



Phase 1 Final Report

Project Dreamcatcher (Project Pal202022)

Revision Record

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1.0 Contents

2.0 EXECUTIVE SUMMARY	6
3.0 UNDERLYING TECHNOLOGY	8
3.1 OVERVIEW OF THE CARBON ENGINEERING DIRECT AIR CAPTURE TECHNOLOGY	8
3.2 PROCESS CHEMISTRY AND THERMODYNAMICS	9
3.3 TECHNOLOGY READINESS LEVEL	12
3.4 ENVIRONMENTAL IMPACT	13
4.0 CHANGES DEVELOPED IN THIS PROJECT	15
4.1 PROJECT OBJECTIVES	15
4.2 SELECTION OF ALTERNATIVE FUEL.....	15
4.3 DEVELOPMENT OF KEY DESIGN PARAMETERS.....	15
5.0 FEED ENGINEERING SUMMARY.....	17
5.1 OVERVIEW	17
5.2 DESIGN PHILOSOPHY.....	18
5.3 PROCESS DESCRIPTION	18
5.4 MECHANICAL DESIGN & LAYOUT.....	22
5.4.1 <i>Container Unloading and Loading Concept</i>	22
5.4.2 <i>Silo Filling Concept</i>	24
5.4.3 <i>Conveying Philosophy</i>	25
5.5 CIVIL ENGINEERING	32
5.6 C&I OVERVIEW	35
5.6.1 <i>Control and Instrumentation Summary</i>	35
5.6.2 <i>Operating Philosophy</i>	36
5.6.3 <i>Monitoring and Control Criteria</i>	37
5.7 ELECTRICAL OVERVIEW.....	38
5.7.1 <i>Design Approach and Facilities</i>	38
5.7.2 <i>Power Supply and distribution</i>	38
5.7.3 <i>Equipment Design</i>	39
5.7.4 <i>Single Line Diagram</i>	40
5.8 MATERIALS ENGINEERING	41
5.9 TECHNICAL SAFETY	42

5.9.1	<i>HSE Hazard Risk Management</i>	42
5.9.2	<i>Hazard Identification</i>	43
5.9.3	<i>Process Safety and Loss Prevention Design Principles</i>	43
5.10	ENERGY AND FUEL REQUIREMENTS	46
5.10.1	<i>Hydrogen</i>	46
5.10.2	<i>Start-up Fuel Gas</i>	46
5.10.3	<i>Cooling Water</i>	47
5.11	ENVIRONMENTAL IMPACT	47
5.11.1	<i>Emissions</i>	47
5.11.2	<i>Effluent discharge</i>	51
5.11.3	<i>Waste</i>	51
5.11.4	<i>Use and Storage of Chemicals</i>	51
5.12	UK SUPPLY CHAIN	52
6.0	PILOT PLANT LOCATION	53
6.1	PERMITS AND CONSENTS	54
7.0	EXECUTION PLAN	55
7.1	DELIVERY TEAM STRUCTURE & RESPONSIBILITIES	55
7.1.1	<i>Steering Board</i>	55
7.1.2	<i>Project Direction</i>	55
7.1.3	<i>Project Management Office</i>	55
7.1.4	<i>Project Organisation</i>	56
7.2	SCHEDULE	56
7.3	PROJECT MANAGEMENT	58
7.3.1	<i>Petrofac Project Management Framework</i>	58
7.4	PROCUREMENTS AND SUBCONTRACTS	58
7.4.1	<i>Procedures and Processes</i>	60
7.4.2	<i>Mechanical Packages</i>	60
7.4.3	<i>Subcontract strategy</i>	60
7.4.4	<i>UK Construction Partner</i>	60
7.5	PROJECT CONTROLS AND SERVICES	60
7.5.1	<i>Risk & Opportunity Management</i>	61
7.6	INTERFACE MANAGEMENT	61
7.7	CONSTRUCTION AND INSTALLATION	61
7.7.1	<i>Procedures and Processes</i>	64

7.7.2	<i>Constructability</i>	64
7.7.3	<i>Construction Strategy and Construction Execution Plan</i>	65
7.8	COMMISSIONING	66
7.8.1	<i>Commissioning Processes and Procedures</i>	66
7.9	OPERATION AND MAINTENANCE	67
8.0	PILOT PLANT LOCATION	68
8.1	PERMITS AND CONSENTS	69
9.0	COSTED PROJECT PLAN	70
9.1	COST ESTIMATE FOR PILOT PLANT	70
9.1.1	<i>Approach to Estimate</i>	70
9.1.2	<i>CAPEX</i>	72
9.1.3	<i>OPEX</i>	74
9.1.4	<i>Decommissioning</i>	76
9.1.5	<i>Assumptions</i>	77
9.1.6	<i>Overall Cost Summary</i>	78
9.2	COST ESTIMATE FOR COMMERCIAL SCALE DEPLOYMENT	79
9.2.1	<i>Approach to Estimate</i>	79
9.2.2	<i>CAPEX Estimating Methodology and Results</i>	80
9.2.3	<i>OPEX</i>	83
9.2.4	<i>Decommissioning</i>	84
9.2.5	<i>Overall Cost Estimate</i>	85
10.0	FURTHER DEVELOPMENT OF THE TECHNOLOGY	86
10.1	STOREGGA DIRECT AND CARBON REMOVAL CREDITS.....	86
10.2	THE MARKET FOR CONVENTIONAL DIRECT AIR CAPTURE	86
10.3	THE FIRST STOREGGA DIRECT DAC PLANT.....	88
10.4	FUTURE GROWTH.....	89
10.4.1	<i>Storegga Direct: overall strategic ambitions to 2030:</i>	90
10.4.2	<i>Storegga Direct: overall strategic ambitions beyond 2030:</i>	91
10.5	MARKET FOR ALTERNATIVE FUEL DIRECT AIR CAPTURE	91
10.6	TIMELINE	92
10.7	POTENTIAL CHALLENGES	93
10.7.1	<i>Technology scale-up</i>	93
10.7.2	<i>Steps to Commercialisation</i>	93

10.7.3	<i>Business Model Development</i>	94
10.7.4	<i>Additional revenue streams</i>	94
10.7.5	<i>Societal acceptance</i>	94
10.7.6	<i>Consents and licences</i>	95
11.0	ATTACHMENT 1	96

2.0 Executive Summary

The regenerative DAC technology for Project Dreamcatcher has been developed by Carbon Engineering (CE) over the last 10 years. It removes CO₂ from the atmosphere through two integrated chemical reaction loops in 4 major units (see Section 3.1). The current state of development of the technology is that Front End Engineering Design (FEED) is being undertaken on a 1 million tonnes per annum CO₂ capture plant in the US, and pre-FEED is being undertaken on a 0.5 million tonnes per annum CO₂ capture plant in the UK. We anticipate these full-scale industrial plants to be operational by 2024-2026.

The Calciner technology that releases the captured CO₂ currently uses an oxygen-fired Circulating Fluidised Bed (CFB) with natural gas injected directly into the fluidised bed for combustion. The CO₂ created from the natural gas combustion is captured.

This Project sought to eliminate this hydrocarbon from the process. Two options were identified: -

1. Electric heating of the solid calcium carbonate.
2. Using hydrogen as an energy vector in place of natural gas.

Both approaches were found to be technically viable, and the hydrogen option was developed in this project as the overall project costs were significantly lower.

Computational fluid dynamics work undertaken by the University of Cambridge determined that the existing calciner design was suitable for use with hydrogen. This finding was unexpected, as the hydrogen density is significantly lower than natural gas, and hence the volume of hydrogen required to deliver sufficiently energy for the calcination is significantly greater. Modelling found that this increased volume of hydrogen was offset by a reduction in the volume of produced CO₂ as the combustion of hydrogen produces water rather than CO₂. The volumes offset one another, and the calciner was found to operate as effectively with hydrogen as with natural gas.

The cost of building a full chain DAC system in the UK was found to be prohibitive, under the rules of the competition. A FEED study has therefore been undertaken on a trial of the hydrogen-fuelled calciner only, importing CO₂ laden pellets from the Carbon Engineering Innovation Centre in Canada, and processing them at a test facility in the UK.

This study has found that the cost of the trial will be ~£14.3 million. This is significantly greater than the GGR Competition Phase 2 budget which limits Phase 2 projects to £5 million.

We have discussed several options to reduce cost with BEIS

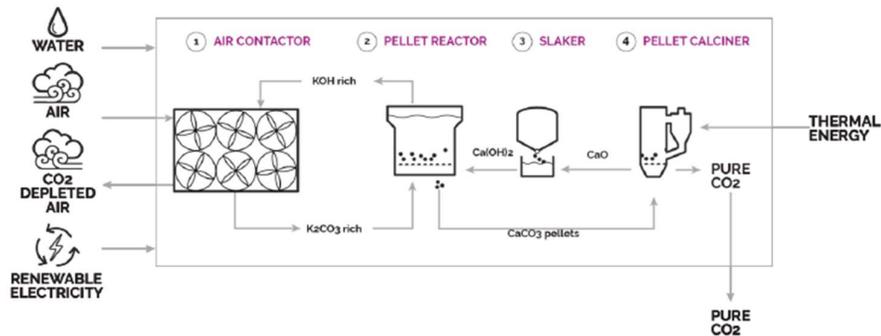
- Options to test the hydrogen fuel concept with existing calciners in Canada and USA are not permissible under the GGR Competition Guidance as >75% costs would be incurred outside UK
- Options to test the hydrogen fuel concept with Storegga Direct calciners following commissioning of the UK's first DAC plant in 2026 / 2027. This approach is also not permissible under the GGR Competition Guidance Costs as the timing is outside the Phase 2 window which ends March 2025.

We have therefore concluded that we will not submit an application for a Phase 2 grant. Instead, Carbon Engineering will progress this technology as part of its low-carbon research programme.

3.0 Underlying Technology

3.1 Overview of the Carbon Engineering Direct Air Capture Technology

The regenerative DAC technology for Project Dreamcatcher has been developed by Carbon Engineering (CE) over the last 10 years. It removes CO₂ from the atmosphere through two integrated chemical reaction loops in 4 major units - as illustrated below:



The process operates as follows: -

- Step 1 – Air is drawn through the Air Contactor by fans where CO₂ in the air reacts with an aqueous solution of potassium hydroxide (KOH) to form water (H₂O) and an aqueous solution of potassium carbonate (K₂CO₃)
- Step 2 - The potassium carbonate rich solution is then fed to the Pellet Reactor where it reacts with calcium hydroxide (Ca (OH)₂) to form solid calcium carbonate (CaCO₃) and regenerate the aqueous KOH which is recycled back to the Air Contactor.
- Step 3 – This solid calcium carbonate is heated (using natural gas and pure oxygen) in a Calciner where CO₂ is released as a gas, and solid calcium oxide (CaO) is formed. The purified CO₂ can then be compressed and stored in a CCS project.
- Step 4 - The calcium oxide passes to a Lime Slaker where it is combined with water to regenerate the calcium hydroxide which is fed back into the Pellet Reactor.

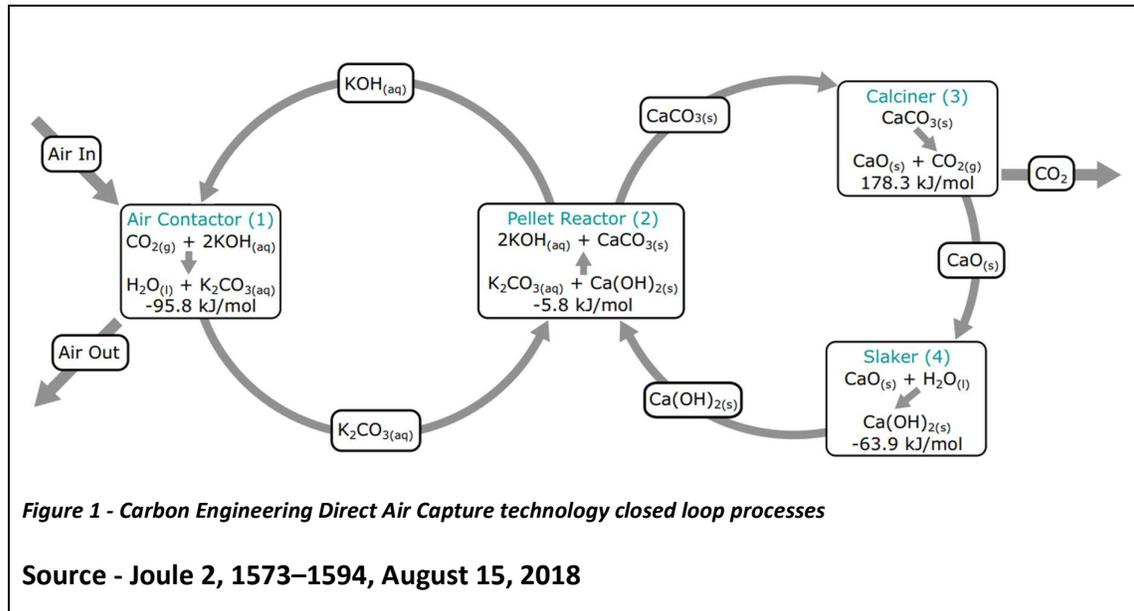
The technology is described in detail in an article entitled “A Process for Capturing CO₂ from the Atmosphere” by David W. Keith, Geoffrey Holmes, David St. Angelo and Kenton Heidel of Carbon Engineering in Joule, published 15 August 2018. This is included as Attachment 1.

CE has been operating this DAC process since 2015 at their pilot facility in Squamish, near Vancouver in British Columbia, Canada. CE has also built a facility which is a fully integrated Innovation Center located in Squamish. Construction is complete and start-up planned for Q4 2021. Carbon Engineering's (CE) business model is to license its technology to development partners to finance, build and operate facilities utilizing its technology. In the US, CE has signed a licensing agreement with 1PointFive, a development company formed by Oxy Low Carbon Ventures, LLC, a subsidiary of Occidental. 1PointFive and Carbon Engineering are engineering the first large-scale commercial facility to utilise CE's technology, with construction expected to begin in 2022 and operations targeted for 2024. Located in the Permian Basin, US, the facility is expected to capture one million tonnes of carbon dioxide from the atmosphere annually when complete, and is the first step toward the partners' aspiration to deliver this technology on an industrial scale throughout the United States..

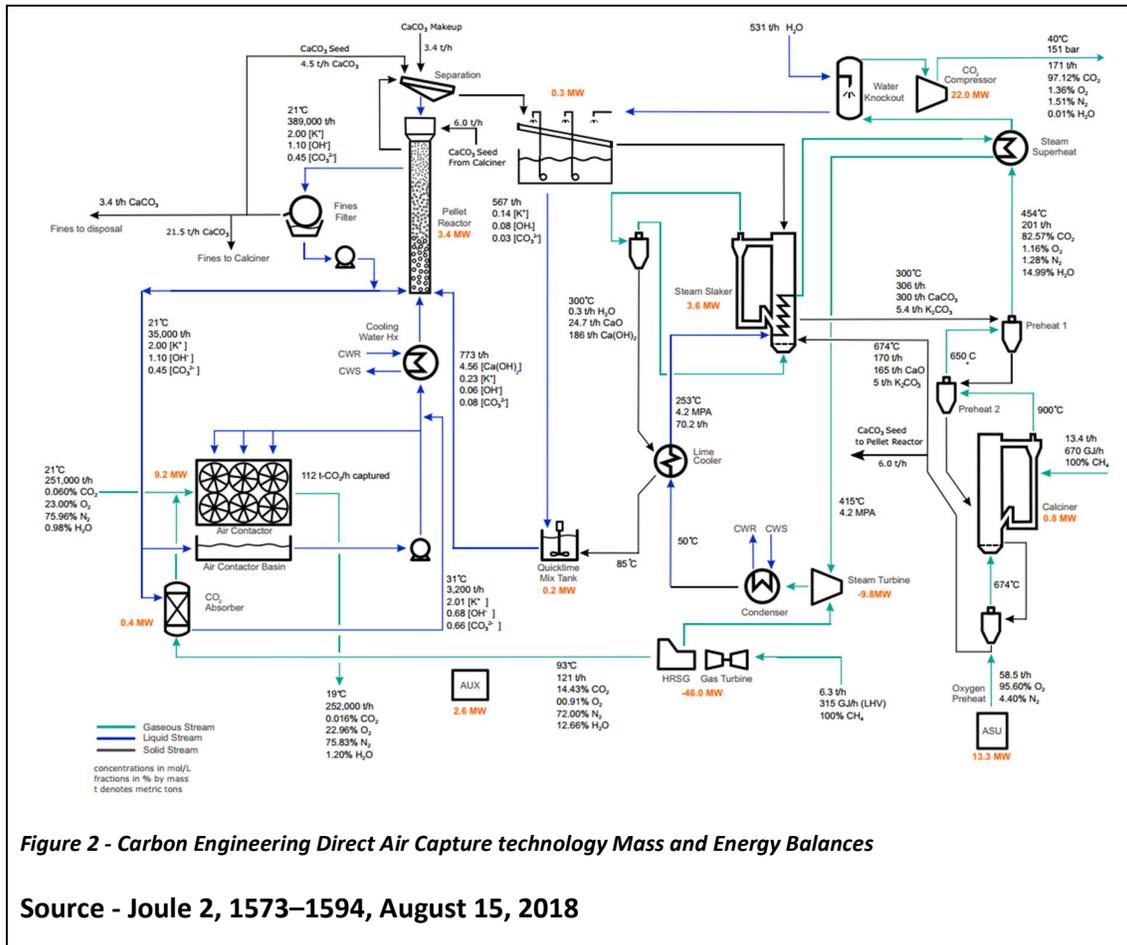
In the UK Carbon Engineering and Storegga have begun the engineering and design of a DAC facility that will permanently remove between 500,000 and one million tonnes of carbon dioxide from the atmosphere annually. Targeted for North-East Scotland, the proposed facility will be the first large-scale facility of its kind in Europe and the partners are aiming towards it being operational by 2026.

3.2 Process Chemistry and Thermodynamics

The Carbon Engineering Direct Air Capture technology involves two closed loop processes as illustrated in the following diagram.



A calcium loop (right) drives the removal of carbonate ion and thus the regeneration of the alkali capture fluid (left). Boxes with titles show the names of the four most important unit operations. Each box shows the chemical reaction with reaction enthalpy at STP in kilojoules per mole of carbon and the reaction number for reference elsewhere in the paper. Note that water is liberated in reaction 1 and consumed in reaction 4, balancing the process. The full process has evaporative losses, as shown in Figure 2



Electricity demands are indicated in orange as MW. Selected gas and liquid streams show the most important constituents using mass fraction as for gaseous streams and molar concentration for aqueous. Mixed phase streams with substantial solid-phase mass flow are color-coded based on the phase of the gas or liquid transporting the solid. Units are indicated with graphical representations that suggest a schematic physical design of the unit. Many minor streams, such as cooling water to the multistage CO₂ compressor, are not shown. As described in the text, there are several options for introducing the fines stream back into the calciner, these are omitted for simplicity, and this heat and mass balance reflects fines being treated identically to the pellet stream leaving the washer.

3.3 Technology Readiness Level

The underlying technology discussed in this document is the established Carbon Engineering DAC system, which has been operating at a pilot scale in Canada since 2015 (see Figure 3 – the Carbon Engineering Direct Air Capture Innovation Centre at Squamish, near Vancouver in British Columbia, Canada.) It uses natural gas as the source of energy in a major process step, and the CO₂ generated from this natural gas combustion is captured and stored along with the CO₂ extracted from the atmosphere.

At present we assess the technology as being at TRL 7 (“System prototype demonstration in operational environment”)



Figure 3 – the Carbon Engineering Direct Air Capture Innovation Centre at Squamish, near Vancouver in British Columbia, Canada.

3.4 Environmental impact

The following table summarises the environmental impacts, consumables and effluents for the Carbon Engineering DAC plant based on a 0.98 million tonnes per annum CO₂ capture plant

Component	Quantities
Inputs	
Air to be processed	251,000 tonnes per hour
Oxygen (manufactured on site)	Consumption of 58.2 tonnes per hour Requires 13.3 Mw electrical power
Water	Consumption of 531 tonnes per hour 4.7 tonnes water per tonne of CO ₂ captured = 4.7 million tonnes per annum
Natural Gas (Calciner)	Consumption of 13.4 tonnes per hour / 670 GJ/hour
Natural Gas (Local power generation)	Consumption of 6.3 tonnes per hour / 315 GJ/hour
Calcium Carbonate	Makeup of 3.4 tonnes per hour
Outputs	
Processed air	252,000 tonnes per hour
Calcium Carbonate	Production of 3.4 tonnes per hour
CO₂ for storage	Production of 171 tonnes per hour 97.12% CO ₂
Source - Joule 2, 1573–1594, August 15, 2018	

4.0 Changes developed in this project

4.1 Project objectives

The objective of this Project Dreamcatcher was to investigate the possibility of replacing natural gas in the Calciner in the Carbon Engineering DAC process with a clean energy solution (wind-generated electricity or low-carbon hydrogen).

The Calciner technology currently used by CE is an oxygen-fired circulating fluidised bed (CFB) with natural gas injected directly into the fluidised bed for combustion. This provides efficient heating of solid calcium carbonate because of direct convective heat transfer from the combustion gases as well as radiative heating from the flame. The CO₂ created from the natural gas combustion is captured, such that the amount of CO₂ to be stored is approx. 40% greater than the amount of CO₂ captured from the atmosphere. To eliminate this hydrocarbon from the process we identified two options: -

1. Efficient electric heating Existing electric heating technologies require an intermediate medium (metal surface, molten salt or liquid metals) which is heated through induction or resistance. Indirect heating is more complex than direct heating.
2. Using hydrogen as a fuel in place of natural gas. This innovation involves researching the combustion profile of injected hydrogen in a fluidised bed and the subsequent development of a modified calciner system.

4.2 Selection of alternative fuel

Deliverable 1 of this project assessed the potential to replace natural gas in the Calciner with either direct electrical heating or with hydrogen.

Both approaches were found to be technically feasible, but this project has determined that the use of hydrogen is the preferred solution on the grounds of cost.

4.3 Development of key design parameters

Deliverable 2 of this project modelled the use of hydrogen in a Calciner to determine whether any modifications would be required to achieve acceptable calcination.

The primary conclusion from the CFD simulations and 1D Phenomenological model is that H₂/O₂ combustion in a circulating fluidized bed calciner can stably support CaCO₃ calcination with comparable in-bed combustion, particle hydrodynamics, temperature, and reaction profiles to CH₄/O₂ combustion. The results indicate that switching from CH₄ to H₂ combustion will not bring significant changes to reactor design or operation and a hydrogen-oxygen combustion CFB calciner will be the most cost-effective solution to demonstrate.

5.0 FEED Engineering Summary

This study has found that the cost of the trial will be ~£14.3 million. This is significantly greater than the GGR Competition Phase 2 budget which limits Phase 2 projects to £5 million.

We have discussed several options to reduce cost with BEIS

- Options to test the hydrogen fuel concept with existing calciners in Canada and USA are not permissible under the GGR Competition Guidance as >75% costs would be incurred outside UK
- Options to test the hydrogen fuel concept with Storegga Direct calciners following commissioning of the UK's first DAC plant in 2026 / 2027. This approach is also not permissible under the GGR Competition Guidance Costs as the timing is outside the Phase 2 window which ends March 2025.

As a result the project participants will not proceed with a Phase 2 application. This conclusion was reached after the results of the FEED work were completed. As a result this section refers to the Phase 2 prototype as if it is intended to go ahead. This is no longer the case.

5.1 Overview

Deliverable 3 presented a techno-economic assessment for two calciner demonstration options for verifying the operation of a prototype calciner using hydrogen rather than natural gas for calcination:

- Integrated End-to-End DAC System
- Separated End-to-End DAC System with Calciner Demonstration in the UK and atmospheric-CO₂ laden pellets shipped from CE's Canadian innovation centre

Both options for calciner demonstration are capable of achieving an end-to-end DAC demonstration with a capacity of 1,000 tonnes CO₂ per year.

CE has previously demonstrated the end-to-end DAC process at the CE Pilot and soon to be Innovation Centre in Squamish British Columbia, Canada. A full chain integrated DAC system is not required for testing the prototype Hydrogen Fired Calciner.

A separated end-to-end Calciner option will be able to effectively demonstrate the hydrogen gas heating alternative in the most capital and land efficient way aligning with the objectives of this study

A FEED study has been completed for the pilot plant for the hydrogen fired Calciner. This section summarises the outcomes of the work completed. A listing of the discipline deliverables produced and the main areas to be developed in the detailed design and execution phase are provided in the separate FEED report.

5.2 Design Philosophy

CE advised that sufficient data to prove and optimise the design of the Hydrogen Calciner can be obtained with 20 days of unit operations. This is envisaged to be split over 3 campaigns with periods for data analysis between campaigns. Each campaign will be operated as a batch process for 4 to 10 days to optimise and test the Calciner, building on previous prototype data that CE acquired for natural gas fired calcination process. The Calciner will be pre-heated with natural gas prior to the data acquisition runs.

The pilot plant facilities shall be designed for a minimum 1-year life and the equipment items and/ or packages will be leased wherever possible.

5.3 Process Description

The block flow diagram for the pilot plant is presented in Figure 5.2.

Dry CaCO₃ pellets are transported to site in 1 tonne bags, stacked 2 wide and 2 high, in 20ft shipping containers. The bags are initially loaded via tele-handler-type forklift of the type shown in Figure 5-1, onto a Flexible Intermediate Bulk Container (FIBC) unloader X-001.



