of the decomposition, with a second stage gas stripping system as described in the Koppers patent to achieve high conversion of the sulphate, taken as 98% overall. Unconverted sulphate in the product bisulphate has to be recycled through the process, and consequently imposes an energy penalty, principally by increasing the energy requirement in the sulphate crystallization stage. The required residence time in the agitated melt vessel is taken as a nominal 30 min.

It should be possible to run the sulphate decomposer at pressure, although it should be expected that higher pressures will reduce rate of decomposition. Operation at pressure has particular advantages for the recovery of the ammonia. The present flowsheet assumes that this is possible at modest pressure without any significant adverse impact on the rate of decomposition. The operating pressure in the melt vessel is taken as 10 bara. Solid sulphate is transferred into the decomposition reactor from a feed hopper via a rotary valve.

The ammonia liquor is used at low pressure. If the decomposition can be run at pressure then it is possible to generate power from expansion of the product vapours. For example, the vapour stream from the melt vessel at 10 bara could generate approximately 22 MW if let down to low pressure (e.g. the flue gas scrubber operating pressure). This option has not been further evaluated.

SEPARATION OF LEACHED MINERAL

The density of serpentine is about 2600 kg/m^3. Leaching removes 56% of the weight of the ore, leaving behind silica. It is understood that, after leaching, the original particle dimensions are retained, with the dissolution process behaving as a shrinking core, where a reaction front moves progressively into the particle. The degree of Mg extraction assumed for design is 75%, so nominally 25% of the feed ore is unreacted, and this material is the residual core of the leached particles. The feed material has a range of sizes (150-300 μm), so the smaller particles will have a higher level of extraction, and the larger particles less. The smallest particles will therefore be more or less only silica, the largest with significantly more than 25% unreacted mineral remaining. A size distribution of the material is not available.

The density of silica (as quartz) is 2200 kg/m^3. Based on this figure, the residual solid matrix has a voidage fraction of 48%, assuming that the initial material has a voidage of zero. An 'average' particle, with 25% of the original material unreacted, will thus have an outer ring with of voidage 48%, which will be filled with solution, of density c. 1.1 kg/m^3. The effective particle density is 1543 kg/m^3 based on the overall particle dimensions. The overall voidage is 35%. The volume of liquid within the particle is 0.23 litre per kg of residual solid, approximately 0.25 kg solution per kg solids, or 20% liquor by weight.

The dewatering behaviour of this material is not known. The mass balance assumes that liquor within the leached matrix mixes freely with the bulk fluid and hence that the material can be washed effectively. The residual liquor in the leached ore stream after filtration is taken as c. 15% by wt.

Efficient washing will be essential to limit the loss of ammonia sulphate. If the liquor is held tightly within the pores then it may be necessary to use reslurrying to affect efficient washing.

A belt filter has been provisionally selected for this duty for the following reasons:

- high wash efficiencies can be achieved, by using multiple stages of counter current washing - wash efficiencies are typically up to 90% but can be as high as 99.8%
- the mineral may be too dense and/or too large to form a good cake on a drum type filter
- belt filters can be optimised in use, for example by adjusting the position of trays, the belt speed ; the belt design can also be changed

The residual wetness can be reduced to perhaps 0.1% w/w with hot air/gas/steam drying but as there is no requirement for the product to be dried, the cake will be left wet. Without a drying step, the residual wetness for a belt filter is typically 10-15%, although it might be as high as 25%.

It is assumed that the particles are sufficiently robust that there is no breakage either in the extraction reactors or in the filtration stage.

The washing efficiency is calculated using the simple approximation of full mixing of the residual cake liquor with the applied wash water.

SEPARATION OF PRECIPITATED MAGNESITE

A rotary vacuum filter is taken to be suitable for filtration of the magnesite slurry, in that the solids are assumed to form a reasonable cake. Washing of the magnesite cake is important, and this may be an issue for this type of filtration device. It may be necessary to use reslurrying and re-filtration to affect efficient washing.

MATERIALS OF CONSTRUCTION

The operating conditions in the extraction vessels are highly corrosive, with a pH of around 1, and the slurry may be abrasive. Serpentine has a hardness of 3-4 on the Mohs scale, steel is 4.5, rising to 8 if hardened. No direct information on the suitability of materials for ammonium bisulphate has been found, but if the solution is judged to be similar to dilute sulphuric acid then the availability of suitable materials may limit the either the operating temperature or solution strength, or both.

In sulphuric acid corrosion, the level of O_2 present is a significant factor. The presence of Fe^{3+} is also understood to significantly improve the corrosion resistance of alloys. It appears that there are suitable alloys available capable of operating at the extraction temperature with acceptable rates of corrosion, but these will be expensive. For cost estimating purposes, the extraction reactors are assumed to require alloys such as LEWMAT alloy 66 or ZERON 100 (which is a superduplex). If further development of the process is envisaged then a thorough review of the materials options for the reactor should be undertaken.

Downstream of the extraction reactors the use of standard stainless steels or carbon steel is assumed.

CONTROL AND OPERABILITY

Control of the process raises some significant issues, particularly with regard to the management of sulphate and ammonia inventories within the process. There are large recycles of these materials, recycles which are essentially ‘closed’ in that there are only minimal losses of sulphate and ammonia from the process. It is also necessary to match leaching rates of magnesium with carbonate production in the flue gas scrubber and the performance of the magnesite precipitation step.

No provision is made in the process design for intermediate storage of chemicals.

Other metals will also be leached from the ore. The fate of minor components within the process and the possible build-up of impurities within the process have not been considered.

EXTRACTION KINETICS

The extraction data for ET16 samples was analyzed with the intention of deriving a simple kinetic model. However, within the time available this has not proved possible to satisfactorily fit such a model to the existing body of data. Work on possible alternative models was not undertaken.

The kinetic model evaluated is that of a shrinking core, with the active area equal to the surface area of the core. The reaction rate at the surface was taken to be a simple power law in the bulk bisulphate concentration. This assumes that there is no significant mass transfer limitation either within the particle or external to the particle, and also that the material can be characterized as a sphere with a diameter equal to the arithmetic mean of the given size range. Although this can give a very good fit to individual sub-sets of the experimental data, a reasonable overall fit to the full body of the data is not possible, and it is particularly noticeable that such fits tend to significantly over predict extraction levels at high extraction times; the level of extraction in the data seems to flatten off at a surprisingly modest level (typically c. 80% Mg extraction). The maximum extraction reported is 87% after 3 hours using a 3M solution. Only around 30% of the initial bisulphate has been used at this degree of extraction and the residual bisulphate concentration is 2.1 M (i.e. higher than the starting concentration used in other runs in that set).

