DACMin Project

Contract Ref: 4696/11/2020(2)

Final Project Report

14th November 2024

Authors: Michael Evans, CEO Cambridge Carbon Capture Ltd
Craig Webster, CTO/COO Cambridge Carbon Capture Ltd

Table of Contents

1 Summary

This evidence-based final project report details the design and development of the pilot system, demonstration and trials results, key successes, lessons learned, remaining uncertainties and next steps.

The main conclusion drawn from this project was that the majority of DAC technologies are reliant on the future market for carbon credits and higher values for credits generated by carbon removal. International support for carbon credits to date has been patchy and the current market has been driven by philanthropy from only a small number of corporations and family offices. Although this is enabling early projects and building demand and investor confidence, there is a significant risk that this will not translate into an internationally accepted marketplace, especially at the high carbon pricing needed to support engineered DAC. The other revenue streams offered by CCC's technology mitigate this risk, making future projects economically viable at current market prices for carbon of around $50 USD.

The DACMin project has led us to the conclusion that the most impactful way to achieve both Gigaton levels of carbon dioxide removal and to maximise revenues (thereby reducing the cost of CDR), was to leverage the vast contact surface area between sea and the atmosphere and dispense with the air contactors and high energy requirements needed to shift large quantities of air through scrubbers or filters, not to mention the capital cost of the huge machines needed to achieve this. This led to the decision to pivot the company technology. The conclusions of our pilot project indicate that using our magnesium hydroxide for Passive DAC (e.g. Enhanced Weathering or Ocean Alkalinity Enhancement), rather than engineered DAC, doubled the amount of CO2 we could extract from the atmosphere. Our models of our process showed that Active DAC contributes less than 10% to revenue and less than 50% of the total CDR capacity. This led us to remove the Active DAC functionality of our system, take a small hit on revenues and create a simpler, lower capital solution focusing on Passive CDR and revenues from Co-Products. It also allows us to revert to a previous version of our technology using a sodium hydroxide digestion which is lower temperature and has a much higher chance of success than the previous high temperature sodium carbonate route which this project has shown to be problematic.

2 Detailed description of the design and development of the pilot system

2.1 The design and development process

Phase 1 of the project resulted in the development of flow sheets and costing of a pilot system for the DACMin process. This was developed in collaboration with project partners, Otto-Simon Ltd for the general engineering and design of the overall system, and Parsons Clean Air Ltd for the design and specification of the air scrubbing and CO2 mineralisation systems. Securing of Phase 2 of the project enabled CCC to engage Otto-Simon Ltd (OSL) as CDM Principle Designer and Principle Contractor, and Parsons Clean Air Ltd (Parsons) as our scrubber system designer and supplier to carry out the detailed design and specification of the pilot plant.

OSL undertook the overall Project Management for the design phase. This was on the basis that the Parsons supplied equipment package was standalone and without engagement from OSL and that Parsons defined all its required external interfaces such as foundation requirements and power supply. In summary the equipment that was supplied by Parsons included the following:

- Air Scrubbing system, comprising:
    - Two spray column absorber scrubber towers
    - Air extract fan
    - Discharge stack
    - Interconnecting ductwork between the scrubbers, fan and stack
    - Control panel for the system
- Mineralisation scrubber system, comprising:
    - CO2 absorber scrubber tower
    - Control panel for the system

The Parsons scope included the design, fabrication, installation and commissioning of its equipment, including making the final interface connections.

Equipment other than that supplied by Parsons was defined technically by OSL and quotations were obtained from suitable suppliers. CCC used this as the basis for purchase orders for this equipment. Similarly, OSL scoped the various installation contracts (civil works, EC&I installation), obtain proposals from suitable contractors and CCC entered into agreements with the chosen contractors.

Once equipment suppliers were appointed, OSL integrated their designs into the overall plant design.

At the construction site CCC ensured that the installation contractors perform to specification and in accordance with all relevant health and safety regulations.
The OSL role ceased at “Mechanical Completion”, i.e. equipment is installed, power is on, safety systems have been tested, motors turn, instrument loops have been tested for continuity, etc. Air will have been moved through the scrubbers (under Parsons’ supervision) but no process chemicals, e.g. magnesium silicate or sodium carbonate, will had been introduced. The final commissioning stage, referred to as “Hot Commissioning”, was undertaken by CCC personnel.

2.2 Outputs of any modelling used to inform the design and descriptions of the models used, including identification and justification of any assumptions made in the models.

Carbon Capture Utilisation and Storage (CCUS) and Direct Air Capture (DAC) technologies were identified as the core focus for the pilot plant project. Based on preceding research and experience, the mineral feedstock “olivine” was selected for

the process. A combination of Olivine and Sodium Carbonate was subsequently determined as the optimal feedstock for the calcining stage. To establish a foundational framework for the requisite P&ID design, an initial mass balance for each phase of the pilot plant was constructed.
The project is designed to capture 100 tonnes of CO2 annually through pilot-scale operations. Consequently, the mass balance commences at the Direct Air Capture (DAC) column. To facilitate comprehension for the engineering team and optimize pilot plant design, each process stage has been delineated as a distinct unit. As such, the process is divided into the following stages:

1. Digestion
2. Dissolution
3. Separation
4. DAC
5. Mineralisation

Each stage will encompass mass balance calculations, laboratory analyses, relevant technology integration, experimental design, process modelling using HSC software, pilot-scale results, and subsequent conclusions.

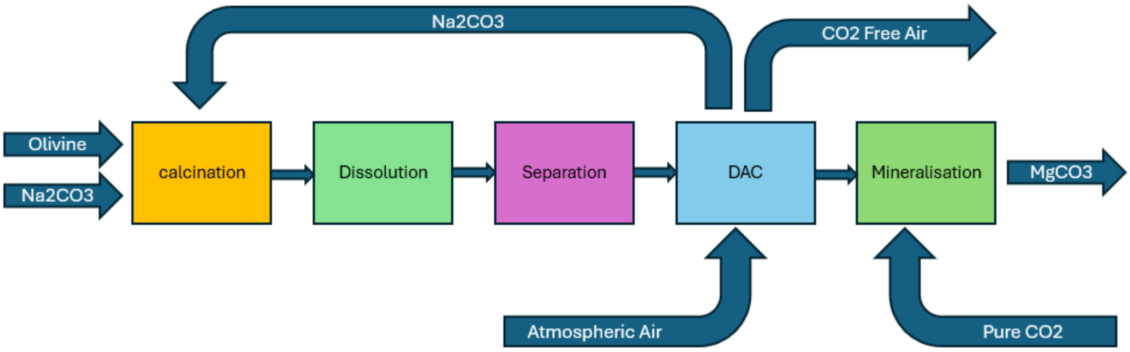


Figure 1: Process diagram

Talc was initially selected as the primary mineral feedstock, serving as the foundation for the initial mass balance and pilot plant design. Subsequent research and process development elevated olivine, previously considered an alternative, to the primary mineral for pilot plant implementation. Comparative analysis of talc and olivine revealed minimal performance disparities in terms of overall efficiency and outcomes. However, olivine demonstrated a higher yield of target elements within our project scope, coupled with abundant availability. Consequently, olivine was determined to be the optimal mineral feedstock for the pilot plant.

Table 1: A general chemical composition differences between Talc and Olivine

Chemical Composition	Talc	Olivine
SiO2	63.5 %	41.0 %
MgO	31.7 %	45.2 %

FeO	0.18 %	7.2 %
NiO	Trace	0.4 - 1.2 %
Other	2.8 %	5.4 %

The pilot's design was informed by mass and energy flow models developed by the R&D CCC team, with further improvements made using HSE software. This data was used by OS and Parson to generate a P&ID document for the pilot plant. This document identified the data / basis that Otto Simon Limited (OSL) used for the design the pilot plant.

2.3 Relevant data informing the design in tabular form.

As previously outlined, the initial mass balance for the pilot plant's black box design, based on an annual CO2 capture capacity of 100 tonnes, is presented below (Tables 2-5). However, since the digestion stage is outsourced to IBU-tech and the resulting material is transported and stored at our pilot plant, the P&ID and studies for the pilot plant begin at the transfer of IBU material from storage to the initial mixing tanks for dissolution in water.

Table 2: Mass balance of the digestion process (Pilot plant)

	Reactants		Products (Solid)				Products (gas)	
Material	Olivine	Sodium Carbonate	SMS	Magnesium Oxide	Iron Oxide	Unreacted Material	CO2	Steam
Material flow kg/h	55.55	42.03	54.73	9.25	2.18	14.64	14.83	1.95

Table 3: Mass balance of dissolution process (Pilot plant)

	Reactants		Products (Solid)					
Material	SMS	Water	Na2SiO3	MgO	Iron Oxide	Unreacted Material	Unreacted Sodium Carbonate	Insoluble SMS residue
Material flow kg/h	54.73	195.45	39.09	22.15	2.18	8.33	6.30	2.74

Table 4: Balance of DAC (Pilot plant)

	Reactants			Products (Solid)		
Material	Carbon Dioxide	Sodium Silicate	Water	Sodium Carbonate	Silica	Unreacted Material

Material flow kg/h	11.41	39.09	195.45	36.85	17.32	3.91

Table 5: Mass balance of mineralisation of high concentration of CO2 (Pilot plant)

	Reactants			Products (Solid)			
Material	Carbon Dioxide	Magnesium Oxide	Water	Magnesium Carbonate	Water	Unreacted Material	Iron Oxide
Material flow kg/h	14.83	12.91	51.64	36.85	34.33	3.91	2.18

Therefore, our initial black box model now includes the input and output flow rates of materials for each process.

By employing the material balance and sharing it with suppliers like Parsons and OSL, collaborative discussions regarding design ensued. Leveraging the initial P&ID design (Figure 2), OSL, as the consultant process engineer, completed the pilot plant design. Additional details were subsequently provided by Parsons several months later.

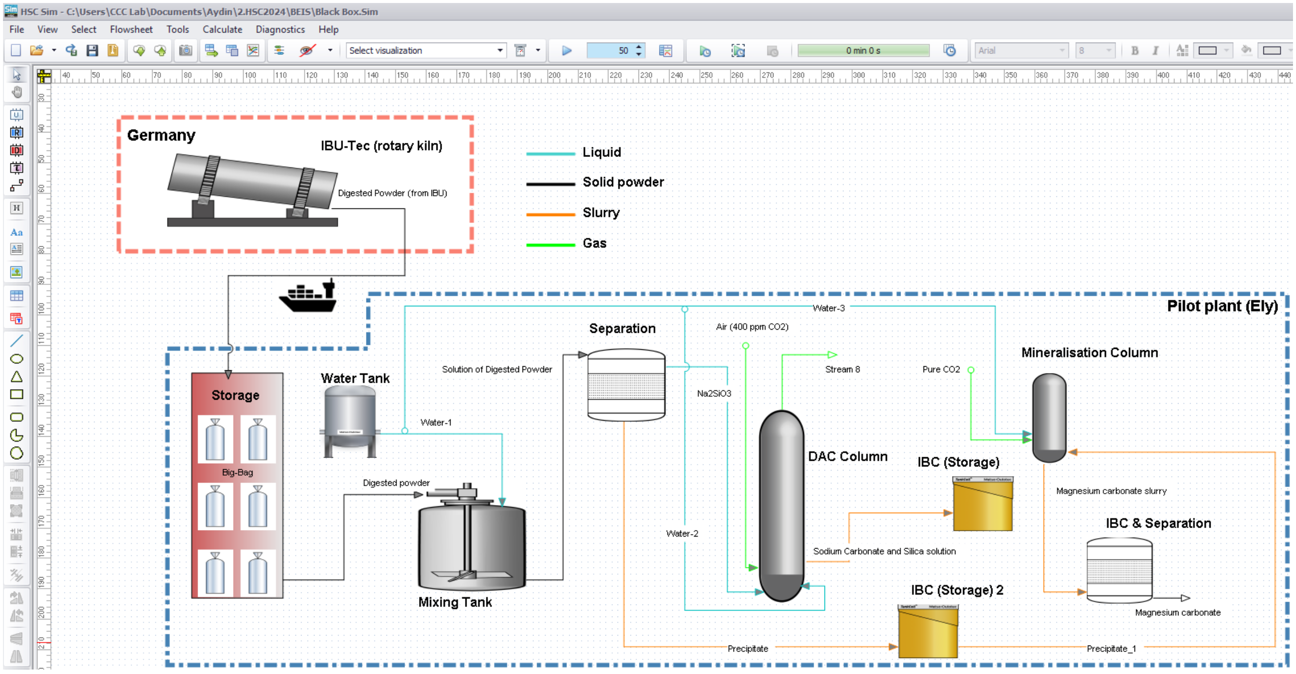


Figure 2: The P&ID for the pilot plant was developed by CCC utilizing HSC Chemistry software.

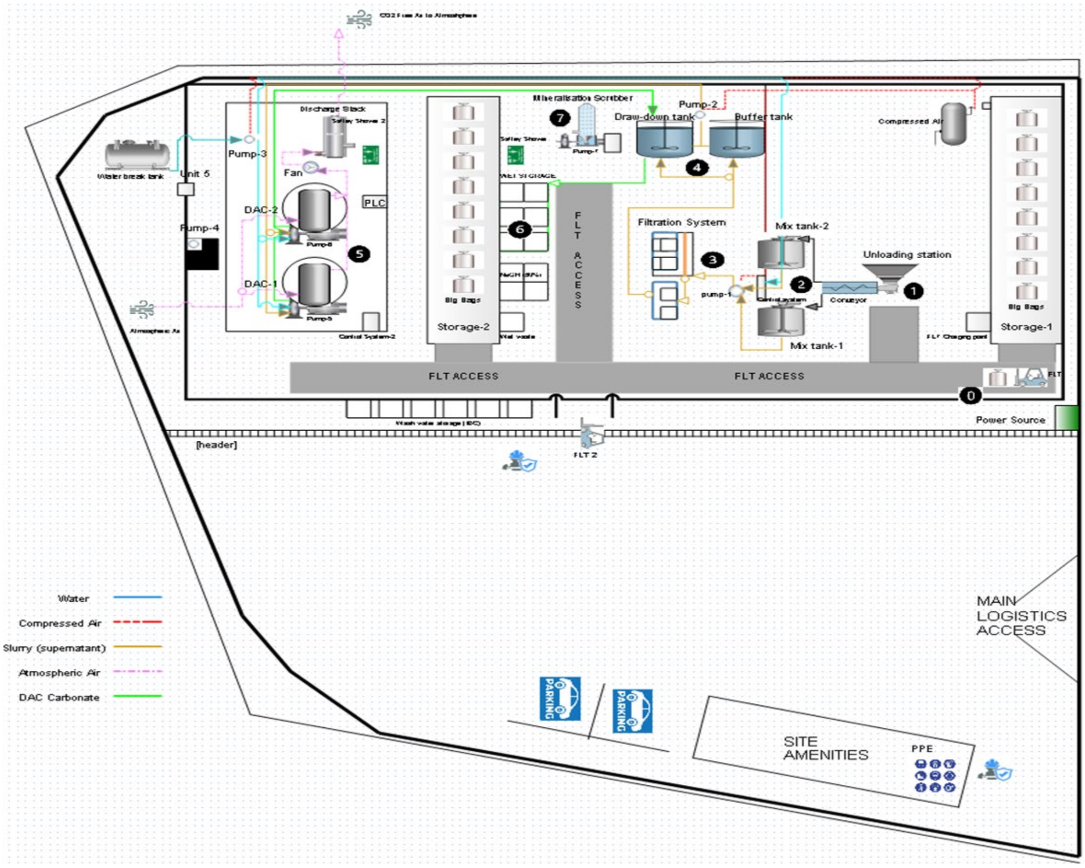


Figure 3: The layout of the CCC pilot plant at Ely designing with HSC Chemistry software.

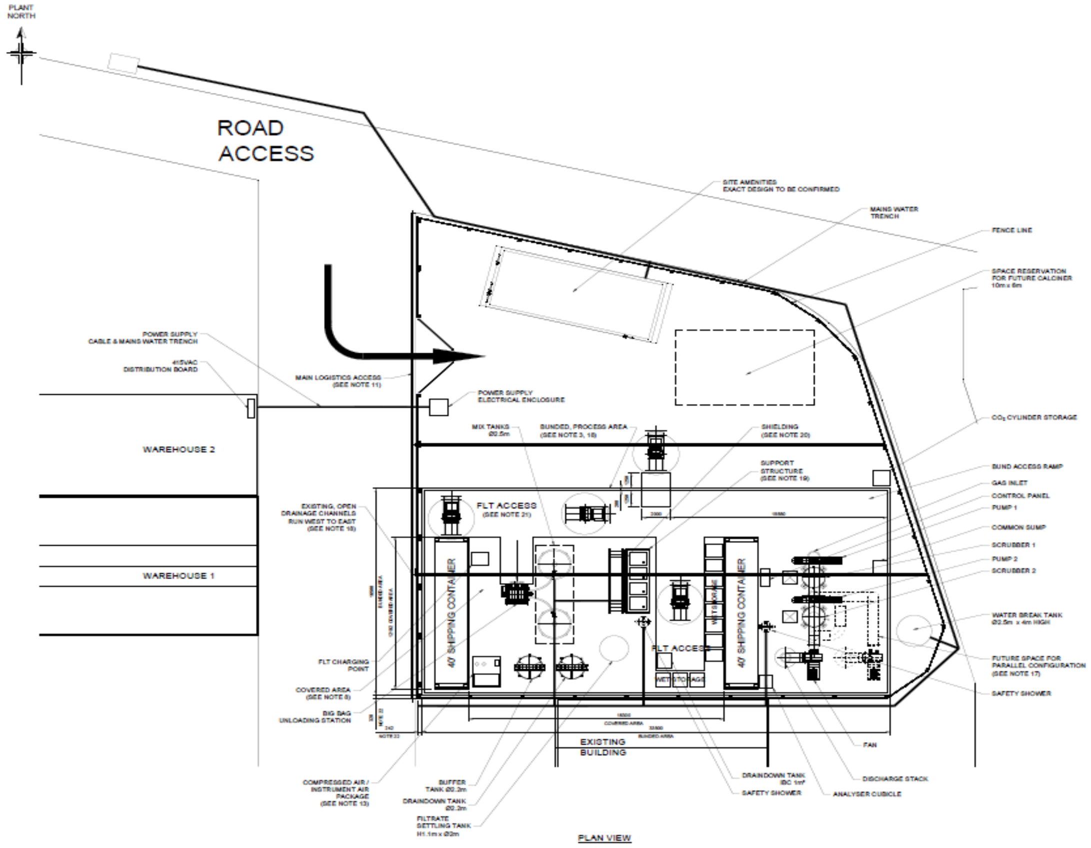


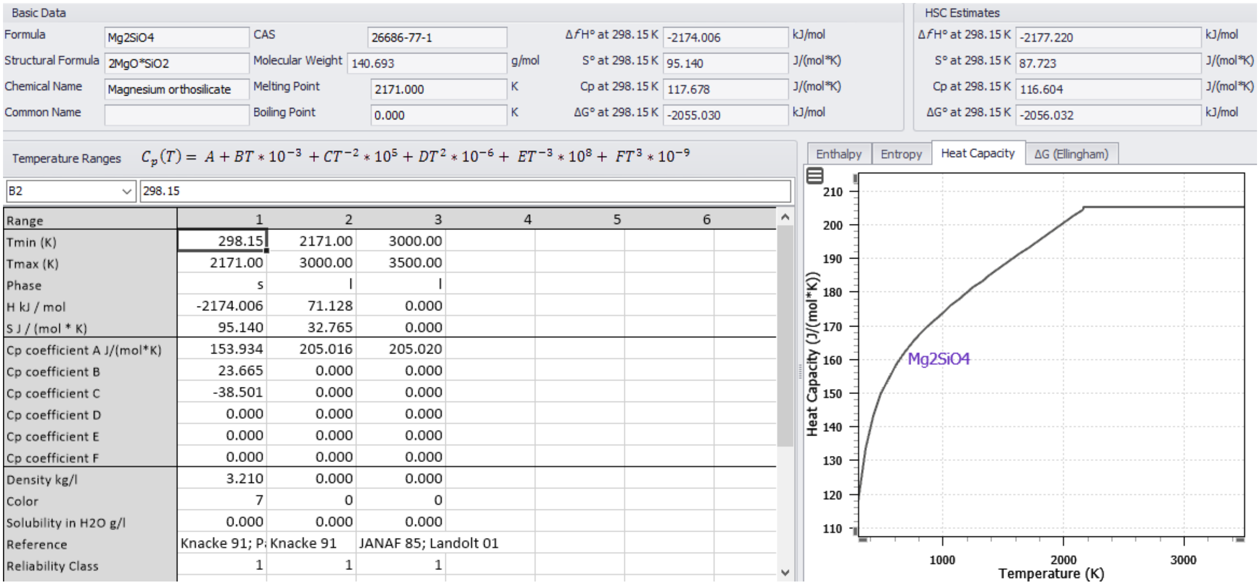
Figure 4: The last version of P&ID of the CCC Pilot Plant at Ely by Otto Simon Limited

2.3.1 Thermodynamic and Process Simulation

Metso Outotec's HSC Chemistry is a versatile thermochemical software package equipped with a flowsheet simulation module. Designed for a wide range of chemical reactions and equilibrium calculations, as well as process simulation, HSC Chemistry is a valuable tool for optimising reaction conditions, yields, mass and energy balances, and gaining a comprehensive overview of processes through the HSC Sim module. While HSC Chemistry cannot account for the kinetics of chemical reactions, it remains a useful tool for many applications. In the following section, a general overview of the process's thermodynamic profile is presented.

2.3.2 Reaction Equations

Given the decision to utilise olivine as the primary mineral and sodium carbonate as the secondary reactant, a calculation of enthalpy is necessary to assess the reaction's energy requirements. However, due to the variability in olivine types and elemental composition across different mineral forms, magnesium orthosilicate (Mg_2SiO_4) was chosen as a representative olivine type to simplify the model and improve clarity. This selection was informed by data from HSC Chemistry, as depicted in Figure 5.



Basic Data

Formula	Mg2SiO4	CAS	26686-77-1		ΔfH° at 298.15 K	-2174.006	kJ/mol
Structural Formula	2MgO*SiO2	Molecular Weight	140.693	g/mol	S° at 298.15 K	95.140	J/(mol*K)
Chemical Name	Magnesium orthosilicate	Melting Point	2171.000	K	Cp at 298.15 K	117.678	J/(mol*K)
Common Name		Boiling Point	0.000	K	ΔG° at 298.15 K	-2055.030	kJ/mol

HSC Estimates

ΔfH° at 298.15 K	-2177.220	kJ/mol
S° at 298.15 K	87.723	J/(mol*K)
Cp at 298.15 K	116.604	J/(mol*K)
ΔG° at 298.15 K	-2056.032	kJ/mol

Temperature Ranges $C_p(T) = A + BT * 10^{-3} + CT^{-2} * 10^{5} + DT^{2} * 10^{-6} + ET^{-3} * 10^{8} + FT^{3} * 10^{-9}$

B2 298.15

Range	1	2	3	4	5	6
Tmin (K)	298.15	2171.00	3000.00			
Tmax (K)	2171.00	3000.00	3500.00			
Phase	s	l	l			
H kJ / mol	-2174.006	71.128	0.000			
S J / (mol * K)	95.140	32.765	0.000			
Cp coefficient A J/(mol*K)	153.934	205.016	205.020			
Cp coefficient B	23.665	0.000	0.000			
Cp coefficient C	-38.501	0.000	0.000			
Cp coefficient D	0.000	0.000	0.000			
Cp coefficient E	0.000	0.000	0.000			
Cp coefficient F	0.000	0.000	0.000			
Density kg/l	3.210	0.000	0.000			
Color	7	0	0			
Solubility in H2O g/l	0.000	0.000	0.000			
Reference	Knacke 91; P:	Knacke 91	JANAF 85; Landolt 01			
Reliability Class	1	1	1			

Figure 5: Basic thermodynamic data of Mg2SiO4

A comparison of the enthalpy reactions between Mg_2SiO_4 and $Mg_{1.778}Fe_{0.222}SiO_4$ was conducted to assess the similarity between the two reactions (Figure 6). While $Mg_{1.778}Fe_{0.222}SiO_4$ more closely resembles the type of olivine utilised in the CCC process, incorporating this specific olivine composition in HSC Chemistry would complicate the process, making it challenging to manage. However, the results indicate that the difference in enthalpy between these two reactions is minimal (28 KJ), supporting the use of Mg_2SiO_4 as a reliable representation of the process. Therefore, the primary simulation is conducted without the inclusion of metallic content.