Figure 5-1: Telehandler Forklift

The FIBC incorporates an iris valve which is closed around the FIBC outlet tail prior to untying the tail. Once the tail is untied and the operator has confirmed that a rotary valve, Roots Blower and silo filter fan are all energised, the iris valve is opened manually, and the free-flowing pellets are transferred through the rotary valve into a lean phase pneumatic conveying line and delivered to Silo S-001. Conveying air is provided by the Roots blower. A filter is provided to remove dust from air vented from the silo, assisted by a fan, integral to the filter.

The FIBC discharge rotary valve and Roots Blower trip on high level to prevent it from overflowing when the CaCO_3 Silo is full.

The CaCO_3 pellets are fed into the Calciner package via a rotary valve at the outlet of the CaCO_3 Silo S-001 using a motor driven tubular cable conveyor. The CaCO_3 feed flowrate is controlled into the pellet feeder using a metering hopper S-003 and associated conveyor.

The dry Calcium Carbonate is heated and fluidised in a vertical calciner using CE's proprietary Calciner package. All the feed streams to the Calciner are manually controlled using a rotameter with a transmitter that trips the Calciner low flow.

The proprietary Calciner package contains Trade Secrets of Carbon Engineering with only data essential to perform a robust FEED being made available for the FEED study. In summary, the Calciner operates at atmospheric pressure and $\sim 900^\circ\text{C}$. Calcination of the CaCO_3 produces CO_2 in the hydrogen fired catalytic fluidised bed. Heat integration within the Calciner increases the energy efficiency using heat recovery cyclones. Cooling water to remove the heat load in the Calciner is provided by a Cooling Water Package PK-004. Conversion of the Calcium Carbonate to Calcium Oxide is $\sim 98\%$. The Calcium Oxide is transferred to the CaO Silo S-002 via Calciner discharge rotary valve RV-003 and CaO Conveyor CV-002. A filter, F-002, is provided to remove dust from air vented from the silo S-002.

Low feed flows or malfunction of the Calciner due to high temperature, or high or low levels will trip the Calciner closing ESD and XV valves within the package boundary. It will also stop the CO₂ Vent Blower BL-002 and the CaO discharge and trip the CaO conveyor.

CaO is discharged from the silo via manually started rotary valve RV-004, into FIBC's suspended beneath the rotary valve outlet chute in a loading frame designed to minimise dust egress at the loading point. The FIBC's are removed by forklift trucks when full.

Wet CO₂ from the Calciner is withdrawn into CO₂ Baghouse F-001, within the CO₂ Baghouse Package PK-001, for discharge to the atmosphere by the Vent Blower BL-002. The CO₂ baghouse F-003 utilises long cylindrical bags made of woven or felted fabric as a filter medium. Dust laden CO₂ enters the baghouse by suction and is directed into the baghouse compartment. The heavier CaCO₃ dust particles fall off as it enters the baghouse, while the lighter dust particles along with CO₂ gas is carried upward to the bags. The gas is drawn through the bags and dust accumulates on the filter media. The filter is cleaned periodically using timed reverse jet air pulses, while the baghouse is on-line. The dust accumulated on the bags is removed from the fabric surface and deposited in the hopper which is emptied periodically into FIBC's for subsequent transport back to Canada or, if deemed more appropriate, disposed of in the UK. The CO₂ Baghouse receives wet CO₂ at 175°C and removes the dust prior discharge into the atmosphere at a safe location.

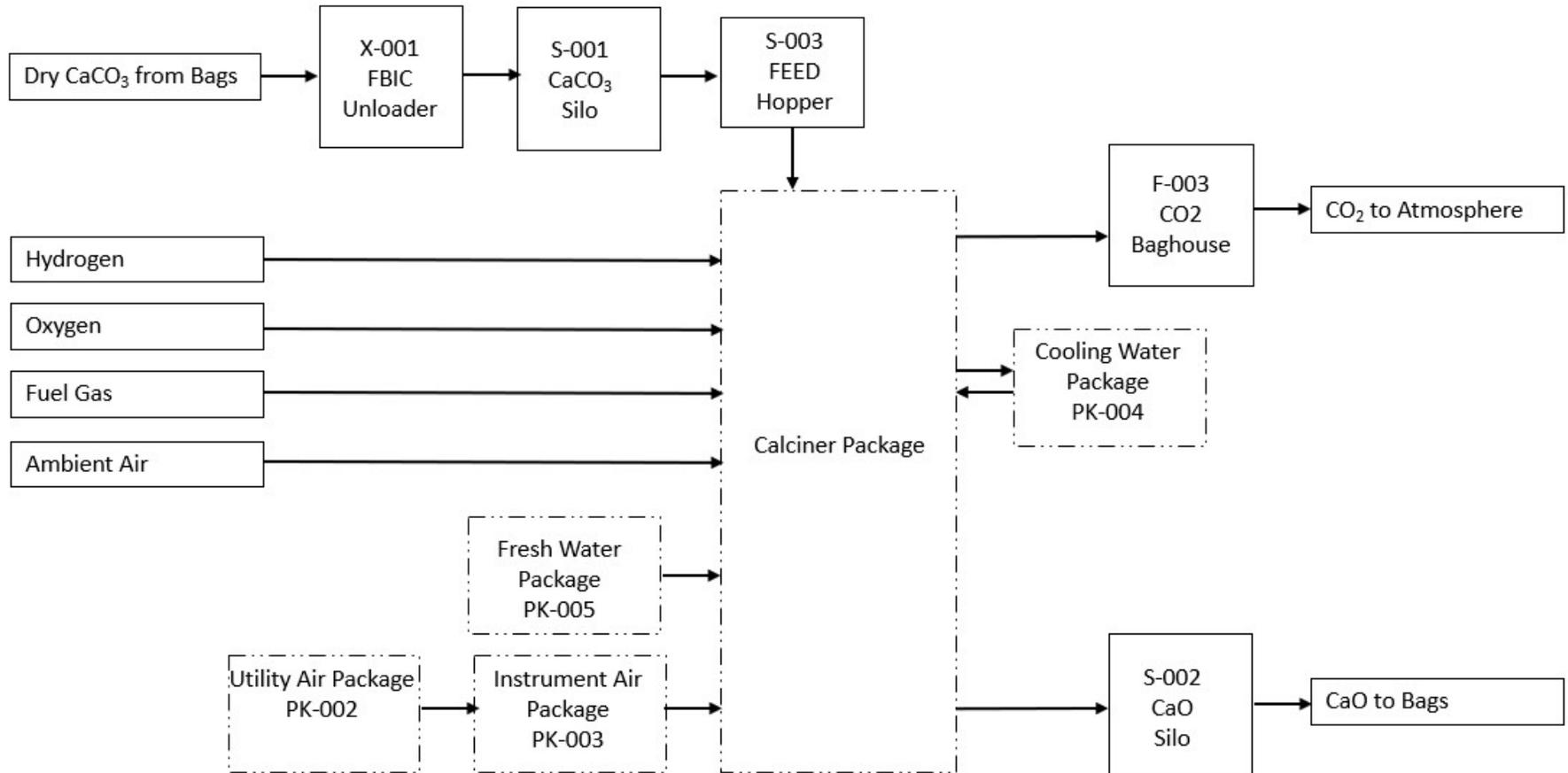


Figure 5-2: Block Flow Diagram for Pilot Scale Hydrogen Fired Calciner Plant

Phase 1 Final Report FEED Engineering Summary

Instrument air is provided from an Instrument Air Package PK-003 to condition compressed air from the Utility Air Package PK-002. An instrument air receiver within the package will also provide the air required for cleaning the three filters F-001, F-002 and F-003.

5.4 Mechanical Design & Layout

5.4.1 Container Unloading and Loading Concept

CE advised that Flexible Intermediate Bulk Containers (sometimes referred to as 'big bags') will be delivered to site stacked two high and two wide in shipping containers. It was considered that to unload containers safely it is preferable to avoid human entry into the containers and as such the FIBCs should be mounted on pallets in the container (two bags high on a pallet, appropriately strapped) and that each pallet should be unloaded using a fork-lift. To reach all pallets in the container a tele-handler would have to be employed, and the container length limited to a 20' standard container size.

Based on this, the physical properties of the materials and the operational period it was calculated that 8 containers, each holding 16 bags of CaCO_3 would be required.

For the loading of the CaO , it is again assumed that there is space on site for 20 days continuous running and that it will be removed from site as a single consignment. It was calculated that 7 containers, each holding 16 bags of CaO would be required.

The internal layout of the containers is presented in Figure 5.3.

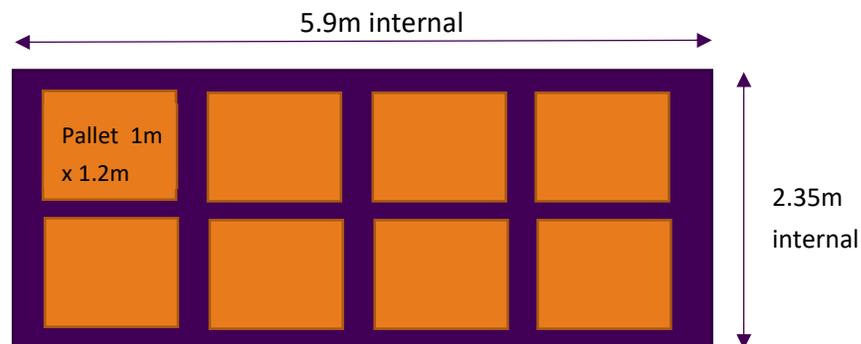


Figure 5-3: Envisaged Layout of FIBCs in Standard 20' Shipping Container

Figure 5-4 shows an extract from the layout drawing 200259-LYT-L-0001 showing the layout of the storage containers on site. It also shows:

- Space for 18 off 20' containers, allowing for 8 + 7 + margin.
- Turning space for lorries and manoeuvring space for tele-handler type fork-lift have been allowed.

Phase 1 Final Report FEED Engineering Summary

In the detailed design phase the actual number of FIBCs per container will be finalised with consideration for safety during unloading/loading, and the final site spatial requirements determined and incorporating the details of the final selected site.

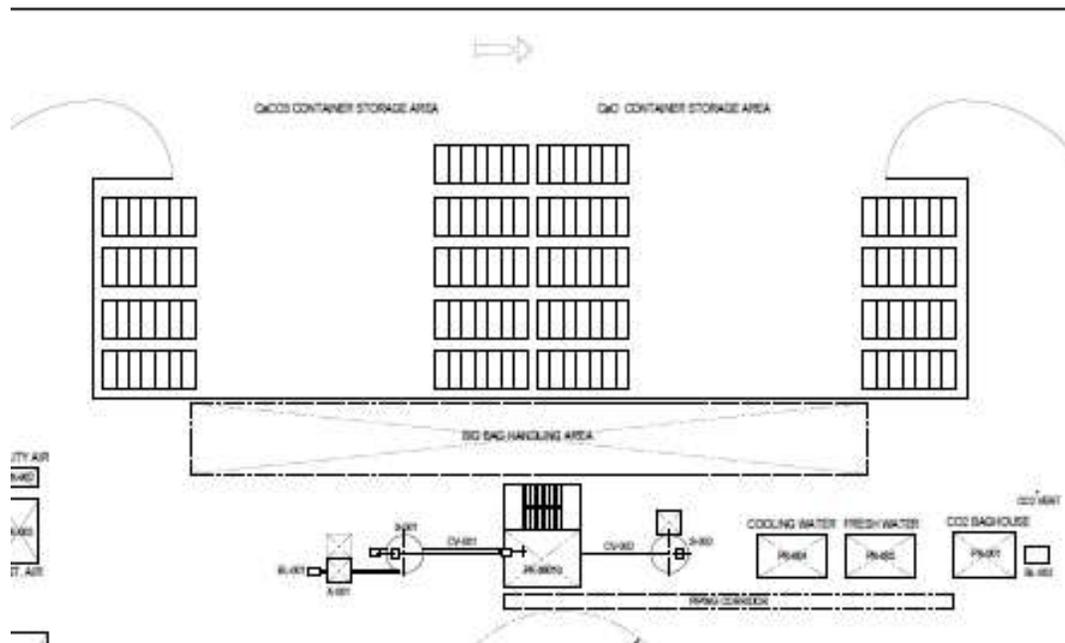


Figure 5-4: Extract from Drawing 200259-LYT-L-0001 Showing Container Storage area

5.4.1.1 Silo Sizing

Early in the design it was determined that the volumetric requirements of the two silos were similar, so it would be convenient if both the silos were dimensionally the same (reduction in design time). The similar handling properties of the materials pre- and post-Calciner are also similar which supports this design premise.

The silos have been sized based on 3 days continuous operation. Longer operation periods can be supported through concurrent filling and emptying operations.

S-001

Required Volume of $\text{CaCO}_3 = 18 \text{ m}^3$

The silo proposed is 2.4m diameter x 5.5m roof to outlet height, with a 60° conical base and a force-conveyed tangential inlet, which provides a volumetric capacity of 18.5 m^3 .

S-002

Required Volume of $\text{CaO} = 15 \text{ m}^3$

Phase 1 Final Report FEED Engineering Summary

The same size silo is proposed i.e. 2.4m diameter x 5.5m roof to outlet with a 60° conical base and a gravity-fed top centre inlet, which provides volumetric capacity 23% greater than that required.

For CaO, an angle of repose of 55 degrees was assumed, based on data from the National Lime Association website <https://www.lime.org/>. This will be confirmed as appropriate for the Calciner produced pellets from Squamish during detailed design.

5.4.2 Silo Filling Concept

UK Health and Safety regulations, and particularly the Construction (Design and Management) Regulations 2015, require us to consider in our design the health and safety of those constructing, operating, maintaining, cleaning, decommissioning and dismantling any structure within Great Britain. This plant in its entirety counts as a structure.

Loading bags directly into the silo would require an open inlet, potentially allowing unwanted moisture ingress into the silo (and possible foreign matter which could damage downstream equipment), and fugitive dust egress from the silo. It would also require personnel to be operating continually at high level, requiring a full staircase to the silo top, proper level platforming and handrails, a lifting davit with a hoist and potential local exhaust ventilation. High winds and rain at the potential test sites in the Northwest of England would make for poor ergonomic conditions and as such it was considered that such an arrangement would not be acceptable in the UK.

To satisfy the requirements for the UK, FIBC unloader will be provided, with a removable top frame from which FIBCs are suspended. The bag-lifting frame is lifted by a fork-lift to the top of the FIBC unloader and mounted there with the FIBC suspended over the discharge pan, and then lowered so the container (bag) sits on the pan. The outlet spout (tail) of the container (bag) is passed manually through the centre of the pan through an iris valve to an untying chamber, and the iris valve is then closed around the tail. Once the iris valve is closed the tail can be untied for CaCO₃ pellets to flow into the downstream equipment.

Variations to the above theme occur from supplier to supplier. Figure 5.5 shows Flexion's FIBC Unloader Frame. Note the options for FIBC (bag) massagers (Flexicon calls these 'flow flexers') in the unloading pan, amongst other options offered.

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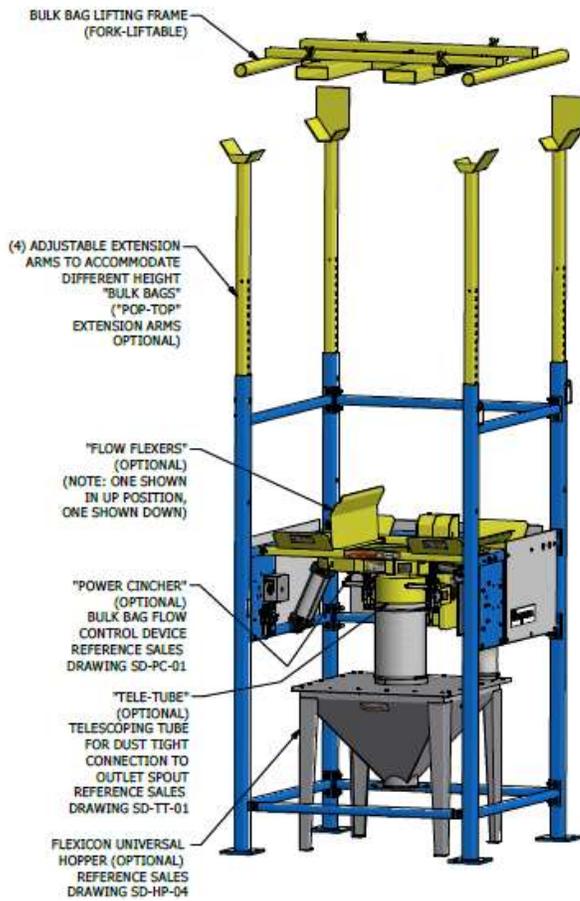


Figure 5-5: Typical FIBC (Flexicon)

Various methods were considered for transfer of the CaCO_3 from the FIBC unloader to the silo and these are explained in the Section 5.4.3.

5.4.3 Conveying Philosophy

There are three principal areas requiring consideration: transfer from the FIBC unloader to silo S-001, transfer from the silo to the Calciner and transfer from the Calciner to silo S-002. All three have different considerations.

5.4.3.1 Transfer from FIBC Unloader to Silo S-001

The main considerations in the design are as follows:

Phase 1 Final Report FEED Engineering Summary

- Unloading time for a FIBC – ideally unload a bag in around 15 minutes – assuming a FIBC holds a tonne of material this sets the unloading rate at 4 tonnes per hour.
- Fully enclosed method of conveying – eliminates belt conveyors
- The small particle size requires close tolerances on moving parts, eliminating enclosed chain conveyors, side-wall conveyors and bucket elevators.
- Remaining options include screw conveyors, pneumatic conveying and cable/chain and tube conveyors. Due to the desire to make both silos the same size it was deemed appropriate to pneumatically convey the material into the silo via a tangential inlet, which reduces the material profile height at the top of the silo. Additionally, lean phase pneumatic conveying is common for dry materials of relatively homogenous particle sizing, so this was chosen. A rotary valve is required to be incorporated between the untying chamber of the FIBC unloader and the inlet to the pneumatic conveying pipeline. The motive air for the pneumatic pipeline would be provided by a Roots blower, also commonly used for this application. The silo would have to incorporate a vent filter together with a forced-vent fan to remove the conveying air.

Phase 1 Final Report FEED Engineering Summary

5.4.3.2 Transfer from Silo S-001 to the Calciner

The vertical distance from the base of the silo to the top of the Calciner is over 16m, which, for the small flow (less than 1 tonne per hour) is not appropriate for a screw conveyor whose length may be limited by torque and bending considerations. Conveying may be achieved by blowing pneumatically into a receiver mounted on top of the Calciner and from there dosing the pellets via a small screw conveyor into the Calciner inlet hopper. However, this would introduce another filter into the train (which requires maintenance at a high level) and a fan to remove the air conveyed, and the filter hopper would have to be set on weigh cells to be able to measure (via loss in weight) material into the Calciner. Whilst a small hopper and dosing screw is required for any method of feeding the Calciner, the conveying may be more simply achieved using a tube and cable conveyor or tube and chain conveyor, consisting of a conveying tube in which the motive force is provided by discs of the inner diameter of the tube, joined by cable or chain which is driven by a sprocket at the head end of the conveyor. This can convey material and drop it into a small (100 litre) hopper from which it is dosed into the Calciner inlet using a small variable speed screw (~100mm diameter) which can be calibrated to feed the correct quantity. High- and low-level switches in the 100-litre hopper give a signal to the control system to start or stop pellet feed to the hopper.

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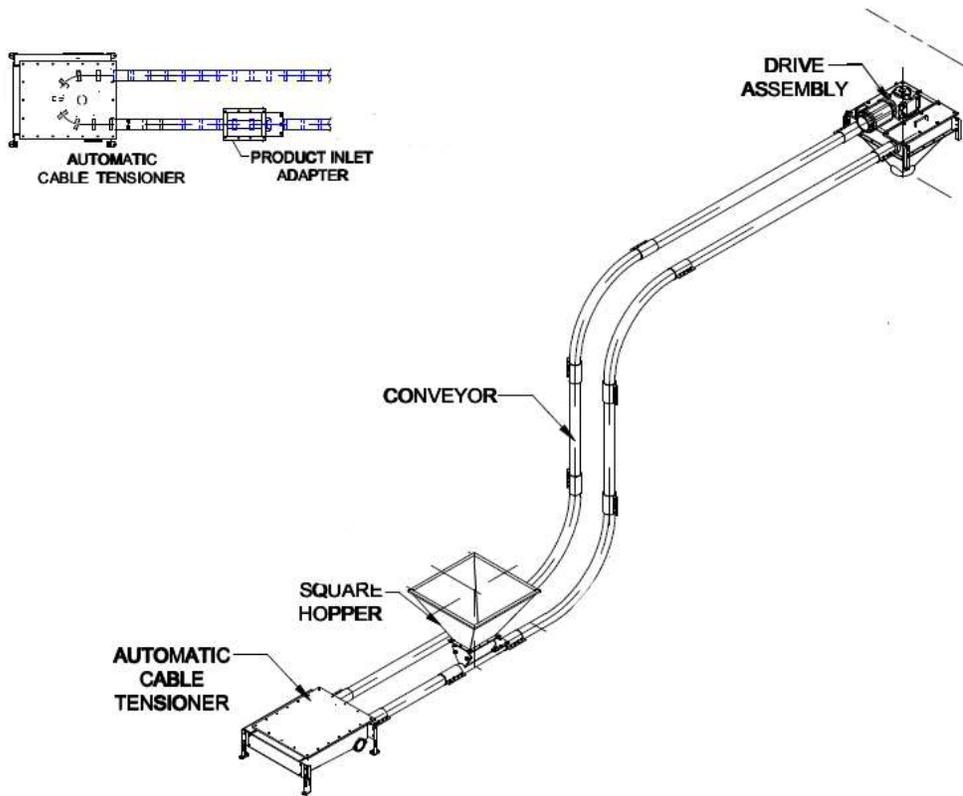


Figure 5-6: Typical Tube and Cable Conveyor (Flexicon)

A further advantage of the tube and cable conveyor over pneumatic conveying is that blowers are typically noisy and require noise attenuation, especially if they are to be used continuously. The FIBC unloading will occur in infrequent 15 minute bursts, whereas the Calciner feed would be more likely to be continuous.

Phase 1 Final Report FEED Engineering Summary

5.4.3.3 Transfer from Calciner to Silo S-002

The high temperature and potential need for further cooling of the product negate the use of pneumatic conveying. Tube and cable conveying generally uses polymer parts which also may be susceptible to the high temperature. In the absence of further information on the Calciner layout, the outlet from the Calciner was assumed to be some 4m above grade, allowing a screw conveyor to be used to convey the 7 to 8m to the top of the CaO silo S-002. The screw conveyor may also be water cooled if necessary, using an external water jacket. CE advised that based on their experience on their Canadian pilot unit cooling was not considered necessary and has thus not been included in the cooling water calculations or costing at this stage. However, storing micron-sized pellets at 165°C in a silo could present health and safety issues, such as loading hot material into FIBCs. This issue will be reviewed in the detailed design phase of the project and the need for any requirements for cooling confirmed. Although the distance of conveying is short enough for a screw conveyor, the assumed conveying angle of 45 degrees and the relatively low flow (considered a 'trickle feed') may require a specially designed screw conveyor. Furthermore, it is possible that a 'tube and chain' conveyor (like the tube and cable conveyor but chain instead of cable) may be able to be adapted for the hot material and cooled on its horizontal run as well. These aspects will be investigated further during the next stage of the project. They are not considered material to the feasibility, cost or schedule of the design.

The layout resulting from the mechanical considerations is shown in Figure 5-7: Elevation and Plan on Receipt, Conveying, Storage and Calciner extracted from drawing 200259-LYT-L-0001 which has been extracted from the preliminary site model showing the layout of the main process equipment.