There are a number of possible factors at work here, namely:

- the tests use a spread of particle sizes
- the shape factor of the particles is not defined (for example, they may be more plate-like than spheres)
- the material may be anisotropic, for example with differing properties along different planes within the mineral structure
- there may be a significant diffusion resistance to mass transfer within the particles
- it is assumed that all the Mg can be leached out. If some of the Mg is locked up this could explain the limited extraction at long reaction times and might also explain the differing extraction behaviour of otherwise similar minerals
- there may be limitations in the experimental technique (for example, the withdrawal of representative liquid samples)

The performance of the extraction step is based on the predicted behaviour using the kinetic model as fitted to one set of data for the selected size range (150-300 µm) coupled with the temperature dependency as derived from another set (at 50-100°C). The extraction conditions were selected to retain a residence time of 60 min and to keep the Mg extraction = < 80%, due to concerns that extraction levels higher than this may be questionable.

Appendix 3　Process Flow Diagrams

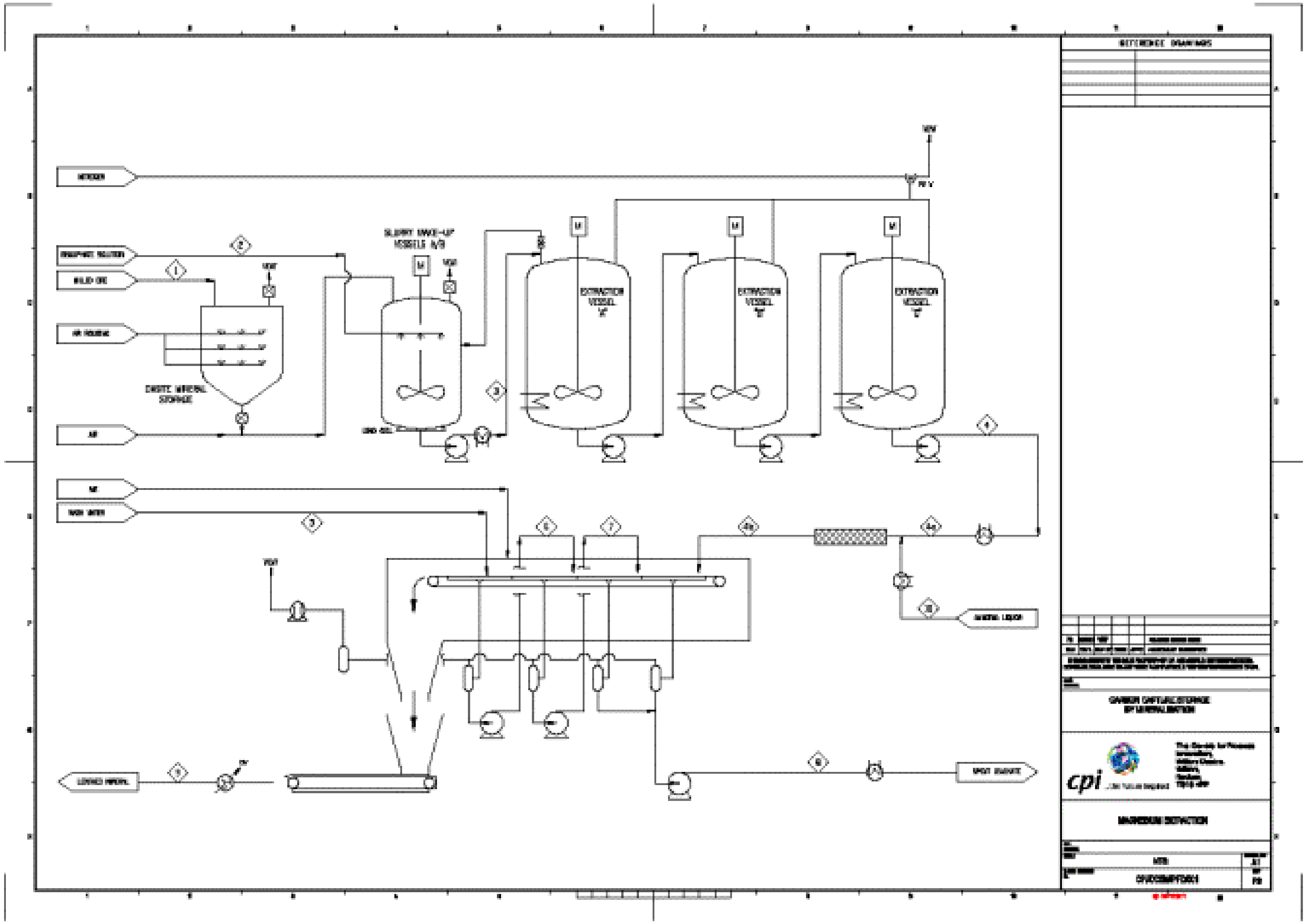

Figure 7. Process Flow Diagram - Magnesium Extraction.