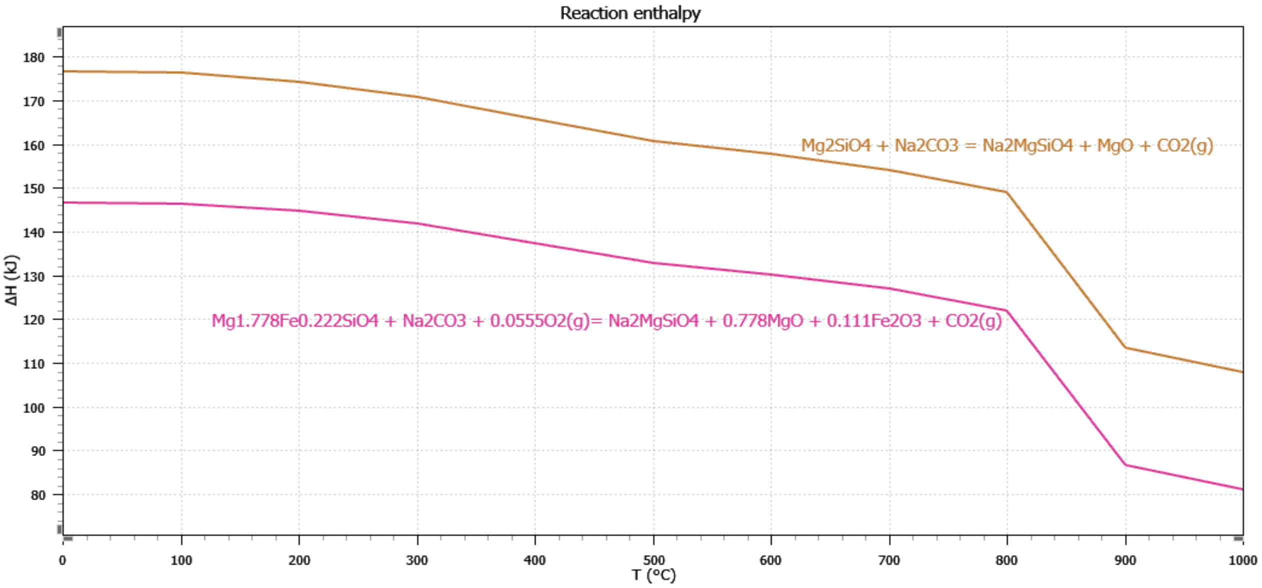


Figure 6 ΔH of Mg_2SiO_4 and $Mg_{1.778}Fe_{0.222}SiO_4$ with Sodium carbonate.

Additionally, other chemical reactions occurring within the process were simulated using the reaction equation module of HSC Software, and a comprehensive reaction data table (Table 6) was generated.

Table 6: Reaction Data for different stages of CCC Process

Digestion Process				**Dissolution**			
Mg2SiO4 + Na2CO3 = Na2MgSiO4 + MgO + CO2(g)				Na2MgSiO4 + H2O = Na2O*SiO2 + MgO + H2O			
T(°C)	ΔH (kJ)	ΔS(J/K)	ΔG(kJ)	T(°C)	ΔH (kJ)	ΔS(J/K)	ΔG(kJ)
0	176.696	175.027	128.887	0.000	-22.530	-22.374	-16.418
100	176.229	173.729	111.402	10.000	-22.621	-22.702	-16.193
200	174.24	169.069	94.245	20.000	-22.688	-22.935	-15.964
300	170.858	162.631	77.646	30.000	-22.733	-23.087	-15.734
400	165.83	154.584	61.772	40.000	-22.759	-23.172	-15.503
500	160.777	147.491	46.744	50.000	-22.768	-23.201	-15.271
600	157.887	143.988	32.165	60.000	-22.762	-23.182	-15.039
700	154.006	139.789	17.97	70.000	-22.742	-23.124	-14.807
800	149.013	134.914	4.229	80.000	-22.710	-23.032	-14.577
900	113.561	103.339	-7.672	90.000	-22.667	-22.912	-14.347
1000	107.814	98.639	-17.768	100.000	-22.614	-22.768	-14.119
Direct Air Capture				**Mineralisation of CO2**			
Na2O*SiO2 + CO2(g) = Na2CO3 + SiO2				MgO + CO2(g) + 3H2O= MgCO3*3H2O			
T(°C)	ΔH (kJ)	ΔS(J/K)	ΔG(kJ)	T(°C)	ΔH (kJ)	ΔS(J/K)	ΔG(kJ)
0.000	-94.258	-152.573	-52.583	0.000	-104.882	-183.460	-54.770
5.000	-94.230	-152.473	-51.820	10.000	-123.531	-251.694	-52.264
10.000	-94.203	-152.374	-51.058	20.000	-124.158	-253.869	-49.736
15.000	-94.175	-152.278	-50.296	30.000	-124.782	-255.961	-47.187
20.000	-94.147	-152.182	-49.535	40.000	-125.403	-257.976	-44.617
25.000	-94.119	-152.088	-48.774	50.000	-126.021	-259.918	-42.028
30.000	-94.086	-151.978	-48.014	60.000	-126.635	-261.792	-39.419
35.000	-94.054	-151.871	-47.255	70.000	-127.247	-263.602	-36.792
40.000	-94.022	-151.768	-46.496	80.000	-127.858	-265.355	-34.147
45.000	-93.990	-151.667	-45.737	90.000	-128.467	-267.057	-31.485
50.000	-93.958	-151.567	-44.979	100.000	-129.078	-268.715	-28.806

2.3.3 Equilibrium Compositions

The HSC Equilibrium module offers a user-friendly approach to calculating multi-component equilibrium compositions within heterogeneous systems. By defining the chemical reaction system, its phases, and species, the module determines the equilibrium quantities of products under specified isothermal and isobaric conditions.

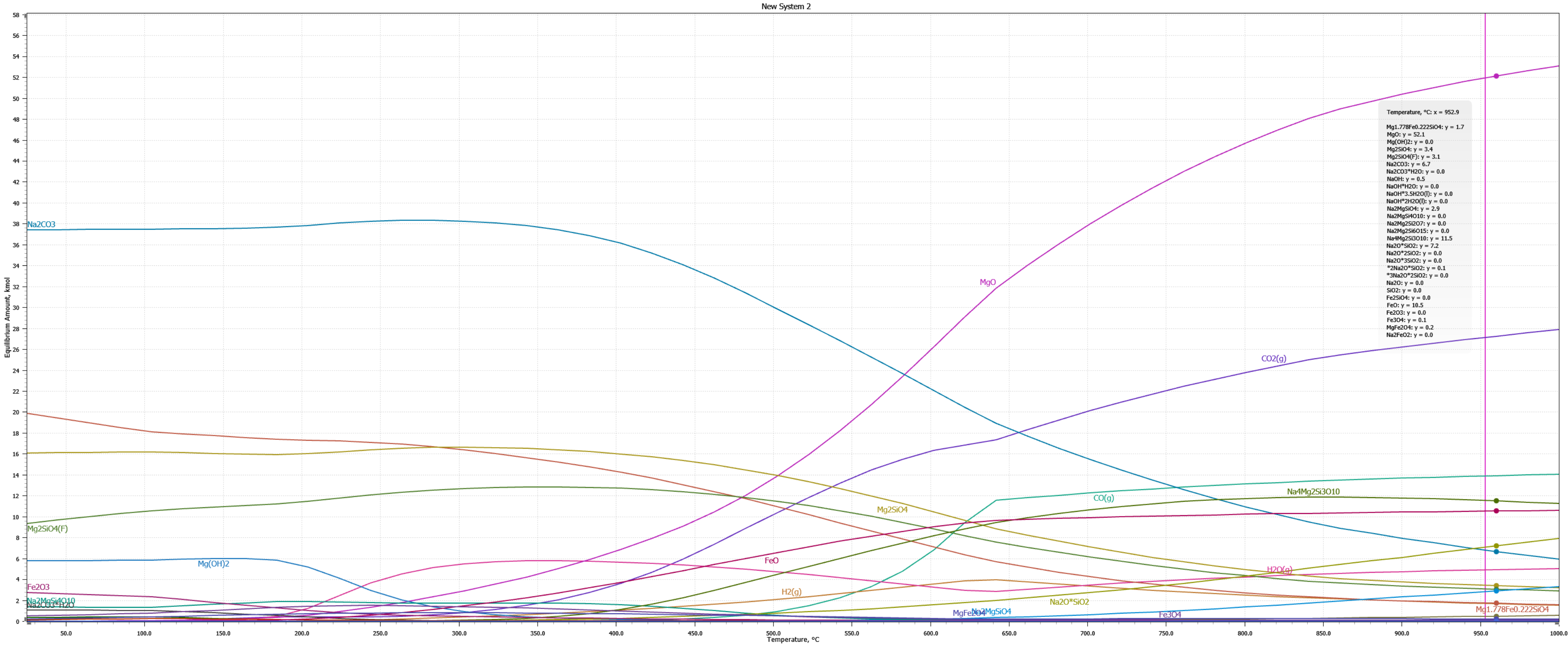


Figure 7: The equilibrium curve for the digestion process is presented across a temperature range of 0-1000 °C.

As illustrated in Figure 7, the equilibrium curve depicts the potential phases of each component across a range of temperatures. This curve is based on an ideal mixture of gases and oxides, excluding other phases. A vertical line drawn at 900 degrees Celsius indicates the probable products at this temperature. The data table reveals that the primary products are MgO and various forms of sodium-magnesium silicate, alongside some unreacted components and iron-containing compounds such as FeO. While this provides a general overview of the reaction's progression at different temperatures, more precise results can be obtained through laboratory and larger-scale experiments.

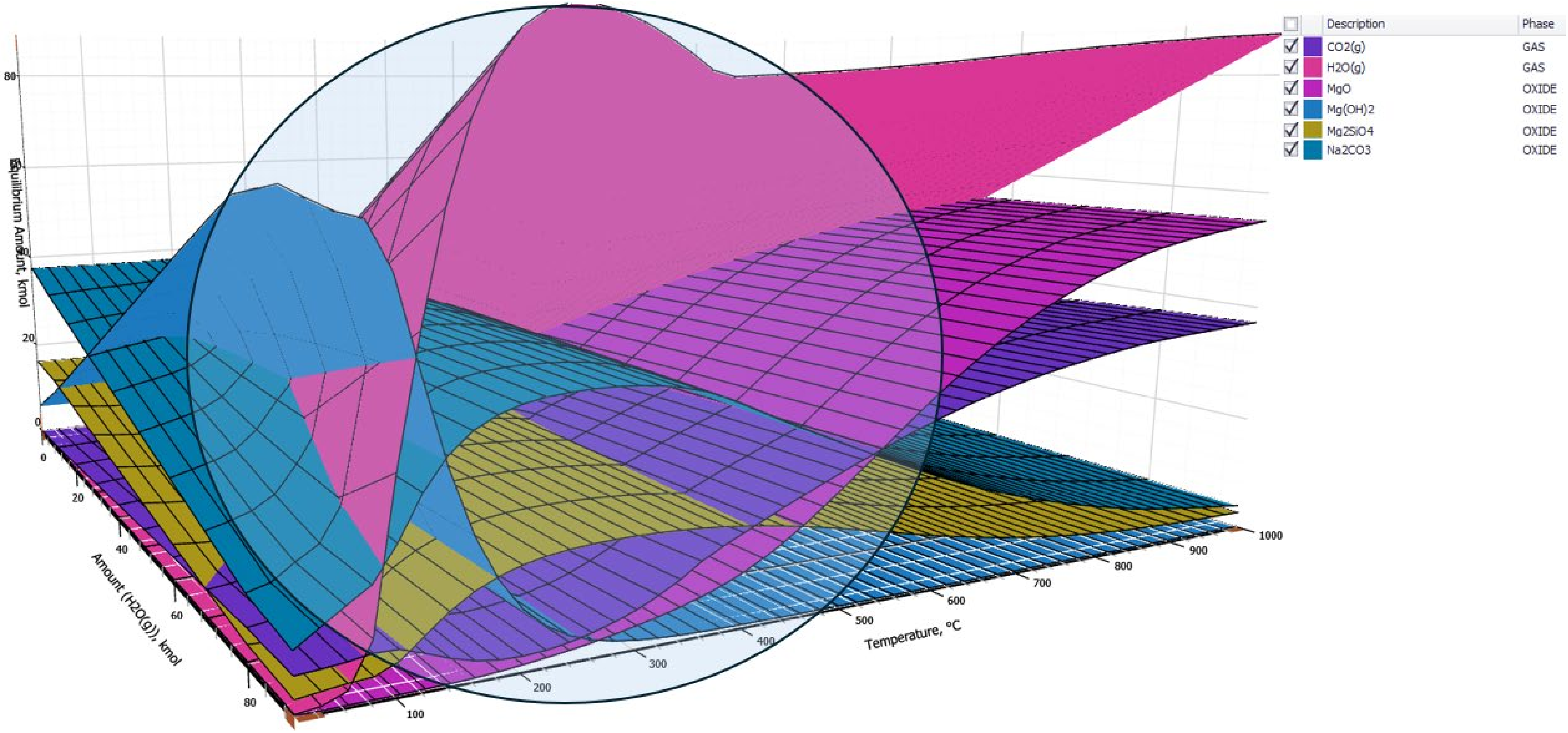


Figure 8: 3D equilibrium curve. The impact of Steam injection is presented across a temperature range of 0-1000 °C

The 3D equilibrium module was employed to investigate the impact of steam on the digestion process by incrementally introducing steam across a range of temperatures. As depicted in Figure 8, the results indicate that steam injection significantly influences digestion efficiency, particularly within the temperature range of 100 to 550 degrees Celsius. Additionally, various models suggest that steam can act as a catalyst, facilitating the reaction at lower temperatures. Experimental validation confirmed this observation. However, the decision to exclude steam from the digestion process in our pilot project will be discussed in subsequent chapters.

2.3.4 Heat and Material Balance

Heat balance calculations are fundamental to the development and optimisation of chemical processes. By quantifying heat inputs and outputs at each process stage, including temperatures, quantities, and overall heat balance, engineers can assess the process's energy efficiency and identify potential areas for improvement. This approach provides a comprehensive view of both heat and material flows within the process.

Table 7: Heat and Material Balance using HSC Chemistry

BALANCE	Temper. °C	Pressure bar	Amount kg	Heat Content kWh	Total H kWh	EXERGY kWh
Preheating Stage						
IN1	25	1	56450.850	0.00	-202453.67	6857
OUT1	400	1	56450.850	6997.20	-191838.93	9940
BALANCE			0.000	6997.20	10614.74	3083
Digestion Stage						
IN1	400	1	56450.85	6997.2	-192838.93	9940
OUT1	1000	1	56450.85	19134.09	-159853.1	23858
BALANCE			0	12136.89	32985.82	13918
Dissolution Stage						
IN1	25	1	42461.34	0	-154971.22	11030
OUT1	25	1	42461.34	0	-156062.19	10268
BALANCE			0	0	-1090.97	-761
Direct Air Capture (DAC) Stage						
IN1	25	1	111950.24	0	-462391.03	5612
OUT1	25	1	111950.24	0	-465781.76	3854
BALANCE			0	0	-3390.72	-1757
Mineralisation of CO2 Stage						
IN1	25	1	40873.32	0	-165712.96	4398
OUT1	25	1	40873.32	0	-169452.28	2241
BALANCE			0	0	-3739.32	-2157

To assess energy requirements, exergy, and potential work for each stage, the Heat and Material module in HSC software was employed to analyse thermodynamic data. The results are summarised in a table detailing the balance of heat content, total enthalpy, and exergy for each process stage. Table 7 provides an overview of this thermodynamic data, along with key data points that inform the thermodynamic process for a 50 kt plant feasibility study.

The critical data from the Heat module primarily focuses on the preheating and digestion stages, where heat input is necessary. Based on HSC thermodynamic modelling, the collected information excludes energy requirements for auxiliary equipment like pumps and stirrers.

2.3.5 Preheating Stage

Heat Content: 6997 kW is required to preheat olivine and sodium carbonate, assuming only water vaporisation occurs.

Exergy: Increases from 6857 kWh to 9940 kWh, indicating a 3083 kWh increase in work potential.

2.3.6 Digestion Stage

Heat Content: 12,136.89 kW is required.

Energy Input: 32,985.82 kWh is needed, assuming 95% efficiency.

Exergy: 9940 kW of exergy from the preheating stage is transferred here.

This analysis provides a foundational understanding of the energy requirements for the preheating and digestion stages of the process.

The thermodynamic processes for dissolution, DAC, and CO_2 mineralisation are also represented, although mechanical energy is a primary factor in these stages. Thus, these three data points will be discussed in the following chapters.

2.3.7 Simulation of Process

The CCC process is divided into five key stages: mineral grinding, thermal process, dissolution, direct air capture (DAC), and CO2 mineralisation. Mineral grinding is a well-established technology, and energy consumption and device size data have been provided by Sibelco, a Norwegian olivine supplier. While DAC and CO2 mineralisation scrubber technologies, provided by Parsons, have been designed, the available information and data were limited at the time. Therefore, the simulation and modelling of the process in HSC Chemistry focused on the thermodynamic aspects of the thermal process (preheating and heating), which accounts for approximately 80% of the total energy consumption of the entire process.

The following simulation assumes 100% efficiency for the thermal process. The primary goal of using the HSC Sim module was to simplify the process and identify areas requiring further attention. Additionally, all previous HSC modules were integrated within the HSC Sim module, which is specifically designed for flowsheet design and simulation, including material and energy balance calculations.

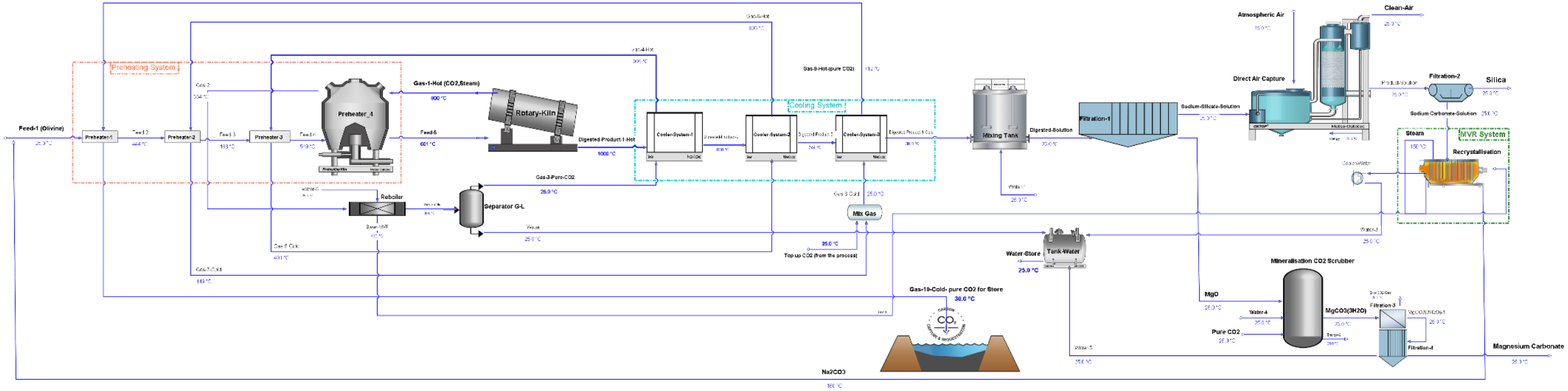


*Figure 9: illustrates the **temperature** profile of each stream*

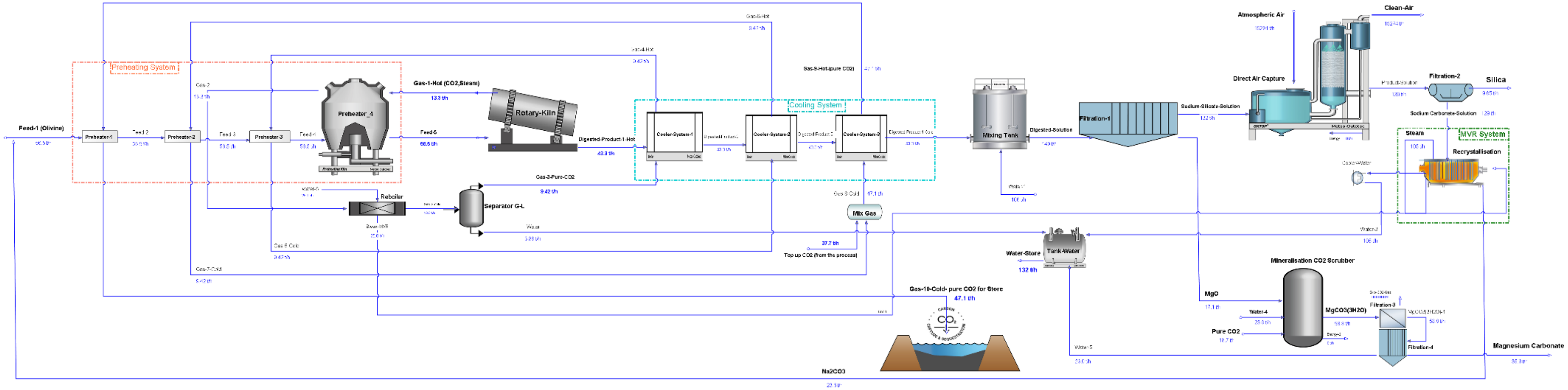


*Figure 10: illustrates **the flowrate** profile of each stream.*

Figures 9 and 10 illustrate the temperature and mass flow rate profiles of each stream within the flowsheet, respectively. The equipment is interconnected, representing an idealised process. To achieve a more realistic efficiency, further detailed calculations are required, which will be presented in the following chapter. Consequently, the data and information provided here do not extend to the intricate details of scrubber design at various stages. Instead, the focus is on the thermodynamic profile to facilitate a comprehensive understanding of the process.

The majority of this modelling was conducted prior to the operation of the pilot plant. Therefore, data derived from these preliminary stages was not incorporated into the design and simulation processes.

The preheating system was divided into four series of heat exchangers. A counter-current flow arrangement allowed the feed stream to increase in temperature before entering the rotary kiln, thereby enhancing heat recovery efficiency.

Thermodynamic simulations indicated that the feed temperature could reach up to 601°C (Figure 9) before entering the rotary kiln. Table 8 presents the energy profile of each heat exchanger within the preheating system. Pure CO2, produced from a thermochemical process, served as the heat transfer medium in this series of heat exchangers. This enabled the maximum efficiency of heat recovery to be achieved.