Phase 1 Final Report FEED Engineering Summary

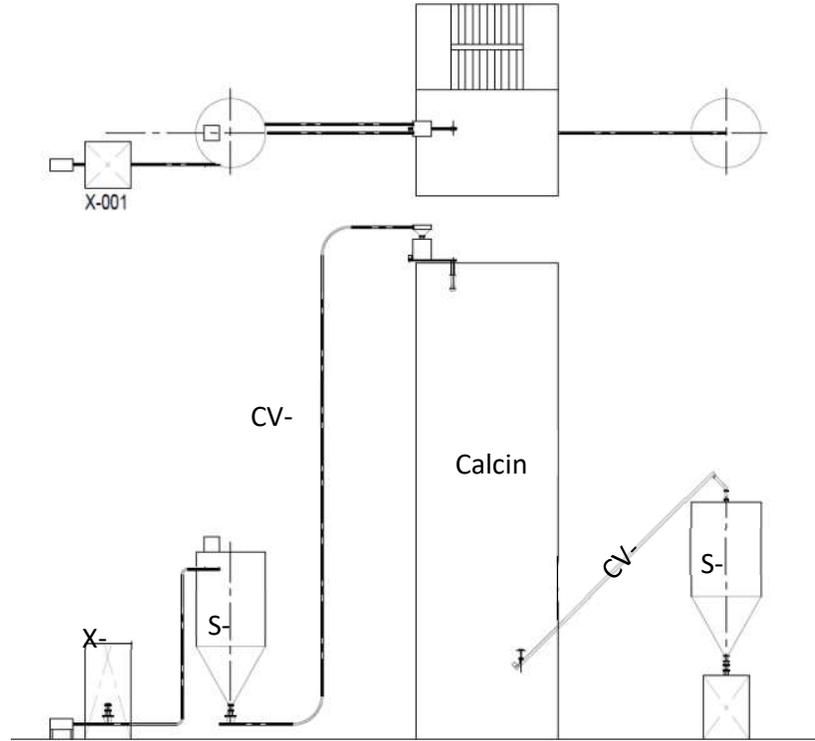


Figure 5-7: Elevation and Plan on Receipt, Conveying, Storage and Calciner

Phase 1 Final Report FEED Engineering Summary

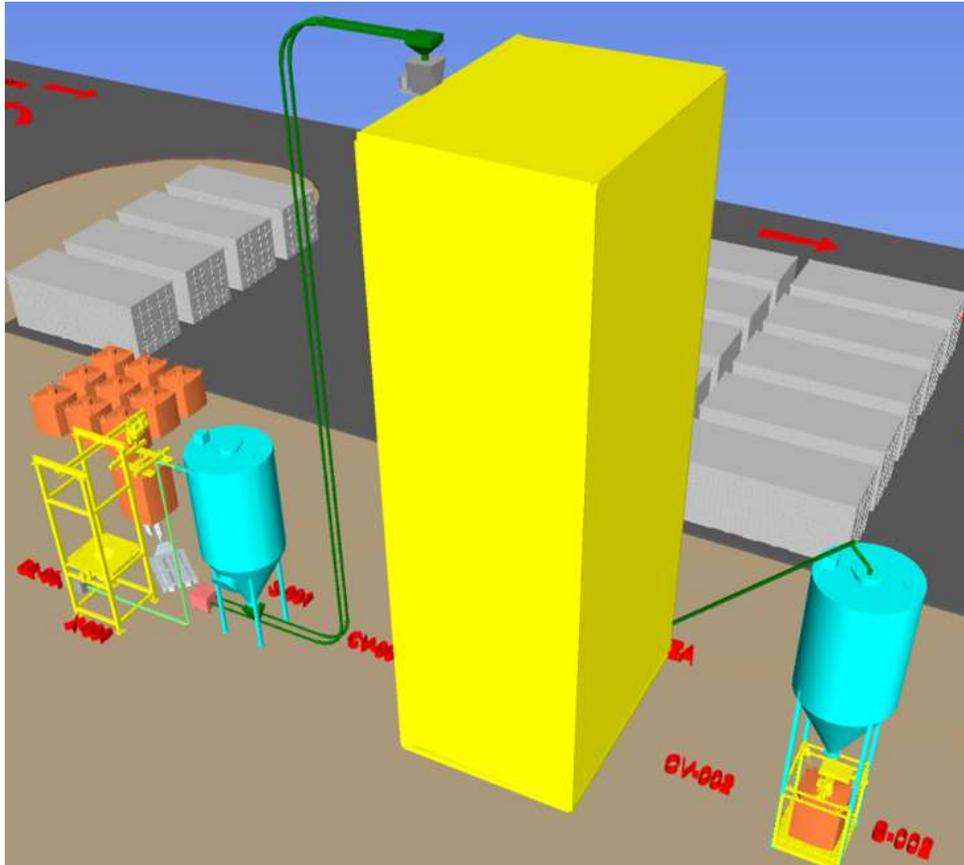


Figure 5-8: Model of Conveying, Storage and Calciner

Other considerations on the layout included the following:

- Between the hydrogen delivery manifold and any source of ignition there needs to be a distance of 8m
- The road layout incorporated a one-way system around the plant
- Space will be needed around the container store for a crane to lift the containers off and onto the vehicles
- A piping corridor was incorporated for piping and ducting between the utility skids (cooling water and fresh water) and CO₂ Baghouse. These items of equipment, and the main plant also, were all assumed to be one side of the piping corridor. Relative positions of the cooling water skid, freshwater skid and CO₂ baghouse would be finalised during detailed design.

Recognising these points an overall site lay out was developed. The layout will be refined further in the next phase of the project once the site has been finalised and any additional constraints known. Figure 5-9 and Figure 5-10 show the preliminary plant layouts.

Phase 1 Final Report FEED Engineering Summary

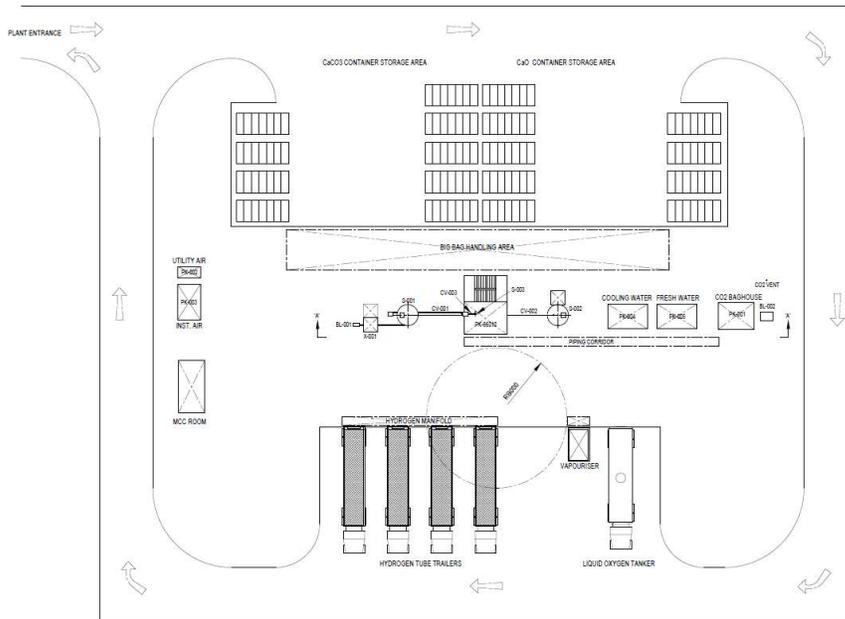


Figure 5-9: Preliminary Overall Site Layout

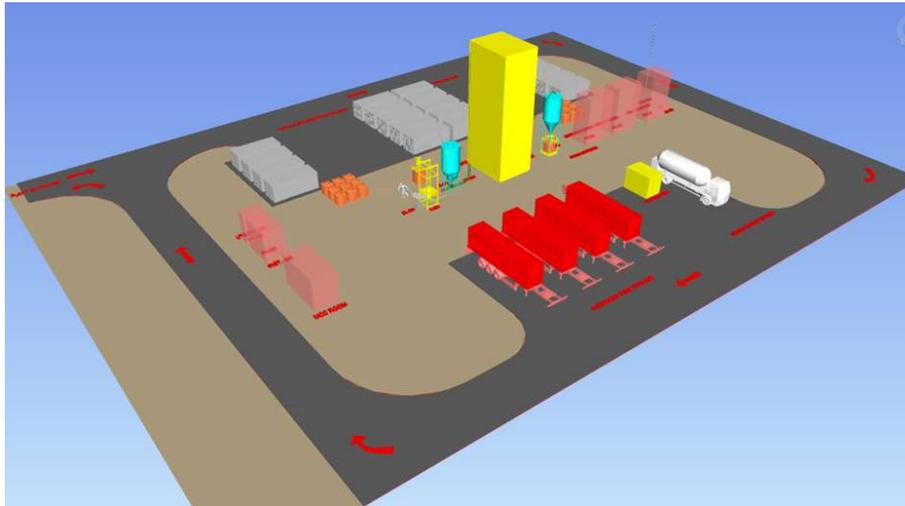


Figure 5-10: Model of Preliminary Overall Site Layout

5.5 Civil Engineering

The site selection activities ran in parallel with the FEED engineering and for the purposes of engineering studies the site has been assumed to be located on a brown field site such as Spadeadam in Cumbria or the HSE Science and Research Centre at Buxton.

Phase 1 Final Report FEED Engineering Summary

Whilst it is acknowledged that the process design may not be sensitive to UK site location, the civil design is directly related to the site-specific conditions. Consequently, for the development of the civil FEED engineering a number of assumptions have been made and parameters selected to reflect a standard representative site as closely as possible.

It has been assumed that the facilities will be located within an existing flat and level paved area able to accommodate regular operational vehicle movements without requiring re-surfacing or paving. Demolition or removal of existing buildings, structures or foundations extending above ground is assumed not to be required. Localised breaking out of paved areas to enable the construction of foundations has been allowed. It is also assumed that facilities can be supported off shallow (non-piled) foundations comprising either strip footings or pad foundations.

Site foundations are required for the major equipment items specifically: CaCO₃ Silo, CaO Silo Calciner Package and piping & cabling T posts. A diagram showing the foundations is presented in Figure 5-11.

Phase 1 Final Report FEED Engineering Summary

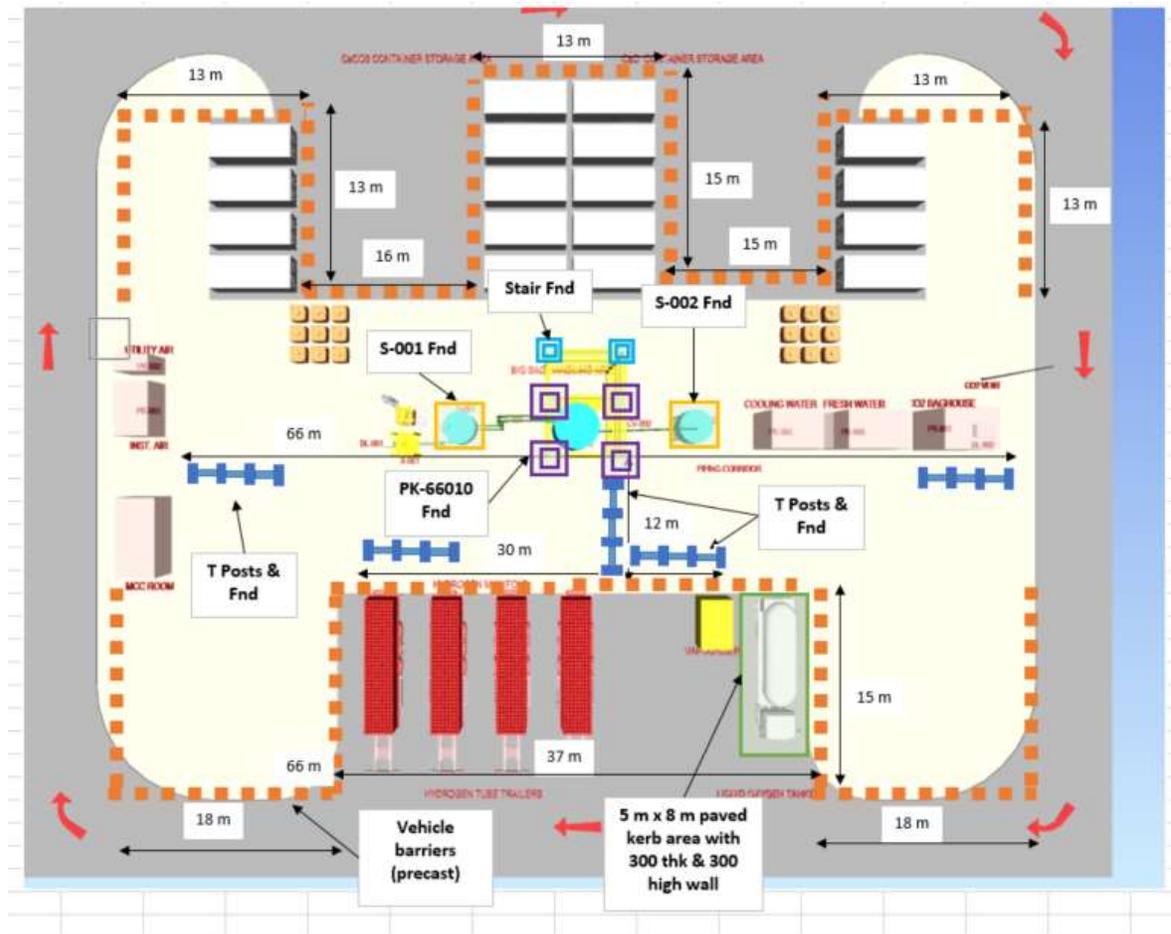


Figure 5-11: Diagram showing Civil Foundations

The Facility is assumed to be within an existing paved compound laid to falls with existing surface water drainage suitable for container storage, forklift and HGV movements with no additional surface water drainage requirements required.

All interconnecting small-bore piping and cabling are assumed to be mounted on above ground tray / racking supported at minimum 2.2m above grade on steelwork T-posts (spaced at 3 m c/c) and longitudinal interconnecting beams.

It is assumed that the site will be within a fenced secured area with suitable access for HGV deliveries. No provision or allowance has been made for the provision of new roadways or general surfacing. A new concrete paved LOX tanker unloading area has been allowed for due to the risk of potential LOX spillages from tanker hook ups and deliveries.

Phase 1 Final Report FEED Engineering Summary

Allowance has been made for pre-cast concrete vehicle barriers located around the plot to delineate vehicle access routes and tanker / container loading parking areas.

No allowance has been made for trenching for buried utility services such as electrical supply / lighting cables, potable water supply or fuel gas piping. Similarly, no allowance for erection or foundations for area lighting poles or columns has been made.

A nominal allowance shall be made for the provision of vehicle directional and warning signage and road markings.

The civils requirements shall be revisited in the detailed engineering phase following the selection of the site and once the associated details are available.

5.6 C&I Overview

A control and instrumentation design specification was developed for the project reflecting current best practice. The principals and philosophy contained within it were applied to the process design. The design approach and operating philosophy are summarised below.

5.6.1 Control and Instrumentation Summary

The Calciner will be continuously manned from a Control Room (CR), from where the operator will monitor all operations, initiate control actions, and manage start-up and shutdown of the facility, units, and packages.

Local Instrument and Electrical Equipment Rooms (LERs) will house Control, Safety and Package system and marshalling cabinets based on control and cable optimisation. Under normal operating conditions the operator will monitor and control the whole plant from Integrated Control and Safety System (ICSS) and associated Operator Workstations (OWSs) within the CR.

The ICSS will be designed to keep the plant within safe operating limits. It should automatically correct disturbances caused by process conditions. The ICSS will be capable of controlling the plant during start-up, normal shutdown, and emergency shutdown. Manual controls will generally be limited to special cases such as:

- During maintenance/repair of field equipment
- Start-up of packaged equipment
- Infrequent and simple operations

Phase 1 Final Report FEED Engineering Summary

Major process equipment or package start-up activities will normally be from the CR with field support dictated by the requirements of the mechanical equipment as required. Package auxiliaries (lube and seal oil, cooling water systems) will normally be started locally and automatically as part of the automatic start sequence from package Unit Control Panel (UCP) however these will also have the capability for remote start from the CR.

For packaged equipment which have vendor supplied dedicated control and safety systems, the ICSS will function mainly as a monitor, using network data links to collect, display, and archive/trend operating data. Master control set points and sequenced starts will however be controlled by the ICSS.

It is preferred that major process package units and equipment uses the same manufacturer and platform as the ICSS for their dedicated control and safety systems to allow seamless integration and share OWSs, this would provide a single operating system software window and tools for all main plant areas.

Distributed remote Process Control System (PCS), Safety Instrumented System (SIS), and Fire and Gas System (FGS) Inputs/ Outputs (I/O) with Universal I/O cards where available will be implemented to reduce field instrument cable runs and costly installation. The main ICSS equipment will be in the Control Building housing the CR and main Instrumentation Equipment Room, with remote I/O located in appropriately determined and economically assessed, suitability rated remote outdoor Junction Box (JB) Panels.

The basis will look for the Control Building with CR and LER to be modular build design, with the building and rooms built offsite and fitted out with the ICSS equipment, smoke and fire detection systems, fire extinguishing systems and HVAC, all wired and tested before being transported to site.

5.6.2 Operating Philosophy

Continuous on-site monitoring of Dreamcatcher Pilot Scale Hydrogen Fired Calciner Plant will be required. The equipment will be configured to allow control either via local control panels, or remotely from a CR.

Minimum manning of the Plant will be considered. Operators will be in the CR on a continuous 24/7 basis during periods of plant operation. Requirements to monitor the process units locally will be kept to a minimum. The plant units and layout support the maximum use of a distributed ICSS for process control and plant safety and display of information for the operators in the CR.

Phase 1 Final Report FEED Engineering Summary

The ICSS will maintain production of on-specification product, safely, efficiently, and with minimum impact on the environment, nearby community, and countryside. Monitoring and control of process units will be performed from the CR. Control and shutdown should be automatic and operator intervention will be minimised where practical, with most functions being provided from the CR ICSS OWSs.

5.6.3 Monitoring and Control Criteria

The ICSS will be designed and configured to deliver a stable, efficient, safe plant operation. It will be capable of operating between the minimum and the maximum design conditions. The general control requirements are:

- Control facilities will be designed such that manipulation of all controlled variables is performed via OWSs located in the CR. Field or local panels may be required for package units, however these will be avoided where possible with all master control from the CR
- Operator manipulations during normal operation will generally be adjusting set points, changing control modes (auto, manual, cascade etc), activating remote commands (open/close, run/stop, on/off), performing an emergency shutdown, resetting shutdown events, acknowledging alarms, testing instruments, etc.
- The 'Fail Safe' concept will be applied, i.e. the de-energised state of any actuator device or controlled device will result in its safest overall action
- Trend functions, indicators, and alarms will be provided in the CR to enable the detection of abnormal operation. While operating the plant, the operator will be informed about any incoming safety information such as fire and gas detection
- The control system will be designed to minimise the effect of operational variation disturbances in one section of the plant affecting other sections of the plant
- Packaged units will interface seamlessly with the ICSS and preferably use the same make, model, and technology platform, to allow full plant control, operation, and shutdown from common OWSs in the CR

Phase 1 Final Report FEED Engineering Summary

5.7 Electrical Overview

An electrical design specification was developed for the project reflecting current best practice and UK regulatory standards. The design approach and preliminary system design is summarised below.

5.7.1 Design Approach and Facilities

Recognising the short operating life requirement, electrical equipment will be designed to minimise the cost. The selection of electrical equipment will be governed by fitness for purpose, safety in operation and suitability for environment. The intent will be to use supplier standard equipment suitable for the environment.

The design will comply with all UK regulatory standards and equipment, services and installations applied to the project will meet all applicable regulations relating to health, safety, and environmental issues.

5.7.2 Power Supply and distribution

5.7.2.1 General

The electrical system and equipment within it will be designed to ensure:

- Safety to all personnel
- Continuous and reliable service
- Convenience of operation
- Suitability for the environmental conditions

5.7.2.2 Main Power Supply

It is estimated that the power requirement for the pilot plant is 175 kVA, 245 Amps (including 20% design margin). It is assumed that the host location has adequate spare capacity to feed this power and that it will be derived from a feeder off an existing 415V switchgear. Some loads in the pilot plant are required to operate under emergency condition. It is assumed that the 415V switchgear which provides power to the pilot plant is connected to the existing facilities emergency supply.

The electrical design will be revisited in detailed engineering following selection of the site and, if necessary, a temporary generator utilised for the plant.

Phase 1 Final Report FEED Engineering Summary

5.7.2.3 Power Distribution

A new 415V distribution board will be provided in the pilot plant to feed the power to the loads. The 415V switchgear will have a single bus.

UPS power is assumed to be supplied from the existing site distribution system.

5.7.2.4 Control and Protection.

Direct-on-line starting will be used for motors. The starter will consist of an adequately rated fuse, thermal overload relay and contactor unit. The thermal overload relay will be of manual reset type.

Each motor will be provided with a local Emergency Stop, 'stay-put', push-button station that will be hard wired to the motor contactor trip circuit.

5.7.2.5 Metering

Metering Instruments will be provided on the main incomer of the 415V Switchgear to maintain records of power consumption and allow supervision of all necessary parameters, including but not limited to, current, voltage, power, etc. All instruments will be flush mounted.

5.7.3 Equipment Design

5.7.3.1 Low Voltage Switchgear

415V switchgear will comprise metal enclosed, free standing, vertical enclosures housing insulated copper phase bus bars, earth bus bar and of fixed pattern construction. An incomer will be provided with Switch fuse unit.

Switchboard sections will be combined in a single line-up with a common fully rated bus-bar system. Motor starters and distribution feeder units will be fixed type construction.

Switchboards will be configured with single bus bar section. Motor circuits will be controlled by fused protected contactors. Motor starters will be of the single speed, non-reversing, full voltage 'direct-on-line type'.

5.7.3.2 Lighting Design

All luminaires will be located such that maintenance and lamp changing can be safely carried out to the maximum practical extent without the use of ladders or other equipment.

Phase 1 Final Report FEED Engineering Summary

Lighting circuit design minimise the use of junction boxes, wherever possible, circuit connections will be made by looping through luminaires. Lighting luminaires should be installed in such a way that in the event of one circuit failure, adequate illumination is provided by luminaires powered by other phase circuit.

Generally, lighting in open areas will utilise LED type floodlights and, where practical, these will be mounted on adjacent plant structures, beams, columns or supports. LED type floodlight luminaires suitable for pole mounting will be considered if required. Outdoor lighting should be automatically controlled by light sensitive (photo electric) switches, aimed in a north direction, and supervised with hand-off-auto switch located at controller location to allow for manual control of the lighting.

Escape lighting fixtures will be provided along defined escape routes to lead personnel out of the building or plant area in case of black-out. Escape lighting will be supplied with integral batteries with an autonomy time of 90 minutes providing no delay to illuminate upon loss of normal lighting.

The area flood lighting luminaires as far as possible will be mounted on the adjacent plant structures, beams, columns or supports.

5.7.3.3 Equipment Installation and Layout

LV distribution boards and control panels will be installed in a factory constructed, insulated, heated, and ventilated substation building.

Electrical equipment and components located in prefabricated buildings, or on packaged machinery, will be installed, connected, and tested to the maximum practical extent prior to transportation to site.

Suitably rated BS standard rubber mats will be provided in front of panels for their entire length.

5.7.4 Single Line Diagram

A Single Line Diagram has been developed for the demonstration unit and is presented in Figure 5-12

Phase 1 Final Report FEED Engineering Summary

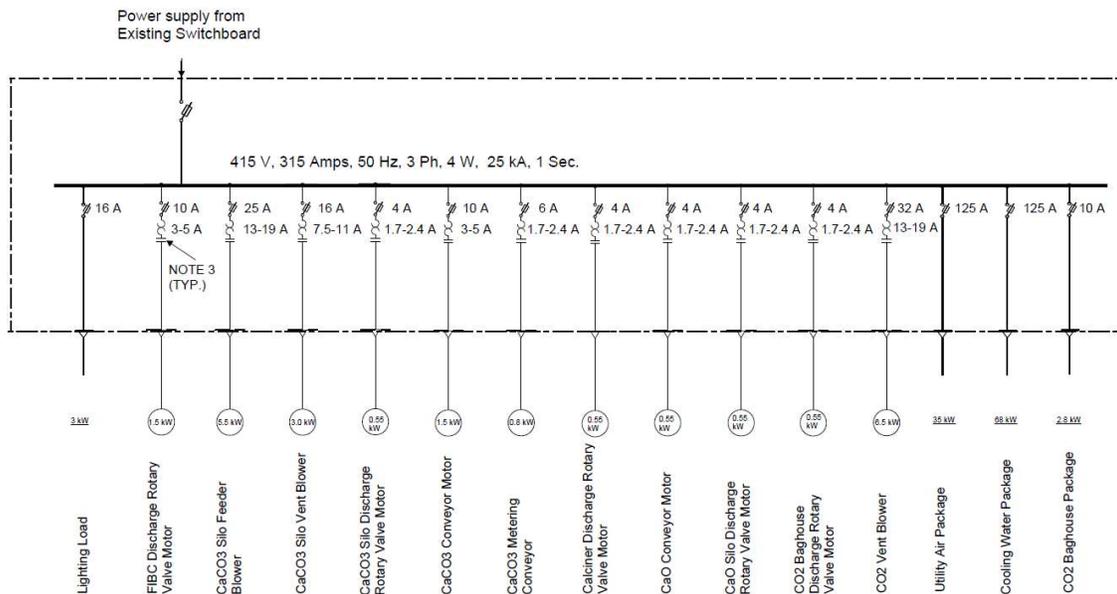


Figure 5-12: Single Line Diagram for Pilot Demonstration Plant

5.8 Materials Engineering

A materials selection study was conducted for the pilot plant. It specifically considered both the 1-year design life and the expected intermittent use over 2-3 months within that year. As stagnant conditions are likely for periods, higher corrosion rates than predicted may be experienced. To mitigate this a higher corrosion allowance has been selected and purge valves have been added to allow for purging using an inert media, such as nitrogen.

Predicted corrosion rates are low with the highest being for the wet CO₂ gas exiting the Calciner at 0.1 mm/yr. As stagnant conditions in this stream could increase the corrosion rate further, a selection of Carbon Steel + 3 mm corrosion allowance has been made. A purge valve has been added to allow for purging with an inert gas, such as nitrogen, to prevent stagnation if the system will be unused for an extended time.

Hydrogen lines have been selected in accordance with EIGA 121/14 and ASME B31.12.

Selection for the plant can be in Carbon Steel with a corrosion allowance of 1.5mm for all lines except for wet CO₂ which is recommended to be Carbon Steel with a corrosion allowance of 3 mm and oxygen which is selected in 316L Stainless Steel in accordance with EIGA 13/20. Equipment and vessels material selection has been made on the same basis.

Phase 1 Final Report FEED Engineering Summary

5.9 Technical Safety

A Safety and Loss Prevention Philosophy was developed that outlines the process safety, loss prevention and environmental philosophy to prevent, or reduce to a minimum, consequences to life, health, the environment, and damage to the project facility. This will be achieved by reducing the magnitude of hazardous events or reducing the probability of the consequences. The key points are summarised below.

5.9.1 HSE Hazard Risk Management

HSE Hazard Risk Management includes the identification, assessment and minimisation of the hazards and their likelihood. The main objectives of this philosophy are to reduce the risk to personnel and the environment to a level that is ALARP (As Low As Reasonably Practicable).