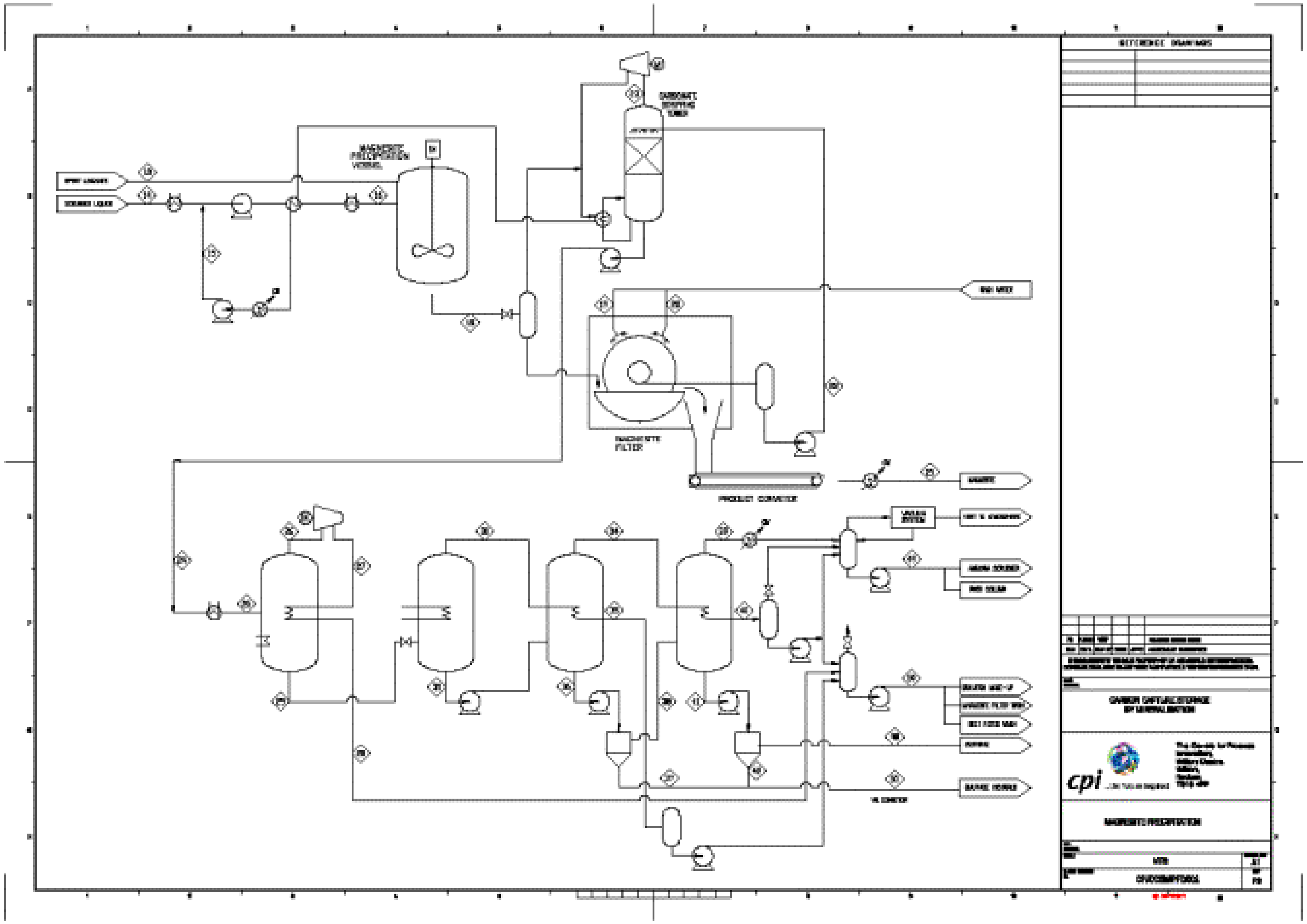

Figure 8. Process Flow Diagram - Sulphate Decomposition and Flue Gas Scrubbing.

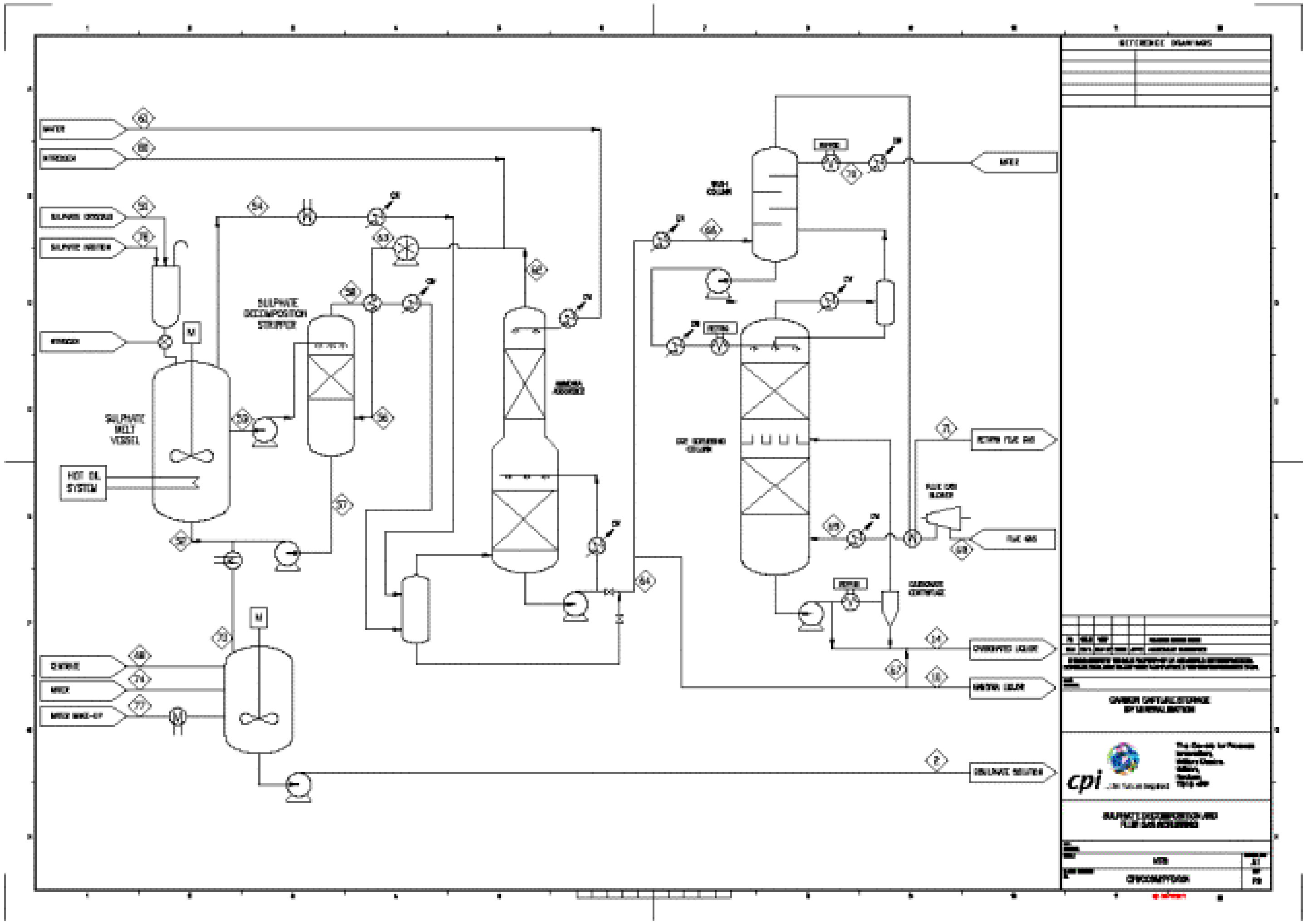

Figure 9. Process Flow Diagram - Sulphate Decomposition and Flue Gas Scrubbing.