Table 8: Preheating Energy Table

Equipment	Heat Content Flow (KW)	Heat Duty (KW)	Exergy Flow Out (KW)	Log Mean Temperature Difference (^{o}C)
Preheater 1	9693.4	9638	10043.6	65.99
Preheater 2	1133.8	1118.9	10710.2	75.50
Preheater 3	1643.1	1628	11724.3	98.16
Preheater 4	1407.5	1380.4	12619.5	52.57

A rotary kiln is an industrial device primarily used for producing materials such as cement, lime, and other minerals. It consists of a rotating cylindrical vessel operating at high temperatures. Olivine and sodium carbonate are introduced at one end of the kiln and progress towards the other end, where they are discharged. The kiln is slightly inclined to facilitate the downward flow of material as it rotates, enabling efficient heating and reaction along its length.

Table 9 presents the mass and energy balances for the thermochemical reaction. Under ideal conditions with 100% reaction efficiency, thermodynamic data from HSC Chemistry software estimates a required energy input of 16.5 MW. To meet this energy demand, approximately 1605 Nm^3 of natural gas would be consumed (Table 10).

Table 9 Thermochemical Process

Mass In	Flow Rate (kg/h)	Mass Out	Flow Rate (kg/h)	Energy Profile	Value
Mg2SiO4	29896	Na2MgSiO4	34770	Temperature (°C)	1000
Na2CO3	22697	MgO	8498	Energy (KW)	16499
H2O(g)	3858	CO2(g)	9424	Exergy Out (KW)	20151
O2(g)	27	H2O(g)	3858	Efficiency (%)	100

Table 10: Natural gas consumption and mass balance

Input	Amount (Nm^3)
Natural Gas	1650.107
Air	16015.653
Output	Amount (Nm^3)
CO2(g)	1817.607
H2O(g)	3431.792
N2(g)	12508
Energy (KWh)	**17587**

In the CCC 50kt plant, the cooler system is a critical component of the production process, serving both as a heat exchanger and a heat recovery system. After exiting the rotary kiln at extremely high temperatures (around 1000°C), the product powder must be cooled to manageable levels (< 100°C) prior to further processing. The cooler system efficiently reduces this temperature while recovering valuable heat, which is then reused within the process, enhancing the overall energy efficiency of the operation.

Pure CO2 plays a crucial role as the heat transfer medium, transporting thermal energy from the kiln's output to the preheater system. Simulations indicate that for a cooling product powder flow rate of 43.3 t/h from the kiln, 47.1 t/h of pure CO2 is required. This implies that after 5 hours of kiln operation, a sufficient quantity of CO2 is available to fully operate the heat recovery system. Moreover, the kiln produces 9.42 t of pure CO2 per hour of operation. After extended operation, excess CO2 can be directed towards geologic carbon storage or mineralisation in our scrubbers. Table 11 provides a detailed energy profile of the cooling system.

Table 11: Cooler System Energy Table

Equipment	Heat Content Flow (KW)	Heat Duty (KW)	Exergy Flow Out (KW)	Log Mean Temperature Difference (°C)
Cooler 1	2893.6	2876.7	2836.5	154.26
Cooler 2	1009.8	993.4	2363.5	63.09
Cooler 3	9750.1	9693.4	10737.6	14.45

Direct Air Capture (DAC) scrubbers are technologies designed to extract carbon dioxide (CO_2) directly from the atmosphere. DAC systems operate at low CO_2 concentrations, typically around 400 parts per million, rendering the process challenging but potentially significant for mitigating climate change. Table 12 presents relevant thermodynamic data.

While the simulation tool employed could not provide the necessary units or equations for simulating the DAC column, fundamental thermodynamic information about the reaction is outlined. A more in-depth mathematical model will be explored in the subsequent chapter.

Table 12: Direct Air Capture

Mass In	Flow Rate (Kg/h)	Mass Out	Flow Rate (Kg/h)	Energy Profile	Value
Na2SiO3	26.1	Na2CO3	17	Temperature (ºC)	25
H2O	105.5	SiO2	10	Energy (KW)	-146189
Atmospheric Air	15281	Clean Air	15274	Efficiency (%)	75

Mineralisation of CO_2 through scrubbers involves capturing and chemically converting carbon dioxide (CO_2) into stable, solid minerals like magnesium carbonate. This process permanently sequesters CO_2, preventing its release back into the atmosphere. As part of carbon capture, utilisation, and storage (CCUS), mineralisation offers a promising long-term solution for carbon sequestration.

According to the mass balance and flowsheet from HSC, the thermochemical process (rotary kiln) can produce 17.1 t/h of MgO with 100% reaction efficiency, sufficient to capture 18.7 t/h of CO2 in the form of magnesium carbonate (Table 13). Simultaneously, the kiln produces 9.4 t/h of pure CO_2 while consuming natural gas, emitting a further 7.1 t/h of CO_2. By utilising geological CO_2 storage, the excess MgO can be employed to capture additional CO_2 emissions from the CCC process, potentially transitioning the process to negative emissions.

Table 13: Mineralisation of CO2

Material In	Flow Rate (t/h)	Material Out	Flow Rate (t/h)	Energy Profile	Value
MgO	17.1	MgCO3	36	Temperature (ºC)	25
H2O	23	H2O	23	Energy (KW)	-14694
CO2(g)	18.7	CO2(g)	0	Efficiency (%)	100

The digestion reaction requires approximately 16.7 MWh of energy. This energy is supplied by a natural gas-powered engine plant, which provides electricity to the process. The plant offers an electrical efficiency of over 50%, necessitating 3,296 Nm^3/h of natural gas to generate the required electrical energy for the digestion process. Preheating and heat recovery systems, including heat exchangers and coolers, are employed to minimise energy requirements effectively.

Additional energy is required for support equipment such as fans, cyclones, and pumps. It is estimated that this equipment produces approximately 1 t/h of CO_2. However, this CO_2 can be captured using the MgO produced in the process.

The CCC process produces 17.1 t/h of MgO. This MgO is divided into three portions: 8.6 t/h, 6.49 t/h, and 0.9 t/h. These portions can capture 9.42 t/h of pure CO_2 which produced from the reaction, 7.1 t/h of CO_2 from electricity generation for running the rotary kiln, and 1 t/h of

CO_2 from other equipment in thermal stage, respectively. This leaves a surplus of 1.1 t/h of MgO (Figure 11). If the pure CO_2 from the rotary kiln is not stored underground, it can be captured in a mineralisation stage of the process. This approach ensures efficient and sustainable CO_2 management within the process.

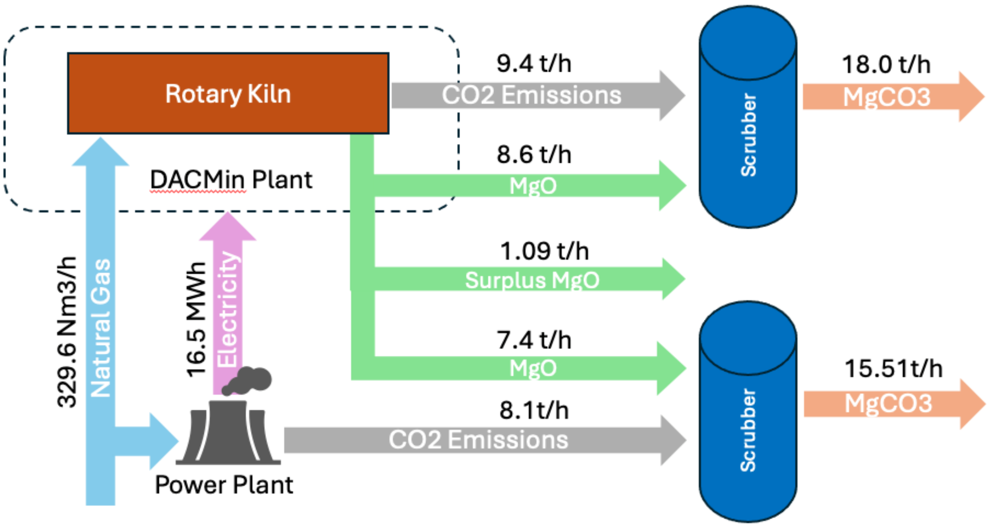


Figure 11: Option for dedicated power plant with CCM using surplus MgO

The total water requirement for the entire process is approximately 157 tonnes per hour. A portion of this water is generated during the thermochemical process from moisture and crystalline water present in olivine. However, a significant challenge arises in separating 81.75 tonnes per hour of water via evaporation from a solution containing 18.25 tonnes per hour of unreacted sodium carbonate and sodium silicate. To address this challenge and enhance energy efficiency, a Mechanical Vapour Recompression (MVR) system is proposed.

An MVR system recovers and reuses energy typically lost during the evaporation process by compressing the evaporated vapour and reusing it as a heat source within the same cycle. This closed-loop approach has been successfully applied in various industries, such as the dairy sector, to reduce energy consumption and greenhouse gas emissions. Calculations indicate that further efficiency improvements can be achieved by harnessing the heat content from the output gases of Preheater 4, which contain steam and CO_2 at a temperature of 554°C. By integrating this heat source with the MVR system, overall energy efficiency can be maximised.
For a plant with a capacity of 50,000 tonnes CO2 captured per year, a blower motor with an approximate power requirement of 3 MW is necessary. This setup supports optimal MVR performance by enabling the system to operate at a high Coefficient of Performance (COP), potentially reducing primary energy demands from conventional heating sources and enhancing the overall sustainability of the process.

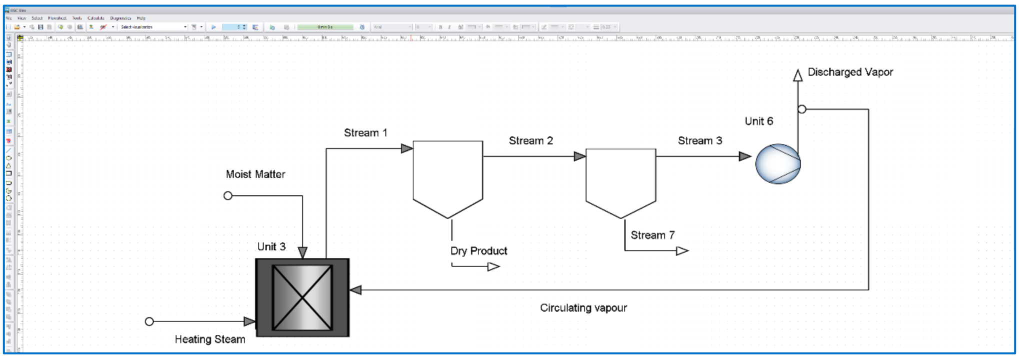


Figure 12: A simplified MVR loop for an MVR drying operation by HSC Chemistry software

2.3.8 Conclusion

The thermodynamic simulations conducted using HSC Chemistry software provide a robust foundation for understanding the energy requirements and potential efficiencies of the CCC process. By analysing the various stages, including preheating, digestion, cooling, and energy recovery, valuable insights have been gained.

Key findings from the simulations include:

- Energy Requirements: The thermal process, particularly the preheating and digestion stages, requires significant energy input.
- Heat Recovery: The efficient design of the preheating system, coupled with the utilisation of CO2 as a heat transfer medium, enables significant heat recovery.
- Process Efficiency: The integration of a Mechanical Vapour Recompression, (MVR) system for water evaporation can further enhance energy efficiency by recovering and reusing energy.
- Thermodynamic Limitations: While HSC Chemistry provides valuable insights, it is essential to consider kinetic limitations and real-world process conditions, which may impact overall performance.
- Future work will involve refining the simulation models to incorporate more detailed process parameters, such as kinetic rates and heat transfer coefficients. Additionally, experimental validation will be crucial to validate the simulation predictions and optimise the process. By combining theoretical modelling with experimental data, a comprehensive understanding of the CCC process can be achieved, leading to the development of a highly efficient and sustainable carbon capture solution.

2.4 A description of the installation of the pilot system.

The lease on the pilot site in Ely was signed on 1st March 2023. Prior to this, surveys were carried out and initial site layouts developed by Otto Simon (OSL) and planning agents, The Energy Workshop, were contracted to manage the planning process. Planning consent was applied for, and it was granted on 1st March 2023, triggering the signing of the lease for the site. The site was cleared and security fences erected. Drainage surveys were carried out and local contractors invited to the site to prepare quotes.

The civils contract was placed with Bowie Construction and work started on site on 7th June 23 with installation of water and electricity supplies to the site.

The plinth to carry the weight of the scrubbing towers was designed by OSL with input from Parsons for weights and support frame locations. The plinth was cast on 11th July 23, and the bund completed 24th July 23, ready for delivery and location of the shipping containers on the 25th, see above photos.

By 1st Aug 23, deliveries of plant and equipment started to arrive on site, these included, water break tank, mixer tanks and buffer drawdown tanks and the site office.

By early September 23 the water and power supplies had been installed and site office was operational. The scrubber system arrived onsite ready for installation.


19th September 23, cranes were hired and scrubber system erected. Bad weather resulted in the loss of the canopy and delays in the installation of the scrubber system. This resulted in additional crane rental costs.


Tanks placed in position and canopy re-installed and cranes back on site for completion of the scrubber install, 3rd Oct 23.

Installation of the big bag loading system, load cells and mix tank system, pipework and electrical connections to plant made throughout Oct to Jan 24 with first switch on of the scrubbing system for leak testing on 18th Jan 24.

Mineralisation scrubber installed early Feb 24 and pneumatic pumps and compressed air system installed and commissioned, trials started late March 24.

Modifications to the filtration system made mid-June 24 to accommodate ionic beads filtration and regeneration. Trials completed end August 24.

2.5 Challenges overcome during the design and development of the pilot system.

At the beginning of the project the plan was to locate the pilot plant within the HyNET project, in the grounds of the University of Chester's campus at Thornton Park. Despite guarantees during the Phase 1 of the project, Chester University were unable to host the pilot site due to internal issues surrounding the management of their site. This meant that CCC had to find a new site location and obtain planning permissions. A new site was found in Ely 15 miles North of CCC's labs in Cambridge.

Another setback was that the delays in the placing of the contract for Phase 2 meant that the calciner we had planned to use in the pilot plant was sold to someone else, an alternative was not found for the available budget so CCC located a toll manufacturer in Germany, called IBU-tech, offering a service to scale up heat processes and process pilot quantities of materials. The budget set aside for the calciner was reallocated enabling 1 scale up trial and a production run to supply bulk materials for the pilot trials.

CCC lab digestions were showing digestion rates in excess of 90% but elemental analysis suggested the formation of hexavalent chrome, a highly toxic substance. Methods of suppressing its formation were explored and the CCC team found that the addition of carbon into the mix created a reducing atmosphere within the kiln, suppressing its formation. This had an additional benefit to further reducing the metals contained within the Olivine, making them easier to extract.

IBU-tech scale-up trials were successful, but the learning did not translate to a successful bulk material run in the larger kilns they had. This resulted in only partial digestion of around 30%. Many different parameters were tried during the course of the production run, but it became clear that the kiln selected did not have sufficient heat capacity for the reaction we wanted. A clear path to a future successful run was suggested by IBU-tech's experts but unfortunately, there was no additional budget to run further material. Consequently we had to make do with materials for the trial which were well below the expected digestion levels. This would have an adverse impact on the performance of the pilot plant, but we would still be able to demonstrate a level of CO2 capture and model performance for scale up based on the knowledge gained.

Another challenging issue was that SMS samples produced both in the lab and by IBU-tech did not dissolve at the expected rates. There was a fast initial dissolution which quickly slowed. This resulted in significant time needed in the mixing tanks to fully dissolve the SMS. CCC's scientific team carried out in-depth analysis of samples at various stages of dissolution, accessing analytical techniques such as QEMSCAN and XRD to look at particle surface composition and developed titration and colourimeter tests to assess silica in solution, a key marker to levels of dissolution. It was determined that the observed effect was caused by the formulation of a silica passivation layer around the SMS particles blocking further dissolution. Various techniques were explored, ranging from elevation of temperature and pressure to surface modification but most success was achieved through the use of ionic beads used in the water purification industry to remove silica. These beads attracted the silica released by the dissolution, preventing it building to levels within the solution that lead to the formation of the passivation layer around the SMS particles. The beads could then be filtered out and reprocessed, releasing the trapped silica and re-activating the beads for further use. Modifications were made to the pilot plant filtration system and CCC tried to secure samples of beads for the pilot trials. Unfortunately, the suppliers were not able to provide stock within the timescales of the trials so we were not able to include this step in the trials so had to rely on lab experiments to inform the modelling of the process and live with the slow dissolution rates in the pilot trials.

2.6 Final costs of the pilot system.

The scope of the DACMin project was to pilot the main stages of CCC's process, as such the pilot system did not include much of the integration of the various stages. This was covered by the use of modelling, whereby the overall performance was predicted using the data gained from the piloting of the various stages. The cost of the pilot built in the DACMin project can be broken down as follows:

Project Management	£1,098,261.41
Detailed Pilot Design	£167,662.63
Equipment Procurement	£1,234,025.27
Pilot Build	£231,848.94
Pilot Trials	£98,948.99
Supporting Research Activities	£94,083.51
50kt Plant Planning	£27,407.56
Business and Commercial Planning	£37,725.34
Decommissioning	£10,000
TOTAL	**£2,999,963.65**

3 The results of the demonstration and trials

3.1 Laboratory tests using TGA

The Perkin Elmer TGA7 thermogravimetric analyser accurately monitors the mass of a small crucible while heating it at a programmable ramp rate. Tests with about 30mg of Olivine mixed with sodium carbonate showed that the rate of reaction increased rapidly as the temperature approached the sodium carbonate melting point of about 850ºC.

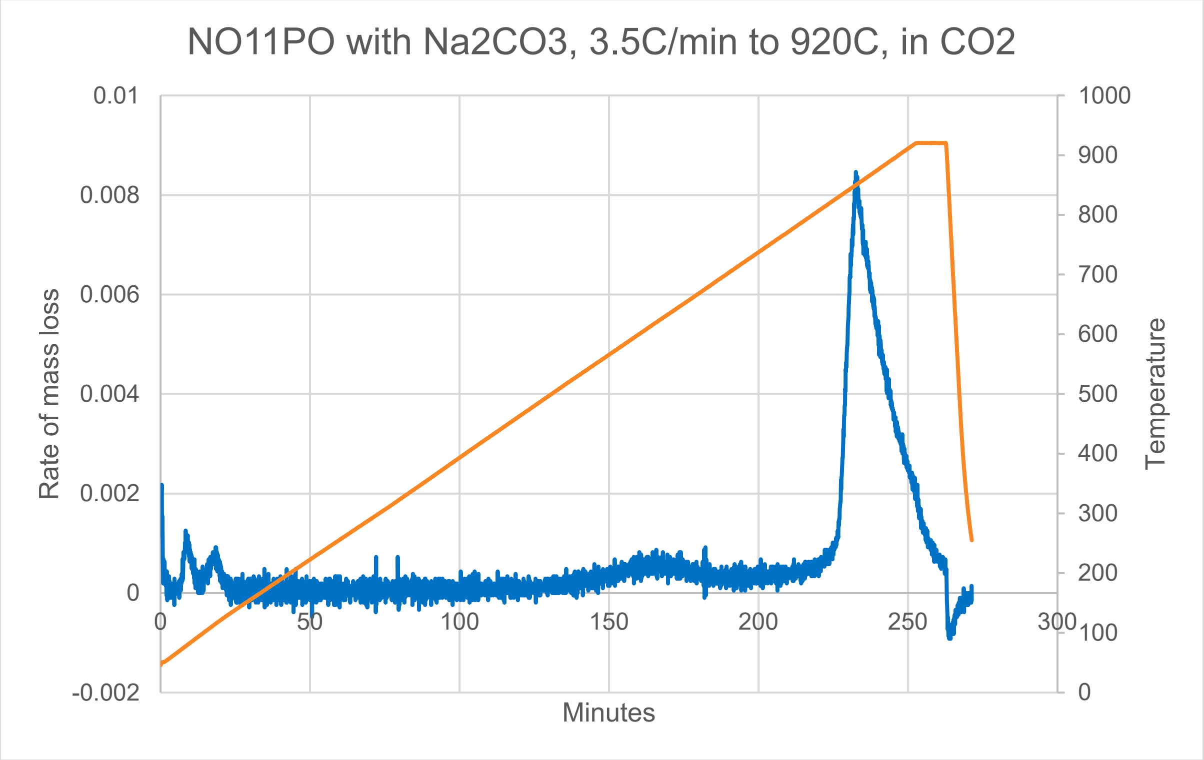


Figure 13: TGA results using fine Olivine powder

Fig 1. TGA008 result using the fine Olivine powder NO11PO and sodium carbonate. Heated slowly at 3.5ºC/min to 920ºC in CO2 atmosphere. Rate of reaction in blue (fraction of original mass lost per minute) shows a peak at about 850ºC. The slower mass loss at about 650ºC is due to dehydration of the Olivine.

A 32mg sample of the powder mix used at IBU-tec was heated to 941ºC at 5ºC/min. with nitrogen cover gas. This mix contained activated carbon to reduce the transition metal oxides. The peak in reaction rate occurred at 834ºC and the mass loss completely stopped within 5 minutes of reaching 941ºC. Mass loss was 22.3%. The first sample from the large-scale trial at IBU-tec was also tested and the comparative mass loss suggested that the digestion had only progressed 29.5% of the way to completion.

3.2 Laboratory rotary kiln tests

A Carbolite rotary kiln (Model TS01-1100) with an Inconel heated tube was used to help determine the optimum digestion conditions. One end of the Inconel tube was either closed off, or used to introduce steam or nitrogen. The other end was connected to a bag for accurately measuring the volume of gas released. Some of this measured gas volume was due to expansion of gas within the furnace tube, but most was due to the release of CO2 as the Na2CO3 reacted with the Olivine. Tests with about 450 g of the same input mixture used in the TGA required about 30 minutes at 1000ºC to attain near complete reaction. It was concluded that a higher temperature was needed compared to TGA because of the heat transfer limitation with the 10 000 fold increase in sample mass.

Tests were also performed with a small rotary kiln at IBU-tec. Previous laboratory tests had been performed where the powders were heated under a steam atmosphere and these showed that the digestion reaction proceeded more quickly at a lower temperature when steam was introduced. This is understood to be because the steam reacts with the sodium carbonate to form carbon dioxide and sodium hydroxide which reacts very readily with the Olivine. The kiln in our lab could exclude the introduction of air, but the IBU-tec kiln could not. It was found that the air and steam combination caused some of the chromium to form chromium trioxide which is toxic. To avoid this it was decided to add a small proportion of activated carbon to act as a reducing agent. This stops the chromium oxidizing to the toxic form, but it also partially or fully reduces the iron oxides which makes them more magnetic. This is useful because it enables magnetic separation.

3.2.1 Laboratory XRD tests.

An Aeris model XRD from Panalytical was used to determine the crystal structure of the digestion products. The Rietveld refinement method was used to determine the quantity of each component.

It was originally thought that the digestion products would be predominantly sodium silicate and magnesium oxide. Magnesium oxide is insoluble in water and sodium silicate dissolves very well, so separating the 2 would be easy. However, the XRD results proved that sodium magnesium silicate (SMS) was the dominant product. Tests showed that SMS reacts slowly with water producing mostly a precipitate of silica and magnesium oxide, and a solution of sodium hydroxide. The solids settle to form a sticky mud that reacts with pure carbon dioxide at a moderate rate. Once settled the supernatant, composed of sodium hydroxide solution with a little sodium silicate, reacts readily enough with carbon dioxide to be used for direct air capture (DAC).