Once hazards have been identified, they are to be managed according to the following hierarchy in decreasing order of preference:

- Elimination and minimisation of hazards by using options with a lower impact on HSE
- Substitution by using products/ processes with a lower impact on HSE
- Isolation/ separation of hazards and targets
- Engineering controls – prevention
- Engineering controls – mitigation
- Organisational controls e.g. competence and communication
- Procedural controls
- Personal Protective Equipment (PPE)

Hazardous properties of process and utility stream handled in the facility are tabulated in Table 5-1

Table 5-1: Hazardous Properties of Materials

Process Stream	Hazardous properties
Calcium Carbonate	H315 Causes skin irritation. H320 Causes eye damage. H335 May cause respiratory irritation
Calcium Oxide	H315 Causes skin irritation. H318 Causes serious eye damage. H335 May cause respiratory irritation

Phase 1 Final Report FEED Engineering Summary

Hydrogen	H220: Extremely flammable gas. H280: Contains gas under pressure; may explode if heated.
Liquid Oxygen	H270: May cause or intensify fire; oxidiser. H281: Contains refrigerated gas; may cause cryogenic burns or injury.

5.9.2 Hazard Identification

5.9.2.1 HAZOP Study

A Hazard and Operability study (HAZOP) was carried out to identify any potential hazards and operability issues for the pilot plant facility. The review was conducted by a multi-disciplinary team comprising personnel from Petrofac and CE. For each design element, the team considered possible deviations from the design intent to determine whether appropriate means of protection have been provided.

The principal process changes were reflected in the Issued for Design (IFD) issue of the Process & Instrumentation Diagrams (P&IDs). All actions will be reviewed in detail, addressed and closed out as part of the detailed design phase.

5.9.3 Process Safety and Loss Prevention Design Principles

5.9.3.1 Plant Layout

The layout of the facility is a primary means of preventing incidents and preventing initiating incidents from escalating. The design considers adequate provision for operations and maintenance and minimise risk to site personnel and property.

The prevailing wind directions will be considered when establishing spatial arrangements and orientations of facilities once the site location has been selected.

Hydrogen and oxygen gas cylinders banks are protected against major mechanical impact. They are separated from the main process area and each other and kept in a naturally ventilated area.

5.9.3.2 Pressure Containment

Where practicable, process vessels, equipment and pipework are designed for the maximum credible pressure excursions and shall minimise the extent and capacity of the overpressure protection and disposal systems.

Phase 1 Final Report FEED Engineering Summary

5.9.3.3 Emergency Shut Down (ESD)

The emergency shutdown system (ESD) system is designed to prevent the escalation of abnormal conditions into a major hazardous event and limit the extent and duration of the hazardous event.

5.9.3.4 Pressure Relief System

A Pressure Relief system is provided where fire or a process system or equipment failure can cause the pressure in an equipment item or pressure system to exceed the maximum allowable working pressure.

5.9.3.5 Ignition Control

A hazardous area classification schedule has been prepared in accordance with EI 15 Model Code of Safe Practice Part 15: Area Classification for Installations Handling Flammable Fluids to minimise the likelihood of flammable gas ignition from electrical equipment in hazardous areas.

The selection of electrical equipment for use in hazardous areas shall be appropriate for the zone in which it is located. This will be addressed in the detailed design phase.

Neither CaCO_3 nor CaO are combustible so fine particles suspended in the air cannot cause dust explosion. Hence, dust hazardous area classification does not apply to the facility.

In the detailed design phase earth bonding will be specified to non-electrical equipment to minimise the likelihood of ignition from static in hazardous areas.

5.9.3.6 Fire Protection system

An active and passive fire protection system is not envisaged for the project facility handling non-combustible Calcium carbonate (CaCO_3) and Calcium Oxide (CaO).

Calcium oxide will not burn. Calcium oxide reacts with water and generates sufficient heat to ignite combustible material. Hence, the application of water is not recommended on calcium oxide.

Accidental ignition of hydrogen gas and fuel gas used in the Calciner should be allowed to burn until the fuel source can be isolated. An emergency isolation valve should be considered in the hydrogen and fuel gas piping to the Calciner.

Portable fire extinguishers shall be considered for Class A and Class E fire hazards.

Phase 1 Final Report FEED Engineering Summary

A further review of fire-fighting provision will be conducted in the context of the selected site and recognising both any neighbouring equipment or activities and any fire-fighting provisions already in place at the site.

5.9.3.7 Escape Routes and Muster Points

Escape routes leading to a muster from all locations will be identified following the selection of the site. Escape routes will be well marked, including signs. Marking shall show the preferred direction of escape. There shall always be a minimum of two (2) access/escape routes from any location.

A place where personnel can muster safely in the event of an emergency will be identified.

Emergency escape lighting will be provided along designated emergency escape routes and muster points.

5.9.3.8 Emergency Communication

An emergency communication system will be provided to warn and guide plant personnel in an emergency. This will be fully specified in the detailed design phase following selection of the site.

5.9.3.9 Emergency Power

An uninterruptible power supply will be provided to ensure continuous supply to the escape lighting, emergency alarm and communication system and emergency shutdown system. At this stage it is assumed that this will be present on the selected site. This will be confirmed in the detailed design phase.

5.9.3.10 Personnel Protective Equipment

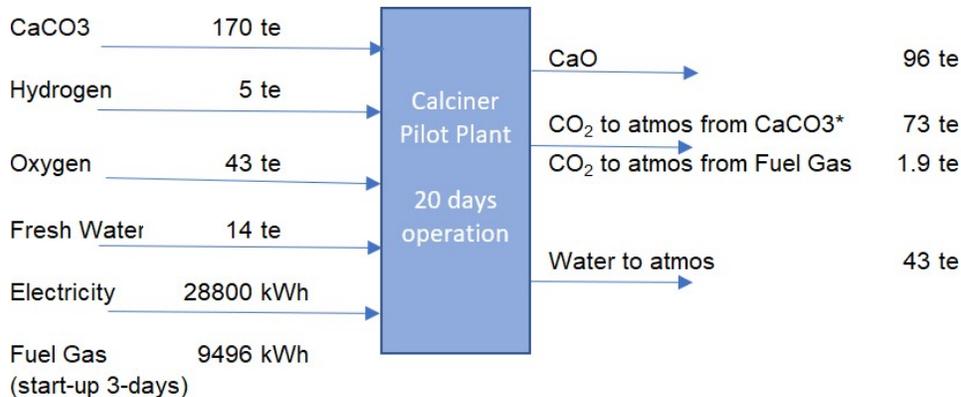
Appropriate protective equipment will be provided to plant personnel to prevent direct exposure of calcium carbonate and calcium oxide to plant personnel.

Emergency safety showers with eyewash stations shall be provided in areas where calcium carbonate and calcium oxide solids are being handled. Plant personnel may be exposed to the product through operating or maintenance tasks.

Phase 1 Final Report FEED Engineering Summary

5.10 Energy and Fuel Requirements

The overall consumables and outputs for the pilot scale Hydrogen fired Calciner were evaluated and are presented Figure 5-13. They are based on 20 days continuous operation and 3 start-ups (warm-up periods) i.e. they account for the full data gathering period. Fluidisation air is not included in the figure above as it goes straight through the Calciner.



*CO₂ from decomposition of CaCO₃ was captured from the air in Squamish

Figure 5-13: Calciner Pilot Plant Consumables and Products

5.10.1 Hydrogen

The Calciner uses hydrogen for combustion during normal operation test runs. Gaseous Hydrogen will be supplied from 4,000 Nm³ trailer tubes at 200 barg at 15°C as the safest and most cost-effective means to source Hydrogen for the pilot plant. Each hydrogen tube trailer will provide ~1.5 days supply. Provision is made to accommodate 4 hydrogen trailers on site to be connected to a manifold to supply the Calciner with hydrogen at the desired let-down pressure of 250 kPag. Hydrogen tube trailers will be on order to replace the depleted hydrogen trailers to ensure adequate supply is continuously available for the desired batch test duration.

5.10.2 Start-up Fuel Gas

Fuel gas is required at start-up for approximately 1 day to preheat the Calciner. Natural gas is used for this duty in the pilot phase due to its lower cost compared to hydrogen. The flowrate required is ~12 Nm³/h based on natural gas from the UK national grid having a calorific value of 39.4 MJ/m³. Assuming 3 start-up campaigns of the pilot plant, the fuel gas flowrate required is 868 Nm³/h. A fuel gas connection supplying gas from the UK National Grid is assumed to be available at the UK pilot facility site.

Phase 1 Final Report FEED Engineering Summary

5.10.3 Cooling Water

A cooling water package PK-004 will provide cooling water for the required duty for the Calciner. The closed loop system will supply water to the Calciner at 20-30°C and 324 kPag.

5.11 Environmental Impact

5.11.1 Emissions

Continuous emissions generated by the facility will be limited to CO₂. Total CO₂ emissions to air from the Calciner package are expected to arise from normal operation of the demonstration facility over a total of 20 days and a total of 3 days for CO₂ emissions released from fuel gas combustion during start-up.

The CO₂ produced from the decomposition of the CaCO₃ is that which was captured from the air at the plant in Squamish. Therefore, although it is quantified in this section, is not considered to be an emission of the Dreamcatcher project.

There are no regulatory discharge limits for venting CO₂ to the atmosphere for facilities of research and development (including testing new products and processes) within the UK as per Paragraph 2, Schedule 2 of UK ETS (UK Emissions Trading Scheme) under the greenhouse gas emissions trading scheme Order 2020.

Ambient dust levels of CaCO₃ and CaO are controlled under the air quality standards regulations 2010. The bag filter downstream of the Calciner receives wet CO₂ at 175°C and removes the dust before discharging safely into the atmosphere. As a result, there will be less than 850mg/Nm³ emission of dust emitted to atmosphere which is considered negligible.

5.11.1.1 Carbon Dioxide Equivalence (CO₂e)

For air emissions an additional metric, the carbon dioxide equivalence (CO₂e), has also been presented. This equivalence represents the greenhouse gas (GHG) of certain components present in the gaseous air emissions other than CO₂ and expresses them as a CO₂ equivalent (CO₂e).

Phase 1 Final Report FEED Engineering Summary

5.11.1.2 Calculation Methodology

The air emissions for Carbon Dioxide (CO₂) were calculated based on the UK Government GHG Conversion Factors and blue Hydrogen intensity factor from the Global CSS Institute. Table 5 2 shows the emission factors used.

Table 5 2: Emissions Factors

Pollutant Source	Emissions Factor	Unit
Blue Hydrogen	2	kg CO ₂ e/kg
Fuel Gas (start-up)	0.20297	kg CO ₂ e/kWh
Water	0.14900	kg CO ₂ e/m ³
Electricity	0.21233	kg CO ₂ e/kWh
Transport: Road	1.05913	kg CO ₂ e/km
Transport: Rail	0.02556	kg CO ₂ e/km
Transport: Ship	0.003539	kg CO ₂ e/km

Fuels conversion factors should be used for primary fuel sources combusted at a site. The UK Government GHG system is useful in cases where a facility is not yet operational or where other specific performance specifications are not known for a specific piece of equipment, and an estimation of the air emissions is necessary.

¹ <https://www.gov.uk/government/publications/greenhouse-gas-reporting-conversion-factors-2021>

² Circular-Carbon-Economy-series-Blue-Hydrogen, Global CCS Institute, Blue Hydrogen April 2021

Phase 1 Final Report FEED Engineering Summary

Air emissions were calculated based on fuel consumption rates (or the power output for each source or distance travelled) and the relevant Emission Conversion Factor (CF), based on the following equation:

$$E = (X) \times (CF) \times (T)$$

Where:

- E** = emissions rate (CO₂ emitted by the source)
- X** = rate of fuel consumption, power output or distance travelled
- T** = annual operating period (expressed in hour)
- CF** = emissions conversion factor (expressed in kg/kWh, kg/kg fuel or kg/km)

An assessment of the total CO₂ produced from the process and from the transport of CaCO₃ feedstock and CaO produced are included below in Table 5-3..

The following assumptions were made for the assessment:

- Test runs operating continuously totalling 20 days
- start-up campaigns (i.e. 3-days fuel gas combustion)
- Hydrogen emission intensity based on production by steam methane reformation with carbon capture (Blue hydrogen)
- The air emissions calculations assume 100% combustion efficiency

Table 5-2: CO₂ Emissions from Plant Operation

Emissions Sources	Consumption	Unit	Emissions, kg CO ₂ e
Hydrogen	4732	kg	9,463
Fuel Gas (start-up)	9496	kWh	1,927
Water	14.4	m ³	2
Electricity	28800	kWh	6,115

Phase 1 Final Report FEED Engineering Summary

Total	-	-	17,507
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Three alternative routes are considered for shipping the CaCO₃ to Spadeadam and return the CaO to Squamish in Canada (the values to Buxton will be very similar). Option A, which makes the most extensive use of shipping has the lowest overall emissions.

Table 5-3: Materials Transport Routes

Transport Route of CaCO ₃ and CaO
Route A: Squamish-Vancouver (Road) ↔ Vancouver-Liverpool (Ship via Panama Canal) ↔ Liverpool-Spadeadam (Road)
Route B: Squamish-Vancouver (Road) ↔ Vancouver-New York (Rail) ↔ New York-Liverpool (Ship) ↔ Liverpool-Spadeadam (Road)
Route C: Squamish-Vancouver (Road) ↔ Vancouver-Toronto (Rail) ↔ Toronto-New York (Road); New York-Liverpool (Ship) ↔ Liverpool-Spadeadam (Road)

Table 5-4: Total CO₂e from Transport of CaCO₃ Feedstock and Produced CaO

Transport	kg CO ₂ e
Option A	101,729
Option B	124,527
Option C	341,367

Table 5.6 shows the amount of CO₂ captured in Squamish that will be returned to the atmosphere during the demonstration period.

Table 5-5: Captured CO₂ from Squamish Returned to Atmosphere

Emissions Sources	CO ₂ Returned, kg

Phase 1 Final Report FEED Engineering Summary

CO ₂ emission from Calciner (CaCO ₃ decomposition)*	72,816
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*CO₂ captured in Squamish

5.11.2 Effluent discharge

The facility will not produce any aqueous discharges.

5.11.3 Waste

The production of waste will be limited to municipal waste generated from office and welfare facilities on site which will be managed in accordance with waste management regulations.

No waste materials will be produced from the process. CaO produced from the process will be collected and is assumed to be shipped back to regenerate the Calcium Hydroxide / water slurry in the Pellet Reactor at the Squamish site in Canada.

In the event that CE deem the produced CaO to be similar enough in its properties (i.e. particle size, bulk density) to that produced in the natural gas fired unit that there is no research and development value in returning it to Canada, it may be disposed of in the UK. CaO, or quicklime, is widely used in industrial processes so routes into this supply chain for re-use will be evaluated in this eventuality. This would reduce both carbon footprint and cost of the pilot.

5.11.4 Use and Storage of Chemicals

Storage of the CaCO₃ pellets, arriving in 1 tonne bags stacked 2 wide and 2 high, are stored onsite in 20 ft shipping containers. The produced CaO will be collected from the silo for storage onsite in 20 ft shipping containers prior to transfer back to Squamish BC, Canada.

Hydrogen will be brought in trailer tubes, 4000 Nm³ each providing 1.5 days supply and oxygen in a 38 te liquefied oxygen truck. Both the gaseous hydrogen and oxygen are stored at 200 barg. Fuel gas for start-up is assumed to be available at site.

The pilot Calciner will be operated for short data gathering campaigns. The plant will operate 4 to 10 day campaigns and storage capacity is provided in 20 ft containers for 20 days.

Phase 1 Final Report FEED Engineering Summary

5.12 UK Supply Chain

In phase 1 of the Dreamcatcher project the UK supply chain was engaged to identify suppliers of the equipment required for the pilot scale plant. The same suppliers are also suitable for the planned commercial scale plant where the equipment type is the same. Engagement with suppliers for the full commercial scale plant with its significantly longer equipment list is beyond the scope of this report.

The initial engagement suggests that the majority of the equipment can be procured through the UK supply chain.

6.0 Pilot Plant Location

Due to the planned short operating life of the pilot plant the intent is to utilise an existing dedicated test facility. Two suitable facilities have been identified:

- The DNV Spadeadam facility in Cumbria
- The HSE Science and Research Centre site at Buxton

These sites offer a number of advantages to the Dreamcatcher project:

- The sites are specifically designed for testing of equipment utilising hazardous substances such as natural gas and hydrogen.
- The undertaking of activities such as this is normal operating activity for the sites and reduces the requirement for new and bespoke permits and consents
- They offer existing access roads, security, utilities and administrative buildings
- They have existing, proven, procedures for managing flammable materials
- Their staff have expertise in data gathering and experimental procedures that could be leveraged in the detailed design and/or data gathering phase
- They are already accepted in their locations for this type of work thus eliminating the need for community engagement that may impact schedule

Both sites have indicated that they are open to accommodating the pilot unit. More detailed conversations are ongoing and will be concluded should the project be selected for phase 2.

Phase 1 Final Report Pilot Plant Location

Table 6-1: Addresses of Potential Pilot Plant Locations

DNV facility at Spadeadam	HSE Science and Research Centre at Buxton
Spadeadam Test Site Mod R5, Brampton CA8 7AU UK	Harpur Hill Buxton Derbyshire SK17 9JN UK

6.1 Permits and Consents

It is clearly an advantage to use a pre-existing test facility as they are set up and have the relevant consents in place to support these processes as part of their normal operating regime. This reduces the permitting requirements for the project. The outcome being as follows:

- Town and Country Planning – not expected to be required given within existing test facility and temporary in nature.
- Environmental Permitting - not expected to be required given within existing test facility and temporary and benign in nature.
- Emissions Trading – not required for test facilities and emissions not of a sufficient level to require reporting.
- Health and Safety Regulations – Health and Safety at Works would apply and be followed as part of normal operating procedures. Construction (and Design) Management Regulations would be applicable and taken forward in conjunction with site owners, who have established operating practices to reduce risks. Control of Major Accident Hazards (COMAH) is not expected to be required; however, the project will benefit from the sites experience in managing test programmes utilising flammables.

Phase 1 Final Report Execution Plan

7.0 Execution Plan

The following sections provide a summary of the Project Execution Plan for Phase 2 that was developed as part of the FEED. The section provides a high-level work methodology for the Phase 2 execution of the Pilot Plant.

7.1 Delivery Team Structure & Responsibilities

7.1.1 Steering Board

The delivery of Project Dreamcatcher Phase 2 will be overseen by a Steering Board made up of senior representatives from Pale Blue Dot Energy Limited, Carbon Engineering Limited and Petrofac Limited.

In its capacity as Lead Developer Pale Blue Dot Energy will chair the Steering Board.

The Steering Board will meet every month to monitor progress and address issues that might arise in the project.

7.1.2 Project Direction

The project will be led by:

- A Pale Blue Dot Energy Project Director who will have ultimate accountability for delivery of the project
- A Petrofac Project Manager who will be accountable to the Project Director and responsible for the delivery pilot plant
- A Carbon Engineering Technical Project Director will be accountable for the design and execution of the data gathering programme and the delivery of the Calciner for integration into the plant

7.1.3 Project Management Office

Petrofac will provide the Project Management Office (PMO) support for the project.

They will deploy their proven procedures, processes and tools for executing Pre-FEED projects of this nature. Their ISO-9001 Quality Management System includes

- Project Management Roadmap and Manual
- Project Execution Plan Development Guidelines
- Project Controls Procedure
- Engineering Design Integrity Manual

Phase 1 Final Report Execution Plan

Additionally, they will utilise Petrofac’s detailed engineering procedures, workflow processes, design software and design manuals, along with UK and international codes and standards to undertake work to the highest standards.

7.1.4 Project Organisation

Petrofac will supply the majority of the project execution organisation but will work closely with CE and the Site Team. The project organisation chart is provided in Figure 7-1below.

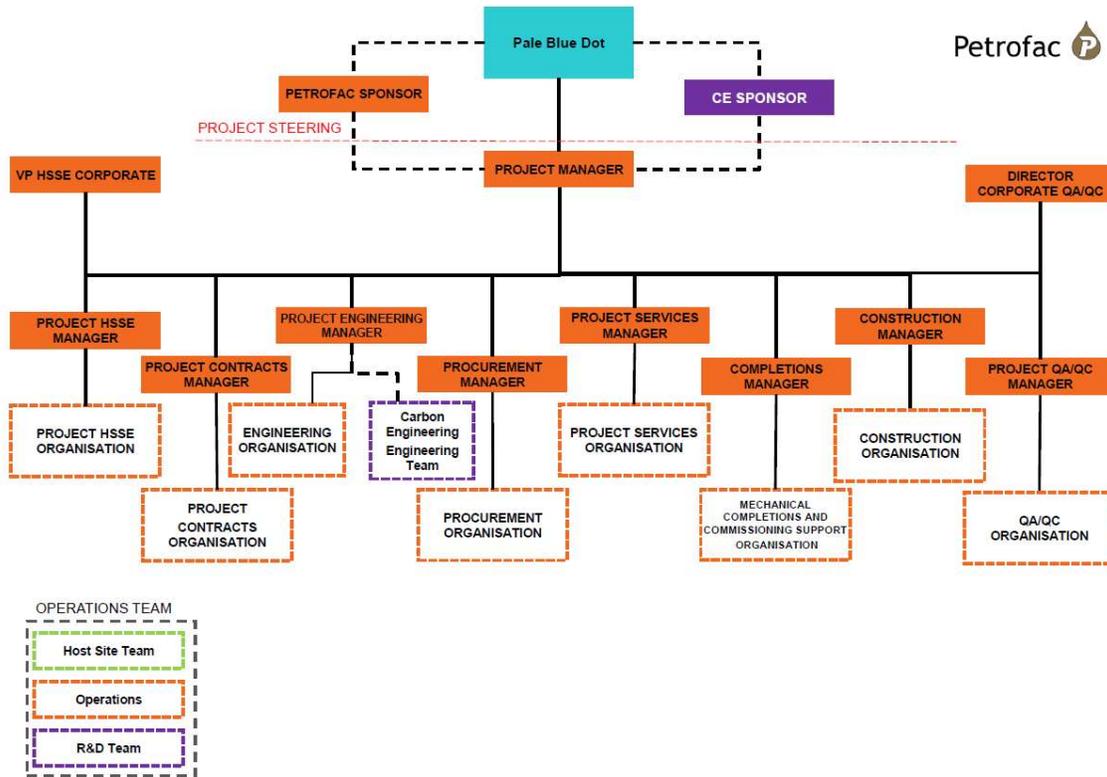


Figure 7-1: EPC Organisation chart

7.2 Schedule

The Phase 2 duration is 27 months from Award to complete Engineering, Construction, Operational Testing and Decommissioning.

A summary GANTT chart is presented in Figure 7-2

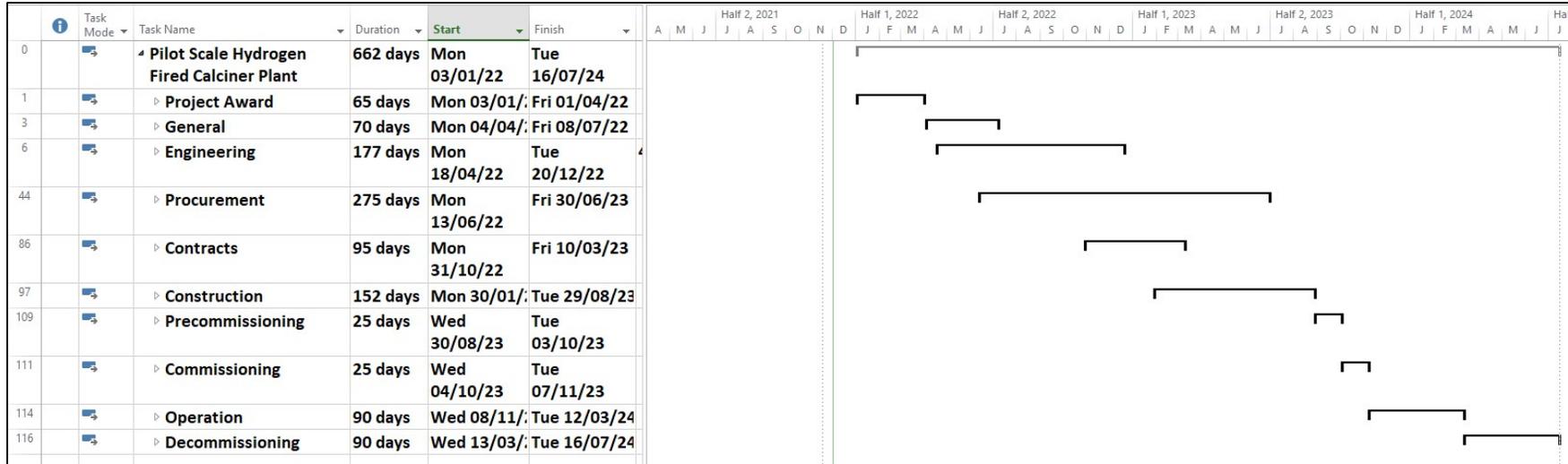


Figure 7-2: Summary Gantt Chart

Phase 1 Final Report Execution Plan

7.3 Project Management

7.3.1 Petrofac Project Management Framework

Petrofac will use its suite of Project Controls and Project Management processes and procedures to provide accurate and timely key information to Petrofac and Pale Blue Dot Project Management Teams, allowing control of key activities within the project. AN overview of these procedures and tools is provided in Figure 7-3 below.