Appendix 4 Process Stream Lists

Stream Number		1	2	3	4	4a	4b	5	6	7	8	9	10	13	14	15
Stream Description		Ore	Bisulphate	Heated Feed	Leached Exit	Cooled Slurry2	BF Feed	BF Wash Water	BF S2 Wash Rec	BF S1 Wash Rec	HP Leachate	Export Mineral	Fe AmmLiq	Reheated Leachate	Mixed CarbLiq	Recycle Carbonate
H_2O	Te/hr	0.0	3513.5	3513.5	3658.1	3658.1	3728.7	70.0	64.4	58.9	3733.0	65.7	49.9	3733.0	641.6	646.4
N_2	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O_2	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO_2	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	1.3	0.0
NH_3	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$(NH_4)_2SO_4$_AQ	Te/hr	0.0	148.1	148.1	806.0	806.0	1012.0	0.0	3.4	6.7	1008.5	3.4	0.0	1008.5	0.0	0.0
NH_4HSO_4_AQ	Te/hr	0.0	1278.7	1278.7	132.6	132.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH_4OH	Te/hr	0.0	0.0	0.0	0.0	0.0	1.2	0.0	0.0	0.0	1.2	0.0	70.1	1.2	128.2	147.0
$MGSO_4$_AQ	Te/hr	0.0	96.2	96.2	646.5	646.5	646.5	0.0	2.2	4.3	644.3	2.2	0.0	644.3	0.0	0.0
$FESO_4$_AQ	Te/hr	0.0	0.0	0.0	15.5	15.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$FE_2(SO_4)_3$_AQ	Te/hr	0.0	0.0	0.0	40.7	40.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$(NH_4)_2CO_3$_AQ	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	87.1	131.7
NH_4HCO_3_AQ	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	160.7	108.3
SERPENTINE	Te/hr	563.0	0.0	563.0	140.8	140.8	140.8	0.0	0.0	0.0	0.0	140.8	0.0	0.0	0.0	0.0
QUARTZ	Te/hr	12.2	0.0	12.2	195.3	195.3	195.3	0.0	0.0	0.0	0.0	195.3	0.0	0.0	0.0	0.0
MAGNESITE	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$FE(OH)_3$_PPT	Te/hr	0.0	0.0	0.0	0.0	0.0	21.8	0.0	0.1	0.1	21.7	0.1	0.0	21.7	0.0	0.0
$FE(OH)_2$_PPT	Te/hr	0.0	0.0	0.0	0.0	0.0	9.1	0.0	0.0	0.1	9.1	0.0	0.0	9.1	0.0	0.0
FE_3O_4	Te/hr	50.2	0.0	50.2	26.6	26.6	26.6	0.0	0.0	0.0	0.0	26.6	0.0	0.0	0.0	0.0
$(NH_4)_2SO_4$_SOLID	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH_4HCO_3_SOLID	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	126.0	0.0
NH_4HSO_4_MELT	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$(NH_4)_2SO_4$_MELT	Te/hr	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL	Te/hr	625.4	5036.6	5661.9	5661.9	5661.9	5781.9	70.0	70.0	70.0	5417.8	434.0	120.0	5417.8	1144.9	1033.4
Vapour fraction	w/w	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Solids %	w/w	100.0	0.0	11.0	6.4	6.4	6.8	0.0	0.1	0.3	0.6	83.6	0.0	0.6	11.0	0.0
Temperature	°C	15.0	113.8	122.5	127.5	95.0	100.9	103.4	102.9	102.3	101.6	40.0	40.0	122.5	26.0	65.1
Pressure	bara	1.00	2.00	3.00	3.00	2.80	2.80	2.00	1.30	1.30	20.00	1.00	5.00	19.80	5.00	5.00
Fluid Mol. Wt	kg/kmol	0.0	24.2	24.2	24.5	24.5	24.5	18.0	19.3	20.8	24.5	19.3	25.2	24.5	24.1	24.1
Density	kg/m³	2010.8	1109.6	1161.0	1165.0	1189.5	1183.0	956.0	999.4	1046.0	1148.8	1778.8	891.5	1132.8	711.0	1061.5
Volume	m3/hr	311.0	4539.0	4876.8	4860.2	4759.8	4887.3	73.2	70.0	66.9	4716.1	244.0	134.6	4782.6	1610.2	973.6
Enthalpy	MW	-2580.8	-18890.8	-21410.8	-21570.4	-21732.3	-22146.5	-302.1	-293.1	-284.3	-20680.0	-1775.7	-419.3	-20575.3	-4278.0	-3904.3

Notes :

1. enthalpy basis elements in std states

Status			**Client**	ETI
Revision		**Date**	**Project**	Mineralisation of CO_2
Prepared	ICJeffery	21-Mar-12	**Case**	Magnesite Product
Checked			**PFD**	
Approved			**Sheet**	1 of 6