The table below shows the XRD analysis results for the first 3 big bags run through the pilot plant. The feed material was 55.7% Olivine, 42.2% sodium carbonate and 2.1% activated charcoal. In a fully reacted digestate these 3 starting materials would all be consumed. The fact that there was so much left illustrates that the digestion achieved at IBU-tec was far from complete.

Table 14: Analysis by XRD and Rietveld refinement of composition of IBU-tech digestate

Big bag number.	Sodium carbonate %	NaMgSiO4 (SMS) %	MgFeO4 %	MgO %	Magnesium silicate hydroxide %	Olivine %
2	37.56	19.34	3.76	3.47	0.27	34.71
4	39.5	17.3	1.06	4	1.09	38.6
10	30.185	39.62	0.96	8.27	2.29	18.42

3.3 IBU-tech thermal treatment trials in Germany

3.3.1 Scope

As part of an R&D project, the CCC collaborated with IBU, a company specialising in rotary kiln technology. IBU provided equipment and expertise to heat a mixture of olivine and sodium carbonate within the kiln, facilitating the digestion of olivine. The primary objective of this trial was to upscale the process from laboratory/technical scale to a more production-oriented level by utilising IBU-tec's largest indirectly heated rotary kiln, the IDO9. This scaling-up step aimed to produce a sufficient quantity of digested material to operate the CCC pilot plant in Ely.

15.75 Tonnes of Olivine type NO11PO was added to 11.94 t of sodium carbonate and 0.599 t of charcoal and then mixed. The mix was fed into IBU-tec rotary kiln IDO9 for calcination. As mentioned above, the first sample suggested the reaction had only gone 29.5% of the way towards completion. The temperature was increased to the maximum (1010ºC at the outside wall of zone 3), and flow rate was decreased, but yield remained poor. We believe that because the reaction is strongly endothermic, insufficient heat was transferred through the powder bed, and the temperature achieved by the bulk of the powder only averaged 820 to 840ºC.

Of the 28.289 t of material mixed, 20.245 t was calcined and the product mass was 18.3415 t. This represents a mass loss of 9.40% compared to the 22.3% observed with the TGA which suggests that on average the digestion achieved was 42% towards completion. This partly digested powder was packaged into 20 big bags and shipped to the pilot plant at Ely.

Table 15: Process Parameters

Temperature (ºC)	approx. 1000
Residence time at reaction temperature (min)	approx. 45
Gas Atmosphere	reducing (CO, N2)
Exhaust Gas	H2O, CO, CO2

3.3.2 Operation:

For this trial, the FKM1600 mixer and the IDO9 indirectly heated rotary kiln were employed. Their technical specifications are outlined in Tables 2 and 3, respectively.

Table 16: Technical parameters of IDO9

Mixer	FKM1600
Type	Ploughshare mixer
Volume	1600 l
Batch Size Used:	~40-75 % filling ratio **898 kg**

Table 17: Technical parameters of IDO9

Kiln	IDO9
Tube	Metallic (1.4841)
Tube Diameter	1 m
Heated Kiln Length	7m
Total Length (inlet, heating zone, cooler, discharge)	Approx. 14.8 m
Heating type	indirect, natural gas, 5 heating zones
Possible Temperature Range Used:	300 – 1,100 °C **1000-1100 °C**
Common (standard) through-put Used:	100 – 1,000 kg/h **100-500 kg/h**
Internals	Cage with small lifting bars

Exhaust Gas Cleaning	Cyclone/ Afterburner / Dust filter
Rotation speed Used:	Rotation speed adjustable: 0.5 – 6 rpm **0.5-1.1 rpm**
Inclination	1.5 ° (Not adjustable)

The mixed raw material was fed from a big bag into a hopper, then conveyed via a screw to a gravimetric dosing unit. Subsequently, the material was introduced into the kiln through a size 6 kiln inlet screw. To enhance material mixing, contact with the reducing atmosphere, and prevent material accumulation on the kiln tube wall, internals equipped with lifting bars were employed. After passing through the cooler, the product was conveyed via a water-jacket-cooled screw into another big bag.

The kiln operated in counter-current mode under inert (reducing) conditions. The primary nitrogen flow was introduced at the end of the cooler (2-24 m³/h). Additionally, the kiln inlet and outlet heads, the dosing unit, and the product screw were also purged with nitrogen. The exhaust line was fitted with a cyclone, a thermal oxidizer (post-combustion unit), and a process filter with filter bags. Material separated by the cyclone was returned to the kiln inlet screw for direct refeeding. A flow chart illustrating the kiln setup is presented in Figure 14.

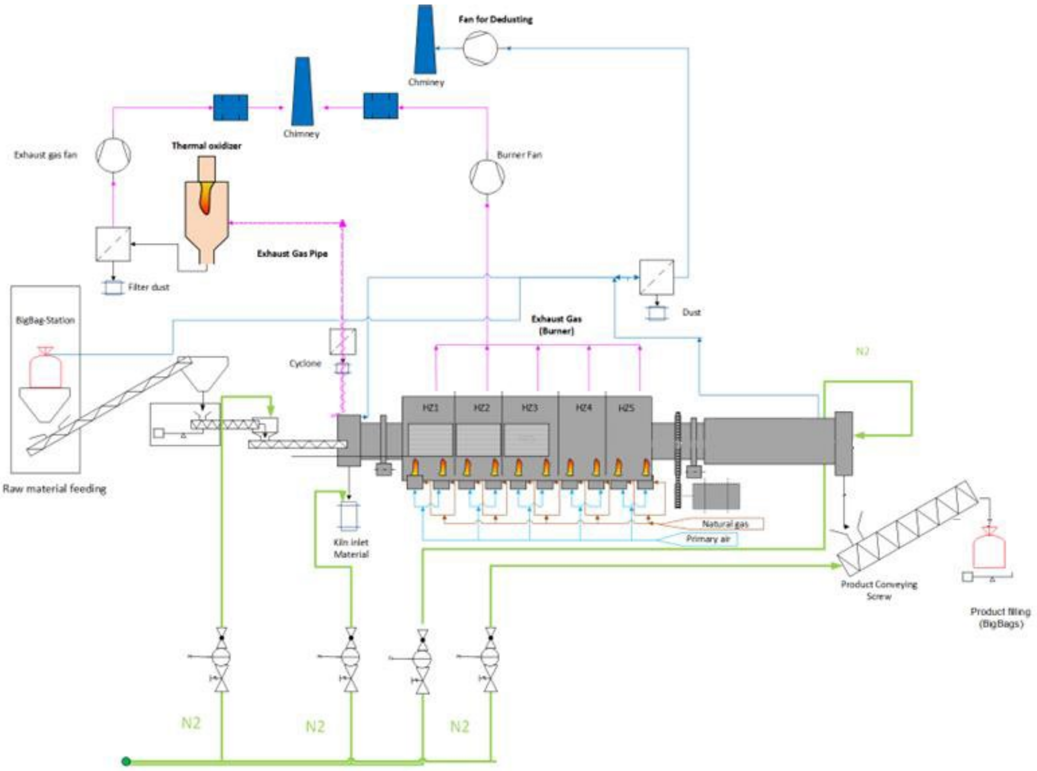


Figure 14: Flow chart IDO9

3.3.3 Results and discussions

The material's residence time within the kiln's heated section, along with the filling ratio, is influenced by factors such as feed rate and kiln rotation speed. These parameters typically result in residence times ranging from 35 to 90 minutes and filling ratios between 3.0% and 10.2%.

The burner temperature was gradually increased to 1100°C by the second trial day. The corresponding pyrometer temperatures for heating zones 2 to 4 ranged from 1000°C to 1020°C. Significant temperature deviations were observed in the first and last heating zones. The first zone, influenced by the inflow of cold raw material, exhibited a lower temperature of approximately 960°C. The fifth zone, due to its proximity to the cooler, maintained a temperature of around 870°C.

The kiln inlet and outlet temperatures (Figure 6.3) varied with the feed rate. For an empty kiln, the inlet temperature was approximately 500°C. As the feed rate increased to 300 kg/h, this temperature decreased to around 340°C. A slight increase was observed for feed rates of 100 kg/h. When the cooler was not in operation, the temperature rose to approximately 400°C.

Table 18: Kiln feed parameters

Feed rate [kg]	Rotation speed [rpm]	Calculated residence time (heated zones) [min]	Calculated filling ratio [%]
500	1.1	40	7.6
300	1.1	35	4.1
300	0.7	60	7.1
300	0.6	75	8.4
300	0.5	90	10.2
100	0.5	80	3.0

X-ray diffraction (XRD) analysis was employed to evaluate the product material. Quantitative analysis of the XRD results provided insights into the raw material digestion level, serving as a primary quality control tool. Ideally, the content of sodium magnesium silicate (SMS) and periclase (MgO) should be maximized, while minimizing the presence of natrite (Na2CO3), magnesioferrite (MgFe2O4), forsterite (olivine), and clinoclore.

To assess digestion success, the combined SMS and MgO content was compared to the combined soda and olivine content. Throughout the trial, the digestion level fell short of expectations, ranging from 20% to 50% SMS+MgO. While increased temperature and residence time led to slight improvements, a more significant increase was observed after reducing the feed rate to 100 kg/h. However, even with this reduction, the SMS+MgO share barely reached 50%. Notably, the post-run material, obtained after raw material feed cessation and kiln cooldown, exhibited a significantly higher SMS+MgO share exceeding 75%. This discrepancy may be attributed to a thinner material bed due to the halted feed or potential material accumulation on the kiln tube wall or internals, resulting in prolonged residence time. Figure 15 illustrates the digestion efficiency.

The bulk density of the product was slightly lower than that of the raw material mixture, while the specific surface area was notably higher than that of either olivine or soda ash. This disparity may be attributed to the presence of small amounts of unburned activated carbon within the product, which boasts a specific surface area exceeding 1100 m²/g. The particle size distribution of the product generally aligned with expectations based on the raw material components.

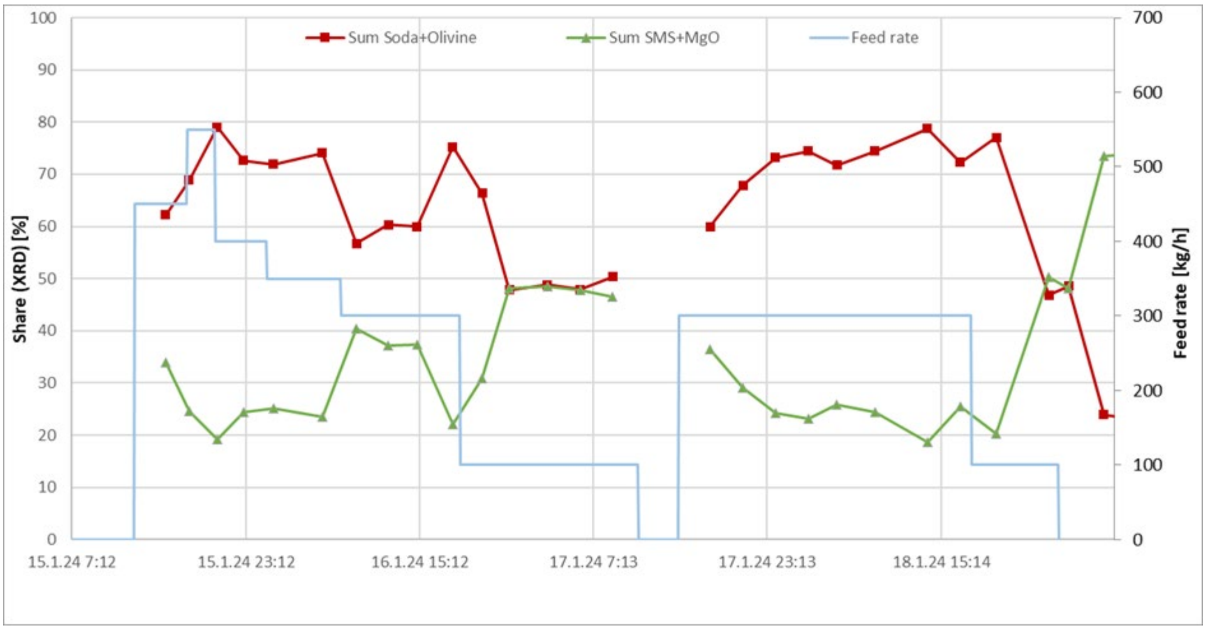


Figure 15: Level of digestion

A key focus was the comparison of water-soluble chromium levels with those found in the acidic digestate. Typically, only hexavalent chromium species, known for their high toxicity and carcinogenicity, are water-soluble. Conversely, while still classified as hazardous, trivalent and metallic chromium exhibit significantly lower exposure limits.

Although a substantial amount of chromium (up to 1600 ppm) was detected in the acidic digestate of the samples, the level of water-soluble chromium remained at or below the instrument's detection limit of 10 ppm. This finding extended to the post-run material, whose yellowish colour initially raised concerns about potential chromate content. These results suggest that the reducing CO atmosphere, facilitated by the use of carbon in the thermal process, effectively prevented the formation of hazardous hexavalent chromium species.

Figure 3 illustrates the measured concentrations of O_2, CO_2, and CO during the trial, with data points recorded every 15 minutes. The release of carbon dioxide from the soda component added to the raw material was anticipated during the digestion reaction:

$$Na_2CO_3 + Mg_2SiO_4 \rightarrow Na_2MgSiO_4 + MgO + CO_2$$

A high concentration of CO was sought to establish a reducing atmosphere. CO is generated through the reaction between activated carbon and the transition metal oxides present in olivine.

$$C + M_xO_y \rightarrow CO_2 + x\,M$$

or the Boudouard equilibrium: $C + CO_2 \rightarrow 2\,CO$

The concentrations of CO and CO_2 were primarily influenced by the feed rate. For higher feed rates of 300 kg/h or more, the upper detection limit of approximately 11% for CO and 22% for CO_2 was frequently exceeded. Throughout the trial, the oxygen content was maintained below 2%.

Another factor affecting the kiln atmosphere was the nitrogen flow rate. For instance, the CO and CO2 concentrations observed at a feed rate of 100 kg/h and a nitrogen flow rate of 11 m³/h were similar to those measured later at a feed rate of 300 kg/h, but with a higher nitrogen flow rate of 24 m³/h.

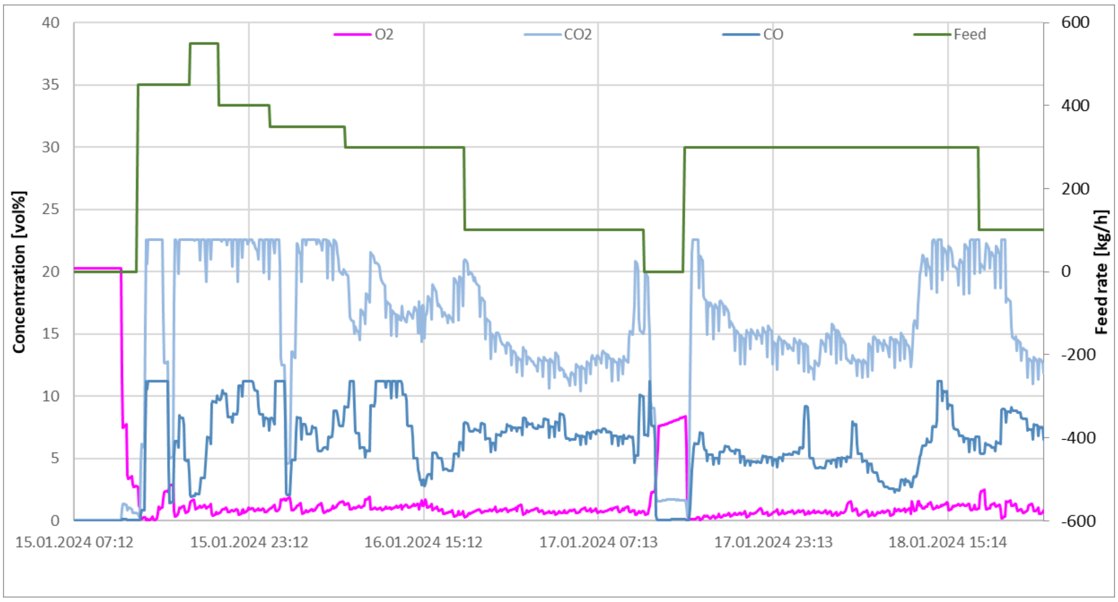


Figure 16: Gas analysis results

3.3.4 Energy

The total energy consumption (natural gas and electricity) was recorded for the entire trial and is presented in Table 19. To estimate the actual gas usage during specific trial points, the relative change in the burner control position, illustrated in Figure 14, can serve as a proxy. The highest heating power is required during the initial heating phase, temperature increases, and periods of high feed rates. Conversely, lower heating power is necessary for low feed rates, particularly evident in the second and third heating zones.

Table 19: Kiln consumption data

Total natural gas consumption	8,793.5 m3
Total electricity consumption	4,594.0 kWh
Operation Time (including heat up etc.)	99 h

3.3.5 Observation

Following the trial and subsequent cool down, the kiln was opened to inspect for any irregularities, such as material adhering to the tube wall or internals. As depicted in Figure 17 (left), the kiln tube walls appeared free of attached or sticking material. However, a thin layer of yellowish material, likely calcined product material, was found adhering to the internals. Upon closer examination, a significant quantity of hard, metallic shards was observed mixed with the calcined material (Figure 17, right). These shards are believed to have originated from a layer that was attached to the tube wall during the trial and detached during the cooldown period due to thermal contraction of the tube. The thickness of this layer was estimated to be between 1 and 2 mm. Some shards exhibited an additional, loosely attached layer of yellowish material, measuring at most 5 mm in thickness.

Figure 17: Left: View inside the kiln tube after the trial. Right: Detailed view on the shards forming the layer on the kiln tube wall.

3.4 Direct Air Capture (DAC) Trials.

3.4.1 DAC column commissioning

Figure 18: DAC scrubber system

The two columns each have their own air intake port and then both outlets are connected to a single fan that sucks the air through both columns in parallel. The velocity of air flowing through the duct leading to the fan was measured at different positions within the duct to determine the flow rate created by the fan. The power consumption, and the pressure drops created by that flow, were measured.

A carbon dioxide sensor (Uras26 from ABB) was connected to a dehumidifier (sample gas feed unit from ABB) to dry the gas and then pipes were connected to different points in the air flow circuit of the DAC columns. The readings from the CO2 sensor were recorded with ABB software called Optimus TCT. Before use it was calibrated with calibration gas. It was found that the CO2 content of the ambient air changed considerably during the course of each day. The CO2 levels increased after sunset and then decreased again after sunrise, presumably because of the all the surrounding plants starting to absorb it. Setting up a system that constantly compared the ambient CO2 levels, and the CO2 concentration of the processed air was difficult and there was insufficient time in the trial period. The CO2 concentration of the processed air was therefore only a rough guide to the quantity of CO2 absorbed. Titration of samples of the slurry was used as a back-up measure of the amount of CO2 absorbed.

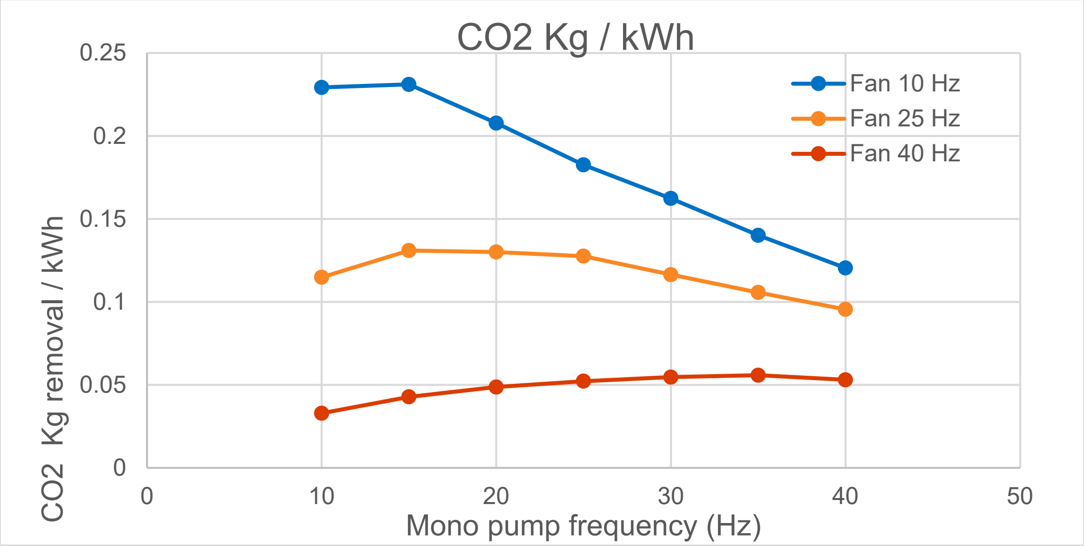


Figure 19: Graph showing optimum operational point

The supernatant fed into the columns contained a small amount of dissolved sodium silicate but was mostly a sodium hydroxide solution. The sodium silicate reacts with the CO2 and solid silica is formed and this creates a slurry. At times the sodium carbonate and sodium bicarbonate products became saturated and precipitated. The slurry in the columns was circulated with a mono pump on each column. The pumps fed the slurry to 12 spray nozzles positioned at 3 levels within the column. Air is drawn up the columns where it contacts the spray of slurry created by the nozzles. The flow rate of slurry was estimated from the manufacturer's data. The pump and fan motors were powered by inverters that could fix the frequency of the electricity supply between 5 and 50Hz.

There was concern that the fan would require so much power that the carbon footprint of running it would exceed the carbon dioxide absorbed. It was found that the column performed most efficiently when the fan was run at about half speed, and at that rate, using the average UK carbon footprint of UK electricity, the system was carbon negative. See PPdac003.xlsx. The most efficient operating point was determined to be fan speed set at 10 Hz and circulation pumps set at 15 Hz. See figure 19.

3.4.2 Test of DAC columns with sodium hydroxide.

In order to measure baseline performance of the scrubbing system, sodium hydroxide was used. This resulted in foaming which caused overflow of the slurry and vibration in the recirculation pipes. Adding anti-foaming agent stopped the foaming for a few minutes, but the foam kept reforming until more agent was added. The foaming was probably caused by the formation of sodium carbonate (washing soda) due to the take up of CO2 from the air, and because this was the dominant product of the reaction an excessive amount of agent would be needed to maintain high CO2 absorption rates. If the drive frequency of the mono pumps circulating the solution was kept to 15Hz, or below, and the slurry concentration was kept low, foaming did not occur. There was concern that vibration of the recirculation pipes could cause them to break, and result in the dangerous release of the all the slurry, so the decision had to be made to use the columns well below their full capacity.

Comparison of the measured performance with the design specification showed that, at the specified operation conditions, the scrubber performance was only 23% of that expected. The limited capture rate was due to the spray nozzles being badly positioned so that they sprayed most of the slurry onto the column walls, and the problem of foaming. The low energy efficiencies at high operating rates was another factor.

3.4.3 Digestate dissolution

Olivine was digested using the rotary kiln at IBU-tec, and then shipped to the pilot plant at Ely in big bags. The big bags were then mounted onto a screw feeder which fed the digestate powder into a stirred tank. About 3012.9 kg of tap water was added to a 3700 litre stirred tank and then about 757.1 kg of digestate powder from big bag 10 was added with a screw feeder. The slurry was stirred for 6 days and then pumped into IBCs where the solids were allowed to settle. Most of the solids settled within a few hours, but several days were usually allowed. The clear supernatant was then pumped into the DAC columns.