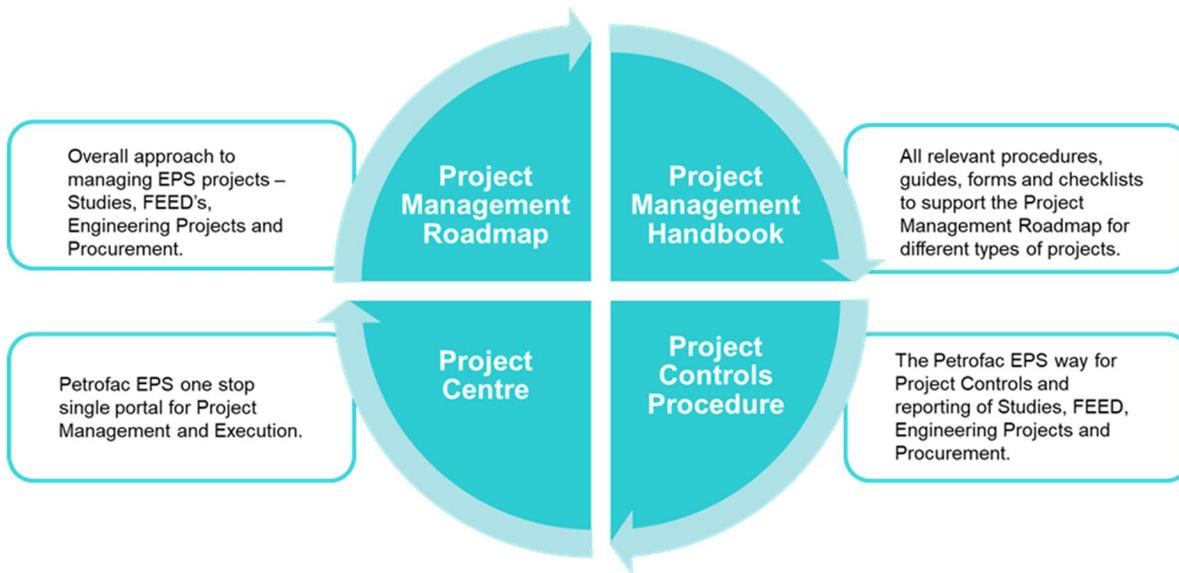


Figure 7-3: Project Management Framework

7.4 Procurements and Subcontracts

This section describes the supply and the contracting strategy that will be followed in the EPC phase for the Pilot Plant to unlock value and reduce risks.

Supply Chain Principles

Petrofac will follow its Supply Chain Principles which set out best practices for business conduct and establish our principles and expectations on how employees and the wider supply chain will conduct business.

Phase 1 Final Report Execution Plan



Figure 7-4: Petrofac Supply Chain Principles

In Phase 1 of the Dreamcatcher project the UK supply chain was engaged to identify suppliers of the equipment required for the pilot scale plant. The initial engagement suggests that the overwhelming majority of the equipment can be procured through the UK supply chain.

A list of potential UK manufacturers and suppliers of new equipment items and packages was developed for the 1,000 te-CO₂/year scale clean Calciner pilot plant. For a plant operating for such a limited period of time it may be possible to hire some equipment. Due to its relatively specialist nature this option is supply and timeframe specific which will be evaluated further in detailed engineering.

In Phase 2 Petrofac will engage further with the supply chain to:

- Evaluate used equipment market for mechanical handling equipment that could be re-purposed to reduce overall cost
- Confirm the packaging of Equipment
- Confirm the Approved Suppliers list for each Package
- Further assess the rental of packages
- Identify Long lead items to prioritise early Enquiry and Award to reduce schedule

Phase 1 Final Report Execution Plan

7.4.1 Procedures and Processes

Petrofac will follow its proven procedures and processes for Procurement and Subcontracts in the execution of the works.

7.4.2 Mechanical Packages

The objective is to minimise the number of procurement packages, based on vendor capabilities, and maximise the prefabrication of package to facilitate installation at site. The Package contents and proposed level of packaging/prefabrication will be confirmed during detailed design.

7.4.3 Subcontract strategy

The Subcontracting Strategy will be finalised once the Scope of Work has been finalised. It is envisaged that the following subcontracts will be awarded:

1. Site Preparation/Civil Works
2. General Contractor – Structural, Mechanical, Electrical and Instrumentation

7.4.4 UK Construction Partner

Petrofac will select the most appropriate construction partners for delivery of the Pilot Plant from a small group of preferred contractors. The selected construction partner will be involved during the Engineering phase to gain their insight, knowledge to ensure the design and schedule of the project are robust and achievable, as well as early plans for a project Industrial Relations strategy, logistics, constructability etc.

7.5 Project Controls and Services

Following the award of the EPC phase and building on the FEED phase, the Project Controls and Services Manager will build up and expand the Project Services and Controls System to accommodate the Detailed Engineering, Procurement, Construction Installation and Commissioning Phases of the project. The team will manage:

- Planning and Scheduling
- Risk Analysis
- Progress Reporting
- Progress Measurement
- Manpower Planning and forecasting
- Cost Control

Phase 1 Final Report Execution Plan

7.5.1 Risk & Opportunity Management

The objective of the Risk Management process is to support the successful execution of the project through the effective implementation of Risk and management practices. Project Risk will be managed in accordance with a Risk Review Procedure generally in accordance with ISO 31000 “Risk management – Principles and Guidelines”.

On identification of sources of risk, their potential impact is calculated in terms of individual and combined risks. The impact of risk calculation will follow the typical formula where:

- Impact of risk = likelihood of risk occurring x consequence of risk
- Identified risks will be added to the Risk Register
- Risk reduction measures will be detailed for each risk including:
 - Avoidance where plans are redrawn to eliminate the risk
 - Deflection by passing the risk to other contractors, or vendors if they are better positioned to control the risk
 - Mitigation measures that are planned and subsequently managed to prevent occurrence of the risk.

The overall control document for project risk will be the risk management register. The risk management register will log the risks identified for the project, how they have been assessed, the strategy for their reduction or mitigation and the risk owner who is responsible for managing the risk reduction approach.

Risks will be monitored on a regular basis to determine the effectiveness of any mitigation measures and the need or otherwise to revise the risk reduction approach. The incorporation of newly identified risks will be an on-going process.

7.6 Interface Management

Interface management will continue to be critical to ensure that all the identified interfaces continue to be managed. An interface register will be maintained ensuring that the required communication plan is implemented.

7.7 Construction and Installation

Petrofac recognises that the construction success for the Pilot Plant will be ensured by our understanding and implementation of the following performance standards.

Phase 1 Final Report Execution Plan

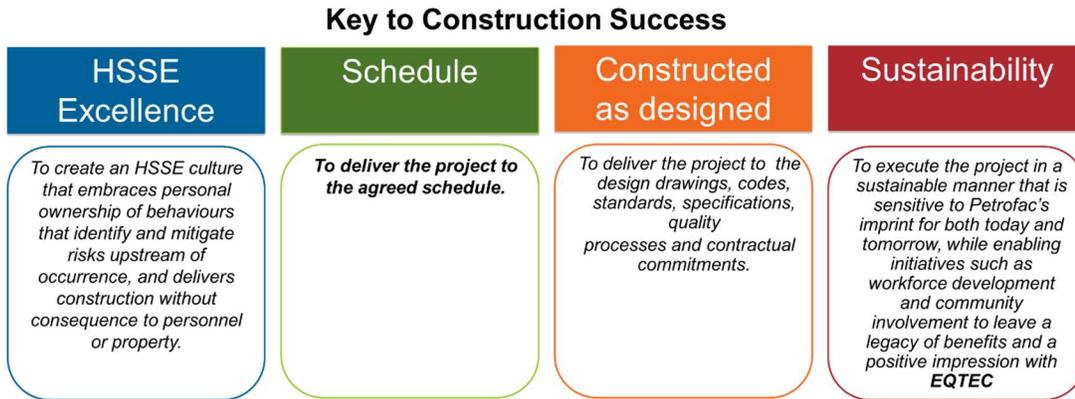


Figure 7-5: Key to Construction Success

The construction key objectives are detailed and included below in Figure 7-6.

Phase 1 Final Report Execution Plan

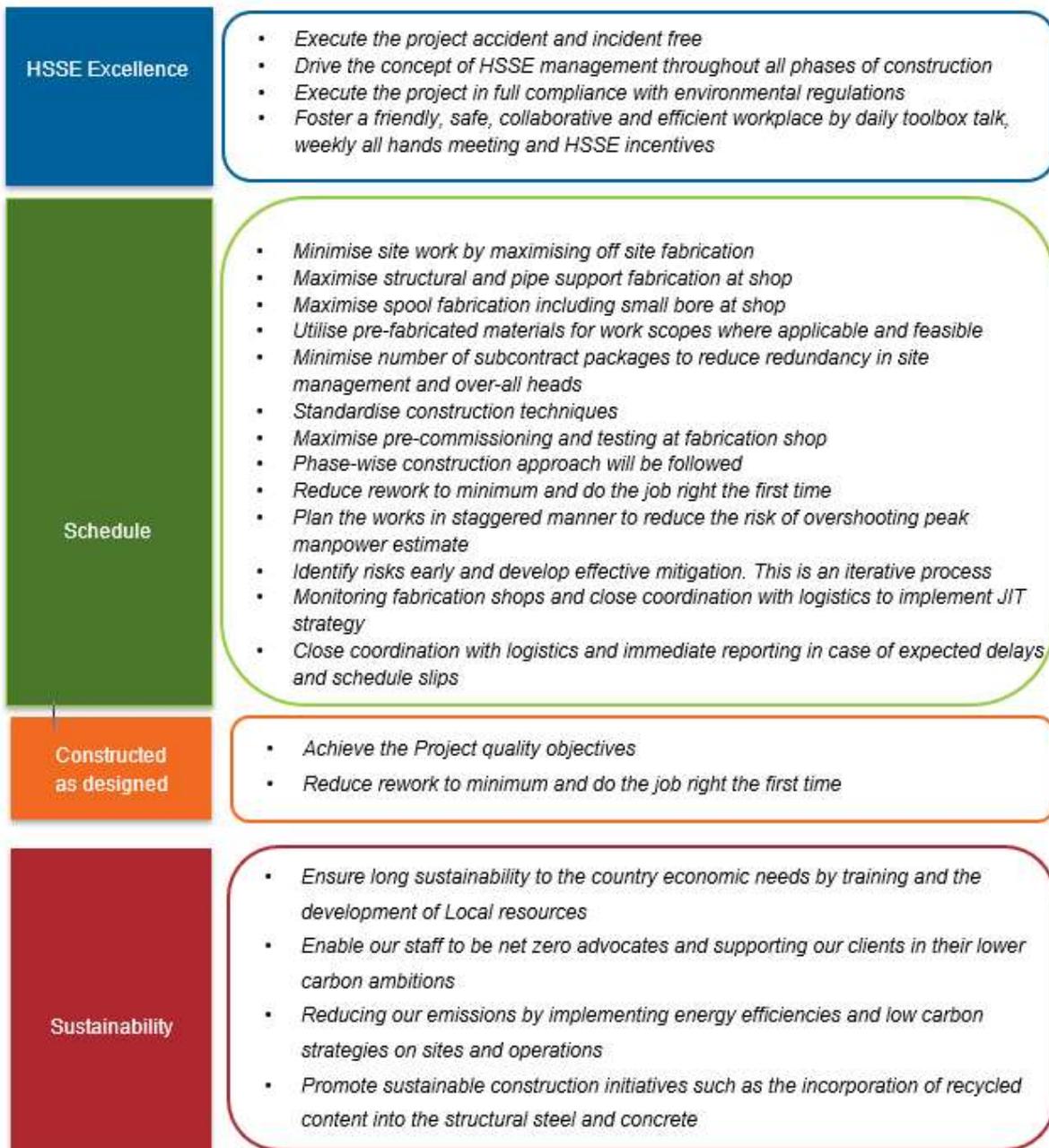


Figure 7-6: Key Construction Objectives

Phase 1 Final Report Execution Plan

1.1.1 Organisation and Responsibilities

Petrofac will work with a selected construction subcontractor to deliver the main project construction works. Petrofac will also appoint other subcontractors, as required, to support the overall construction strategy.

Petrofac will, along with our construction partner, staff the site Construction Manager, Superintendents, Safety and Materials Management positions to use the team's regional knowledge, familiarity with the type of facility, relationships with local labour and relationships with the community.

The roles and responsibilities of the above key positions will be clearly defined in the construction execution plan.

7.7.1 Procedures and Processes

Petrofac will follow its proven procedures and processes for Construction and Pre-Commissioning.

7.7.2 Constructability

Constructability, as a work process, will continue to be an integral part of project execution via dedicated construction personnel involvement in engineering design, planning, logistics and subcontracting philosophy.

This work process will be maintained throughout the detailed engineering phase of the project. Using a set of phased checklists, the design will be continuously subjected to systematic review.

The constructability team will closely interface with Project Management team to finalise:

- Transport routes
- Underground protection
- Drawings issuance sequence
- Subcontractor technical validation
- Construction methodologies
- Construction team member presence during model review sessions and package vendor clarification discussions.
- Reviewing of inter-discipline interfaces in order to maximise parallel activities (e.g., underground foundations, pipe work and trenches)
- Generating final optimised construction procedures

Phase 1 Final Report Execution Plan

7.7.3 Construction Strategy and Construction Execution Plan

Petrofac's construction execution strategy recognises:

- The critical necessity for seamless interface between engineering, procurement, construction, pre-commissioning and commissioning
- Working closely with our construction partner

Engineering and procurement will collectively support construction with timely deliverables. Construction will phase into the lead during the early works programme but continue to interface with engineering and procurement to ensure clear understanding of sequencing and deliverable priorities. One of the important construction strategies is minimising site activities by maximising off site fabrication and the supply of Equipment as skid mounted packages. Construction phases envisaged for the Pilot Plant are illustrated below:

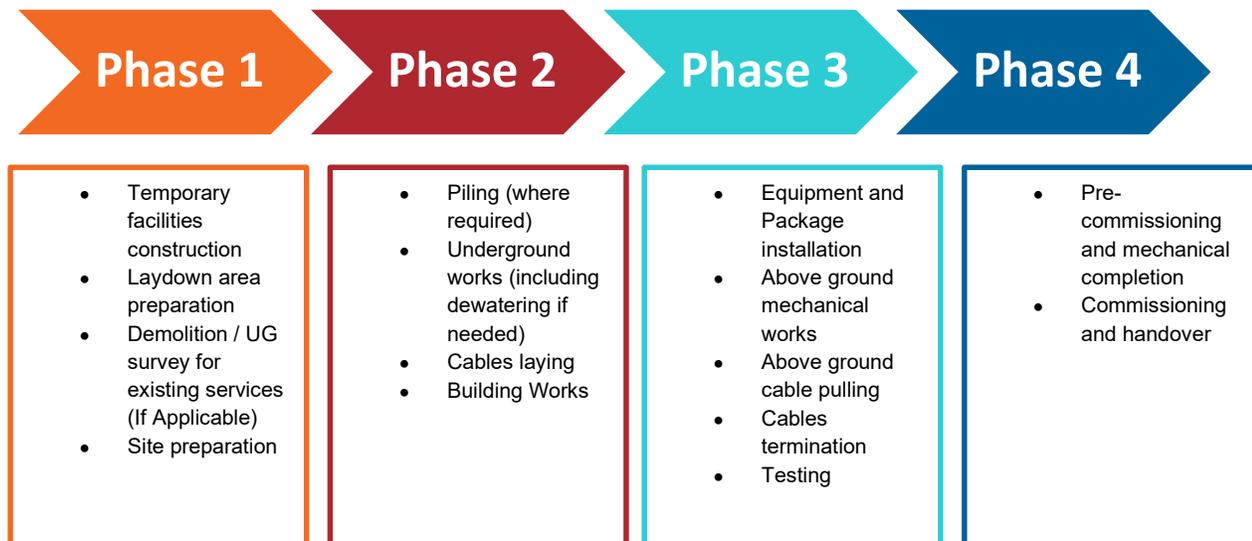


Figure 7-7: Construction Phases

Phase 1 Final Report Execution Plan

7.8 Commissioning

The commissioning of the Pilot Plant will be in the EPC scope of work. The base assumption is that Petrofac will manage the entire process providing continuity throughout and will ensure that this happens through the timely application of appropriate resources. In the event that the selected site's policies demand that it manages the commissioning process, Petrofac will remain available to provide any support necessary.

The key Commissioning Support staff will be part of the Construction team during the EPC phase and will move through the various phases of mechanical completion, pre-commissioning, commissioning and start-up in order to provide continuity and ensure that the knowledge gained during the earlier stages is carried through into the operation phase. The support of key engineering, including CE, personnel will be available throughout the commissioning period.

7.8.1 Commissioning Processes and Procedures

Petrofac will use proven processes for Commissioning and develop detailed procedures prior to execution.

A detailed Pre-Commissioning and Commissioning Plan (schedule) for systems and subsystems will be developed including the commissioning and start-up sequence during the detailed engineering phase of the EPC phase of the Pilot Plant. This will identify the sequence and duration of all the activities required to adequately complete the process and handover completed systems to Operations.

Petrofac's Commissioning/Completions Group will be actively involved in the following phases of the project:

- Engineering/Procurement Phase (Home Office Preparation Phase)
- Mechanical Completion and Pre-Commissioning
- Commissioning and Ready for start-up
- Start-up / Initial Operation / Performance Testing

The project's mechanical completion and commissioning execution methodology will be developed during the EPC phase.

Phase 1 Final Report Execution Plan

7.9 Operation and Maintenance

The operations team will comprise representatives from host site, CE and Petrofac.

It is expected that personnel from the selected host site will ultimately own the site operating procedures and take responsibility for the pilot plant operation. This is expected to include provision of the plant manager and key operating and maintenance personnel. This will support seamless integration with the host site facilities and operating systems, leverage their experience of similar test work and to satisfy any site policies.

CE personnel will own the test programme and provide expertise on operation and optimisation of the Calciner unit. They will also act as the link with the Calciner vendor's technical team.

Petrofac will provide any additional operating and maintenance labour, supported by its UK Technical Support Hub to provides the knowledge and capabilities in the areas of recruitment, supply, mobilisation and management of people, logistics management, equipment supply, training, competence development and assurance. It will also leverage is operations experience to develop the operating/test procedures with input from CE and the host site operators.

8.0 Pilot Plant Location

Due to the planned short operating life of the pilot plant the intent is to utilise an existing dedicated test facility. Two suitable facilities have been identified:

- The DNV Spadeadam facility in Cumbria
- The HSE Science and Research Centre site at Buxton

These sites offer a number of advantages to the Dreamcatcher project:

- The sites are specifically designed for testing of equipment utilising hazardous substances such as natural gas and hydrogen.
- The undertaking of activities such as this is normal operating activity for the sites and reduces the requirement for new and bespoke permits and consents
- They offer existing access roads, security, utilities and administrative buildings
- They have existing, proven, procedures for managing flammable materials
- Their staff have expertise in data gathering and experimental procedures that could be leveraged in the detailed design and/or data gathering phase
- They are already accepted in their locations for this type of work thus eliminating the need for community engagement that may impact schedule

Both sites have indicated that they are open to accommodating the pilot unit. More detailed conversations are ongoing and will be concluded should the project be selected for phase 2.

Phase 1 Final Report Pilot Plant Location

Table 8-1: Addresses of Potential Pilot Plant Locations

DNV facility at Spadeadam	HSE Science and Research Centre at Buxton
Spadeadam Test Site Mod R5, Brampton CA8 7AU UK	Harpur Hill Buxton Derbyshire SK17 9JN UK

8.1 Permits and Consents

It is clearly an advantage to use a pre-existing test facility as they are set up and have the relevant consents in place to support these processes as part of their normal operating regime. This reduces the permitting requirements for the project. The outcome being as follows:

- Town and Country Planning – not expected to be required given within existing test facility and temporary in nature.
- Environmental Permitting - not expected to be required given within existing test facility and temporary and benign in nature.
- Emissions Trading – not required for test facilities and emissions not of a sufficient level to require reporting.
- Health and Safety Regulations – Health and Safety at Works would apply and be followed as part of normal operating procedures. Construction (and Design) Management Regulations would be applicable and taken forward in conjunction with site owners, who have established operating practices to reduce risks. Control of Major Accident Hazards (COMAH) is not expected to be required; however, the project will benefit from the sites experience in managing test programmes utilising flammables.

9.0 Costed Project Plan

Two cost estimates were developed as part of the study:

1. A Cost Estimate for the pilot plant to inform decisions regarding Phase 2 of the project
2. A Cost Estimate for a commercial scale plant to support the development of a roadmap to commercial deployment

9.1 Cost Estimate for Pilot Plant

9.1.1 Approach to Estimate

9.1.1.1 Estimate Accuracy

The Class 3 CAPEX estimate was developed in line with AACEI guidelines with an expected accuracy of +30%/-20%. The OPEX and Decommissioning estimates are both considered Class 4 with an accuracy of + 40%/-20%. As they make up a small proportion of the cost the lower level of maturity of these estimates is considered acceptable and suitable for decision making at this stage.

Phase 1 Final Report Costed Project Plan

Table 9-1: AACEI Estimate Classes and Key Characteristics

	Level of Project Definition	End Usage	Methodology	Expected Range of Accuracy	Preparation Effort
Class	Expressed as a % of complete definition	Typical Purpose of Estimate	Typical Estimating Method	Typical variation in low and high ranges at an 80% confidence interval	Typical effort relative to least cost index of 1
Class 5	0% to 2%	Conceptual Planning	Capacity Factored, parametric Models, judgement or Analogy	L: -20% to -50% H: +30% to +100%	1
Class 4	1% to 15%	Screening options	Equipment Factored or Parametric models	L: -15% to -30% H: +20% to +50%	2 to 4
Class 3	10% to 40%	Funding authorisation	Semi-Detailed Unit Costs with Assembly Level Line Items	L: -10% to -20% H: +10% to +30%	3-10

9.1.1.2 CAPEX Estimate Date and Currency

The estimate has a base date of 4Q2021 and is presented in GBP Sterling. No forward escalation has been applied. Where applicable, the estimate used exchange rates of:

Phase 1 Final Report Costed Project Plan

- \$1.41 = £1.00
- €1.18 = £1.00

9.1.2 CAPEX

9.1.2.1 Material and Equipment Supply

Petrofac's in house cost estimating tool BE\$T uses equipment lists generated by discipline engineering as basic inputs to calculate equipment costs. A blend of inhouse historical cost data and vendor information was used for the equipment items.

BE\$T utilises equipment weights (as specified by discipline engineering) to derive the weights for bulk items using factors. Bulks are priced based on a £/tonne basis and are obtained from recent EPC projects. The following elements of the estimate are calculated from these bulk factors:

- Piping
- Electrical (non-equipment)
- Instrumentation (non-equipment)
- Primary Steelwork
- Secondary Steelwork (T posts MTO)
- Tertiary Steelwork (T posts MTO)
- Safety Equipment
- Architectural
- Other - includes items such as insulation, paint and cathodic protection

Instrumentation - The control system was estimated using the number of envisaged I/Os using a company historical cost norm.

Structural allowances - Posts supporting cable trays and small-bore pipes were factored and added to the provided MTO from Civils and Structural.

Phase 1 Final Report Costed Project Plan

9.1.2.2 Fabrication

Fabrication costs were derived by allocating man-hours to each of the bulk material elements based on their weights. These man hour allocations are based on Petrofac's in-house data. The fabrication cost is then obtained by multiplying the weight to the calculated manhours.

9.1.2.3 Installation

Installation costs were derived by allocating man-hours to each of the equipment and bulk material elements based on their weights. These man hour allocations are based on Petrofac's in-house data. The installation cost was obtained by multiplying the weight to the calculated manhours.

9.1.2.4 Civils

The foundations quantities were provided by preliminary Material Take Off. The allowance for the construction was estimated using norms generated from company historical costing data for each activity,

Note: The MCC building was assumed to be housed in a purchased Portacabin.

9.1.2.5 Indirect Costs

In addition to direct costs (equipment and materials supply, fabrication, installation) indirect costs must also be considered to obtain the total installed costs. These indirect costs were factored from the calculated direct costs.

9.1.2.6 Owner's Costs

- Petrofac has applied a 7% allowance on top of the direct cost to allow for the owner's costs. These are included in the Indirect costs and cover:
 - Owner's project management team
 - Permitting
 - Specialist studies
 - Independent verification

9.1.2.7 Contingency

A contingency of 20% has been applied to CAPEX estimate to capture the known unknowns of the project.

Phase 1 Final Report Costed Project Plan

9.1.3 OPEX

9.1.3.1 Direct Costs

The OPEX costs have considered the quantities in Table 9-2 and the unit rates in Table 9-3.

Phase 1 Final Report Costed Project Plan

Table 9-2: OPEX Quantities

Description	Quantity	Duration	Notes
Manning			
Client Operators	3	3 months	
Site Supervisor	1	3 months	
Site 3 operatives	2	3 months	
Manhour rate include Overheads, Management, Supervision and Safety and Security			
Consumables			
CaCO ₃	170	te	Free Issued 20 days
Hydrogen	5	te	20 days
Oxygen	43	te	20 days
Fresh Water	14	te	20 days
Electricity	28,800	kW/h	20 days
Fuel Gas	9,496	kW/h	Start-up 3 days
OPEX Freight			
CaCO ₃ from Squamish: 170 te in 8no 20' containers	8	-	Trips
CaO to Squamish: 96te in 7 no 20' containers	7	-	Trips
Site access and any host / oversite			
Allowance	-	-	275,000 GBP ⁽¹⁾

Phase 1 Final Report Costed Project Plan

1 - The commercial terms for the site have yet to be confirmed. An allowance of GBP 275,000 has been included in OPEX costs to account for site access and any host / oversite.