Stream Number		16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
Stream Description		HP CarbLiq		Exit Carbonation	Rovac WW1	Rovac WW2	Magnesite Expo	RV Combined Filtrat	Decomp Carb Vap	AS Liquor Feed	Heated Feed	Flash Vapour	Exit Compr	Condensate	Flash Bottoms	Vap1
H_2O	Te/hr	1288.0		5064.4	75.0	75.0	91.6	4952.8	771.3	4498.8	4498.9	2273.6	2273.6	2273.6	2225.3	624.4
N_2	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
O_2	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
CO_2	Te/hr	1.3		1.3	0.0	0.0	0.0	0.0	120.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH_3	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	141.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$(NH_4)_2SO_4$_AQ	Te/hr	0.0		1609.7	0.0	0.0	7.8	1601.9	0.0	1601.9	1602.2	0.0	0.0	0.0	1602.2	0.0
NH_4HSO_4_AQ	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH_4OH	Te/hr	275.4		192.2	0.0	0.0	0.7	150.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$MGSO_4$_AQ	Te/hr	0.0		96.6	0.0	0.0	0.5	96.2	0.0	96.2	96.2	0.0	0.0	0.0	96.2	0.0
$FESO_4$_AQ	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$FE_2(SO_4)_3$_AQ	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$(NH_4)_2CO_3$_AQ	Te/hr	218.4		12.7	0.0	0.0	0.1	12.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH_4HCO_3_AQ	Te/hr	394.8		204.4	0.0	0.0	0.4	80.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
SERPENTINE	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
QUARTZ	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
MAGNESITE	Te/hr	0.0		383.6	0.0	0.0	383.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$FE(OH)_3$_PPT	Te/hr	0.0		21.7	0.0	0.0	21.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$FE(OH)_2$_PPT	Te/hr	0.0		9.1	0.0	0.0	9.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
FE_3O_4	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$(NH_4)_2SO_4$_SOLID	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH_4HCO_3_SOLID	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
NH_4HSO_4_MELT	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
$(NH_4)_2SO_4$_MELT	Te/hr	0.0		0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
TOTAL	Te/hr	2177.9		7595.7	75.0	75.0	515.5	6894.0	1033.4	6196.8	6197.3	2273.6	2273.6	2273.6	3923.7	624.4
Vapour fraction	w/w	0.0		0.0	0.0	0.0	0.0	0.0	1.0	0.0	0.0	1.0	1.0	0.0	0.0	1.0
Solids %	w/w	0.0		5.5	0.0	0.0	80.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Temperature	°C	122.5		119.6	103.4	103.4	40.0	92.0	81.2	81.1	92.4	105.3	159.9	113.5	105.3	99.9
Pressure	bara	20.00		19.80	2.00	2.00	1.00	1.00	0.40	2.00	2.00	1.00	1.60	1.60	1.00	0.75
Fluid Mol. Wt	kg/kmol	25.1		23.8	18.0	18.0	19.5	23.5	19.2	23.6	23.6	18.0	18.0	18.0	28.8	18.0
Density	kg/m³	822.5		1025.3	956.0	956.0	2183.1	1088.6	0.3	1110.8	1103.8	0.6	0.8	948.6	1197.1	0.4
Volume	m3/hr	2647.8		7408.2	78.5	78.5	236.1	6333.0	3934188.4	5578.6	5614.4	3939480.6	2814812.8	2396.9	3277.6	1424482.2
Enthalpy	MW	-8015.1		-28590.4	-323.7	-323.7	-1892.8	-26343.7	-3253.2	-23757.9	-23691.9	-8384.9	-8320.1	-9785.7	-13807.0	-2304.4

Notes :

1. enthalpy basis elements in std states

Status			**Client**	ETI		
Revision		**Date**	**Project**	Mineralisation of CO_2		
Prepared	ICJeffery	21-Mar-12	**Case**	Magnesite Product		
Checked			**PFD**			
Approved			**Sheet**	2	of	6

Stream Number		31	32	33	34	35	36	37	38	39	40	41	42	43	44	45
Stream Description		Liq1			Vap2	Condy1	Liq2	Stage2 Xtal	C3 Feed	Vap3	Condy 2	Liq3	Stage3 Xtal			Condy3
H_2O	Te/hr	1600.9			700.9	624.4	900.0	9.2	140.8	750.0	700.9	140.8	7.3			750.0
N_2	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
O_2	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
CO_2	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
NH_3	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$(NH_4)_2SO_4$_AQ	Te/hr	1602.2			0.0	0.0	873.8	8.9	128.5	0.0	0.0	128.5	6.7			0.0
NH_4HSO_4_AQ	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
NH_4OH	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$MGSO_4$_AQ	Te/hr	96.2			0.0	0.0	96.2	1.0	95.2	0.0	0.0	95.2	5.0			0.0
$FESO_4$_AQ	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$FE_2(SO_4)_3$_AQ	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$(NH_4)_2CO_3$_AQ	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
NH_4HCO_3_AQ	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
SERPENTINE	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
QUARTZ	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
MAGNESITE	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$FE(OH)_3$_PPT	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$FE(OH)_2$_PPT	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
FE_3O_4	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$(NH_4)_2SO_4$_SOLID	Te/hr	0.0			0.0	0.0	728.4	728.2	736.6	0.0	0.0	736.7	736.6			0.0
NH_4HCO_3_SOLID	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
NH_4HSO_4_MELT	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
$(NH_4)_2SO_4$_MELT	Te/hr	0.0			0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0			0.0
TOTAL	Te/hr	3299.3			700.9	624.4	2598.4	747.2	1101.2	750.0	700.9	1101.2	755.7			750.0
Vapour fraction	w/w	0.0			1.0	0.0	0.0	0.0	0.0	1.0	0.0	0.0	0.0			0.0
Solids %	w/w	0.0			0.0	0.0	28.0	97.5	66.9	0.0	0.0	66.9	97.5			0.0
Temperature	°C	99.9			84.8	91.8	84.8	84.9	68.7	68.6	77.4	68.6	68.7			60.1
Pressure	bara	0.75			0.43	0.75	0.43	2.00	2.00	0.20	0.43	0.20	2.00			0.20
Fluid Mol. Wt	kg/kmol	32.4			18.0	18.0	32.6	32.6	38.0	18.0	18.0	38.0	38.0			18.0
Density	kg/m³	1261.1			0.3	963.9	1386.7	1744.1	1647.0	0.1	972.8	1647.0	1754.4			982.2
Volume	m3/hr	2616.3			2713608.7	647.8	1873.8	428.4	668.6	5901749.3	720.5	668.6	430.7			763.6
Enthalpy	MW	-11127.7			-2591.9	-2703.3	-8136.7	-1855.3	-3047.2	-2779.8	-3046.2	-3047.2	-1880.9			-3274.9

Notes :

1. enthalpy basis elements in std states

Status			**Client**	ETI		
Revision		**Date**	**Project**	Mineralisation of CO_2		
Prepared	ICJeffery	21-Mar-12	**Case**	Magnesite Product		
Checked			**PFD**			
Approved			**Sheet**	3	of	6