3.4.4 Supernatant in the DAC columns

The concentrations of sodium silicate in the supernatant solutions was low, and by keeping the mono pumps circulating the solution to 15Hz the problem of foaming was largely avoided. The issue of foaming would be a big factor guiding us away from the use of spray columns in future DAC plants.

The DAC columns were operated for a total of 93 days and captured 346 kg of CO2 from the air. At peak performance, 70ppm of CO2 was removed from the air by the system. Assuming 100% plant utilisation 24/7, this would equate to a maximum CO2 capture rate of 2.8 tCO2/year.

3.4.5 Chemical Analysis

In the DAC columns the amount of CO2 reacted was estimated from the record of the CO2 concentration and from doing titrations of samples of the solution. A standard acid/base titration using HCl as the titrant was used to quantify the concentration of the Na2CO3 and NaHCO3 produced in the columns, as well as the remnant, unreacted NaOH in solution. The titrations were completed using an automatic potentiometric titration system from Hanna Instruments. A minimum of 3 titrations were carried out per sample to ensure repeatable results and highlight anomalies. Each titration was run to 2 end points as this allowed for the calculation of the concentration of bicarbonate and the unreacted OH, as well as the carbonate concentration.

3.5 Mineralisation column

3.5.1 Mineralisation column commissioning


Figure 20: Mineralisation Column

The column was designed to hold slurry to a depth of around 3 metres and then above the slurry there was a space that needed to be gas tight. A pressure sensor was connected in that space to check for leaks, and then to judge how fast the CO2 was being absorbed during operation.

The carbon dioxide was delivered from a cylinder of pure gas via a flow meter for measuring the rate of delivery to the column. The target delivery rate was 104 l/min, but above about 10l/min the flow meter iced up. Heaters were added to prevent this icing.

The CO2 was delivered to near the bottom of the column and then the gas bubbled to the top of the slurry. To give the slurry a second chance to absorb the CO2 the gas at the top of the column was fed to a recirculation pump. This recirculation pump delivered 42 l/min at 0 bar and half that at 1 bar.

To keep the slurry well mixed, and to prevent settling and setting of the solids a mono pump was fitted to circulate the slurry. The flow of slurry from the pump also served to break up any foam created by the rising bubbles of CO2.

The reaction between magnesium oxide and carbon dioxide is significantly exothermic, so a thermometer was added. Previous experience has shown us that above 60ºC carbonates other than nesquehonite ($MgCO_3 \cdot 3H_2O$) start forming and these are more difficult to handle (more sticky). A cooler jacket was therefore included in the middle of the column.

3.5.2 Solution in Mineralisation column

Slurries with a high solids content need to be constantly agitated to stop the solids settling. Nesquehonite sets if allowed to settle and it becomes very difficult to remobilise. The sodium carbonate produced can also become saturated and precipitate and the resultant crystals can also cause blockages. Because of this the first test with the Mineralisation Column was performed with supernatant from big bag 10 dissolution.

A test was performed where the mono pump for recirculating the slurry was not turned on. The CO2 flow rate that kept the pressure in the top space constant was 5 l/min. The pressure was always kept within about 10kPa of ambient. Performance increased to 20 l/min when slurry recirculation was started. However, changing the mono pump speed made no measurable difference.

For best performance the top gas space would contain pure CO2, but flushing the air out was not easy because the slurry overflow system prevented drawing a vacuum on that space. The relatively low concentration of the supernatant also contributed to the low CO2 absorption rate of about 20l/min.

3.5.3 Slurry in Mineralisation column

After mixing the digestate with water, and then removing the supernatant to feed the DAC columns, a thick mud was left at the bottom of the IBCs. This mud consisted of undissolved SMS, unreacted olivine and sodium carbonate, and a little magnesium oxide and silica. The SMS and magnesium oxide reacts with concentrated carbon dioxide at a moderate rate The unreacted olivine can also carbonate, but at a very slow rate. The slurry fed into the column was made by mixing the mud with a small amount of supernatant in the IBC and then pumping it into the column. To minimise clogging of the column it was emptied after every day of operation.

Ideally, all the air at the top of the column, above the level of the slurry, would be pumped out and then replaced with pure CO2. The column had an overflow pipe that fed into a water trap and this prevented the pressure in the top space from deviating far from ambient. Therefore, the air above the slurry was replaced by using the recirculation pump to remove this air while CO2 was supplied from a cylinder of the pure gas. This CO2 was directed above the slurry using the pressure sensing line. It was impossible to get 100% CO2 in this gas space using this method. Tests showed that absorption rate was significantly improved as the CO2 level approached 100%. The highest CO2 absorption rate observed was 45 l/min. A larger recirculation pump would possibly enable the target absorption rate of 104 l/min.

During the filling process the slurry with suspended solids entered the sight glass used for measuring the slurry level. Because the slurry in the sight glass was not agitated during the test the solids settled to the bottom of the sight glass and blocked it. Future columns need to eliminate sight glasses, and all other features where solids can settle.

The absorption rate of the CO2 was not high enough to cause a heating problem.

4 Key successes, lessons learnt, remaining uncertainties and how these might be overcome.

Digestion step:

- The raw material mixture can be homogenized very well in a ploughshare mixer. Using the contractors FKM1600 ploughshare mixer, a throughput of roughly 300 kg/h can be achieved.
- The use of activated carbon powder to create a reducing kiln atmosphere was successful. The concentration of CO in the kiln atmosphere was between 5 and >11 % during the whole trial.
- The addition of steam was not tested during this trial. It remains to be seen if this measure can improve the digestion yield without compromising the material handling and behaviour in the kiln and its periphery.
- The material could be handled well during the trial. The dust emissions of the fine and very dry material were significant, but due to the very low gas volume flow through the kiln, mass losses were negligible.
- The digestion reaction did not take place as it would have been expected from the lab scale trials. It appears that the heat flow to the bulk of the material seems to be not

sufficient. Higher set temperatures and longer residence times were not possible due to the technical limitations of the kiln. Some possible explanations were discussed:

  1. The formation of a material layer on the kiln tube wall could have acted as an insulation barrier.
  2. Based on the thermodynamic calculations, the digestion process, as an endothermic reaction, consumes a significant amount of thermal energy. This may have led to an unexpectedly high heat loss across the material bed thickness (100kg/h feed rate and 3% filling ratio), the digestion yield improved only to 50%.
  3. The nitrogen added at the outlet side of the kiln was not preheated. The kiln inlet was slightly lower than expected. Thus, it may have been possible that the gas atmosphere had a cooling effect to the material. However, the nitrogen volume flow rate was very low and the heat capacity of gases is low in general.
- A limited rate of heat transfer is a known limitation of large rotary kilns. Because of this, future digestions on a large scale should use fluidized beds where heat transfer rates can be 2 to 3 orders of magnitude faster. More work is needed on the use of steam to accelerate the digestion reaction.

Dissolution Step:

- At a lab scale, the use of ionic beads was successful in preventing the levels of dissolved silica building up to a point where passivation layers formed on the particles of SMS retarding dissolution. Further work is required to develop a practical way of re-processing the beads, removing and recovering the silica at an industrial scale. This is especially important as the need to replenish damaged or blocked beads is both costly and carbon intensive, high recovery rates are therefore essential to both reduce the cost of carbon removal and the net carbon removed by the process.

Separation of metals:

- Due to the need to create a reducing atmosphere in the kiln, metal oxides within the Olivine were reduced creating alloys with the Iron present within the Olivine. This rendered these alloys magnetic enabling magnetic separation techniques to be deployed, a well-established industrial process. The team are confident that this will enable economic separation of the metals from the SMS, producing a material rich in metals which can be sold to existing metal refiners or refined further as part of the CCC process plant.

DAC Scrubber Step:

- We demonstrated the removal of carbon dioxide from the atmosphere and capturing it as a solid product, keep it out of the atmosphere permanently. This was done with the most reactive components of our digestion product. We also demonstrated that the less reactive components can be fully carbonated with more concentrated carbon dioxide. The pilot plant showed that the technology has huge potential, and make a global impact, but more small-scale laboratory research is needed to refine the process.

- Foaming is one of the reasons a spray tower is not ideal for DAC. For DAC a better system would be to trickle the slurry over an open structure as in a packed column.

- Handling concentrated slurries is difficult because of the particles settling and blocking pipes. To avoid blockages, future pilot plants will need to integrate all steps so that the slurry is transferred between process steps without delay.

5 An assessment of the benefits and challenges of the solution

5.1.1 Benefits:

The main benefit of the DACMin approach is the generation of additional revenues to the carbon credits generated through the sale of low carbon co-products such as critical metals and silica. This makes the Internal Rate of Return (IRR) for investors in future projects very attractive ensuring low financing costs for future projects and health returns for their investors. This will encourage significant investment leading to rapid scale-up of the process and the potential for significant carbon reduction impacts.

Another benefit is that the generation of low carbon critical metals and silica will lead to additional impacts due to the displacement of carbon intensive alternatives. This will be especially important with critical metals needed to enable the widespread electrification of industry and transportation that will be necessary to reduce emissions from those sectors.

The process produces surplus MgO with the benefit of enabling point source CO2 capture and mineralisation or further DAC through its use in either Enhanced Weathering or Ocean Alkalinity Enhancement.

5.1.2 Challenges:

The DACMin plant is capital intensive and there will be high risk in early projects, this will make early-stage investment challenging.

A further challenge will be securing off-take agreements and the cost and time it will take to access established markets for the co-products such as silica and critical metals.

There will be a need for new mining to support scale up. The time and cost to establish new mining sites and to get them on line are high. CCC would need to present a strong case for investment to the mining sector who are very risk averse, making it a challenge with a nascent market. Our approach would be to work with existing olivine miners, such as Sibelco, and partner with them on scale up projects so they witness the opportunity developing and can plan expansion of operations to match the growing demand. This will encourage other miners to explore the market opportunity creating a competitive market for the supply of olivine in the future.

5.2 An estimate of the CAPEX costs of the GGR solution, the OPEX cost and lifecycle costs per tonne CO2e captured/removed, with reference to the final costs of the pilot system.

5.2.1 Project Cost Models

Two financial models have been created, one for a 50kt DACMin plant and one for a Hybrid DAC plus Passive Capture.

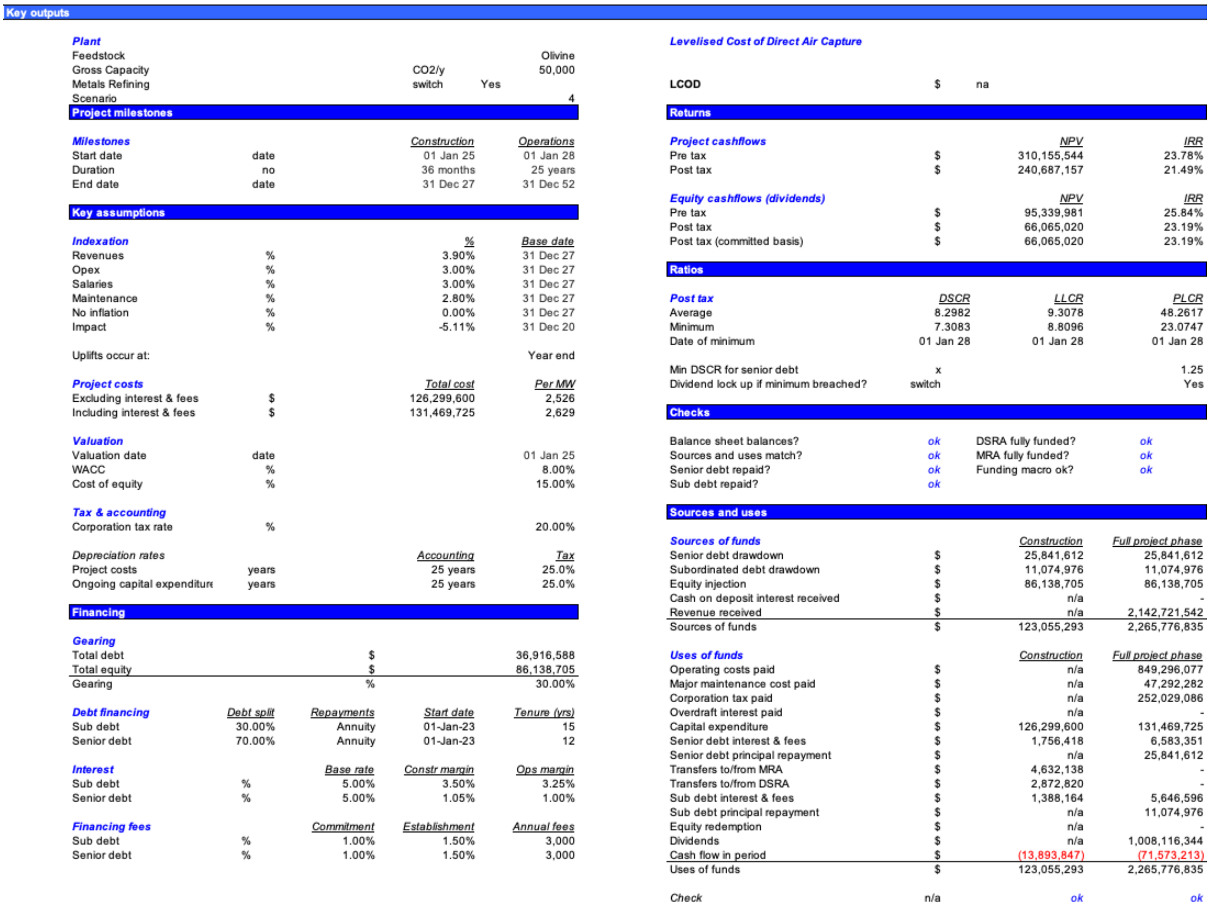
Capital costs for the Hybrid DAC are lower mineralisation plant is not required and carbon capture and removal capacity are increased by the addition of passive carbon dioxide removal using the Magnesium Oxide that would have been used for mineralisation.

The DACMin plant is powered by natural gas, with emissions being mineralised by the excess magnesium oxide remaining after mineralisation of captured CO_2, Most of the onsite electrical load is provide by a gas powered plant sized to match the availability of magnesium oxide for emissions removal, the balance is imported from the grid.

The Hybrid plant is powered by electricity to maximise the availability of magnesium oxide for passive CDR.

5.2.2 Direct Air Capture and Mineralisation Model

Dashboard from DACMin Financial Model



Key outputs

Plant			
Feedstock			Olivine
Gross Capacity	CO2/y		50,000
Metals Refining	switch	Yes	
Scenario			4

Project milestones

Milestones		*Construction*	*Operations*
Start date	date	01 Jan 25	01 Jan 28
Duration	no	36 months	25 years
End date	date	31 Dec 27	31 Dec 52

Key assumptions

Indexation		*%*	*Base date*
Revenues	%	3.90%	31 Dec 27
Opex	%	3.00%	31 Dec 27
Salaries	%	3.00%	31 Dec 27
Maintenance	%	2.80%	31 Dec 27
No inflation	%	0.00%	31 Dec 27
Impact	%	-5.11%	31 Dec 20
Uplifts occur at:			Year end

Project costs		*Total cost*	*Per MW*
Excluding interest & fees	$	126,299,600	2,526
Including interest & fees	$	131,469,725	2,629

Valuation			
Valuation date	date		01 Jan 25
WACC	%		8.00%
Cost of equity	%		15.00%

Tax & accounting			
Corporation tax rate	%		20.00%
Depreciation rates		*Accounting*	*Tax*
Project costs	years	25 years	25.0%
Ongoing capital expenditure	years	25 years	25.0%

Financing

Gearing		
Total debt	$	36,916,588
Total equity	$	86,138,705
Gearing	%	30.00%

Debt financing	*Debt split*	*Repayments*	*Start date*	*Tenure (yrs)*
Sub debt	30.00%	Annuity	01-Jan-23	15
Senior debt	70.00%	Annuity	01-Jan-23	12

Interest		*Base rate*	*Constr margin*	*Ops margin*
Sub debt	%	5.00%	3.50%	3.25%
Senior debt	%	5.00%	1.05%	1.00%

Financing fees		*Commitment*	*Establishment*	*Annual fees*
Sub debt	%	1.00%	1.50%	3,000
Senior debt	%	1.00%	1.50%	3,000

Levelised Cost of Direct Air Capture

LCOD	$	na

Returns

Project cashflows		*NPV*	*IRR*
Pre tax	$	310,155,544	23.78%
Post tax	$	240,687,157	21.49%

Equity cashflows (dividends)		*NPV*	*IRR*
Pre tax	$	95,339,981	25.84%
Post tax	$	66,065,020	23.19%
Post tax (committed basis)	$	66,065,020	23.19%

Ratios

Post tax	*DSCR*	*LLCR*	*PLCR*
Average	8.2982	9.3078	48.2617
Minimum	7.3083	8.8096	23.0747
Date of minimum	01 Jan 28	01 Jan 28	01 Jan 28
Min DSCR for senior debt	x		1.25
Dividend lock up if minimum breached?	switch		Yes

Checks

Balance sheet balances?	*ok*	DSRA fully funded?	*ok*
Sources and uses match?	*ok*	MRA fully funded?	*ok*
Senior debt repaid?	*ok*	Funding macro ok?	*ok*
Sub debt repaid?	*ok*		

Sources and uses

Sources of funds		*Construction*	*Full project phase*
Senior debt drawdown	$	25,841,612	25,841,612
Subordinated debt drawdown	$	11,074,976	11,074,976
Equity injection	$	86,138,705	86,138,705
Cash on deposit interest received	$	n/a	-
Revenue received	$	n/a	2,142,721,542
Sources of funds	$	123,055,293	2,265,776,835

Uses of funds		*Construction*	*Full project phase*
Operating costs paid	$	n/a	849,296,077
Major maintenance cost paid	$	n/a	47,292,282
Corporation tax paid	$	n/a	252,029,086
Overdraft interest paid	$	n/a	-
Capital expenditure	$	126,299,600	131,469,725
Senior debt interest & fees	$	1,756,418	6,583,351
Senior debt principal repayment	$	n/a	25,841,612
Transfers to/from MRA	$	4,632,138	-
Transfers to/from DSRA	$	2,872,820	-
Sub debt interest & fees	$	1,388,164	5,646,596
Sub debt principal repayment	$	n/a	11,074,976
Equity redemption	$	n/a	-
Dividends	$	n/a	1,008,116,344
Cash flow in period	$	(13,893,847)	(71,573,213)
Uses of funds	$	123,055,293	2,265,776,835
Check	n/a	*ok*	*ok*

5.2.3 Lifecycle Cost for Direct Air Capture and Mineralisation

Cashflows over the life of the project

Revenue received Co-Products	$	1,787,167,186
Operating costs paid	$	(849,296,077)
Maintenance costs paid	$	(47,292,282)
Corporation tax paid	$	(180,923,815)
Capital expenditure	$	(131,441,725)
Cashflows for Levelised Cost of Carbon	$	578,213,288

Cashflows include revenues for co-product but do not include revenues for CDR.

5.2.4 Life Cycle Costs (DACMin)

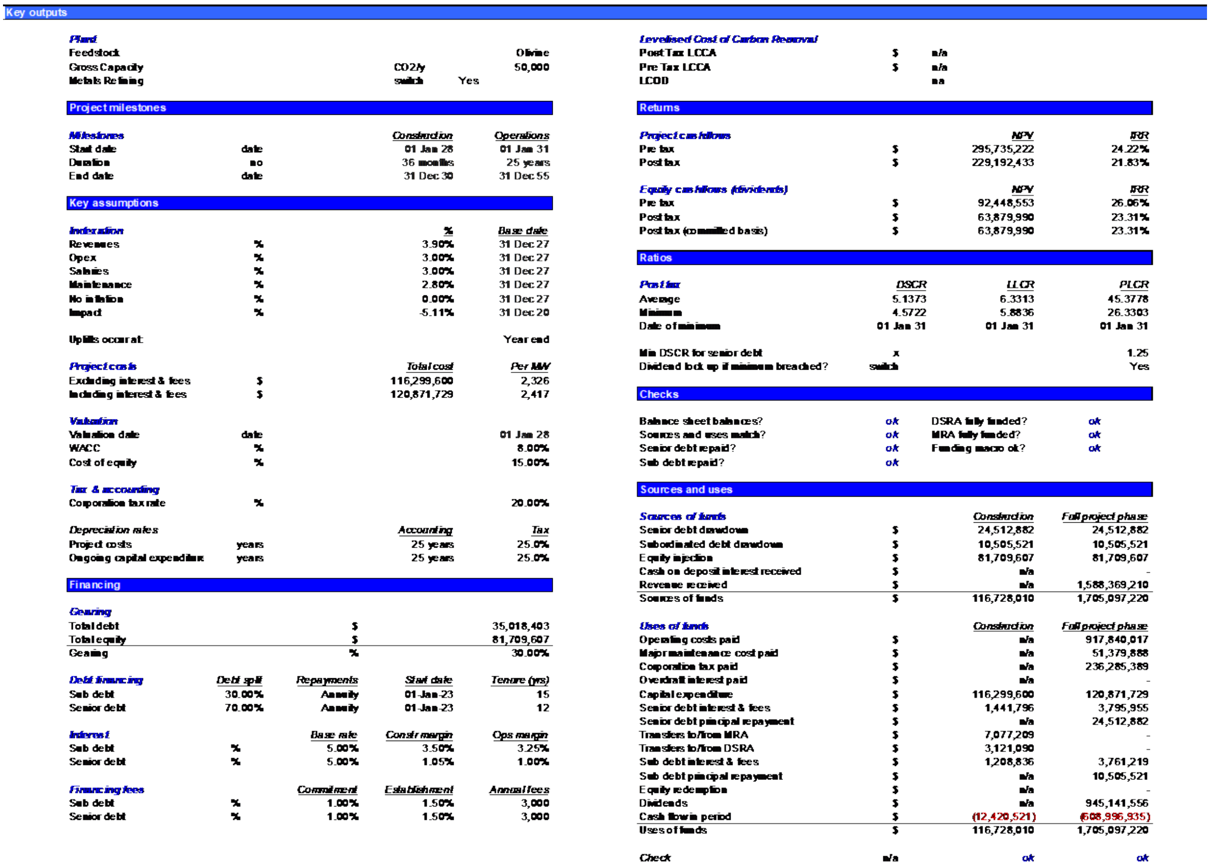
CO_2 Captured over life of project	t/CO2	1,200,000
CO_2 Removed over life of Project	t/CO2	714,375

	Units	Cost per tonne CO_2 Removed	Cost per tonne CO_2 Captured
Lifecycle Cost including Co-Product Revenues	$/t	809	482
Lifecycle Cost Excluding Co Product Revenues	$/t	(1,642)	(1007)

Cost is calculated with and without co-product revenues and per tonne captured and per tonne removed.
Lifecycle costs are positive as revenues received from co-products are greater than the sum of project costs.