Table 9-3: OPEX unit rates

Element	Rates, GBP
CaCO ₃	Free issued by CE
Electricity	0.144 per kWh
Water	3.00 per Tonne
Hydrogen	2.40 per Tonne
Oxygen	0.79 per Tonne
Fuel Gas	0.04 per kWh

9.1.4 Decommissioning

9.1.4.1 Direct Costs

The cost of disposal of the equipment and bulks offset by any resale value are assumed to be cost neutral.

The basis for the deconstruction for the plant is considered to be 50% of the installation hours.

The following indirect costs have been applied.

Phase 1 Final Report Costed Project Plan

Table 9-4: Decommissioning Indirect Costs

Description	Percentage Applied	Assumptions
Detail Engineering	-	Engineering for removal assumed to be 50% of that required for installation
Project Management	-	Project Management for the removal assumed to be 50% of that for installation
Procurement	8.1%	Applied to Direct Cost
Construction Management	-	Construction Management assumed as 50% against the construction cost for installation

9.1.4.2 Decommissioning Contingency

A contingency of 30% was applied to the decommissioning estimate.

9.1.4.3 Owner's Costs

- Petrofac applied a 7% allowance on top of the direct cost to allow for the owner's costs. These are included in the Indirect costs and cover:
 - Owner's project management team
 - Permitting
 - Specialist studies
 - Independent verification

9.1.5 Assumptions

At this time the precise location of the DAC plant has not been identified but it is anticipated that it would be either at HSE Laboratory at Buxton or the DNV facility at Spadeadam.

Petrofac have an extensive amount of historical cost data available from previous projects, including pipeline installation projects as well as brownfield and small scope projects in the UK at sites including Bacton as well as major EPCs in Shetland. This experience has been utilised to provide estimates of the total installed costs.

Phase 1 Final Report Costed Project Plan

The estimate was developed on the following basis:

- Time on site nominally 6 months construction, 3 months operations, 3 months decommissioning
- Feedstock CaCO_3 is 'free issue' from CE
- Shipping costs for materials were included:

CaCO_3 from Squamish: 170 te in 8no 20' containers

CaO to Squamish: 96 te in 7no 20' containers

- Local labour is assumed for the duration of the construction and as a result no site accommodation requirements
- Included in the estimate are allowances for site surveys as well as engineering design and construction management. No man-power plan (MPP) was prepared
- An allowance was included for freight which includes transportation costs Ex Works to the site
- Certification, QA and Inspection allowances were based on a percentage of the total supply cost of equipment/pipeline materials bulks and fabrication/installation costs
- An allowance for commissioning spares, operating spares and strategic spares was included in the estimate as a percentage of the total equipment supply cost. This includes pressure and leak testing of the piping
- Freight insurance and construction risk insurance was included as an allowance against the total cost of the estimate
- Certification and where applicable manufacturer's documentation costs are included

9.1.6 Overall Cost Summary

The overall cost estimate is presented in Table 9-5

Phase 1 Final Report Costed Project Plan

Table 9-5: Class 3/4 Combined Estimate, +/-30%

Element	Million GBP
CAPEX	12.1
OPEX	0.7
Decommissioning	1.5
Total Development Cost	14.3

9.2 Cost Estimate for Commercial Scale Deployment

9.2.1 Approach to Estimate

To support the development of a roadmap to commercial scale deployment a Class 5 cost estimate was prepared through collaboration between Carbon Engineering, Petrofac and PBDE. The principal activities of each party by element are described below. The lead party for each element of the estimate are presented in Table 9-6.

Table 9-6: Lead Party for Estimate Elements

Carbon Engineering	Petrofac	Pale Blue Dot
Core DAC Plant	Offsite Utilities tie-ins	Transport & Storage Costs
	Owners Costs	
	OPEX	
	Decommissioning	

Phase 1 Final Report Costed Project Plan

9.2.2 CAPEX Estimating Methodology and Results

9.2.2.1 Core DAC Plant

At a Class 5 level of accuracy the CAPEX for a hydrogen fired DAC is anticipated to be comparable to that of a natural gas fired unit.

Carbon Engineering and their engineering partner, BBA, developed an estimate by modifying those developed for natural gas fired plants in the USA and Canada. Adjustments were made by CE/BBA to tailor the estimate to the specifics of a UK deployment using norms taken from Petrofac's internal database.

The adjustments specifically addressed:

- Equipment changes due to local site climatic conditions
- Labour rates
- Productivity Rates
- Anticipated capacity i.e. 0.5 MTCO₂/yr

9.2.2.2 Offsite Utility Connections

The offsite cost estimate was prepared in line with AACEI (Association for the Advancement of Cost Engineering International) 18R-97 Cost Estimate Classification System and is considered a Class 5 estimate with a target accuracy range of +/- 50%, i.e. the estimate is based upon conceptual designs and ideas.

At this time the precise location of the DAC plant has not been identified but it is anticipated that it would be in the Peterhead/St Fergus area to provide easy access to the Acorn CO₂ transport and Storage infrastructure. Offsite costs are therefore based on assumed tie-in distances for utilities. This approach means that the estimates are considered equally valid for any site in the UK.

Petrofac have an extensive amount of historical cost data available from previous projects, including pipeline installation projects as well as brownfield and small scope projects in the UK at sites including Bacton as well as major EPCs in Shetland. This experience was utilised to provide estimates of the total installed cost for each of the segments.

Phase 1 Final Report Costed Project Plan

A preliminary offsite scope required for the DAC to function has been identified. This is associated with the supply of utilities and the disposal of wastewater. All pipelines that have been considered have been assumed to be installed in buried trenches. Minor road crossings have been allowed for. Allowances for pipeline tie-ins have been made.

The estimates have been developed on the following basis:

- All pipe has been considered as carbon steel although there potentially is an opportunity for installation of GRE/GRP which should be considered once routing and tie in materials have been confirmed.
- Local labour is assumed for the duration of the construction and as a result no site accommodation has been provided.
- Included in the estimate are allowances for site surveys as well as engineering design and construction management.
- An allowance has been included for freight which includes transportation costs Ex-works to the site.
- Certification, QA and Inspection allowances have been based on a percentage of the total supply cost of equipment/pipeline materials bulks and fabrication/installation costs.
- Minimal vendor costs are assumed for the pipelines but allowances are included which would ensure that any items requiring support have been capture.
- An allowance for commissioning spares, first two years' operating spares and strategic spares was included in the estimate as a percentage of the total equipment supply cost. This includes pressure and leak testing of the lines.
- Freight insurance and construction risk insurance have been included as an allowance against the total cost of the estimate.
- Certification and where applicable manufacturer's documentation costs are included for the offsite costs.
- An allowance of 7% was added to the estimate to account for the contractor's profit and fees. This is included in the shown in the summary table as a below the line cost to give a full a complete Total Installed Cost (TIC).
- The following exclusions to the OSBL CAPEX estimate have been made:

Phase 1 Final Report Costed Project Plan

- Taxes and import duties
- Costs associated with Permits, Permissions, License and Legal Fees (covered under owner's costs)
- Project financing, working capital and interest charges, including those resulting from fluctuations in exchange rates
- Forward Escalation

The main element specific assumptions for each element are presented in Table 9-7.

Table 9-7: Offsite CAPEX Summary and Assumptions

Element	Assumptions
Raw Water	Water will be taken from the water mains. It is assumed that a buffer storage tank is required with a holding volume of 60m ³ . A small 3kW pump (10m ³ /h) has been included with the tank. 3 km, 20" buried pipe
Wastewater Disposal	3 km, 20" buried pipe Wastewater is treated inside of the battery limits to a specification suitable for disposal into the municipal sewer system. No additional storage capacity has been included.
Natural Gas Supply	3 km, 8" buried pipe. No buffer storage capacity has been included.
Electrical Power	3 km 132 kV overhead cable The main substation is included in the ISBL estimate
CO ₂ Pipeline	5km, 6" buried pipeline to 3 rd party CO ₂ compression station for offshore storage.
Air Separation Unit	In common with the approach planned for the USA it is assumed that ASU is installed and operated by the technology provided and supplied on a £/tonne basis. This element is included in the OPEX.

9.2.2.3 Contingency

Based upon the conceptual design an allowance for contingency has been included. As the project evolves and the design matures the contingency can be evaluated utilising a more specific risk-based approach. For the Offsite scope the contingency of 40% against the total Offsite scope cost has been assumed. This is greater than the 20% allowance in the CE estimate and reflects the lower level of definition and the inclusion of land acquisition and UK permitting costs.

Phase 1 Final Report Costed Project Plan

9.2.2.4 *Owner's Costs (Petrofac)*

- Petrofac would recommend the inclusion of a 15% allowance on top of the full project to allow for the owner's costs. This would cover costs associated with:
 - Owner's project management team
 - Land purchase
 - Permitting
 - Specialist studies
 - Independent verification
 - Local investment etc..

9.2.3 OPEX

- The elements of the OPEX estimate and the key assumptions made in them are presented in Table 9-8.

-

Table 9-8: OPEX Assumptions

Element	Assumption
Electricity	Electricity consumption was supplied by CE assuming that CO ₂ is exported from the DAC plant at the anticipated operating pressure of the line feeding CO ₂ to Acorn. The cost of power required to boost the pressure to that required for onward transport to the storage location was covered under CO ₂ Transport and Storage. A price of £56.80 £/MWhr, for grid power (Source BEIS GGR Workbook)
Water	A combined water supply and disposal price if £3/tonne has been assumed.
Hydrogen	A price of £41.70 £/MWh for blue hydrogen (Source BEIS GGR Workbook)
Oxygen	At this stage it is assumed that the Air Separation Unit is installed and operated by the technology provider and supplied to the DAC plant on a £/tonne basis. A price was assumed based on data from recent projects.
Variable Costs	This includes: <ul style="list-style-type: none"> • Chemical costs • Waste disposal costs Data supplied by CE
Fixed Costs	This includes:

Phase 1 Final Report Costed Project Plan

	<ul style="list-style-type: none"> • Annual operation and administration costs • Annual Maintenance • Property Damage Insurance <p>Data supplied by CE</p>
CO ₂ Transport and Storage Costs	<p>The cost of transporting and storing captured CO₂ is a complex calculation with a number of variables that are not yet clear. Depending on what assumptions are made in respect of the transport and storage company's installed capacity versus the forecast CO₂ throughput, the cost can vary considerably.</p> <p>The initial capital cost of the pipeline, the subsea equipment, the wells and the long-term store monitoring can be forecast with some confidence. However, the volume of CO₂ going through the system is variable dependant on customers contracted, the nature of the customers business (e.g. a power station might produce CO₂ intermittently), and the need for additional transport and storage capital expenditure to accommodate increased injection volumes.</p> <p>Having reviewed the range of cost projections, and to avoid complicating the Project Dreamcatcher cost predictions, Storegga provided an aggregate value of £40 per tonne for transport and storage for the first three years, increasing to £45.60 per tonne thereafter</p>
General	Consumables are not used while plant is off-line

9.2.4 Decommissioning

The facility is expected to have a life of at least 25 years and potentially much longer. Due the effect of discounting the abandonment costs is not expected to have a material impact of project economics. A simple assumption of 10% of CAPEX is therefore recommended, this equates to 50 million GBP.

Phase 1 Final Report Costed Project Plan

9.2.5 Overall Cost Estimate

- The overall cost estimate is presented in Table 9-9.

Table 9-9: Class 5 Cost Estimate for 0.5 **MTCO₂/yr** DAC Plant, ±50%

Element	Millions GBP
CAPEX	500
ANNUAL OPEX – Yrs 1-3 / Yrs 4+	119 / 122
Decommissioning Allowance	50

10.0 Further Development of the Technology

10.1 Storegga Direct and carbon removal credits

Pale Blue Dot Energy Limited is part of the Storegga Group of companies.

Another group company, known as Storegga Direct, is currently being developed. The purpose of this company is to develop commercial Direct Air Capture plants.

The product of Storegga Direct will be carbon removal credits aimed initially at the emerging market for voluntary carbon credits. Each ton of CO₂ successfully captured and stored creates a carbon removal credit that can be sold to any organisation wishing to offset its emissions.

As an example, Virgin Atlantic is a long-haul airline who have entered into an MOU with Storegga Direct

- They will continue to emit CO₂ from the combustion of jet fuel in their long -haul airline service.
- These emissions will be offset by the purchase of high-quality carbon offsets from Storegga Direct
- This combination makes the Virgin Atlantic flights carbon neutral.

The first Direct Air Capture plant planned by Storegga Direct will be located in the north-east of Scotland near to the Acorn CCS project which will provide carbon storage services.

This NE Scotland DAC plant, which is planned to be the first of multiple Storegga Direct DAC plants in the UK, will generate carbon credits equivalent to 500kt of carbon removed and stored per year. We anticipate that this will be the first commercial DAC plant in the UK.

8.5% of these credits will be retired by Storegga Direct, to offset the life cycle emissions of this DAC plant itself (these 8.5% account for all embedded carbon emissions of the natural gas supply and the construction activities).

10.2 The market for conventional Direct Air Capture

The remaining carbon removal credits will be sold to UK companies in the emerging voluntary carbon removal market. This market is currently being developed through several initiatives:

Phase 1 Final Report Further Development of the Technology

1. A private sector initiative defining government bodies and structures (the *Taskforce for Scaling Voluntary Carbon Markets*) co-led by Mark Carney and Standard Chartered, who presented their interim report at WEF Davos in January 2021.
2. An Oxford University initiative to develop a taxonomy for net zero offsetting credits, who published their first report 2020
3. Forward-thinking corporations, who have begun buying carbon removal credits (with long-term storage) even at remarkably high price points. (e.g. Microsoft's program, which already buys over 1MT of carbon removal annually)
4. The UK Government is consulting on including engineered GGR's into the UK Emissions Trading Scheme.
5. The UK Government is consulting on developing a robust MRV system to manage carbon credits within and outside the UK ETS.

By the end of the operational period (in the 2050s), this market is expected to reach a global size of ~\$50-100 bn, and will consist entirely of carbon removal credits which can guarantee long-term carbon storage

Within this emerging market, the Storegga Direct carbon removal credits will be sold to 4 segments of UK companies:

1. UK emitters, who cannot access carbon abatement below GB£300 / ton at all. This includes long-range transport (mostly aviation and shipping, who require carbon removal until green hydrogen becomes widely available at low cost) and selected industrial players (who cannot only perform post-combustion capture at 80-95% efficiency and have few cost-effective capture technologies to remove the residual carbon).
2. Companies who made net zero commitments which includes their supply chain, and where this supply chain contains difficult-to-abate sectors from the first segment.
3. Forward-thinking companies, who would like to be publicly known as climate leaders and want to be associated with advancing new technologies required to achieve a net-zero world
4. Companies who would like to acquire carbon removal credits in order to re-sell them to their own clients. This includes financial institutions (e.g. banks), who expect the market for voluntary carbon removal to grow to \$50-\$100bn and would like to secure a market position early on. It also includes companies who offer decarbonisation services (engineering or consulting), and who would like to integrate carbon removal credits into this service offering.

Phase 1 Final Report Further Development of the Technology

Storegga Direct has already signed up 3 clients under MoU and will develop binding commercial offtake agreements with these clients over the next 2 years. In addition to these current prospects, Storegga Direct has a pipeline of further leads (not yet at MoU stage). All these current MoU clients and leads are engaging at a price level which does not assume policy support for Greenhouse Gas Removals.

Storegga Direct will develop this client base further during pre-FEED and FEED of the NE Scotland DAC plant. For FID Storegga Direct will require sufficient off-takers with long-term agreements to sanction the project, or strong support from Government for GGR's.

Assuming that BEIS will develop a policy support mechanism for DACS (similar to that being developed for BECCS), Storegga Direct expect prices to decrease from the current level, and client demand to increase dramatically. This increased client demand will be sufficient to FID multiple DAC plants.

10.3 The first Storegga Direct DAC plant

The first proposed Storegga Direct DAC plant in NE Scotland is currently in pre-FEED "Concept Select" engineering.

Initial feasibility studies were concluded in May 2021, and pre-FEED commenced in June 2021. We expect pre-FEED to conclude in June 2022, and FEED to begin thereafter. A final investment decision will be taken mid-2023. After 2.5 years construction, the plant will begin operations in 2026.

The Storegga Direct DAC plant will be designed for a plant life of a 25-years (from 2026 to at least 2051).

The Storegga Direct DAC plant would result in the following captured CO₂ sources across the operational site:

- 500kt of CO₂ annually, captured from atmospheric air.
- 198kt of CO₂ annually, captured from process emissions of the DAC plant itself (arising from the burning of 11,760 GJ of natural gas in the calciner in an oxy-fuel combustion)

Phase 1 Final Report Further Development of the Technology

Apart from these two sources of CO₂ there are no other sources of uncaptured CO₂ across the whole operational site. The estimated 62MW of electricity required to operate the plant will come by a private wire from a renewable provider, and a back-up grid connection for intermittency (specific design will be selected during ongoing pre-FEED work). Some of this electricity is used to manufacture the ~884 tons oxygen / day required for the oxy-fuel combustion of natural gas in the calciner).
CO₂ disposal

The location of the first proposed Storegga Direct DAC plant in NE Scotland is intended to capitalise on the construction and commissioning of the Acorn CCS Project, based at St Fergus.

The Acorn CCS project comprises of

- Onshore compression / pumping equipment to enable the capacity of the export pipeline to be maximised.
- Repurposing of the former Goldeneye natural gas pipeline for CO₂ export purposes. The Goldeneye pipeline is a 20" diameter pipeline laid to transport natural gas from the Goldeneye gas field located primarily within UK licence block 14/29, with extensions into blocks 14/28b, 20/30b and 20/4b. The pipeline is approx.105km long and has recently been inspected and found suitable for repurposing for CO₂ export purposes.
- New subsea infrastructure to distribute the CO₂ from the pipeline to the injection wells.
- New CO₂ injection wells located to optimise injection of the CO₂ into the former Goldeneye gas field structure.

Storegga holds the CO₂ injection licence for the former Goldeneye Field and a number of adjacent storage sites, and is in discussion with the Oil and Gas Authority to secure a CO₂ injection permit specifically for the former Goldeneye Field (now known as Acorn South)

The Acorn storage Project has reserve status under the UK CCS Cluster Sequencing process and is currently in discussion with UK government to ensure that the project proceeds.

10.4 Future growth

Storegga Direct plans to develop additional 500-1,000kt DAC plants within the UK.

The first of these expansion plants will come online in 2030, and the second in 2032. Acorn is expected to serve as storage location for both of these expansion plants.

Phase 1 Final Report Further Development of the Technology

These expansion plants will use existing Carbon Engineering technology and blueprints, which today already form the basis of FEED work on a US facility, and pre-FEED on the first Storegga Direct UK facility (which focuses on adapting the US blueprint to UK climates and utility supplies). Therefore, we do not expect any uncertainty in terms of capture volumes, capture efficiency or capture costs.

For these future expansion plants, capture volumes and efficiencies will correspond to the UK Carbon Engineering facility already in pre-FEED. Capture costs will decrease slightly, thanks to re-using the UK-specific blueprint (currently being developed), and thanks to profiting from a scaled-up UK supply chain.

The major source of uncertainty connected to these expansion plants is the volume of the UK market for voluntary carbon removal credits. Storegga Direct assumes that a policy support mechanism for carbon removal will be developed during 2021 and 2022 (similar to the mechanism already in development for BECCS), and that such a support mechanism would decrease prices for DAC carbon removal credits sufficiently to increase demand to a level where more than one UK DAC plant become viable.

10.4.1 Storegga Direct: overall strategic ambitions to 2030:

By 2030, the first Storegga Direct DAC plant in NE Scotland (location: Peterhead), will have been operational for 4 years, and will have built a strong client base with multi-year carbon removal agreements.

The development experience and customer base from this first DAC plant in the UK (coupled with a policy support mechanism, which we expect to develop during 2021 and 2022) will allow *Storegga Direct* to develop additional UK DAC plants with its exclusive partner Carbon Engineering. By 2030, we expect the first one of these additional 500kt plants to come online, with Acorn its likely storage location. Also by 2030, we expect several more UK plants to be in development.

Phase 1 Final Report Further Development of the Technology

10.4.2 Storegga Direct: overall strategic ambitions beyond 2030:

Beyond 2030, we expect multiple additional Storegga Direct DAC plants to come online in the UK. The first of these post-2030 expansion plants will likely come online in 2032, also utilising Acorn as storage location.

In parallel to this UK expansion, after 2030 Storegga Direct will develop multiple DAC plants in at least one additional country.

In doing so, Storegga Direct and Carbon Engineering will use the leadership in DAC knowledge and supply chain built in the UK to serve this additional market. These UK and international DAC plants will account for several million tonnes of CO₂ removal annually, which will significantly contribute to Storegga Group's global target of removing 50MT CO₂ per year.

10.5 Market for alternative fuel Direct Air Capture

The current Carbon Engineering DAC system uses natural gas to heat the Calciner and captures the resulting CO₂ which is then stored along with the CO₂ captured from the atmosphere. This CO₂ derived from natural gas combustion adds ~40% to the CO₂ captured from the atmosphere. This process results in a low carbon DAC process.

Project Dreamcatcher is developing the Carbon Engineering DAC technology such that the natural gas consumption is eliminated and replaced with hydrogen. This creates two significant opportunities

- It reduces the volume of CO₂ to be stored. As noted in section 10.3, the capture of CO₂ from the air carries a 40% overhead of CO₂ created by the use of natural gas in the calciner.
- It opens the prospect of deploying the Carbon Engineering technology where natural gas is not readily available – using wind energy to manufacture hydrogen to fuel the calciner.

The additional cost of low carbon hydrogen (whether it is “blue” or “green”) is such that we expect the base DAC technology to result in the lowest cost of capture for the foreseeable future.

As a result we anticipate the deployment of a hydrogen-fuelled Carbon Engineering DAC system to be appropriate in specific circumstances where a natural gas supply is not readily available or is not expected to be available for the operating life of the DAC plant.

Phase 1 Final Report Further Development of the Technology

10.6 Timeline

We anticipate the following timeline for the development and deployment of the first commercial Carbon Engineering DAC plant in the UK: -

Timing	Milestone	Comments
2021-mid 2022	Pre-FEED engineering of the first commercial plant (in progress)	Identification and selection of UK supplier for zero-emission calciner unit. Site selection of clean-energy DAC demonstration plant.
2022 – 2023	Front End Engineering and Design (FEED) on commercial-scale DAC plant	Leading to FID in Q3 2023. The main barrier to build is market certainty on the revenue model in order to secure investment.
2025	Commissioning of Acorn carbon capture & storage (CCS) full chain project and offshore transport and storage infrastructure.	An enabler for commercial-scale DAC plant
2024 - 2026	Construction of commercial DAC plant	
2026	Operation of first commercial clean-energy DAC plant	
2026 onwards	Design, construction and operation of further commercial scale DAC plants	Scale up of Storegga Direct business

Phase 1 Final Report Further Development of the Technology

When available	Carbon Engineering is actively working non-natural gas calcination and will deploy hydrogen if and when available.
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10.7 Potential challenges

10.7.1 Technology scale-up

The baseline Carbon Engineering DAC technology using natural gas for calciner heat is ready for deployment at megatonne scale, as evidenced by the US DAC project currently under development with a new company 1PointFive owned by Oxy Low Carbon Ventures, LLC and Rusheen Capital Management.

Carbon Engineering have deliberately selected components with extensive industrial heritage to create their DAC system to minimise the technology risk associated with the system.

As a result we do not anticipate significant technology scale up issues as we deploy the Carbon Engineering DAC technology.

10.7.2 Steps to Commercialisation

We are clear that to be viable the planned Storegga Direct DAC service has to offer both CO₂ capture (the DAC plant) and CO₂ sequestration (via the Acorn project), and that there has to be an economic driver for customers to buy this service.

We have identified some “hard to decarbonise” sectors such as aviation and shipping and agriculture as the primary markets for DAC with sequestration as their alternatives are currently impractical or more expensive (e.g., changing aircraft fuel). Other industries such as cement or iron and steel may also be willing to offset their emissions to using DAC whilst they build new plants or retrofit CCS to transition to reduce carbon emissions themselves. However, such industries are unlikely to act unilaterally as the additional cost of decarbonisation will be commercially unacceptable unless their competitors in other countries face the same costs.

Our view is that the early deployment of DAC will require some form of government action. We anticipate that models such as the contract for difference approach already being considered for CCS, or regulatory approach requiring specific industries to decarbonise, will be necessary to enable us to reach FID on our first DAC plant in late 2023.

Phase 1 Final Report Further Development of the Technology

10.7.3 Business Model Development

The CE business model is to licence the technology for others to deploy.

Storegga Direct is a developer with an ambition to deploy commercial DAC plants, with investment from our shareholders Macquarie Bank, Mitsui, GIC and M&G.

We anticipate that early plants will require government support for both capital and operating costs to enable them to achieve Financial Investment Decision (FID). Over time we would anticipate the emergence of a clear cost for carbon which eventually makes such support unnecessary.

10.7.4 Additional revenue streams

Although we remain open to any opportunity that might arise, our primary plan is to sequester the captured CO₂ in the Acorn CO₂ storage project.

We will consider the potential for creating additional revenue by utilising the captured CO₂ as an input to some other value-adding process. In particular Carbon Engineering has been developing an AIR TO FUELS™ technology over the last 5 years and we will explore this opportunity.

We are also investigating the potential to sell “CO₂ storage credits” to aggregators who might then sell the service on to thousands of consumers and small businesses seeking to decarbonise. There is considerable interest in this service, but until there is an operational DAC system operating there is no credible service to be sold / purchased.