Stream Number		46	47	48	49	50	51	52	53	54	55	56	57	58	59	60
Stream Description				Final Centrate	HP Cold Condy	Mixed HotCond	Recd Sulphate	Recycle melt	Melt to Stripper	Melt Gas		Stripping gas	Stripped Melt	Stripped NH_3		
H_2O	Te/hr			129.3	850.8	3492.7	25.9	0.0	0.0	25.9		0.3	0.0	0.4		
N_2	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		99.7	0.0	99.7		
O_2	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
CO_2	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
NH_3	Te/hr			0.0	0.0	0.0	0.0	0.0	0.1	159.2		0.0	0.0	29.9		
$(NH_4)_2SO_4$_AQ	Te/hr			117.6	0.0	0.0	24.5	0.0	0.0	0.0		0.0	0.0	0.0		
NH_4HSO_4_AQ	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
NH_4OH	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
$MGSO_4$_AQ	Te/hr			89.1	0.0	0.0	7.1	10.7	17.8	0.0		0.0	17.8	0.0		
$FESO_4$_AQ	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
$FE_2(SO_4)_3$_AQ	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
$(NH_4)_2CO_3$_AQ	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
NH_4HCO_3_AQ	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
SERPENTINE	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
QUARTZ	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
MAGNESITE	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
$FE(OH)_3$_PPT	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
$FE(OH)_2$_PPT	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
FE_3O_4	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
$(NH_4)_2SO_4$_SOLID	Te/hr			0.0	0.0	0.0	1459.8	0.0	0.0	0.0		0.0	0.0	0.0		
NH_4HCO_3_SOLID	Te/hr			0.0	0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	0.0		
NH_4HSO_4_MELT	Te/hr			0.0	0.0	0.0	0.0	1942.7	3019.2	0.0		0.0	3221.1	0.0		
$(NH_4)_2SO_4$_MELT	Te/hr			0.0	0.0	0.0	0.0	46.6	308.9	0.0		0.0	77.2	0.0		
TOTAL	Te/hr			336.0	850.8	3492.7	1517.3	2000.0	3346.1	185.1		100.0	3316.1	129.9		
Vapour fraction	w/w			0.0	0.0	0.0	0.0	0.0	0.0	1.0		1.0	0.0	1.0		
Solids %	w/w			0.0	0.0	0.0	96.2	0.0	0.0	0.0		0.0	0.0	0.0		
Temperature	°C			68.7	40.3	103.4	77.3	333.0	360.0	360.0		300.0	333.0	333.0		
Pressure	bara			2.00	10.00	2.00	2.00	10.00	10.00	10.00		10.10	10.00	10.00		
Fluid Mol. Wt	kg/kmol			38.2	18.0	18.0	34.1	119.4	117.0	17.2		28.0	119.4	24.4		
Density	kg/m³			1455.7	991.1	956.0	1741.0	1789.1	1783.3	3.3		5.9	1789.1	4.8		
Volume	m3/hr			230.8	858.5	3653.6	871.5	1117.9	1876.3	56456.1		16920.3	1853.5	26894.7		
Enthalpy	MW			-1133.8	-3734.5	-15073.8	-3789.9	-4586.8	-7620.7	-176.1		6.8	-7605.2	-8.7		

Notes :

1. enthalpy basis elements in std states

Status			**Client**	ETI		
Revision		**Date**	**Project**	Mineralisation of CO_2		
Prepared	ICJeffery	21-Mar-12	**Case**	Magnesite Product		
Checked			**PFD**			
Approved			**Sheet**	4	of	6

Stream Number		61	62	63	64		66	67	68	69	70	71	72	73	74	75
Stream Description		Absorber Water	Absorber Top	Exit Blower	Amm Liq Cut Stream		Cooled Scrub AmmLiq	Bypass AmmLiq	FG	FG Exit Cooler	Cooled Scrub Water	Return FG		Cooled Bisulphate	Rediss Water	
H_2O	Te/hr	450.8	0.3	0.3	276.8		142.0	84.9	13.1	13.1	400.0	14.8		0.0	3272.7	
N_2	Te/hr	0.0	99.6	99.6	0.0		0.0	0.0	882.0	882.0	0.0	882.0		0.0	0.0	
O_2	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	53.5	53.5	0.0	53.5		0.0	0.0	
CO_2	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	251.5	251.5	0.0	50.7		0.0	0.0	
NH_3	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
$(NH_4)_2SO_4$_AQ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
NH_4HSO_4_AQ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
NH_4OH	Te/hr	0.0	0.0	0.0	389.0		199.6	119.3	0.0	0.0	0.0	0.0		0.0	0.0	
$MGSO_4$_AQ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		7.1	0.0	
$FESO_4$_AQ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
$FE_2(SO_4)_3$_AQ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
$(NH_4)_2CO_3$_AQ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
NH_4HCO_3_AQ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
SERPENTINE	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
QUARTZ	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
MAGNESITE	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
$FE(OH)_3$_PPT	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
$FE(OH)_2$_PPT	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
FE_3O_4	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
$(NH_4)_2SO_4$_SOLID	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
NH_4HCO_3_SOLID	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
NH_4HSO_4_MELT	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		1278.4	0.0	
$(NH_4)_2SO_4$_MELT	Te/hr	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		30.7	0.0	
TOTAL	Te/hr	450.8	99.9	99.9	665.8		341.6	204.2	1200.1	1200.1	400.0	1001.0		1316.1	3272.7	
Vapour fraction	w/w	0.0	1.0	1.0	0.0		0.0	0.0	1.0	1.0	0.0	1.0		0.0	0.0	
Solids %	w/w	0.0	0.0	0.0	0.0		0.0	0.0	0.0	0.0	0.0	0.0		0.0	0.0	
Temperature	°C	40.3	40.0	49.2	40.0		25.0	40.0	50.0	25.0	25.0	50.0		160.0	103.4	
Pressure	bara	10.00	9.50	10.30	5.00		4.80	5.00	1.10	1.20	10.00	1.10		10.00	2.00	
Fluid Mol. Wt	kg/kmol	18.0	28.0	28.0	25.2		25.2	25.2	30.3	30.3	18.0	28.5		119.4	18.0	
Density	kg/m³	991.1	10.2	10.8	891.5		893.3	891.5	1.2	1.5	996.6	1.2		1790.0	956.0	
Volume	m3/hr	454.9	9765.8	9275.0	746.8		382.4	229.0	966518.3	817072.5	401.4	857643.0		735.3	3423.5	
Enthalpy	MW	-1978.8	-0.9	-0.6	-2326.4		-1199.8	-713.5	-665.3	-673.7	-1762.8	-173.9		-3144.2	-14124.3	

	Notes : 1. enthalpy basis elements in std states	**Status**			**Client**	ETI		
		Revision		**Date**	**Project**	Mineralisation of CO_2		
		Prepared	ICJeffery	21-Mar-12	**Case**	Magnesite Product		
		Checked			**PFD**			
		Approved			**Sheet**	5	of	6