5.2.5 DAC Hybrid Model

Dashboard from Hybrid Financial Model



Key outputs

Plant			
Feedstock			Olivine
Gross Capacity	CO2/y		50,000
Metals Refining	switch	Yes	

Project milestones

Milestones		*Construction*	*Operations*
Start date	date	01 Jan 28	01 Jan 31
Duration	no	36 months	25 years
End date	date	31 Dec 30	31 Dec 55

Key assumptions

Indexation		*%*	*Base date*
Revenues	%	3.90%	31 Dec 27
Opex	%	3.00%	31 Dec 27
Salaries	%	3.00%	31 Dec 27
Maintenance	%	2.80%	31 Dec 27
No inflation	%	0.00%	31 Dec 27
Impact	%	-5.11%	31 Dec 20
Uplifts occur at:			Year end

Project costs		*Total cost*	*Per MW*
Excluding interest & fees	$	116,299,600	2,326
Including interest & fees	$	120,871,729	2,417

Valuation		
Valuation date	date	01 Jan 28
WACC	%	8.00%
Cost of equity	%	15.00%

Tax & accounting		
Corporation tax rate	%	20.00%

Depreciation rates		*Accounting*	*Tax*
Project costs	years	25 years	25.0%
Ongoing capital expenditure	years	25 years	25.0%

Financing

Gearing		
Total debt	$	35,018,403
Total equity	$	81,709,607
Gearing	%	30.00%

Debt financing	*Debt split*	*Repayments*	*Start date*	*Tenure (yrs)*
Sub debt	30.00%	Annuity	01-Jan-23	15
Senior debt	70.00%	Annuity	01-Jan-23	12

Interest		*Base rate*	*Constr margin*	*Ops margin*
Sub debt	%	5.00%	3.50%	3.25%
Senior debt	%	5.00%	1.05%	1.00%

Financing fees		*Commitment*	*Establishment*	*Annual fees*
Sub debt	%	1.00%	1.50%	3,000
Senior debt	%	1.00%	1.50%	3,000

Levelised Cost of Carbon Removal		
Post Tax LCCA	$	n/a
Pre Tax LCCA	$	n/a
LCOD		na

Returns

Project cashflows		*NPV*	*IRR*
Pre tax	$	295,735,222	24.22%
Post tax	$	229,192,433	21.83%

Equity cashflows (dividends)		*NPV*	*IRR*
Pre tax	$	92,448,553	26.06%
Post tax	$	63,879,990	23.31%
Post tax (committed basis)	$	63,879,990	23.31%

Ratios

Post tax	*DSCR*	*LLCR*	*PLCR*
Average	5.1373	6.3313	45.3778
Minimum	4.5722	5.8836	26.3303
Date of minimum	01 Jan 31	01 Jan 31	01 Jan 31

Min DSCR for senior debt	x	1.25
Dividend lock up if minimum breached?	switch	Yes

Checks

Balance sheet balances?	ok	DSRA fully funded?	ok
Sources and uses match?	ok	MRA fully funded?	ok
Senior debt repaid?	ok	Funding macro ok?	ok
Sub debt repaid?	ok		

Sources and uses

Sources of funds		*Construction*	*Full project phase*
Senior debt drawdown	$	24,512,882	24,512,882
Subordinated debt drawdown	$	10,505,521	10,505,521
Equity injection	$	81,709,607	81,709,607
Cash on deposit interest received	$	n/a	-
Revenue received	$	n/a	1,588,369,210
Sources of funds	$	116,728,010	1,705,097,220

Uses of funds		*Construction*	*Full project phase*
Operating costs paid	$	n/a	917,840,017
Major maintenance cost paid	$	n/a	51,379,888
Corporation tax paid	$	n/a	236,285,389
Overdraft interest paid	$	n/a	-
Capital expenditure	$	116,299,600	120,871,729
Senior debt interest & fees	$	1,441,796	3,795,955
Senior debt principal repayment	$	n/a	24,512,882
Transfers to/from MRA	$	7,077,209	-
Transfers to/from DSRA	$	3,121,090	-
Sub debt interest & fees	$	1,208,836	3,761,219
Sub debt principal repayment	$	n/a	10,505,521
Equity redemption	$	n/a	-
Dividends	$	n/a	945,141,556
Cash flow in period	$	(12,420,521)	(608,996,935)
Uses of funds	$	116,728,010	1,705,097,220
Check	n/a	ok	ok

5.2.6 Lifecycle Cost for Hybrid DAC plus Passive CDR

Revenue received Co-Products	$	1,773,355,943
Operating costs paid	$	(917,840,017)
Maintenance costs paid	$	(51,379,888)
Corporation tax paid	$	(166,565,916)
Capital expenditure	$	(120,843,095)
Cashflows for Levelised Cost of Carbon	$	516,727,028

5.2.7 Lifecycle Costs Hybrid

CO_2 Captured over life of project	t/CO2	3,363,836
CO_2 Removed over life of Project	t/CO2	2,782,668

CO_2 capture and removal estimates assume 100% reaction of Magnesium Oxide used for passive CDR over time.

	Units	**Cost per tonne CO_2 Removed**	**Cost per tonne CO_2 Captured**
Lifecycle Cost including Co-Product Revenues	$/t	186	154
Lifecycle Cost Excluding Co Product Revenues	$/t	452	(374)

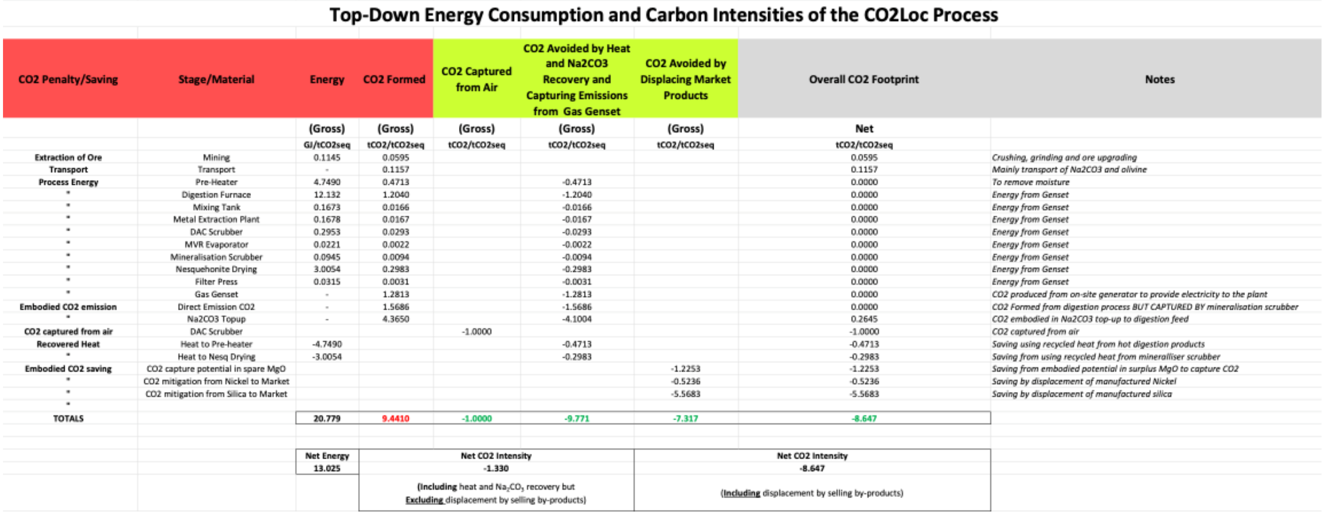
5.3 A life cycle analysis (LCA)

The data for the calculations in the LCA were taken from the heat and mass balances. The aim of this LCA is to highlight stage by stage, the CO_2 intensities of a 50kt/year DAC plant based on the technology of the CO_2Loc Process. Due to the numerous ways of interpreting LCA data, CO_2 intensities are presented under three conditions:

1. Based solely on CO_2 emissions within the process.
2. “1” plus CO_2 avoidance measures such as heat recovery and recycling of digestant.
3. “1” and “2” plus CO_2 avoidance based on product displacement from the marketing of by-products from the process.

The mass and heat flows were modelled in spreadsheets, but for comparison, the process heat balance was also modelled by thermo-chemical software (HSC™ Chemistry) which is presented in section 2.3.

Table 20: Energy and Carbon Intensity by Stage



Top-Down Energy Consumption and Carbon Intensities of the CO2Loc Process

CO2 Penalty/Saving	Stage/Material	Energy	CO2 Formed	CO2 Captured from Air	CO2 Avoided by Heat and Na2CO3 Recovery and Capturing Emissions from Gas Genset	CO2 Avoided by Displacing Market Products	Overall CO2 Footprint	Notes
		(Gross)	(Gross)	(Gross)	(Gross)	(Gross)	Net	
		GJ/tCO2seq	tCO2/tCO2seq	tCO2/tCO2seq	tCO2/tCO2seq	tCO2/tCO2seq	tCO2/tCO2seq	
Extraction of Ore	Mining	0.1145	0.0595				0.0595	*Crushing, grinding and ore upgrading*
Transport	Transport	-	0.1157				0.1157	*Mainly transport of Na2CO3 and olivine*
Process Energy	Pre-Heater	4.7490	0.4713		-0.4713		0.0000	*To remove moisture*
"	Digestion Furnace	12.132	1.2040		-1.2040		0.0000	*Energy from Genset*
"	Mixing Tank	0.1673	0.0166		-0.0166		0.0000	*Energy from Genset*
"	Metal Extraction Plant	0.1678	0.0167		-0.0167		0.0000	*Energy from Genset*
"	DAC Scrubber	0.2953	0.0293		-0.0293		0.0000	*Energy from Genset*
"	MVR Evaporator	0.0221	0.0022		-0.0022		0.0000	*Energy from Genset*
"	Mineralisation Scrubber	0.0945	0.0094		-0.0094		0.0000	*Energy from Genset*
"	Nesquehonite Drying	3.0054	0.2983		-0.2983		0.0000	*Energy from Genset*
"	Filter Press	0.0315	0.0031		-0.0031		0.0000	*Energy from Genset*
"	Gas Genset	-	1.2813		-1.2813		0.0000	*CO2 produced from on-site generator to provide electricity to the plant*
Embodied CO2 emission	Direct Emission CO2	-	1.5686		-1.5686		0.0000	*CO2 Formed from digestion process BUT CAPTURED BY mineralisation scrubber*
"	Na2CO3 Topup	-	4.3650		-4.1004		0.2645	*CO2 embodied in Na2CO3 top-up to digestion feed*
CO2 captured from air	DAC Scrubber			-1.0000			-1.0000	*CO2 captured from air*
Recovered Heat	Heat to Pre-heater	-4.7490			-0.4713		-0.4713	*Saving using recycled heat from hot digestion products*
"	Heat to Nesq Drying	-3.0054			-0.2983		-0.2983	*Saving from using recycled heat from mineralliser scrubber*
Embodied CO2 saving	CO2 capture potential in spare MgO					-1.2253	-1.2253	*Saving from embodied potential in surplus MgO to capture CO2*
"	CO2 mitigation from Nickel to Market					-0.5236	-0.5236	*Saving by displacement of manufactured Nickel*
"	CO2 mitigation from Silica to Market					-5.5683	-5.5683	*Saving by displacement of manufactured silica*
"								
TOTALS		**20.779**	**9.4410**	**-1.0000**	**-9.771**	**-7.317**	**-8.647**	
		Net Energy	**Net CO2 Intensity**			**Net CO2 Intensity**		
		13.025	**-1.330**			**-8.647**		
			(**Including** heat and Na_2CO_3 recovery but **Excluding** displacement by selling by-products)			(**Including** displacement by selling by-products)		

Table 20 shows a detailed energy and carbon intensity stage-by-stage breakdown of the CO_2Loc process. A positive CO_2 intensity indicates a CO_2 penalty, and a negative one indicates CO_2 avoided:

- Columns 3 and 4 show the energy consumption and associated carbon intensities.
- Column 5 shows CO2 captured from the air.
- Column 6 shows CO_2 avoided by heat and mass recovery: i.e., recovery of heat from the hot solid products of digestion and from carbonation in the mineralization scrubber.
- Column 7 shows CO_2 avoided by selling by-products and subsequently displacing the equivalent currently manufactured product.
- Column 8 shows the overall carbon intensity of each process stage.

The CO_2 emission factors used in the LCA are sourced from the DESNZ website.

The CO_2 intensity of the process without CO_2 avoidance measures is 9.4410 tCO_2/tCO_{2seq}. When heat and mass recovery are factored in, the value is -1.330 tCO_2/tCO_{2seq} and if product displacement is factored in the value is -8.647 tCO_2/tCO_{2seq}. Table 2 summarizes this data:

Table 21: LCA Summary

$(tCO2e/tCO2_{seq})$	**CO_2 Emissions**	**Net CO_2 with Recycling Heat and Na_2CO_3**	**Net CO2 with Recycling and Product Displacement**
Including CO2 embodied in Na2CO3	9.4410	-1.330	-8.647

Based on the data and calculations in the "Data Pool and Calcs" worksheet the process is carbon negative even without the CO_2 mitigation from manufactured products displacement. It should be noted that the displacement of manufactured products in current use is complex and that the associated CO_2 credits depend on the impact of the by-products on future markets.

The revenue by-products are nickel and amorphous silica, but of the 12.5t/h magnesium oxide (MgO) produced, only 46wt% is required to capture the CO_2 from the digestion. The rest is surplus and can be sold as a CO_2 sequestrant or for feedstock for making magnesium.

A schematic of the process used in the LCA is shown in Figure 21.

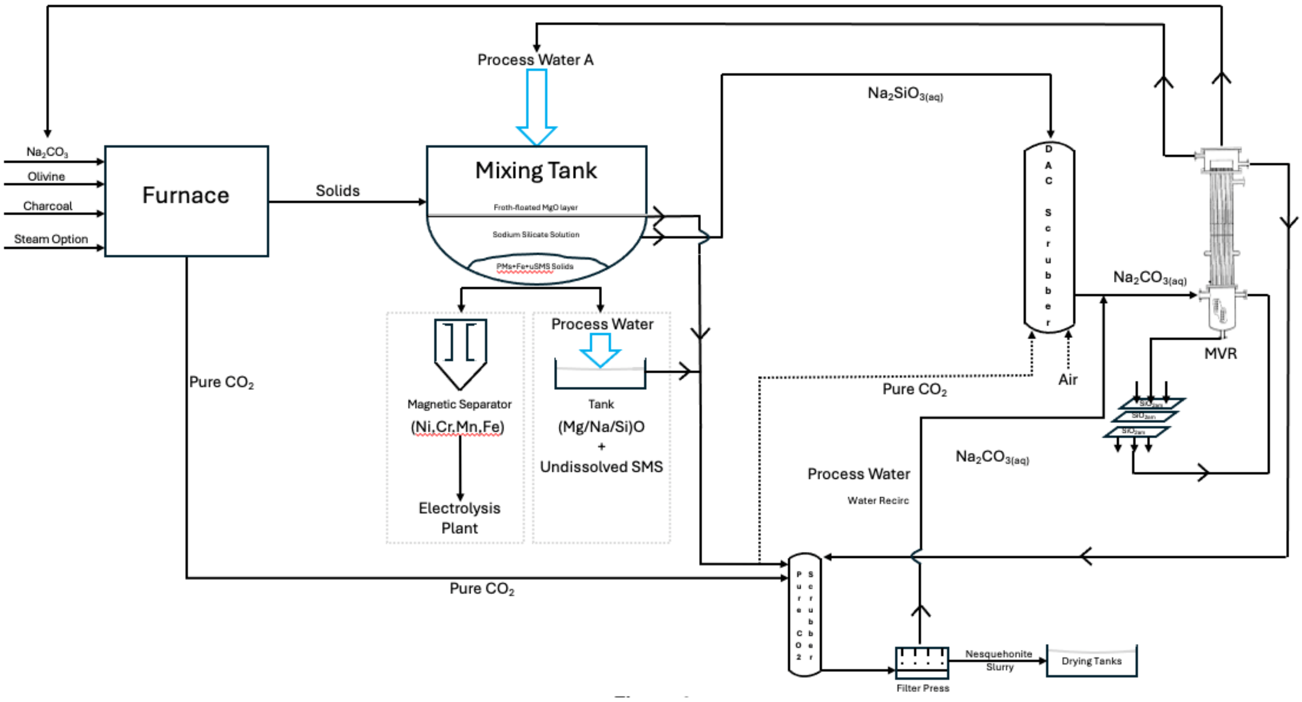


Figure 21: LCA Process Schematic

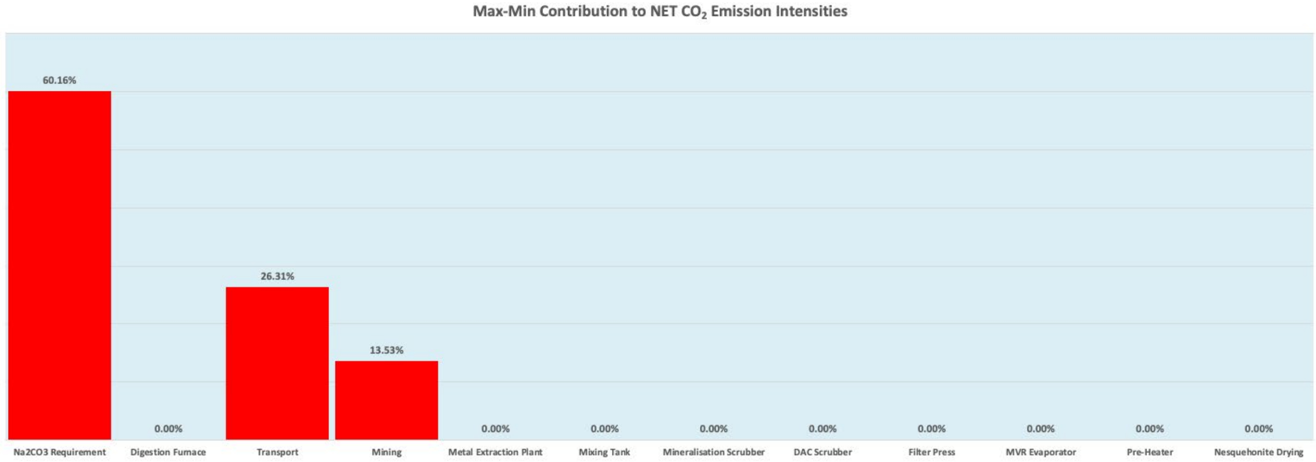


Figure 22: Summary by process step

The graph in Figure 22 clearly shows that use of top-up Na_2CO_3 is the main contributor to the net CO_2 intensity of the process. Note that digestion emissions and all electrically driven processes are now zero due to the use of carbon capture from the onsite natural gas powered genset using surplus MgO.

The graph in figure 23 shows clearly that heat and Na_2CO_3 recovery and capture of the CO_2 from digestion and genset avoid most gross CO_2 emissions in the process.

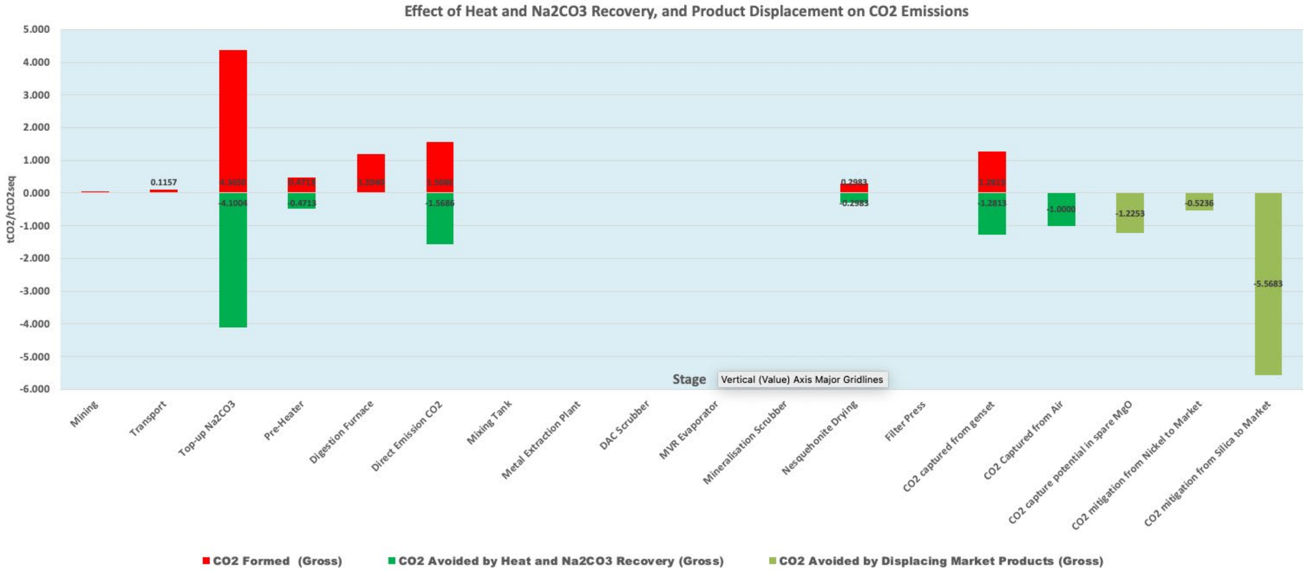


Figure 23: CO2 Emitted vs Recovered

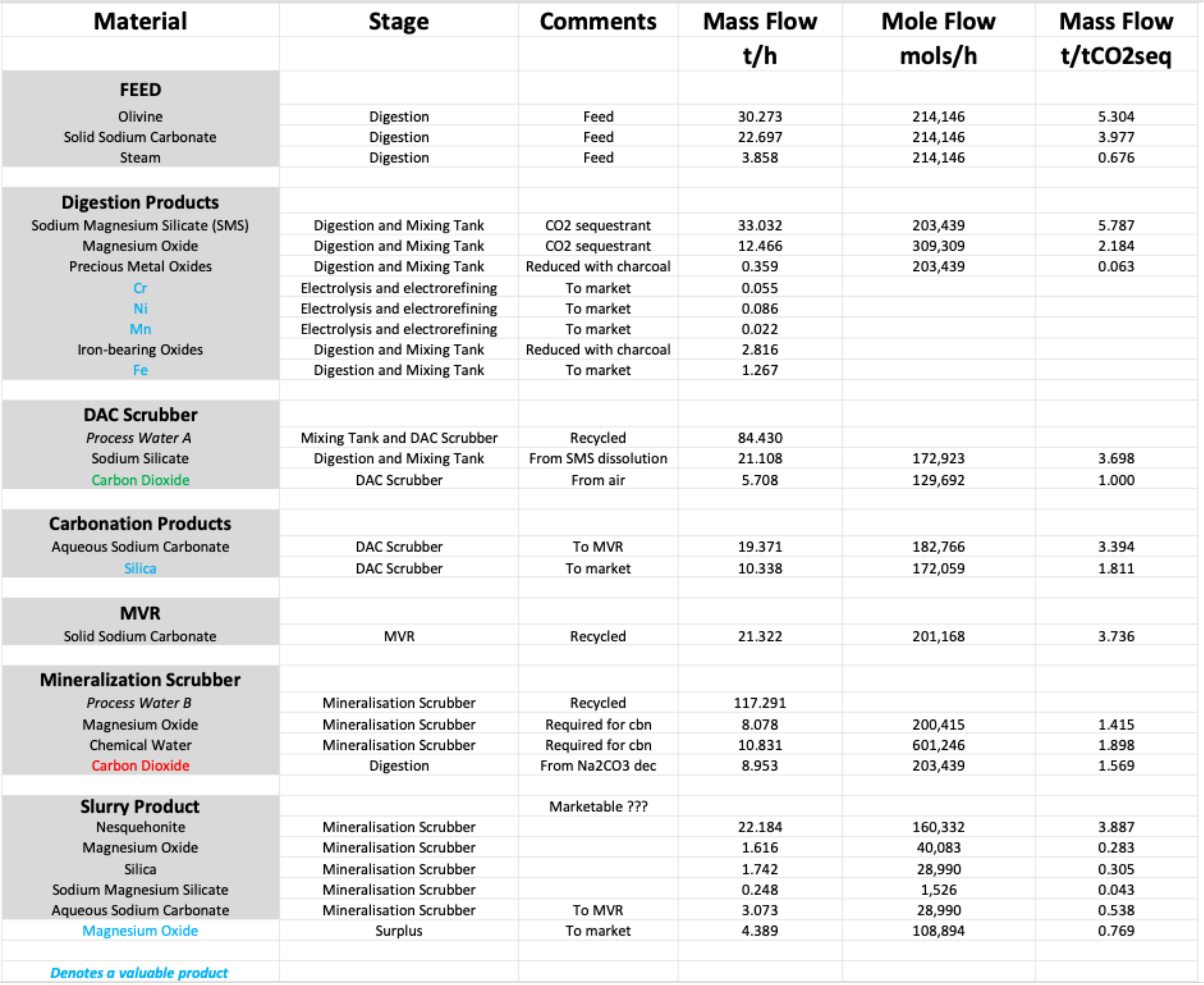
The CO_2 intensity of the Solvay Process to produce Na_2CO_3 is 1.0977 tCO_{2seq}/tNa_2CO_3 (See Data Pool and Calculations). The mass balance summary in Table 23 below shows the CO_2Loc process requires a total of 3.9765 t/tCO_{2seq} with 3.736 t/tCO_{2seq} being recycled, leaving 0.2410 t/tCO_{2seq} to be provided by top-up and meaning 94% of the Na_2CO_3 is recycled. The top-up Na_2CO_3 corresponds to a CO_2 intensity of 0.2645 tCO_2/tCO_{2seq}.