10.7.5 Societal acceptance

Similar to Carbon Capture and Storage (CCS), some environmentalists and segments of society may make a moral argument that the use of natural gas in DAC technology, by proxy, endorses the continued use of hydrocarbons and provides reward, thus incentive for the fossil fuel industry to invest in future developments. Conversely other commentators with a more economic policy view could argue investments in renewable energy infrastructure should be given priority over investments that continue the use of hydrocarbons. Furthermore, there is the debate that centers around the concern that the supply and cost of natural gas in the UK may become too unreliable and expensive. While the scientific and economic basis of these arguments are debatable on their validity and longevity, they are nonetheless potential barriers to the social acceptance of DAC being deployed in the UK.

Phase 1 Final Report Further Development of the Technology

The technical thrust of project dreamcatcher is to provide alternative input energy pathways that eliminate the use of natural gas within the Carbon Engineering DAC process, thus providing configuration options that would address these barriers to adoption. In Phase 1 we have asked SCCS researchers to design a societal attitudes survey work programme to help understand these issues, and to inform on our future public outreach for the DAC development. This work will be undertaken in Phase 2.

In addition, we have identified a possible engineering configuration of the DAC process to take advantage of surplus renewable energy when available, this potential further enhances the integration of a DAC plant into the broader UK energy system. In Phase 1 we have asked the SCCS researchers to design an energy system optimisation work programme, to be undertaken in Phase 2, to explore this opportunity further.

10.7.6 Consents and licences

Deliverable 4 (FEED) involves an allocation of resource to consider the consents and licencing requirements.

We do not anticipate any unusual concerns to be raised in connection with the consents and licencing requirements.

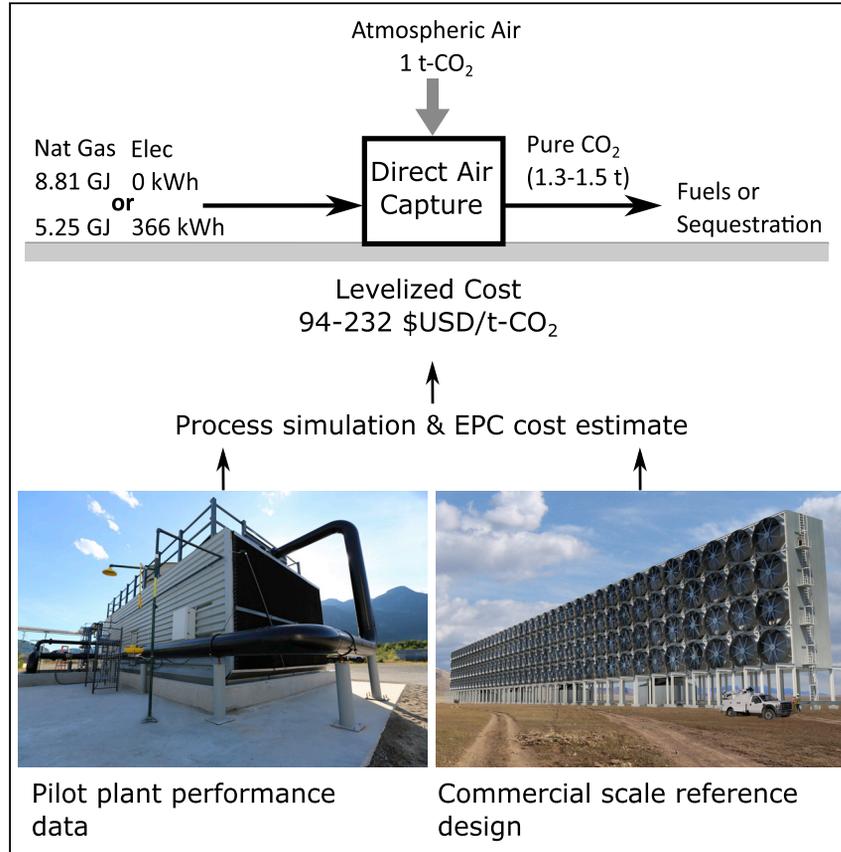
Phase 1 Final Report Attachment 1 – JOULE Article

11.0 Attachment 1 – JOULE Article

The Carbon Engineering Direct Air Capture technology is described in detail in an article entitled “A Process for Capturing CO₂ from the Atmosphere” by David W. Keith, Geoffrey Holmes, David St. Angelo and Kenton Heidel of Carbon Engineering, published in Joule, published 15 August 2018. This is appended below.

Article

A Process for Capturing CO₂ from the Atmosphere



First direct air capture paper for which all major components are either drawn from well-established commercial heritage or described in sufficient detail to allow assessment by third parties. Includes energy and materials balances, commercial engineering cost breakdown, and pilot plant data. When CO₂ is delivered at 15 MPa, the design requires either 8.81 GJ of natural gas, or 5.25 GJ of gas and 366 kWhr of electricity, per ton of CO₂ captured. Levelized cost per t-CO₂ from atmosphere ranges from 94 to 232 \$/t-CO₂.

David W. Keith, Geoffrey Holmes, David St. Angelo, Kenton Heidel

keith@carbonengineering.com

HIGHLIGHTS

Detailed engineering and cost analysis for a 1 Mt-CO₂/year direct air capture plant

Levelized costs of \$94 to \$232 per ton CO₂ from the atmosphere

First DAC paper with commercial engineering cost breakdown

Full mass and energy balance with pilot plant data for each unit operation

Keith et al., Joule 2, 1573–1594

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Article

A Process for Capturing CO₂ from the Atmosphere

David W. Keith,^{1,2,3,4,*} Geoffrey Holmes,³ David St. Angelo,³ and Kenton Heidel³

SUMMARY

We describe a process for capturing CO₂ from the atmosphere in an industrial plant. The design captures ~1 Mt-CO₂/year in a continuous process using an aqueous KOH sorbent coupled to a calcium caustic recovery loop. We describe the design rationale, summarize performance of the major unit operations, and provide a capital cost breakdown developed with an independent consulting engineering firm. We report results from a pilot plant that provides data on performance of the major unit operations. We summarize the energy and material balance computed using an Aspen process simulation. When CO₂ is delivered at 15 MPa, the design requires either 8.81 GJ of natural gas, or 5.25 GJ of gas and 366 kWhr of electricity, per ton of CO₂ captured. Depending on financial assumptions, energy costs, and the specific choice of inputs and outputs, the leveled cost per ton CO₂ captured from the atmosphere ranges from 94 to 232 \$/t-CO₂.

INTRODUCTION

The capture of CO₂ from ambient air was commercialized in the 1950s as a pre-treatment for cryogenic air separation. In the 1960s, capture of CO₂ from air was considered as a feedstock for production of hydrocarbon fuels using mobile nuclear power plants.¹ In the 1990s, Klaus Lackner explored the large-scale capture of CO₂ as a tool for managing climate risk,² now commonly referred to as direct air capture (DAC).

Estimates of the cost of DAC vary widely. Cost estimates based on simple scaling relationships yield results^{3–7} from 50 to 1,000 \$/tCO₂. Uncertainty might be reduced if detailed specifications of individual DAC technologies were available. Yet, despite growing interest in carbon removal as a component of climate strategy, one thorough review,⁸ many papers on DAC-to-CCS (carbon capture and storage) comparison,^{9–13} specific absorbers,^{14–17} or components of plausible DAC systems,¹⁸ no prior paper provides a design and engineering cost basis for a complete DAC system for which all major components are (1) drawn from well-established commercial engineering heritage, or (2) described in sufficient detail to allow assessment by third parties. This paper aims to fill that gap.

Plausible DAC processes¹⁹ use solid sorbents^{20,21} or aqueous basic solutions²² as the capture media. Solid sorbents offer the possibility of low energy input, low operating costs, and applicability across a wide range of scales. The challenges of solid sorbent designs are first, the need to build a very large structure at low cost while allowing the entire structure to be periodically sealed from the ambient air during the regeneration step when temperature, pressure, or humidity must be cycled. And second, the inherently conflicting demands of high sorbent performance, low cost, and long economic life in impure ambient air.

Context & Scale

An industrial process for large-scale capture of atmospheric CO₂ (DAC) serves two roles. First, as a source of CO₂ for making carbon-neutral hydrocarbon fuels, enabling carbon-free energy to be converted into high-energy-density fuels. Solar fuels, for example, may be produced at high-insolation low-cost locations from DAC-CO₂ and electrolytic hydrogen using gas-to-liquids technology enabling decarbonization of difficult-to-electrify sectors such as aviation. And second, DAC with CO₂ sequestration allows carbon removal.

The feasibility of DAC has been disputed, in part, because publications have not provided sufficient engineering detail to allow independent evaluation of costs. We provide an engineering cost basis for a commercial DAC system for which all major components are either drawn from well-established commercial heritage or described in sufficient detail to allow assessment by third parties. This design reflects roughly 100 person-years of development by Carbon Engineering.



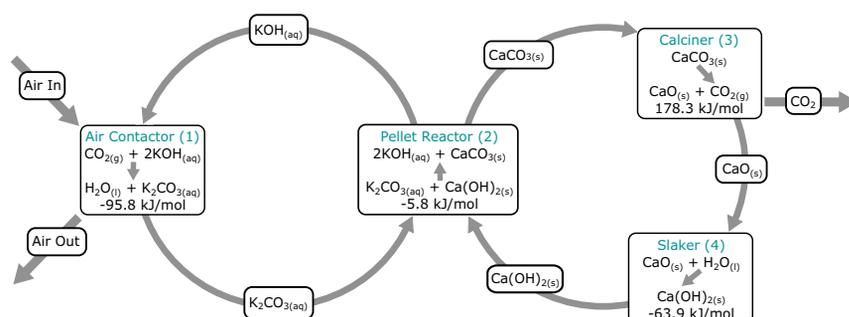


Figure 1. Process Chemistry and Thermodynamics

A calcium loop (right) drives the removal of carbonate ion and thus the regeneration of the alkali capture fluid (left). Boxes with titles show the names of the four most important unit operations. Each box shows the chemical reaction with reaction enthalpy at STP in kilojoules per mole of carbon and the reaction number for reference elsewhere in the paper. Note that water is liberated in reaction 1 and consumed in reaction 4, balancing the process. The full process has evaporative losses, as shown in Figure 2.

Aqueous sorbents offer the advantage that the contactor can operate continuously, can be built using cheap cooling-tower hardware, and the (liquid) surface is continuously renewed allowing very long contactor lifetimes despite dust and atmospheric contaminants. Once captured, CO₂ can be easily pumped to a central regeneration facility allowing economies of scale and avoiding the need to cycle conditions in the inherently large air contactor. Disadvantages of aqueous systems include the cost and complexity of the regeneration system and water loss in dry environments.

Carbon Engineering (CE) has been developing an aqueous DAC system since 2009. In 2012, we described our air-liquid contactor,²³ the front end of the process. Here, in the next section, we provide an end-to-end overview of our baseline DAC system, proceeding from a high-level description of and heat and mass balance down to descriptions of individual unit operations. The following section provides results from a 1 t-CO₂/day pilot plant operated since 2015. CE's capital cost estimating process is described in the section on Process Economics along with the levelized cost of capture under various plant configurations and economic assumptions. Finally, the Discussion provides comparison with prior literature and a discussion of options for improving the technology.

Process Description

Our process comprises two connected chemical loops (Figure 1). The first loop captures CO₂ from the atmosphere using an aqueous solution with ionic concentrations of roughly 1.0 M OH⁻, 0.5 M CO₃²⁻, and 2.0 M K⁺. In the second loop, CO₃²⁻ is precipitated by reaction with Ca²⁺ to form CaCO₃ while the Ca²⁺ is replenished by dissolution of Ca(OH)₂. The CaCO₃ is calcined to liberate CO₂ producing CaO, which is hydrated or "slaked" to produce Ca(OH)₂.

CE has developed a process to implement this cycle at industrial scale. Figure 2 provides a simplified energy and material balance of the complete process (and Figure S1 shows a rendering of one possible configuration of plant equipment to perform this process). At full capacity, this plant captures 0.98 Mt-CO₂/year from the atmosphere and delivers a 1.46 Mt-CO₂/year stream of dry CO₂ at 15 MPa. The additional 0.48 Mt-CO₂/year is produced by on-site combustion of natural gas to meet all plant thermal and electrical requirements. Alternate configurations with electricity and gas input are described in the section on Heat and Mass Balance

¹Harvard School of Engineering and Applied Sciences, 12 Oxford Street, Cambridge, MA 02138, USA

²Harvard Kennedy School, 79 John F. Kennedy Street, Cambridge, MA 02138, USA

³Carbon Engineering, 37321 Galbraith Road, Squamish, BC V8B 0A2, Canada

⁴Lead Contact

*Correspondence: keith@carbonengineering.com

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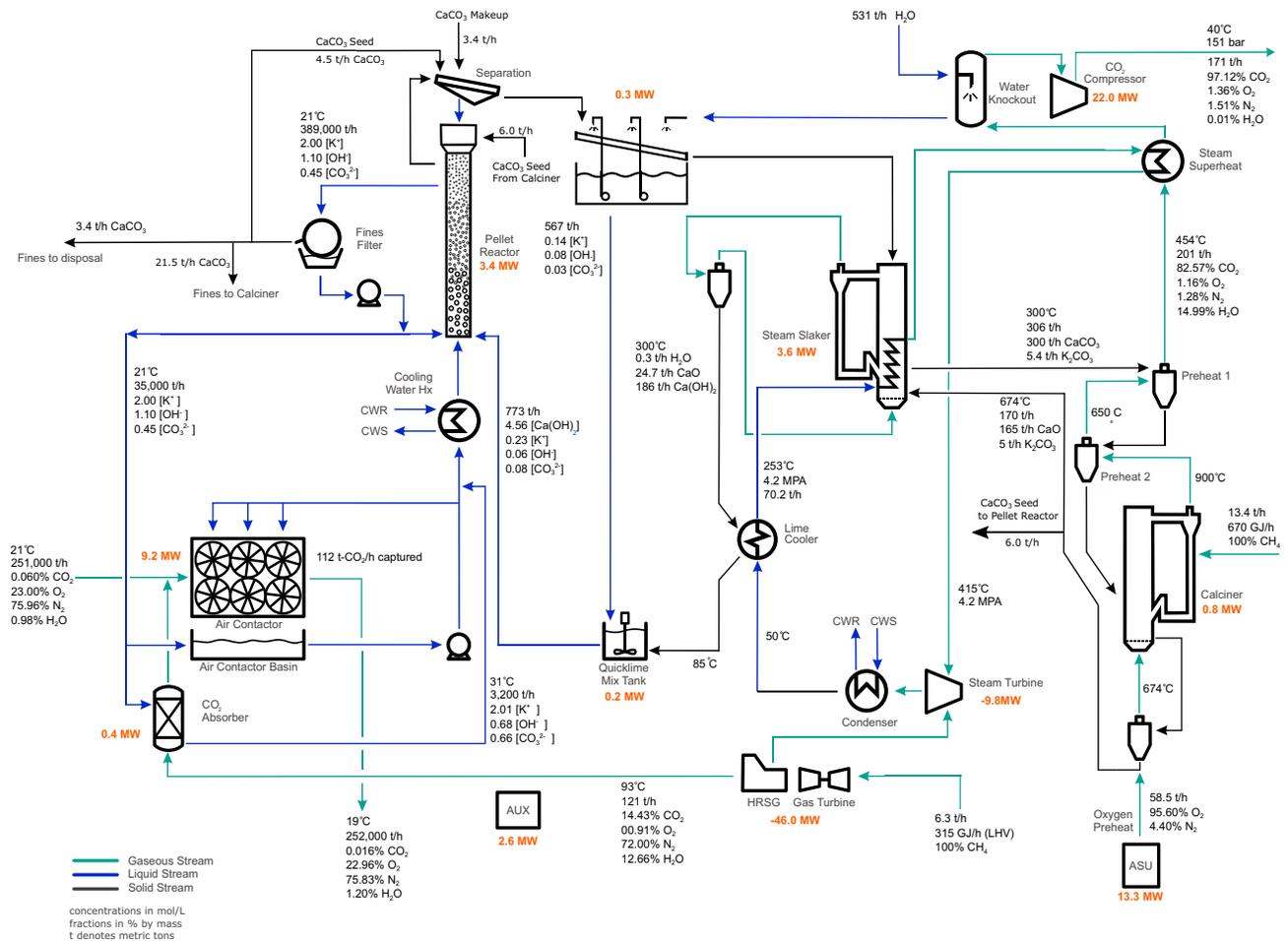


Figure 2. Overview of Process Showing Mass and Energy Balances

Electricity demands are indicated in red as MW. Selected gas and liquid streams show the most important constituents using mass fraction as for gaseous streams and molar concentration for aqueous. Mixed phase streams with substantial solid-phase mass flow are color-coded based on the phase of the gas or liquid transporting the solid. Units are indicated with graphical representations that suggest a schematic physical design of the unit. Many minor streams, such as cooling water to the multistage CO₂ compressor, are not shown. As described in the text, there are several options for introducing the fines stream back into the calciner, these are omitted for simplicity, and this heat and mass balance reflects fines being treated identically to the pellet stream leaving the washer.

and Alternative Configurations and life cycle carbon balance in the section on Avoided Emissions and Life Cycle Accounting.

Energy and material balances come from an Aspen Plus simulation. That simulation depends on performance models of individual unit operations; and these models depend, in turn, on a combination of vendor data and data from the pilot plant described later.

As with any industrial technology, there is a sharp distinction between the ease of developing “paper” designs and the difficulty of developing an operating plant. To paraphrase Rickover: an academic plant is simple, cheap, and uses off-the-shelf components; whereas, a practical plant is complicated, expensive, and “is requiring an immense amount of development on apparently trivial items.”²⁴ CE has now spent roughly 100 person-years on such apparently trivial items to develop a process proposed almost two decades ago by Klaus Lackner and collaborators.²

For each unit, we have either identified a vendor of commercial hardware that meets the process specifications or identified commercial hardware that can be adapted to perform the process. In the latter case, we have typically entered into a formal collaboration with a vendor and then tested the unit at a scale the vendor deems necessary to allow specification of commercial-scale hardware. For the major unit operations, this process has involved several cycles of testing at progressively larger scales working with equipment vendors to de-risk the technology. Consider the pellet reactor as an illustrative example of our development process. The idea originated from a paper that suggested use of a Crystalactor²⁵ developed for wastewater treatment by Royal HaskoningDHV (RHDHV). Working with Procorp Enterprises, RHDHV's American licensee, CE developed a different process configuration.^{26,27} The first tests with CE's process conditions were performed in 2011 using Procorp's existing 5-cm-diameter lab unit. CE then contracted with Procorp to build and operate a larger, 30-cm-diameter custom-built system with more appropriate lime injection technology at Procorp's facility in Waukesha, WI. CE then worked with RHDHV and Procorp to design a 1.2-m-diameter system with up to 11 m of fluidized bed depth as part of CE's Squamish pilot plant. Finally, CE built an additional, smaller 14-cm-diameter system at the pilot plant, to speed up testing of alternative operating conditions that are then implemented on the main pilot pellet reactor.

In this section, we first describe the four major unit operations: the contactor, pellet reactor, calciner, and slaker, corresponding to the four reactions depicted in [Figure 1](#). Performance estimates are based on a combination of data from vendors and from our pilot plant (discussed later), along with data from the minor unit operations (see below). These unit performance estimates then drive a chemical process simulation (see below) that provides the values reported in [Figure 2](#).

Contactor

The contactor brings ambient air in contact with the alkali capture solution. Capture of CO₂ from the air occurs at the surface of a ~50 μm film of solution flowing downward through structured plastic packing through which the air flows horizontally (cross-flow configuration). The transport of CO₂ into the fluid is limited by a reaction-diffusion process occurring in the liquid film with a characteristic e-folding length of ~0.3 μm. The mass transfer coefficient for CO₂ (K_L) is most strongly determined by [OH⁻] and temperature. We use a semi-empirical formula to estimate the mass transfer coefficient on representative well-wetted structured packings (the "effective" K_L) for a range of fluid compositions and ambient temperatures,²⁸ which integrates our own empirical data and modeling and aligns with previous literature values.^{15,25,29-33} A typical K_L is 1.3 mm/s at 20°C and a typical solution composition of 1.0 M OH⁻, 0.5 M CO₃²⁻, and 2.0 M K⁺.

CE's contactor is based on commercial cooling-tower technology, and the design has benefited from close collaboration with SPX Cooling Technologies (SPX), a leading vendor. While the geometry and fluid chemistry differ from conventional cooling towers, CE's design relies on many of the same components, including fans, structured packings, demisters, fluid distribution systems, and fiber-reinforced plastic structural components.

The contactor is the heart of CE's air capture technology. It is the unit that diverges farthest from industrial precedent in that cross-flow cooling-tower components are used for a chemical gas-exchange process, rather than the counterflow vertically oriented tower philosophy typically used for chemical processes. This design choice is a crucial enabler of cost-effective DAC, as designs using vertical packed towers are

Table 1. Summary Data on Major Unit Operations

Parameter	Value	Justification
Contactactor		
Process parameters		
Mass transport coefficient	1.3 mm/s	pilot data and laboratory work ²⁸
Air velocity	1.4 m/s	economic optimization of capital and operating costs ²³
Packing specific surface	210 m ² /m ⁻³	packing parameters are based on Brentwood XF12560 with pressure drop reduced by 30% (see section on the Contactactor)
Packing pressure drop	9.7 Pa/m at 1.4 m/s	
Packing air travel depth	7 m	economic optimization of capital and operating costs ²³
Max liquid flow	4.1 L/m ² s	required for full wetting—manufacturer's specification
Average liquid flow	0.6 L/m ² s	pilot data on flow rate cycling
Performance metrics		
Fan energy	61 kWh/t-CO ₂	ΔP from pilot data and 70% fan efficiency from SPX
Fluid pumping energy	21 kWh/t-CO ₂	pump efficiency 82% from GPSA data book
Fraction of CO ₂ captured	74.5%	performance model validated by pilot data
Capture rate unit inlet area	22 t-CO ₂ m ⁻² /year	determined from velocity and fraction captured assuming 400 ppm CO ₂
Pellet Reactor		
Process parameters		
Fluidization velocity	1.65 cm/s	pilot and benchtop show good performance at 1.65 cm/s for our target pellet size, performance degrades for significantly lower velocities
Bed height	4.5 m	rough optimization of cost of managing fines versus cost of increasing retention; optimization uses empirical performance model driven by pilot data
Calcium loading	20 kg-Ca/m ² hr	
Pellet size	>0.85 mm	pellets removed from bed by passing over a 20 mesh (0.85 mm opening) shaking screen
Performance metrics		
Calcium retention	90%	performance model driven by pilot data
Fluid pumping energy	27 kWh/t-CO ₂	determined from loading rate, fluidization velocity, and pumping efficiency of 82% based on GPSA data
Calciner		
Process parameters		
Bed bulk density	710 kg/m ³	pilot data
Fluidization velocity	0.25–2.5 m/s	minimum and operating fluidization velocity from pilot data
Operating temperature	900°C	reaction thermodynamics and pilot data
Performance metrics		
CaCO ₃ → CaO conversion efficiency	98%	pilot data
Energy consumption	4.05 GJ/t-CO ₂	determined by Aspen Plus simulation in consultation with Technip
Slaker		
Process parameters		
Pellet water carryover	11% by mass	pilot data
Operating temperature	300°C	estimate based on preliminary tests
Performance metrics		
Power produced from slaking heat	77 kWh/t-CO ₂	estimate from simulation, note that the slaker also consumes 32 kWh/t-CO ₂
Conversion to CaO	85%	estimate based on tests conducted by Ben Anthony at CanmetENERGY
Auxiliary Equipment Specifications		
ASU power usage	238 kWh/t-O ₂	quote from major ASU vendor for 95% purity delivered at 120 kPa
CO ₂ absorber—capture frac	90%	Aspen simulation

(Continued on next page)

Table 1. Continued

Parameter	Value	Justification
CO ₂ absorber—pressure drop	1 kPa	
Compressor power usage	132 kWh/t-CO ₂	Aspen simulation, with validation from independent calculations

For each major unit we provide some important process parameters internal to the unit as well as the most important unit performance parameters. Energy consumption values are given for each ton of CO₂ processed by the unit where for calciner, slaker, and compressor, the amount processed is larger than the amount captured from air because of the CO₂ from the power cycle.

far more expensive.^{4,14,25} In this paper, we provide only a short overview because the design is described elsewhere.²³ Major differences between our design and common cooling-tower practice include packing depths of ~ 7 m rather than the ~ 2 – 3 m common in cooling towers with structured packing, and use of cyclic-pulsing solution flow to minimize pumping energy while maintaining good packing wetting.³⁴ The air velocity and packing depth are chosen to minimize combined capital and energy cost,²³ and the resulting design parameters summarized in Table 1.

Working with packing manufacturers and using computational fluid dynamics simulations performed by Professor John Grace's group at the University of British Columbia (UBC), we find that minor changes to packing geometry can significantly reduce pressure drop while retaining similar mass transfer performance. Pressure drop can be reduced by $>30\%$ compared with the Brentwood XF12560 packing we used in the pilot for the same air velocity and surface area density. Improvements on established designs are possible because we are optimizing for different conditions: CO₂ uptake differs from the evaporation and sensible heat exchange in a cooling tower, as does our use of pulsed flow to maintain a largely stagnant surface film. Indeed, changes in the tradeoffs between fan energy and capital cost alter the optimal design.²³ Here, we assume that packing in a commercial plant would have a pressure drop 30% lower than XF12560.