Stream Number		76	77	78	79	80	81	82	83	84	85	86	87	88	89	90
Stream Description			Water MU	Sulphate Feed		N2 MU										
H_2O	Te/hr		111.5	0.0		0.0										
N_2	Te/hr		0.0	0.0		0.1										
O_2	Te/hr		0.0	0.0		0.0										
CO_2	Te/hr		0.0	0.0		0.0										
NH_3	Te/hr		0.0	0.0		0.0										
$(NH_4)_2SO_4$_AQ	Te/hr		0.0	0.0		0.0										
NH_4HSO_4_AQ	Te/hr		0.0	0.0		0.0										
NH_4OH	Te/hr		0.0	0.0		0.0										
$MGSO_4$_AQ	Te/hr		0.0	0.0		0.0										
$FESO_4$_AQ	Te/hr		0.0	0.0		0.0										
$FE_2(SO_4)_3$_AQ	Te/hr		0.0	0.0		0.0										
$(NH_4)_2CO_3$_AQ	Te/hr		0.0	0.0		0.0										
NH_4HCO_3_AQ	Te/hr		0.0	0.0		0.0										
SERPENTINE	Te/hr		0.0	0.0		0.0										
QUARTZ	Te/hr		0.0	0.0		0.0										
MAGNESITE	Te/hr		0.0	0.0		0.0										
$FE(OH)_3$_PPT	Te/hr		0.0	0.0		0.0										
$FE(OH)_2$_PPT	Te/hr		0.0	0.0		0.0										
FE_3O_4	Te/hr		0.0	0.0		0.0										
$(NH_4)_2SO_4$_SOLID	Te/hr		0.0	13.9		0.0										
NH_4HCO_3_SOLID	Te/hr		0.0	0.0		0.0										
NH_4HSO_4_MELT	Te/hr		0.0	0.0		0.0										
$(NH_4)_2SO_4$_MELT	Te/hr		0.0	0.0		0.0										
TOTAL	Te/hr		111.5	13.9		0.1										
Vapour fraction	w/w		0.0	0.0		1.0										
Solids %	w/w		0.0	100.0		0.0										
Temperature	°C		70.0	25.0		20.0										
Pressure	bara		20.00	10.00		20.00										
Fluid Mol. Wt	kg/kmol		18.0	0.0		28.0										
Density	kg/m³		977.0	1772.3		23.2										
Volume	m3/hr		114.1	7.8		4.2										
Enthalpy	MW		-485.7	-34.4		0.0										

	Notes : 1. enthalpy basis elements in std states	**Status**			**Client**	ETI		
		Revision		**Date**	**Project**	Mineralisation of CO_2		
		Prepared	ICJeffery	21-Mar-12	**Case**	Magnesite Product		
		Checked			**PFD**			
		Approved			**Sheet**	6	of	6

Appendix 5 Capital Cost Data

COSTS BY MPI

N.B. Costs are total costs per duty, not per item

Table 16. Pumps.

Item No.	Title	No. Off	MPI Cost (£k)	Installed Cost (£k)
J-101	Slurry Feed Pump	3	78	290
J-102	Reactor Transfer Pumps	27	519	2104
J-103	Leachate Transfer Pump	3	393	955
J-104	Pptn Vessel Feed Pump	3	150	466
J-105	Carbonate Solution Return Pump	3	50	230
J-106	Stripper Feed Pump	3	73	296
J-107	AS Feed Pump	20	261	1264
J-108	2nd Stage Feed Pump	10	92	470
J-109	2nd Centrifuge Feed Pump	50	380	2023
J-110	3rd Centrifuge Feed Pump	50	373	2002
J-111	2nd Stage Condensate Pump	3	19	119
J-112	3rd Stage Condensate Pump	3	44	211
J-113	Cold Water Recycle Pump	3	112	395
J-114	Hot Water Recycle Pump	3	69	285
J-115	AS Stripper Feed Pump	3	42	187
J-116	Melt Recycle Pump	3	54	225
J-117	Bisulphate Return Pump	3	136	436
J-118	Ammonia Liquor Pump	3	76	303
J-119	Wash Column Bottoms Pump	3	24	140
J-120	Scrubbing Column Bottoms Pump	10	83	513
J-301	Hot Oil Recirc Pumps	3	325	840
J-401	CW Circulation Pump	5	911	2157
		Total	**4264**	**15909**

Table 17. Exchangers.

Item No.	**Title**	**No. Off**	**MPI Cost (£k)**	**Installed Cost (£k)**
E-101	Feed Slurry Interchanger	1	3519	4420
E-102	1st Leachate Reheater	1	1498	2102
E-103	AS Recovery Preheater	1	527	875
E-104	Fe Ammonia Liquor Heater	1	34	121
E-105	Leached Ore Cooler	1	122	299
E-106	Magnesite Product Cooler	1	149	345
E-107	Carbonate Stripper Reboiler	8	8663	13701
E-108	1st Recovered Solution Reheater	4	250	717
E-109	Scrubber Liquor Heater	1	209	425
E-110	Carbonate Solution Return Condense	8	765	2080
E-111	3rd Stage Condensate Cooler	1	19	93
E-112	3rd Stage Overheads Condenser	1	2041	2968
E-113	Melt Cooler	2	334	719
E-114	Melt Vapour Cooler	1	95	269
E-115	M/U Water Heater	1	36	124
E-116	Melt Vapour Condenser	1	62	202
E-117	Stripper Interchanger	1	58	171
E-118	Stripper Gas Cooler	1	37	142
E-119	Liquor Recirc Cooler	1	208	441
E-120	Ammonia Feed Cooler	1	56	176
E-121	Wash Water Cooler	1	61	185
E-122	Wash Bottoms Cooler	1	94	249
E-123	FG Cooler	1	116	276
E-124	FG Interchanger	1	91	233
E-125	Scrubbing Column Oheads Refrig Cooler	4	1431	2855
E-126	Wash Water Refrigeration Cooler	1	142	364
E-127	Wash Bottoms Refrig Cooler	1	228	508
E-128	Recycle Liquor Refrig Cooler	4	428	1198
E-129	LP Refrig CW Condenser	1	1425	2014
E-130	HP Refrig CW Condenser	1	767	1270
E-207	2nd Effect Heater	20	2718	6949
E-208	3rd Effect Heater	20	2495	6536
C-101	Leachate Reheat Coils	4	1217	2104

C-102	Recovered Soln Reheat Coils	5	1440	2521
C-103	LP Steam Coils	2	323	650
C-109	Melt Vessel Coils	2	1432	2106
		Total	**33089**	**60407**

Table 18. Vessels.