The improvements being made to the Solvay Process and the possibility of mining sodium carbonate should enable the CO_2Loc process to be even more carbon negative without having to rely on manufacture displacement from our by-products.

5.3.1 Summary of Mass Balance and Assumptions

The mass balance is based on the following criteria and can be followed in the process flow diagram in Figure 21 and the mass balance summary in Table 22.

Table 22: Mass Balance Summary



Material	Stage	Comments	Mass Flow	Mole Flow	Mass Flow
			t/h	mols/h	t/tCO2seq
FEED					
Olivine	Digestion	Feed	30.273	214,146	5.304
Solid Sodium Carbonate	Digestion	Feed	22.697	214,146	3.977
Steam	Digestion	Feed	3.858	214,146	0.676
Digestion Products					
Sodium Magnesium Silicate (SMS)	Digestion and Mixing Tank	CO2 sequestrant	33.032	203,439	5.787
Magnesium Oxide	Digestion and Mixing Tank	CO2 sequestrant	12.466	309,309	2.184
Precious Metal Oxides	Digestion and Mixing Tank	Reduced with charcoal	0.359	203,439	0.063
Cr	Electrolysis and electrorefining	To market	0.055		
Ni	Electrolysis and electrorefining	To market	0.086		
Mn	Electrolysis and electrorefining	To market	0.022		
Iron-bearing Oxides	Digestion and Mixing Tank	Reduced with charcoal	2.816		
Fe	Digestion and Mixing Tank	To market	1.267		
DAC Scrubber					
Process Water A	Mixing Tank and DAC Scrubber	Recycled	84.430		
Sodium Silicate	Digestion and Mixing Tank	From SMS dissolution	21.108	172,923	3.698
Carbon Dioxide	DAC Scrubber	From air	5.708	129,692	1.000
Carbonation Products					
Aqueous Sodium Carbonate	DAC Scrubber	To MVR	19.371	182,766	3.394
Silica	DAC Scrubber	To market	10.338	172,059	1.811
MVR					
Solid Sodium Carbonate	MVR	Recycled	21.322	201,168	3.736
Mineralization Scrubber					
Process Water B	Mineralisation Scrubber	Recycled	117.291		
Magnesium Oxide	Mineralisation Scrubber	Required for cbn	8.078	200,415	1.415
Chemical Water	Mineralisation Scrubber	Required for cbn	10.831	601,246	1.898
Carbon Dioxide	Digestion	From Na2CO3 dec	8.953	203,439	1.569
Slurry Product		Marketable ???			
Nesquehonite	Mineralisation Scrubber		22.184	160,332	3.887
Magnesium Oxide	Mineralisation Scrubber		1.616	40,083	0.283
Silica	Mineralisation Scrubber		1.742	28,990	0.305
Sodium Magnesium Silicate	Mineralisation Scrubber		0.248	1,526	0.043
Aqueous Sodium Carbonate	Mineralisation Scrubber	To MVR	3.073	28,990	0.538
Magnesium Oxide	Surplus	To market	4.389	108,894	0.769
Denotes a valuable product					

Note that recycling efficiencies were estimated and accounted for in the mass balance and can be changed:

1. Olivine sourced from Sibelco whose composition was determined by XRF.
2. Digestion of olivine with sodium carbonate in N_2 or reducing gas mixed atmosphere
    a. *For clarity, the metal oxide reduction step with carbon has been omitted from the mass balance.*
    b. *The digestion can also be performed with or without added steam, although steam can lower the operating temperature and provide a heat transfer medium.*
3. Separation of solid phases:
    a. Precious metals by settling in suspension and electromagnet.
    b. Iron oxides by settling in suspension and electromagnet.

    c. Undissolved SMS and unreacted forsterite residual after magnetic separation.
    d. MgO by froth flotation.
4. Separation of Aqueous Phase:
    a. SMS (Sodium Magnesium Silicate) and unreacted Na_2CO_3 from digestion is dissolved into aqueous solution and sent to DAC scrubber.
5. The sodium silicate solution captures CO_2 from air forming Na_2CO_3 and silica.
6. The heavy metal solids fraction is sent to the metal recovery plant to recover Ni, Cr and Mn.
7. Any remaining unreacted Na_2SiO_3 in the DAC solution is carbonated with 100% CO_2 from the digestion.
8. The spent DAC Na_2CO_3 solution is sent to evaporators to recover solid Na_2CO_3 which feeds back into the digestor, and process water. There is also a ~0.1wt% Na_2SiO_3 fraction which will be recovered with the Na_2CO_3.
9. The mineralizer scrubber carbonates the MgO and undissolved SMS to form a slurry consisting of nesquehonite ($MgCO_3$*$3H_2O$), MgO and any other unreacted solids. It will be more economic not to separate the small amount of MgO (~11wt%) and SiO_2 (~7wt%) from this slurry.
10. Only 54% of the MgO that is formed from the digestion is required to capture CO_2 from the digestion. The remaining MgO is separated prior to carbonation, stored, and sold. Summary of the Process

The following process stage efficiencies shown in Table 23 were used and will of course impact on the figures in the energy balance and the LCA:

Table 23: Assumed Efficiencies

Stage	**Efficiency**
Digestion	95%
Dissolution of SMS	85%
DAC Scrubber	75%
Carbonation of unreacted Na_2SiO_3	98%
Na_2CO_3 Recovery	95%
Mineralization Scrubber	80%
Carbonation of undissolved SMS	95%

The mass balance was verified by comparing magnesium and sodium content from olivine and Na_2CO_3 to the terminal products in the process, namely Fe,Na,Mg,Si oxides (prior to reduction), $MgCO_3$*$3H_2O$ (nesquehonite), MgO, and Na_2CO_3 and unreacted SMS. As can be seen in Figure 5, the error of 0.4% is due to uncertainty in the composition of the remaining unreduced metal oxide fraction.

The following description is based on sequestering CO_2 from air on a scale of 50,000t/yr. The process is continuous and consists of six key stages:

i) Thermal digestion
ii) Dissolution of sequestrant
iii) Separation of solid and aqueous fractions
iv) Direct Air Capture (DAC) in a scrubber
v) Recovery of digestant (Na_2CO_3)

vi) Capture or storage of CO_2 from thermal digestion.

A summary of the mass balance is shown in Table 5.

5.3.2 Thermal Digestion

The deposit of olivine used in the R&D and pilot plant trials was supplied by Sibelco from a Norwegian mine and is remarkable in its low iron content. It is assumed that this olivine would initially be used in a 50,000t/yr plant.

Olivine (Mg - 29.08wt% Fe - 2.4680wt% Si - 20.118wt%), sodium carbonate (Na_2CO_3) and steam are fed into a furnace and mixed at a minimum operating temperature of 900C. As the Na_2CO_3 reacts with the olivine, it emits CO_2 forming Na_2O, a very reactive nucleophile that disrupts the Si—O and M—SiO4 bonds, where M is normally Mg, Ni, Cr, or Mn.

Typically, the yield from digestions performed consisted of sodium magnesium silicate (SMS) (77wt%) and magnesium oxide (MgO) (13wt%), and small residual fractions comprising metal Oxides (~1wt%), Fe phases (~6.5wt%) and sometimes unreacted Na_2CO_3 (~1-2.5wt%). The proportion of the residual phases can vary but rarely exceed 10wt%. If charcoal is used in the feed the transition metal oxides are reduced to metal.

The yield of the digestion is defined on how much of the original forsterite component of the olivine (Mg_2SiO_4) is converted to Na_2MgSiO_4 and MgO as shown in this simplified reaction:

$$Mg_2SiO_4 + Na_2CO_3 = Na_2MgSiO_4 + 0.6704MgO + CO_2$$

The above reaction if gone to completion would represent a 100% yield. However, olivine contains iron (as iron silicate, fayalite) which is reactive and competes with the Mg_2SiO_4 forming various magnesium iron oxides and magnesium iron silicates. To what degree this happens dictates the digestion yield. Without a reducing atmosphere, yields were typically between 75 and 80% at optimal digestion conditions represented in the chemical equation below:

$$\begin{gathered}\text{Sibelco Olivine: } Mg_{1.6704}\,Fe_{0.0617}\,(SiO3_{.8553})\,(H_2O)_{0.3215} \\ + \\ Na_2CO_3 + H_2O(g) \\ = \\ (1-y)Na_2MgSiO_4 + (0.6704-x)MgO \\ + \\ \{ax(Mg, Fe_2)O_4\} + \{bx(Mg, Fe^{2+/3+},)SiO_4\} + \{y(Na_2Fe_2O_4\} \\ \text{where } 2x(a+b) + 2y = 0.0617 \\ + \\ CO_2(g) + 1.3215H_2O(g)\end{gathered}$$

X is the inefficiency caused by magnesium iron oxide and magnesium iron silicate formation, and y is the inefficiency caused by sodium iron oxide formation. If x=0 and y=0, the yield is 100%. The figure 0.0617 is the total Fe in the olivine normalized to $SiO_{3.8553}$, as derived from XRF analysis.

If charcoal is used in the feed, the magnesium iron oxides and most of the magnesium iron silicates are reduced iron allowing the resultant MgO to react with Na_2CO_3 to form more SMS. Typically, yields under reducing conditions are between 90 and 95%.

The precious metal oxide (PMO) fraction was omitted from the above for clarity but can be represented by $\{Al_{l0.0106}, Cr_{0.0042}, Ni_{0.0058}, Mn_{0.0016}, O_{0.0520}\}$, as derived from XRF analysis and represents 1.25wt% of the olivine.

A 1-2wt% addition of charcoal to the feed mix has shown that the transition precious metal oxide fractions can be reduced to their metals making their separation easier. Nickel at the 50kt scale, is a valuable by-product with an exceptionally low carbon footprint tag compared to current industrial extraction routes (see LCA). The nickel could be sold crude, or electrorefined in an electrolysis plant.

The digestion can be performed without steam, however there are two key advantages to using steam:

a) Sodium hydroxide (NaOH) forms by hydrolysis of Na_2CO_3 which can lower the operating temperature by 50-100C.
b) Steam could be used to recover heat from the hot solids exiting the furnace.

The cooled, solid products from the digestion are fed into a mixing/settling tank.

5.3.3 Dissolution of SMS

Process water (84t/h) is added at this stage to dissolve most of the SMS. The volume of water added is controlled to form a 20wt% solution of sodium silicate (Na_2SiO_3).

$$Na_2MgSiO_{4(s)} = Na_2SiO_{3(aq)} + MgO_{(s)}$$

Any unreacted Na_2CO_3 will also dissolve. A slurry is formed comprising ~20wt% solids and ~80wt% aqueous fraction.

5.3.4 Separation of Solids

The MgO formed from dissolution of SMS and MgO already present from the digestion is separated from the heavy iron-containing fractions and undissolved SMS by froth flotation.

The MgO by-product is skimmed off the surface of the tank and stored for use in the mineralization scrubber. There is also a large excess of MgO (5.7t/h) which can be sold.

The precious metals (PMs) and iron phases are allowed to settle to the bottom of the tank and magnetically separated leaving behind a small magnesium iron silicate (forsterite) fraction and undissolved SMS which is sent to the mineralization scrubber to capture CO_2 from the digestion.

5.3.5 DAC

The Na_2SiO_3 solution is pumped into the scrubber and air introduced to capture the CO_2:

$$Na_2SiO_{3(aq)} + CO_{2(g)} = Na_2CO_{3(aq)} + SiO_{2(s)}$$

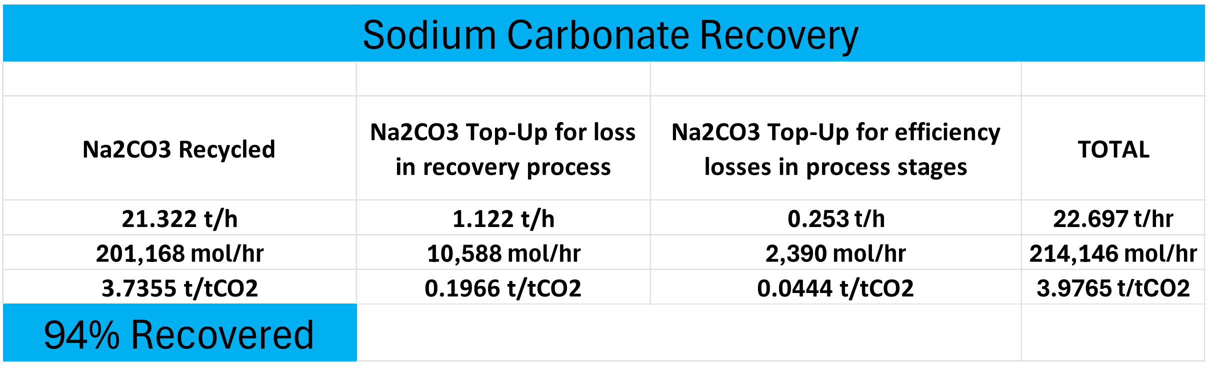
The Na_2SiO_3 is consumed in a batch process where a Na_2CO_3 solution is formed containing dissolved and colloidal silica. A 100% stream of CO_2 from the digestion is then switch in to carbonate any remaining unreacted Na_2SiO_3. The spent solution is pumped out of the scrubber into a tank where a coagulant is added to precipitate out the silica, which is filtered and dried. The silica by-product is a revenue product and can be sold.

The sodium carbonate solution is passed onto the next stage.

5.3.6 Recovery of digestant (Na_2CO_3)

The solution is pumped into a Mechanical Vapor Recompression (MVR) unit to recover the Na_2CO_3 solid and sent back to the digestor. The water is recycled and sent back to the SMS dissolution tank. Table 24 shows the details of the Na_2CO_3 recovery:

Table 24: Sodium Carbonate Recovery



Sodium Carbonate Recovery

Na2CO3 Recycled	Na2CO3 Top-Up for loss in recovery process	Na2CO3 Top-Up for efficiency losses in process stages	TOTAL
21.322 t/h	1.122 t/h	0.253 t/h	22.697 t/hr
201,168 mol/hr	10,588 mol/hr	2,390 mol/hr	214,146 mol/hr
3.7355 t/tCO2	0.1966 t/tCO2	0.0444 t/tCO2	3.9765 t/tCO2
94% Recovered			

5.3.7 Capture of CO_2 from thermal digestion

The undissolved SMS and a portion of the MgO from the mixing tank are added to process water (105t/h) to form a slurry which is pumped into the mineralization scrubber to carbonate the CO_2 from the digestion:

$$MgO_{(\text{from digestor})} + Na_2MgSiO_4 + 3H_2O + CO_{2(g)} = MgCO_3(3H_2O) + Na_2CO_3 + MgO_{(\text{from SMS})} + SiO_2$$

The bulk of the resultant slurry consists of nesquehonite (80wt%), MgO (10wt%), SiO_2 (7wt%) and unreacted SMS (3wt%). This mixture may have uses in the construction industry.

The spent process water consists of a 2.9wt% Na_2CO_3 which is sent to the MVR along with the carbonated DAC solution for recovery.

Other options are to capture some of the digestion CO_2, sell some of it (the CO_2 is very pure) and store the remainder under the sea, freeing up more MgO to be sold. Various combinations of these options could make the process more economic and offer a lower carbon footprint.

5.4 Any improvements between the pilot project and commercialised technology should be identified and justified.

Over the project a number of synergistic manifestations of the DACMin process were identified and explored. These included the realisation that after process emissions had been mineralised, there was a surplus of Magnesium Oxide. In the simplest form, sale of the surplus into existing MgO markets represented a further revenue stream, improving the economics of the process. However, if the electricity used by the plant was generated by a natural gas fuelled genset, the surplus MgO could be used to capture the GHG emissions from the genset further improving the carbon efficiency of the DACMin process. This was the scenario modelled in the LCA in section 5.3. This process variant was a big advantage when looking at options for the siting of the DACMin plant. For example, there is a benefit in siting the DACMin plant near or at the mine site as this would significantly reduce the scope 2 emissions associated with the transportation of the olivine to the plant. The mine sites are usually located in remote locations without access to grid infrastructure and often would have dedicated power generators to power operations at the mine site. In this scenario, the addition of a DACMin plant could be a way of tackling emissions from mining operations, further reducing scope 2 emissions of the DACMin facility.

Siting of the plant has a significant impact on the LCA and TEA of a DACMin plant, another scenario explored was the location of the plant at a port where there was access to CCS infrastructure. In this scenario, scope 1 emissions could be directed into the CCS pipelines liberating significant quantities of Magnesium Oxide. This offered the opportunity for hybridisation of the DACMin plant to offer both active direct air capture and passive air capture through use of the liberated magnesium oxide for enhanced weathering or ocean alkalinity enhancement. This has the effect of more than doubling the carbon drawdown capacity of the plant, more than halving the levelized cost of carbon for the project.

The versatility of the format of the DACMin project would enable selection of early projects offering the best economics and therefore best returns on investment for investors in early projects where technology and commercial risks are highest.

5.5 Process risks.

Energy Consumption: High energy requirements for DAC operations can lead to inefficient processes and increased costs, impacting scalability and economic viability.
Technology Scalability: Challenges in scaling up from pilot plants to full commercial operations may reveal inefficiencies or unforeseen complications.
Material Degradation: Sorbent or filter materials used in DAC systems may degrade faster than expected, reducing the lifespan and efficiency of the technology.
Maintenance and Reliability: Emerging technologies often face higher maintenance needs and unexpected downtime due to unproven machinery and control systems.
Integration Issues: Difficulties integrating DAC systems with renewable energy sources or existing industrial processes could affect consistency and output.
Operational Safety: New technologies may introduce safety risks, such as chemical handling hazards or mechanical failures, especially under high temperatures and pressures.
CO2 Purity and Handling: Ensuring the purity of captured CO2 and developing reliable methods for its transport, storage, or utilization can be problematic in the early stages.
Process Efficiency: Optimizing the capture process to maximize CO2 extraction rates while minimizing energy input remains a significant hurdle for novel DAC systems.
Environmental Impact: Any unforeseen impacts from resource consumption (e.g., water use) or emissions produced by DAC operations themselves.
Supply Chain Dependencies: Dependence on specialized materials or technologies that may not be readily available or could face disruptions.

These risks need to be carefully managed to ensure the successful scaling and commercialization of direct air capture technologies.

5.6 The Monitoring, Reporting and Verification (MRV) methodology.

An MRV (Measurement, Reporting, and Verification) methodology is necessary to ensuring the credibility and transparency of carbon capture and sequestration processes. For the DACMin System, this methodology involves a structured approach to quantify, monitor, and verify the amount of CO2 captured from the atmosphere and subsequently sequestered through mineralisation.

5.6.1 Key Components of the MRV Methodology:

- Measurement:
    - CO2 Captured: Implementing precise sensors and meters to measure the volume of CO2 extracted from the air by the DAC unit. This includes regular calibration of equipment to ensure accuracy.
    - Mineralisation Process: Tracking the conversion of captured CO2 into stable magnesium carbonates through reaction with magnesium hydroxide produced by the DACMin process.
    - Material Input/Output: Monitoring the quantity and type of reactants used and products formed to ensure complete carbon accounting.
- Reporting:
    - Data Collection: Detailed recording of all measurements, including CO2 capture rates, mineralisation yield, and system performance.
    - Transparency Protocols: Adopting standardised reporting formats to ensure data comparability and adherence to international carbon reporting standards.
    - Emission Accounting: Documenting any emissions generated during the operation (e.g., energy consumption, transport of materials) to provide a net CO2 balance.
- Verification:
    - Third-Party Audits: Engaging independent certifiers to verify reported data and confirm the validity of CO2 sequestration claims.
    - Validation of Process Efficiency: Assessing system efficacy through periodic checks to validate that mineralisation is durable and results in long-term carbon storage.
    - Compliance with Regulations: Ensuring that the MRV process adheres to local and international carbon capture and storage (CCS) guidelines, such as those set by the IPCC or ISO standards.

5.6.2 Challenges and Considerations:

- Energy Use: Monitoring the energy source for DAC systems to ensure the carbon footprint is minimal and does not undermine the sequestration benefits.
- Lifecycle Analysis: Conducting comprehensive lifecycle assessments to capture any indirect emissions associated with the construction, maintenance, or end-of-life disposal of the system.
- Scalability and Cost: Developing cost-effective and scalable MRV methodologies that do not impose excessive operational burdens on DACMin facilities.

This MRV methodology ensures that a DACMin system operates transparently, with validated outcomes, contributing to global climate mitigation efforts effectively.

5.7 Environmental and social impacts.

DACMin technology, while promising for carbon reduction, comes with notable environmental impacts that must be managed. One critical concern is the increased mining required to supply olivine used. The selection of mining sites needs to prioritize minimal disruption to ecosystems, with careful assessment of local biodiversity and water resources. Additionally, transportation of these raw materials from mines to processing facilities can contribute to carbon emissions, emphasizing the importance of building eco-friendly infrastructure that minimizes habitat fragmentation and pollution. Early DACMin plants will be supplied by Sibelco’s Aheim mine in Norway which has sufficient annual capacity to supply at least one 1mt/y DACMin plant and has reserves of 150 years.

The scale of DAC systems presents further challenges; to achieve meaningful CO2 removal, extensive air contactors must be constructed, which could lead to significant land use. This raises the issue of site placement—ensuring these facilities do not encroach upon productive farmland, natural habitats, or areas crucial for conservation. The balance between land utilization for DAC and maintaining spaces for wildlife and agriculture is essential. Proper planning and integration into existing industrial zones or non-arable land can help mitigate these impacts and promote a more sustainable approach to carbon capture deployment.

5.8 How the solution could be scaled, any technical challenges related to scaling and potential cost reductions through scaling.