Item No.	**Title**	**No. Off**	**MPI Cost (£k)**	**Installed Cost (£k)**
F-105	Magnesite Flash Vessel	10	470	1657
F-106	Filter C/P	10	1736	4019
F-107	2nd Stage Condy C/P	1	177	407
F-108	Hot Condy Pump Feed C/P	10	347	1654
F-109	3rd Stage Condy C/P	1	58	223
F-110	Cold Condy C/P	1	64	238
F-113	Bisulphate Solution Vessel	2	1051	1708
F-111	Ammonia Liquor C/P	1	61	230
F-112	Scrubber O/Heads C/P	10	1875	4246
D-101	Slurry Make-Up Vessels	2	1879	3079
D-102	Extraction Reactors	9	6496	11555
D-103	Magnesite Precipitation Vessels	10	26198	36057
D-104	Melt Vessel	2	4475	6355
A-101	Carbonate Stripping Tower	8	10704	16954
A-102	AS Decomposition Stripper	1	886	1543
A-103	Ammonia Absorber	1	2887	4045
A-104	Wash Column	1	2178	3168
A-105	CO2 Scrubber	4	6896	10603
V-105	AS MVR Evaporator	12	64975	84170
V-106	AS 1st Effect Evaporator	4	11373	15832
V-107	2nd Effect Crystallizer	20	31280	44496
V-108	3rd Effect Crystallizer	20	31457	44712
V-101	Feed Storage Bin	2	98	182
V-104	AS Lock Hopper	2	55	104
		Total	**207678**	**297238**

Table 19. Compressors / Agitators / Centrifuges / Filters / Conveyors.

Item No.	Title	No. Off	MPI Cost (£k)	Installed Cost (£k)
J-201	Carbonate Stripper Vapour Compress	8	18011	34625
J-202	AS Recovery Vapour Compressor	12	34721	63481
J-203	AS Stripper Gas Blower	1	423	1258
J-204	FG Blower	1	2694	4994
J-205	Refrigeration Compressor Set	1	3376	5996
A-101	Slurry Agitator	2	338	499
A-102	Extraction Agitator	9	1783	2613
A-103	Magnesite Agitator	10	1215	1997
A-104	Melt Agitator	2	287	434
A-105	Bisulphate Agitator	2	228	357
J-207	2nd Stage Centrifuges	20	4780	15223
J-208	3rd Stage Centrifuges	20	4813	15289
G-101	Vacuum Belt Filters	10	28488	37972
G-102	Rotary Vacuum Filters	100	74017	121990
X-101	Feed Ore Conveyor	2	31	113
X-102	AS Feed Conveyor	4	63	226
X-103	Leached Ore Conveyor	1	131	264
X-104	Magnesite Conveyor	1	131	264
		Total	**175530**	**307595**

Table 20. Other.

Item No.	Title	No. Off	MPI Cost (£k)	Installed Cost (£k)
H-301	Hot Oil Heater	6	17180	24503
K-401	Cooling Water Tower	4	11097	11097
		Total	**28277**	**35600**

SIZING DATA FOR KEY ITEMS

Table 21. Heat Transfer Surface Areas (By duty).

Item No.	Title	Heatload (MW)	Area (m2)
E-101	Feed Slurry Interchanger	60.8	14757
E-102	1st Leachate Reheater	43.9	9306
E-103	AS Recovery Preheater	57.8	4345
E-104	Fe Ammonia Liquor Heater	55.0	194
E-105	Leached Ore Cooler	10.5	798
E-106	Magnesite Product Cooler	12.5	995
E-107	Carbonate Stripper Reboiler	360.0	43369
E-108	1st Recovered Solution Reheater	32.6	3608
E-109	Scrubber Liquor Heater	50.3	1765
E-110	Carbonate Solution Return Condense	240.6	6698
E-111	3rd Stage Condensate Cooler	4.4	144
E-112	3rd Stage Overheads Condenser	509.3	16350
E-113	Melt Cooler	125.8	2017
E-114	Melt Vapour Cooler	30.8	821
E-115	M/U Water Heater	12.3	207
E-116	Melt Vapour Condenser	21.0	782
E-117	Stripper Interchanger	7.4	202
E-118	Stripper Gas Cooler	7.2	180
E-119	Liquor Recirc Cooler	62.5	2677
E-120	Ammonia Feed Cooler	6.2	694
E-121	Wash Water Cooler	7.1	794
E-122	Wash Bottoms Cooler	8.2	1283
E-123	FG Cooler	7.2	1040
E-124	FG Interchanger	6.1	792
E-125	Scrubbing Column Overheads Refrig Cooler	66.7	14895
E-126	Wash Water Refrigeration Cooler	5.8	573
E-127	Wash Bottoms Refrig Cooler	10.8	1066
E-128	Recycle Liquor Refrig Cooler	12.2	1489
E-129	LP Refrig CW Condenser	69.9	8925
E-130	HP Refrig CW Condenser	32.5	5502
E-207	2nd Effect Heater	399.0	35013
E-208	3rd Effect Heater	454.0	32143
C-101	Leachate Reheat Coils	60.7	11217

C-102	Recovered Soln Reheat Coils	81.7	8101
C-103	LP Steam Coils	17.2	1647
C-109	Melt Vessel Coils	614.0	23367
V-105	AS MVR Evaporator	1505.3	191292
V-106	AS 1st Effect Evaporator	375.0	20077

Table 22. Vessel Sizes (Each).

Item No.	Title	No. Off	Volume (m3)	Diameter (m)
F-105	Magnesite Flash Vessel	10	60	2.5
F-106	Filter C/P	10	68	3.0
F-107	2nd Stage Condy C/P	1	70	3.0
F-108	Hot Condy Pump Feed C/P	10	39	2.5
F-109	3rd Stage Condy C/P	1	78	3.1
F-110	Cold Condy C/P	1	93	3.3
F-113	Bisulphate Solution Vessel	2	245	4.5
F-111	Ammonia Liquor C/P	1	82	4.0
F-112	Scrubber O/Heads C/P	10	75	3.75
D-101	Slurry Make-Up Vessels	2	460	6.0
D-102	Extraction Reactors	9	493	6.0
D-103	Magnesite Precipitation Vessels	10	427	6.0
D-104	Melt Vessel	2	453	6.0
A-101	Carbonate Stripping Tower	8	340	5.9
A-102	AS Decomposition Stripper	1	215	6.0
A-103	Ammonia Absorber (2 stages)	1	552	4.8 / 6.0
A-104	Wash Column	1	88	3.2
A-105	CO2 Scrubber (2 stages)	4	537	5.7

Table 23. Major Compressors (By Duty).

Item No.	Title	Power (MW)
J-201	Carbonate Stripper Vapour Compressor	29.00
J-202	AS Recovery Vapour Compressor	64.85
J-204	FG Blower	4.82