5.8.1 Getting to scale

The path from pilot to full scale DACMin projects will require our pilot to be further developed, integrating the heat processing with the DAC function of the current pilot. Crucially, this will enable heat recovery between processes and the linking of process emissions with the mineralisation step. The next step would be the first of a kind (FOAK) plant, possibly between 1 -10KtCO2/year scale. This would validate the technology and prove the process economics for a 50ktCO2/year plant and subsequently enable the financing of further scale up projects through normal project finance models.

5.8.2 Technical Challenges

Calciners are one of the biggest CAPEX items in our proposed DACMin plant. As we are proposing electrically heated kilns there is a limit to the size we can go to to achieve the required heat transfer into the digestate. It is therefore necessary to have a larger number of smaller kilns rather than a small number of bigger kilns. On one side, this means that we benefit from the drop in price due to large scale manufacture of similar kilns, on the other, this requires the building up of global production of kilns of this type. Current global production is between a few hundreds to low thousands, a 50kt plant would require 4 and a 1Mt plant 80, so scale up of the technology would require significant scale up of the supply-chains of these kilns as well as other plant.

The efficiency of the process, and therefore the carbon intensity, relies on high levels of heat recovery within the DACMin plant. Much of the recovered heat is from the solids coming out of the kiln so there is a need for heat exchange between solids and gases. Current heat exchanger technology focuses on gas to gas, liquid to liquid or liquid to gas, very few options are currently available for solid to gas or liquid. Therefore, extensive R&D effort will be needed from the supply chain in order to achieve the levels of heat recovery required. This is a technical and commercial risk to scale up projects.

Poor dissolution of the SMS was also a major setback to the project. Although the ionic bead solution was found, this step has the potential to add significant complexity and cost to the process. Further R&D effort would be required to optimise the recovery of the silica from the beads and their regeneration. These ionic beads are routinely used in the water processing industry so the team are confident that the right formulation of beads could be developed to meet our specific needs. There is also an opportunity to investigate alternative solution that there was not time or budget to explore within the DACMin project.

5.8.3 Cost reductions through scaling

Scaling up the DACMin process can significantly reduce costs through operational efficiencies, strategic partnerships, energy-sharing mechanisms, and improved logistics. The potential to increase revenue streams from by-products through vertical integration and leverage government incentives further supports cost-effective scaling. Some examples are:

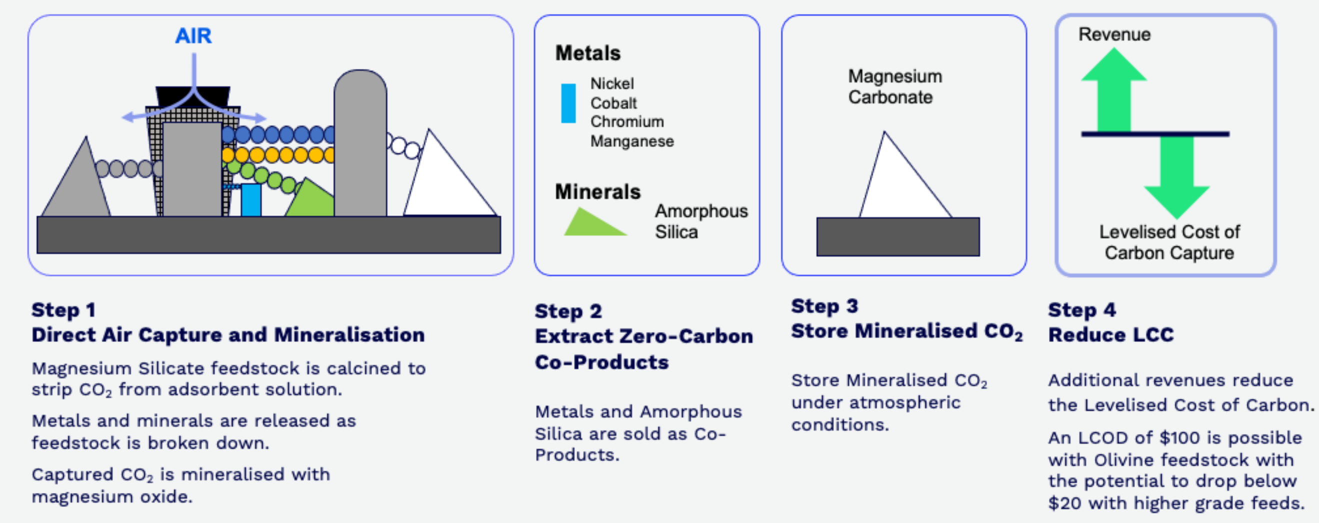
- Economies of scale in equipment and infrastructure.
- Bulk purchase of chemicals and feedstock.
- Further automation and process integration, particularly heat recovery and management.
- Access to cheaper project finance as business models are validated through commercial success.
- Scaling allows for incremental improvements and optimizations based on data, leading to cost reductions through process efficiencies.
- Learning curve and standardisation.
- At scale, dedicated renewables generation could be built to power the DACMin plant, further improving the net carbon removal of the plant.

6 detailed assessment of the business plan for how the GGR solution will continue to be developed after the funding for the Pilot demonstration ends

6.1 Finance Model

The structure of the DACMin project included a work package named Business and Commercial Planning. This was an opportunity for the CCC management team to take the learning from the pilot project and explore business models enabling the raising of investment needed to continue refinement and scale-up of the technology. Central to this activity was the creation of a finance model incorporating the Lifecycle and Techno-Economic Analysis of the scaled-up process, mapping out the journey from pilot, through first of a kind (FOAK), and to first commercial projects. The finance model was built in Excel based on a project model provided by KPMG as part of the business support package offered as part of the GGR programme. The model is designed to test a number of scenarios set by the scenario Manager tab and input options set in the inputs tab. This enabled different scenarios to be modelled and allowed assessment of sensitivities to specific variables such as commodity prices and format of the DACMin plant. Outputs include income statement, balance sheet, cashflow, IRR calculations and Impact statement.

The financial model allowed for various operational formats. The initial scope of the DACMin project was to mineralise all the captured CO2 plus process emissions, see figure 24.



Step 1
Direct Air Capture and Mineralisation

Magnesium Silicate feedstock is calcined to strip CO_2 from adsorbent solution.

Metals and minerals are released as feedstock is broken down.

Captured CO_2 is mineralised with magnesium oxide.

Step 2
Extract Zero-Carbon Co-Products

Metals and Amorphous Silica are sold as Co-Products.

Step 3
Store Mineralised CO_2

Store Mineralised CO_2 under atmospheric conditions.

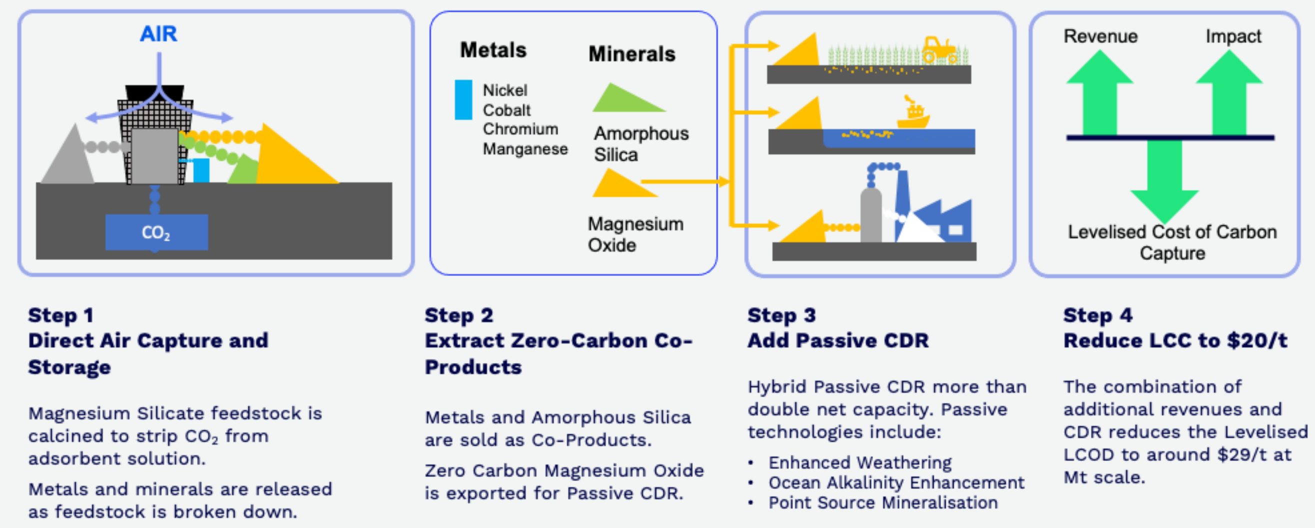
Step 4
Reduce LCC

Additional revenues reduce the Levelised Cost of Carbon.

An LCOD of $100 is possible with Olivine feedstock with the potential to drop below $20 with higher grade feeds.

Figure 24: Direct Air Capture, Mineralisation and Storage Variant

Another variant was possible if the plant was located where CCS infrastructure was available. This enabled the captured CO2 and process emissions could be sent to underground storage, freeing up the Magnesium Oxide, which would have been used to mineralise the CO2, for use for passive CO2 removal via enhanced weathering or ocean alkalinity enhancement. It could also be used for point source CO2 capture and mineralisation. See figure 25.



Step 1
Direct Air Capture and Storage

Magnesium Silicate feedstock is calcined to strip CO_2 from adsorbent solution.

Metals and minerals are released as feedstock is broken down.

Step 2
Extract Zero-Carbon Co-Products

Metals and Amorphous Silica are sold as Co-Products.

Zero Carbon Magnesium Oxide is exported for Passive CDR.

Step 3
Add Passive CDR

Hybrid Passive CDR more than double net capacity. Passive technologies include:

- Enhanced Weathering
- Ocean Alkalinity Enhancement
- Point Source Mineralisation

Step 4
Reduce LCC to $20/t

The combination of additional revenues and CDR reduces the Levelised LCOD to around $29/t at Mt scale.

Figure 25: Hybrid Direct Air Capture

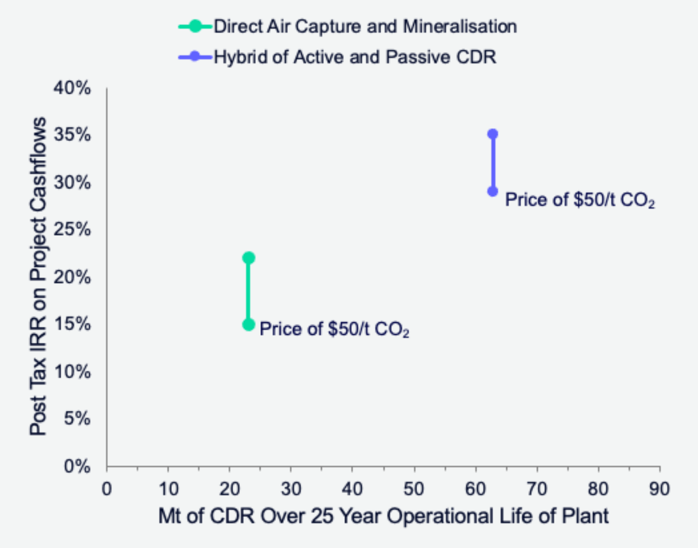


Figure 26: Project Returns at a Carbon Price of $50/tCO2

The financial model was used to assess the financial viability of a 50kt plant project, calculating income, cashflow and balance sheet of a project plus the carbon impact. The analysis resulted in a calculation of the Internal Rate of Return for the project. Figure 26 showing a comparison of likely IRR for straight DACMin vs a Hybridised system. Both offer attractive IRRs for future investors which will attract commercial investment needed to roll out the technology at scale and create a significant CDR impact.

This was used as the basis of an investor pitch which CCC used to approach the investor community for Seed investment to take the technology forward.

6.2 Investor community feedback.

Since the beginning of the project we have been out talking to potential investors about investment in our business based on our DACMin technology.

Their feedback was that while the business model is certainly a step in the right direction, notably in de-risking the capital and financing required, they felt there are too many dependencies:

- CDR market – they recognize the large potential, but CDR adoption has been slow to date, and enhanced weathering and ocean alkalinity CDR technology , while potentially promising, technology readiness is still in the early stages with many regulatory barriers still to overcome.
- Magnesium hydroxide & silica markets risk – particularly in relating to the challenges associated with commodity market offtake pricing and competition.
- Timeline risk – despite reducing pilot plant build out to 1 year, the time to subsequent demonstrations is still a potential challenge to reach the milestones needed for an attractive exit within a 10-year fund life.

Other issues raised were that the high capital cost necessary to get to scale would result in considerable dilution for early investors. Also, the winners had already been picked in the DAC space and were first in line for further support through the IRA in the US. Any new player on the field would be in competition with these companies. (Carbon Engineering sold to Oxy for $1.1bn USD and has a further $1bn in IRA support).

6.3 Pivot to Ocean Alkalinity Enhancement

What we learnt from the DACMin project was that the most impactful way to achieve both Gigaton levels of carbon dioxide removal and to maximise revenues (thereby reducing the cost of CDR), was to leverage the vast contact surface area between sea and the atmosphere and dispense with the air contactors and high energy requirements needed to shift large quantities of air through scrubbers or filters, not to mention the capital cost of the huge machines needed to achieve this. We therefore decided to pivot the company technology. The conclusions of our pilot project indicate that using our magnesium hydroxide for Passive DAC (e.g., Enhanced Weathering or Ocean Alkalinity Enhancement), rather than engineered DAC, doubled the amount of CO2 we could extract from the atmosphere. Our models of our process showed that Active DAC contributes less than 10% to revenue and less than 50% of the total CDR capacity. This led us to ask ourselves, could we simply remove the Active DAC functionality of our system, take a small hit on revenues and create a simpler, lower capital solution focusing on Passive CDR and revenues from Co-Products? This solution could then work alongside, rather than compete with more mature DAC technologies. Our new focus is presented on our updated website on www.co2loc.com. It also allows us to revert to a previous version of our technology using a sodium hydroxide digestion which is lower temperature and has a much higher chance of success than the previous high temperature sodium carbonate route.

6.4 What the next stage in the development of the GGR would be, including its scale and likely location(s).

To pursue this new opportunity, the company is engaging with the Carbon to Sea Initiative, which channels philanthropic funding from organizations such as the Chan Zuckerberg Initiative and the Grantham Foundation. While there has not yet been direct engagement with these foundations, future plans include exploring direct relationships. Additionally, the company is seeking support through advanced purchase agreements for Ocean Alkalinity Enhancement carbon removal credits with Frontier, (see https://frontierclimate.com/apply). These credits would be verified by Isometric using the protocol they developed, which has already been successfully applied in credits procured by Shopify. To further accelerate deployment, Cambridge Carbon Capture is actively identifying partners to develop OAE projects capable of scaling carbon dioxide removal. The company is also in negotiations with the Carbon to Sea Initiative to secure a £60,000 contract to fund a study focused on the sustainability and economics of its OAE technology. If successful, this contract is expected to extend the company’s cash runway to five months, potentially creating further funding opportunities.

Once a OAE project has been identified CCC would be able to secure prepurchase agreements worth up to $500k. This and matching equity investment would enable CCC to build lab based demonstrator for the new low temperature process and produce quantities of materials to support the project. A successful project and investment round will lead to the FEED study of a larger scale project and securing of offtake agreements through Frontier of between $10M and $50M. This will attract further funding to build out the project.

In parallel with the project development activity, CCC will engage with the wider OAE community to help address regulatory barriers and secure licences and permits required to enable the projects.

Initial projects are being explored, the least problematic project location from the regulatory standpoint is an existing industrial outflow such as wastewater or industrial process plant.

Here, licences already exist and much of the infrastructure and monitoring systems are already in place. CCC have been approaching the water industry to find potential partners for a project. There has already been a president set in St Ives with South West Water working with Planetary Technologies. (https://www.planetarytech.com/updated-company-overview-planetary-in-cornwall/).

6.5 How this development(s) would be informed by information gained during Phase 2.

The main learning from the DACMin project has been the fine balance between any future DAC project being net carbon negative or net carbon positive. Underperformance of any part of the DAC process could easily tip the balance. Also, the sheer scale of the air contactors needed and the amount of air that would need to be processed through any DAC device means that a DAC solution is going to be very capital and land intensive, limiting its location to sites where land is cheap and/or un-productive. Our conclusion was that contact with air is best achieved through the use of land or sea surfaces through enhanced weathering (EW) and/or ocean alkalinity enhancement (OAE). Both these approaches offering additional benefits in either improving crop yields in the case of EW, or reducing ocean acidification in the case of OAE. CCC's preference is for OAE as the mechanisms

6.6 Dependencies, uncertainties - describe what your plan depends upon and any assumptions made.

The field of ocean alkalinity is still in a very early stage of development, the main barriers being public perception and the regulatory barriers currently in place. This is an area of great uncertainty but CCC are working with key players in this area and hope to benefit from their work whilst they benefit from our work in pursuing a commercial opportunity in this area. Our assumption is that regulatory frameworks supporting OAE will become established over time as our project work yields results and any perceived harms are countered by scientific evidence gained through our early projects. As a contingency, work is ongoing to explore options around micro dosing of alkalinity in order to keep localised concentrations of alkalinity within the natural pH variations within the ocean.

7 A route to market

The key steps to commercialisation are:

1. Concept Validation and Feasibility Study
    a. Research and Development (R&D): Conduct comprehensive scientific research to prove the concept's viability, including laboratory and pilot-scale testing.
    b. Feasibility Assessment: Analyse technical feasibility, potential carbon sequestration rates, and environmental impact.
2. Regulatory Compliance and Permits
    a. Understand Regulations: Identify and comply with local, national, and international regulations related to ocean interventions, including environmental protection laws.
    b. Engagement with Authorities: Work with regulatory bodies to obtain necessary permits and approvals for field testing and larger-scale deployment.

3. Environmental Impact Assessment
    a. Baseline Studies: Conduct studies to understand existing marine conditions.
    b. Impact Analysis: Assess potential ecological impacts on marine biodiversity and chemistry to ensure minimal disruption and compliance with environmental standards.
    c. Risk Management Plan: Develop a comprehensive strategy to mitigate identified risks.
4. Pilot Projects and Demonstration
    a. Small-Scale Trials: Implement controlled pilot projects to test the technology in real-world conditions and refine the process based on observations.
    b. Data Collection: Gather data on carbon capture efficacy, safety, and scalability from pilot tests.
5. Partnerships and Collaboration
    a. Academic and Industry Partnerships: Collaborate with research institutions, NGOs, and industry experts to enhance credibility and leverage complementary expertise.
    b. Stakeholder Engagement: Engage with local communities, environmental groups, and governmental organizations to build trust and support.
6. Funding and Investment
    a. Secure Funding: Attract investments from venture capitalists, green funds, and public grants tailored for climate technology initiatives.
    b. Public-Private Partnerships: Explore joint ventures with established companies in the marine and environmental sectors for financial and operational support.
7. Technology Optimization and Scalability
    a. Efficiency Improvements: Refine processes to enhance energy efficiency, reduce costs, and improve carbon capture rates.
    b. Scalability Strategy: Develop a clear plan for scaling up from pilot projects to commercial-scale operations, including logistics and supply chain considerations.
8. Intellectual Property and Patents
    a. Protect Innovation: File for patents and secure intellectual property rights to safeguard proprietary methods and technologies.
    b. Licensing: Explore licensing agreements for wider distribution and application.
9. Commercial Strategy and Market Analysis
    a. Market Research: Identify target markets and customers, such as industries needing carbon offsets or compliance with emissions reduction targets.
    b. Business Model Development: Choose a suitable business model, whether it be direct service provision, licensing, or carbon credit sales.
    c. Pricing Strategy: Develop competitive pricing for carbon credits or related services.
10. Monitoring, Reporting, and Verification (MRV)
    a. Transparent Reporting: Establish robust systems for MRV to ensure transparent and accurate tracking of carbon sequestration results.
    b. Third-Party Validation: Work with accredited third-party organizations for validation and certification to build credibility and trust.
11. Market Entry and Commercial Launch
    a. Initial Launch: Deploy initial commercial operations in targeted regions with supportive regulatory environments.

    b. Marketing and Outreach: Raise awareness through targeted campaigns, partnerships, and education about the benefits of OAE.
12. Feedback Loop and Continuous Improvement
    a. Iterative Development: Collect feedback, monitor performance, and make continuous improvements to the technology and business approach.
    b. Sustainability Checks: Regularly assess environmental impacts to adapt to new regulations and ensure long-term viability.

Addressing these steps thoroughly will help navigate the complexities of commercialising OAE technology while balancing environmental responsibility, scalability, and market competitiveness.

The main barriers will be public acceptance of OAE and the biggest risk will be that OAE is rejected as being too damaging to the marine environment without scientific basis. Much of this can be countered through transparent engagement with the public and framing of the technology as a ‘cure’ for ocean acidification reversing the damage already being done by increasing levels of CO2 in the atmosphere and oceans.

7.1.1 potential benefits for other sectors

Aside from the obvious CO2 drawdown potential of the technology, economic growth would be evident not only in job creation but in the boost provided to industries such as marine and maritime operations. The need for specialized ships and equipment for deploying OAE technology would stimulate growth in shipbuilding and marine engineering. Industries involved in the mining and processing of olivine and the offtake of the silica and metal co-products of the CCC process, would see increased demand, adding a further layer of industrial benefit. The renewable energy sector, too, could integrate OAE with offshore installations like wind farms, enhancing revenue streams and fostering collaborative, hybrid projects.

The benefits of OAE technology would extend beyond direct economic impacts, spilling over into other sectors. For instance, the climate finance and insurance industries would find new opportunities to create financial products such as green bonds and specialized insurance for OAE projects. Agriculture and aquaculture might see indirect benefits from improved ocean health and increased biodiversity, leading to healthier fish stocks and more productive aquaculture operations. Coastal tourism, reliant on vibrant marine ecosystems, could also thrive as OAE contributes to reduced ocean acidification and the preservation of vital habitats like coral reefs.

Moreover, the technological and innovative potential of OAE technology would catalyse advancements in various fields. Collaborative efforts between environmental science, technology, and engineering could inspire breakthroughs applicable to water treatment, CO2 capture, and sustainable chemical processes. The development of sophisticated monitoring and data analysis systems for OAE could fuel growth in AI, sensor technology, and data services related to oceanographic research.

On a broader scale, OAE technology aligns with global climate goals, supporting a transition to a greener economy and reinforcing the clean-tech leadership of participating regions. The growing industry would also spark demand for regulatory experts and compliance officers to navigate emerging policies, creating specialized jobs in environmental law and policy.

In essence, commercializing OAE technology offers promising potential not only for direct job creation but also for stimulating economic growth across multiple sectors. It contributes to a more resilient, sustainable future by fostering technological innovation, supporting related industries, and aiding environmental conservation efforts.

8 An assessment of the potential rollout for the technology once successfully commercialised.

The key unique selling point of CCC's OAE technology is that there are revenue streams from co-products above and beyond the sale of carbon credits. The majority of DAC technologies are reliant on the future market for carbon credits and higher values for credits generated by carbon removal. International support for carbon credits to date has been patchy and the current market has been driven by philanthropy from only a small number of corporations and family offices. Although this is enabling early projects and building demand and investor confidence, there is an enormous risk that this will not translate into an internationally accepted marketplace, especially at the high carbon pricing needed to support engineered DAC.

Financial modelling of CCC's OAE projects show that acceptable internal rate of returns, greater than 10%, are generated, even at current average carbon market prices of around $50 per tonne. Once the economics of the process have been demonstrated at a commercially significant scale, this will unlock significant investment from existing ESG focused funds, leading to rapid scale up and subsequent impact of the technology.