Pellet Reactor

Carbonate ion is removed from solution by causticization in the pellet reactor (reaction 2). In this fluidized bed reactor, 0.1–0.9-mm-diameter CaCO₃ pellets are suspended in solution that flows upward at ~ 1.1 – 2.5 cm/s. A slurry of 30% Ca(OH)₂ is injected into the bottom of the reactor vessel (where here and throughout slurry compositions are mass fractions). As Ca²⁺ reacts with CO₃²⁻ it drives dissolution of Ca(OH)₂ and precipitation of CaCO₃, but the fraction of Ca²⁺ that is precipitated onto pellets depends on maintaining a high surface area of pellets relative to the area of circulating fines while minimizing localized high supersaturations of CaCO₃ that form fines. Small seed pellets are added at the top of the bed, and as pellets grow they sink through the reactor until finished pellets are discharged at the bottom. Roughly 10% of the Ca leaves the vessel as fines that must be captured in a downstream filter. The finished pellets are roughly spherical agglomerations of calcite crystals with negligible porosity.

This process is adapted from water treatment technology developed by RHDHV, where it is used to remove multi-valent ions such as CO₃²⁻. The process was reengineered to allow formation of CaCO₃ pellets in high ionic strength solutions. Our process differs from water treatment in that (1) the causticization agent is the limiting reagent, (2) it uses 30% lime slurry rather than the $\sim 2\%$ slurries used in water treatment, and (3) the process parameters are optimized to maximize caustic flux per unit bed area rather than water flux.

As described above and in the section on the Pilot Plant, our process was developed iteratively using several generations of prototypes. The industrial design draws on

RHDHV's experience in engineering and operating large-scale wastewater treatment plants. The high-concentration lime slurry required abandoning the standard dosing racks and adopting a Spiractor configuration,³⁵ but unlike a free-standing Spiractor vessel, conical feed sections form an egg-carton-like bottom for a large concrete reactors. Similar systems have been used at the Groote Lucht Wastewater Plant and at Bahrain Tubli Wastewater Plants. The Bahrain plant, for example, has a flow rate of 66 m³/s providing a solid basis for cost estimates on our plant, which has a flow rate of 166 m³/s.

Our choice of pellet reactor and oxy-fired circulating fluidized bed (CFB) are at the heart of the innovations that reduce the capital and energy cost of the DAC process compared with use of a Kraft caustic recovery loop, which accomplishes the same chemical process in Figure 1. Early process development work at CE and elsewhere considered using the Kraft process followed by a separate CO₂ capture process on the Kraft kiln off-gases as the minimum-risk baseline technology for an aqueous alkaline DAC process.^{4,22,36} (Kraft processes use a Na while our process uses a K to improve CO₂ uptake kinetics.) Performance gains come from the ability to make pellets, rather than "lime mud," which is composed of precipitated 10–50- μ m-diameter calcite crystals. The pellets are washed and dried easily, removing the need for vacuum filtration, and resulting in pellets that are drier and have lower alkali carryover than lime mud, which in turn allows use of a CFB rather than a rotary kiln. The lack of vacuum filtration and low water carryover reduces energy consumption in the kiln. Moreover, the CFB has lower capital cost than a rotary kiln and it can be oxy-fired.

Calciner

Calcination of CaCO₃ to produce CO₂ (reaction 3) is accomplished in an oxygen-fired CFB. Our design has been developed in collaboration with Technip's Dorr-Oliver Fluosolids Systems Division from initial design through laboratory testing, CE's pilot plant, and design of the commercial-scale calciner. Technip has deployed high-temperature fluidized beds at comparable scales, including, for example, two 6.7-m-diameter oxygen-blown CFBs used as gold ore roasters in the Goldstrike mine in Nevada.

The calciner, along with preheat cyclones, are large steel vessels lined internally with refractory brick. Fluidizing gas is supplied into the bottom of the calciner through a distribution plate made from an arch of refractory, and natural gas is injected directly into the fluidized bed just above the distribution plate using a series of lances.

Our conservative heat integration design reduces technical risk compared with alternate designs that maximize energy efficiency at the expense of higher capital cost and technical risk. Incoming pellets, which arrive from the steam slaker at 300°C, pass through two heat recovery cyclones arranged in counter-current configuration with the outgoing gas stream. In the first preheat cyclone, the incoming solids are preheated to 450°C by cooling the outgoing gases from 650°C to 450°C. In the second preheat stage, the solids are further heated to 650°C by cooling the gases from 900°C to 650°C. Following the cyclones, the outgoing gas stream drives a steam superheater, further cooling the gases to 325°C and producing steam for power generation. The outgoing CaO from the calciner is cooled to 674°C in a single cyclone, which in turn preheats the incoming oxygen to the same temperature before the CaO is sent to the steam slaker.

The calciner operates at ambient pressure. Leakage of nitrogen into the system is minimized by using steam-fluidized loop seals at the inlet to the steam slaker and

between the steam slaker and the calciner. Experience with similar dual fluidized bed systems with intermediate loop seals³⁷ suggest that in a worst-case scenario, the steam slaker atmosphere would contain 0.4% air contributing 0.0013% N₂ (by volume) to the calciner off-gases.

As our process is, in some respects, derived from the Kraft pulp process, it is useful to compare this calciner with the rotary kilns used in the Kraft process. A single 6-m-diameter CFB of this design will have a capacity of 2 kt-CaO/day. A typical large Andritz rotary kiln calcining lime mud is 5.5 m diameter × 165 m long and produces 1.6 kt-CaO/day.

Process parameters are summarized in [Table 1](#). The minimum required energy to drive the reaction is 3.18 GJ/t-CaO. Our design requires 4.07 GJ/t-CaO equivalent to 5.25 GJ/t-CO₂ (that is, CO₂ from calcination) to make up for thermal inefficiencies in heating the feed streams and heat losses to ambient air. This makes the calciner approximately 78% thermally efficient, substantially higher than lime mud calciners, which have thermal efficiencies of roughly 39%, though less efficient than limestone calciners, such as the Cimprogetti TWIN-D shaft kilns, which are 89% efficient on the same basis. Our calciner choice and its efficiency are enabled because pellets are easy to dewater and have appropriate fluidization properties.

Steam Slaker

Heat from slaking (reaction 4) is used to dry and preheat the pellets, yielding sufficient steam to sustain the slaking reaction.³³ The thermodynamic advantage of steam slaking over conventional water slaking used in the Kraft process is that the slaking reaction enthalpy is released at higher temperatures. Maximum temperature is 520°C for slaking in 100-kPa steam, whereas we operate at 300°C to achieve fast kinetics.

Designed in partnership with Technip, the slaker is a refractory lined bubbling/turbulent fluid bed that is fluidized by recirculating steam flow.³⁸ It receives ambient temperature CaCO₃ pellets from washing and hot CaO at 674°C from the calciner's oxygen preheat cyclone. Fluidization velocity is 1 m/s, which transports and slakes quicklime (CaO) particles to form Ca(OH)₂. Small quicklime particles are elutriated and recirculated by a primary cyclone and loop seal, while the much finer slaked particles mostly bypass the cyclone and are captured in a dust collector. The outgoing stream is 300°C hydrated lime, from which sensible heat is recovered and, along with heat from the slaking reaction, used to dry and warm the pellets. Dry 300°C pellets are then fed into a closed-loop pneumatic conveyor driven by recirculating CO₂ and steam to deliver them to the top of the calciner stack.

Minor Unit Operations

Beyond the four major units described above, the plant requires many additional unit operations that are "minor" in the sense that they present little or no technical risk. This section summarizes configuration of the power island along with the absorber that captures CO₂ generated from power production, the CO₂ compression and cleanup, and the oxygen plant. Key performance characteristics for each of these units are provided in [Table 1](#).

Power Island. The power island consists of a natural gas turbine, followed by a heat recovery steam generator (HRSG). We model a GE LM 2500 DLE with a 2 × 1 HRSG configuration. The resulting steam is combined with steam from the slaker, passed through the superheater (to extract heat from the calciner off-gases), and then

Table 2. Summary Performance of Various Plant Configurations

Scenario	Gas Input ^a (GJ/t-CO ₂)	Electricity Input ^a (kWh/t-CO ₂)	C-Gas/C-Air	Capital \$ per t-CO ₂ /year	O&M ^b (\$/t-CO ₂)	Levelized ^a (\$/t-CO ₂)	
						CRF ^c	
						7.5%	12.5%
A: Baseline: gas fired → 15 MPa CO ₂ output	8.81	0	0.48	1,146	42	168	232
B: Baseline with N th plant financials	8.81	0	0.48	793	30	126	170
C: Gas and electricity input → 15 MPa CO ₂ output	5.25	366	0.30	694	26	113–124	152–163
D: Gas and electricity input → 0.1 MPa CO ₂ output assuming zero cost O ₂	5.25	77	0.30	609	23	94–97	128–130

^aGas and electrical inputs as well as levelized cost are all per ton CO₂ capture from the atmosphere.

^bNon-energy O&M expressed as fixed per unit of capacity with variable costs including cost of make-up streams included and converted equivalent fixed costs using 90% utilization.

^cCRF is the average capital recovery factor defined in the section on Process Economics. Calculations assume NG at 3.5 \$/GJ and a 90% utilization. For the C and D variants levelized costs are shown as a range using electricity at 30 and 60 \$/MWhr.

used to drive a steam turbine that generates the remainder of the power required by the plant. To reduce complexity, our Aspen simulation approximates this using independent steam cycles for the gas turbine and slaker/superheater. After heat recovery, gas turbine exhaust is sent to the CO₂ absorber.

CO₂ Absorber. The gas turbine exhaust stream is stripped of CO₂ using a conventional counterflow gas-liquid column, using a portion of the fluid stream returning from the contactor. Based on rough optimization using Aspen, we chose a 12 × 7.5 m (height × diameter) column filled with 95 m²/m³ BERL Ceramic packing that captures ~90% of inlet CO₂ with a pressure drop of 1.08 kPa at an average operating gas velocity of 0.75 m/s. The absorber outlet is ducted to main air contactor where ~75% of the remaining CO₂ is captured.

CO₂ Compression and Cleanup. Compression is accomplished using a standard centrifugal compressor. Performance and power demand were simulated on a four-stage centrifugal compressor based on Dresser-Rand data, with a glycol system for dehydration prior to the final stage, and going from atmospheric to 15 MPa at 45°C. The compressor cost estimate included inter-stage coolers and scrubbers, and cost of equipment was generated by Aspen's Capital Cost Estimator and compared with previous vendor estimates.

Oxygen Plant. We use a conventional cryogenic air separation unit (ASU). Large ASUs are available in multi-train complexes that produce over 30 kt-O₂/day. Cryogenic ASUs typically produce oxygen up to 99.8% purity and 10 MPa. Cost was estimated by Solaris (see the section on Process Economics) based on a vendor quote for a 1.5 kt-O₂/day system. Power demand of 238 kWh/t-O₂ for a 120 kPa delivery pressure was estimated by a major ASU vendor.

Heat and Mass Balance and Alternative Configurations

The plant's simplified heat, mass, and power balance are shown Figure 2, with energy inputs summarized in Table 2. At ambient conditions of 20°C and 64% relative humidity, the plant needs 4.7 tons of water per ton CO₂ captured from the atmosphere. This ratio varies with ambient conditions and solution molarity; this relationship is shown in Figure 3F, which was calculated with Aspen data and validated with CE's pilot air contactors. The plant discharges 1% of the circulating Ca each cycle as waste, this discharge serves as a purge that manages the buildup of non-process elements that

enter the cycle by various routes, most importantly as dust ingested into the contactor. At this discharge rate, CE estimates that Ca disposal and make-up contribute a cost of \$0.22/ton-CO₂ to the overall totals presented in this paper.

Process Simulation. Plant performance was computed in Aspen Plus V8.0. We use the ENRTL-RK and RK-SOAVE thermodynamic property packages for aqueous phase and gaseous phase respectively. We used the following APV732 property databanks: ASPENPCD, AQUEOUS, SOLIDS, INORGANIC, and PURE26. Gas solubility and the precipitation of salts were specified manually. Chemical loops were converged in Aspen's sequential modular mode. Many individual unit parameters, including much of the data from [Table 1](#) are computed in a set of linked external spreadsheets.

Scaling. The performance of any large industrial process depends on scale. Both the air contactor and the pellet reactor are modular, so their performance varies little from 1 Mt-CO₂/year down to sizes as small as 10 kt-CO₂/year, and their capital cost per unit capacity is nearly constant down to 100 kt-CO₂/year. In contrast, the calciner is a large refractory lined vessel with complex equipment for thermal integration, which results in both performance and cost scaling strongly with size. This calciner design is appropriate down to an internal diameter of about 1 m corresponding to a capture rate of 15 kt-CO₂/year, but CE judges that—given cost scaling—the smallest economically practical size for the complete process is about 100 kt-CO₂/year. At that scale, for the full DAC process, the energy intensity would very close to the 1 Mt-CO₂/year baseline, but the capital cost per unit of capacity would be ~80% larger.

Alternative Configurations. CE is developing various plant configurations to address specific markets. These configurations all share the four core unit operations described above, but vary in their treatment of power system, oxygen supply, and CO₂ compression.

The baseline plant configuration, "A" on [Table 2](#), is applicable to geologic storage in locations with comparatively low natural gas prices, so this scenario delivers CO₂ at specifications appropriate for pipelines. While actual plants will be grid connected, for convenient analysis we have sized the power system in this baseline configuration so that the facility is electrically neutral with no net power input or output.

An Nth plant variant with the same configuration, "B," is included to reflect improvement in capital and construction costs that vendors and engineering, procurement, and construction (EPC) firms have indicated would be realized following early plant builds as we improve constructability and build supply chain relationships.

We also report two additional process variations upon the baseline. A variant with minimum gas input, "C," has no gas turbine and uses grid electricity to make up all power not supplied by the steam cycle running off the steam slaker. Overall cost and energy requirements are summarized in [Table 2](#). Note that this variant requires a few minor process alterations not shown in [Figure 2](#). This variant is appropriate for locations with low-carbon-intensity low-cost power.

Finally, variant "D" is optimized to provide CO₂ for fuel synthesis. CE is developing air-to-fuel systems in which the hydrogen required as feedstock for the fuel synthesis step is produced by electrolysis.³⁹ In this configuration, the oxygen from electrolysis is sufficient to supply the DAC plant, so in this application we drop the ASU from the DAC process. The fuel synthesis system requires a CO₂ supply pressure of ~3 MPa, reducing the cost and complexity of the CO₂ compression and clean up. CE is

developing methods to integrate the DAC and fuel synthesis, but for simplicity of analysis, here we show (Table 2) the inputs for a plant that receives O₂ and produces atmospheric pressure CO₂.

Pilot Plant

The design or operating conditions of several of the important unit operations in our process differ sufficiently from their industrial counterparts, and as such require process optimization and testing. To address this need as well as the overall integration risk, CE has operated a pilot plant on a 0.5-hectare industrial site in Squamish, BC, since 2015.

The design goal for the pilot were (1) to test each unit operation for which there is significant technical risk at a scale the equipment supplier judged sufficient to allow specification of commercial-scale hardware, and (2) to test the most important units as components of a closed-loop process. The pilot plant builds on previous prototype data that CE acquired for each unit, and on work with SPX, RHDHV, and Technip to design and size the contactor, pellet reactor, and calciner, respectively. CE's pilot data have been used to refine the commercial-scale plant design described earlier.

The pilot is not a complete small-scale version of a commercial plant, as units that were judged low risk are not included. For example, the pilot does not include gas clean up and CO₂ compression downstream of the calciner, as these processes present minimal technical risk. The air contactor and pellet reactor operate as a coupled loop with a capacity of 0.6 t/day of CO₂ captured from the air, a scale at which each unit is large enough for validation of commercial-scale hardware. The minimum scale for the calciner corresponded to roughly double this capacity, so while the contactor, pellet reactor, and slaker can operate continuously, the pellets are accumulated in a storage silo and then the calciner is run in batch mode to process the accumulated pellets and produce lime. Finally, the slaker was chosen to close the overall chemical loop and to produce lime slurry in a stirred tank reactor, rather than in a steam slaker, as discussed earlier.

Pilot Contactor

The pilot contactor tests the performance of CE's cooling-tower-derived packing, drift elimination, and fluid distribution systems. The structure is modified from an SPX commercial unit in which air flows inward through two banks of structured packing and enters a central plenum where it is ejected upward through a vertical-axis fan. Each packing bank has a 3 × 5-m inlet cross section with a 3-m depth of Brentwood XF12560 structured packing. (This depth, although sufficient for pilot testing, is smaller than in the full-scale air contactor design, and thus the pilot capture fraction is lower than at commercial scale.) Standard SPX construction and fluid distribution techniques are used, including extensive use of low-cost fiberglass-reinforced plastic. While the overall geometry differs slightly from CE's single-bank commercial design, the pilot tests the packing and distribution systems as well as the construction methods and materials choices at scales that imply minimal further scale-up risk given the inherently modular nature of the full-scale design.²³ At an inlet velocity of 1.4 m/s the contactor ingests air at 180 t/hr, yielding a 45 kg-CO₂/hr maximum capture rate at 42% capture fraction.

Figure 3 shows selected pilot plant air contactor data, and Figure S2 shows images of CE's pilot air contactor and industrial analogs. Pressure drop closely matches specified performance (Figure 3A) and is stable over ~0.75 year, demonstrating minimal long-duration fouling. Pressure drop increases with liquid flow rate as seen in Figures 3B and 3D, and the sequence of small flow pulses is designed to reduce fluid and air pumping energy. Large pulses are used occasionally to flood the packing, ensuring

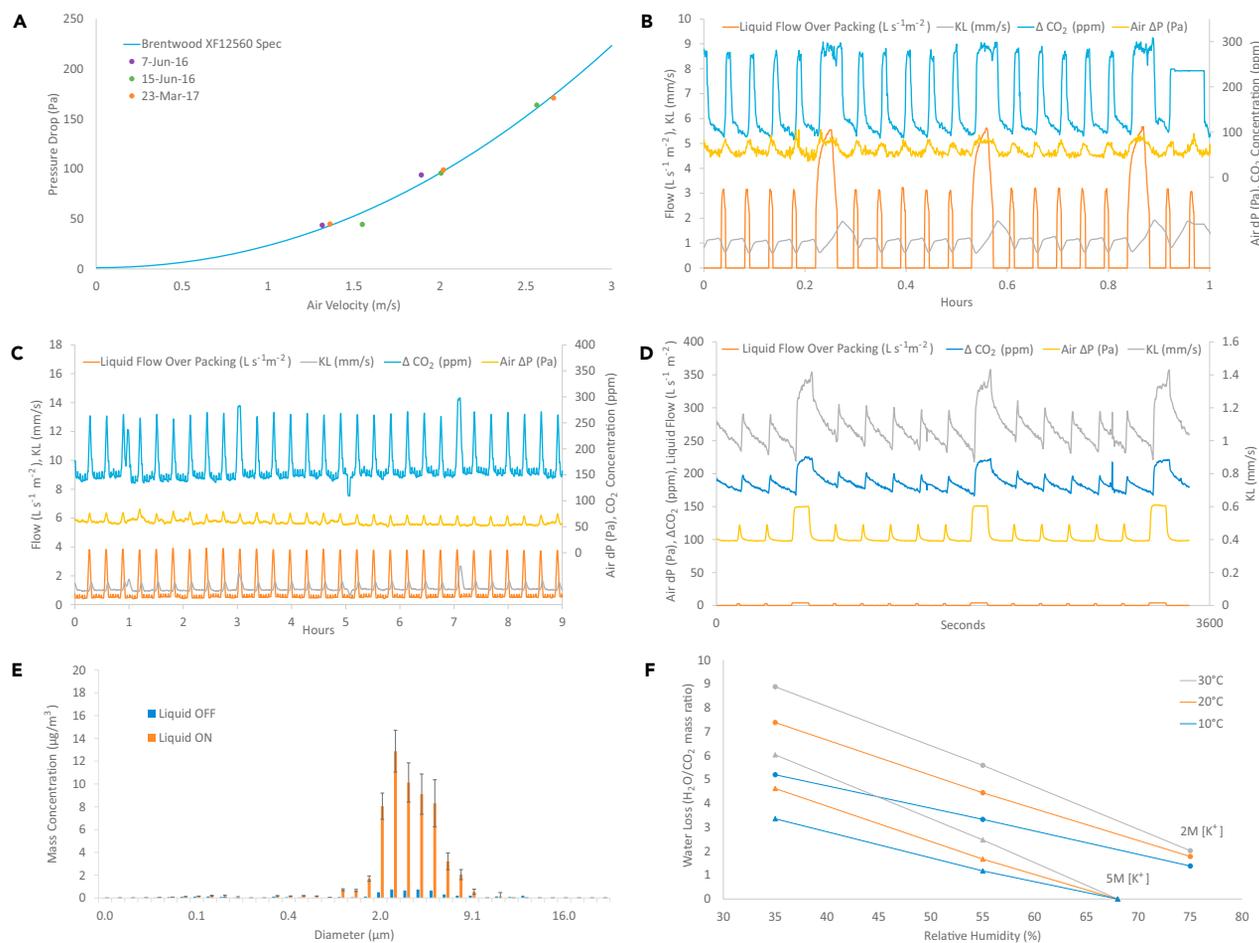


Figure 3. Selected Data from Pilot Contactors

(A) Pressure drop through dry packing and drift eliminator.

(B) Time series pilot contactor operation showing alternation of large and small packing fluid refresh cycles, data are from July 2016 with average air flow velocity of 1.17 m/s at 18 C ambient temperature.

(C) Eight hours of data showing same variables with same color key as in (B).

(D) Time series from a prior CE contactor pilot in 2012 at average air flow velocity of 1.45 m/s showing similar cycling at using a 2.6 M $[\text{OH}^-]$ solution.

(E) Particle size distribution measured at contactor outflow showing contrast between drift with liquid on and off. Values are total mass per unit volume in size bins with centers indicated on the x axis, and error bars are standard deviation across five measurements.

(F) Calculated water loss from evaporation in the air contactor as a function of temperature, relative humidity, and total capture solution molarity.

complete wetting. Using observed air velocity and fraction of CO_2 removed, along with packing depth and surface area density, we use equation 2.1 of Holmes and Keith²³ to derive an “effective” mass transfer coefficient, $K_{L\text{-eff}}$, that is the product of K_L and the surface wetting efficiency. Pressure drop, mass transfer, and drift measurements are all consistent with 2012 results from a previous prototype air contactor⁴⁰ and from laboratory data.²⁸ Note that due to challenges of getting an accurate spatial average of ΔCO_2 measurements across the contactor outflow, we use data from a representative location scaled to match the time-integrated changes in liquid chemistry that provide the most accurate long-term uptake measure.

One potential risk of using cross-flow cooling-tower components is that droplets of contactor fluid can escape into the ambient air, posing a health hazard. The cooling-tower

industry calls such fluid loss “drift.” CE has measured total drift concentrations using standard air quality monitoring vacuum filters followed by quantitative measurement of the alkali content of the filter paper. In addition, we collaborated with Professor Steven Rogak’s group at UBC. They used a TSI NanoScan SMPS optical particle sizer and aerodynamic particle sizing instruments to produce the size distribution data shown [Figure 3E](#). All measurement techniques have shown airborne KOH concentrations less than 0.6 mg m^{-3} of air at contactor outflow, which is well below the 2.0-mg m^{-3} National Institute for Occupational Safety and Health (NIOSH) ceiling recommended exposure limit.⁴¹ This drift performance is consistent with specified performance for the Marley XCEL TU cellular drift eliminator used in our design.

Performance depends on ambient conditions. Water loss is determined by ambient temperature, relative humidity, and molarity of the capture solution ([Figure 3F](#)). K_L is affected by temperature but not significantly by humidity.²³ Fluctuations in K_L caused by diurnal or seasonal temperature changes are material and can be partially compensated for by adjusting the gas and liquid throughputs.

Pilot Pellet Reactor

Our pilot is a steel vessel with internal height of 12 m, a 1.2-m diameter, and a 60° conical base, similar to a single cell of the commercial design. Ancillary equipment allows automated addition of seeds, extraction and processing of fines, and washing/drying of mature pellets. It was designed in collaboration with RHDHV based on a numerical model and on results from tests in 5- and 30-cm-diameter systems by Procorp.

[Figure 4](#) shows selected pilot plant pellet reactor data, and [Figure S3](#) shows images of CE’s pilot pellet reactor and industrial analogs. The initial growth of a mature bed from seeds takes several months under typical process conditions ([Figure 4D](#)), and the cycle from pellet discharge and seed addition to recovery of the bed density profile is about 2 days ([Figure 4A](#)). In addition to the primary pilot, we built two $0.1 \times 5.2\text{-m}$ “benchtop” pellet reactors,¹⁸ to allow testing under a larger number of process conditions and to test in conditions that might cause bed collapse and reactor plugging, which would have been difficult to manage in the primary pilot.

Reactor performance depends on bed height, Ca loading rate, fluidization velocity, pellet size at bed base, and the circulating concentration of fine calcite particles ($<50 \text{ }\mu\text{m}$) that provide nucleation points for calcite precipitation. Our overall objective is to minimize the energy and capital cost of the reactor while maintaining a retention rate above $\sim 85\%$, where retention rate is the fraction of injected Ca that leaves the bed as pellets rather than being lost as fines. The pumping energy per unit of Ca flux is proportional to the fluidization velocity and bed height and inversely proportional to the Ca loading rate.

[Figure 4B](#) illustrates the impact of fines processing on retention, and much of the design optimization amounts to adjusting other parameters to balance minimizing fines production with minimizing energy cost.

Our initial design assumed a benefit to extending the bed height over 10 m and relied on loading rates of over $40 \text{ kg-Ca m}^{-2}/\text{hr}$ to get acceptable energy performance. However, we found no increase in retention for bed heights greater than 4.5 m, probably because dissolution-precipitation kinetics were faster than anticipated in the numerical model. Our current design point is a $20 \text{ kg-Ca m}^{-2}/\text{hr}$ loading rate with a 4.5-m bed height ([Table 1](#)). We anticipate small but significant

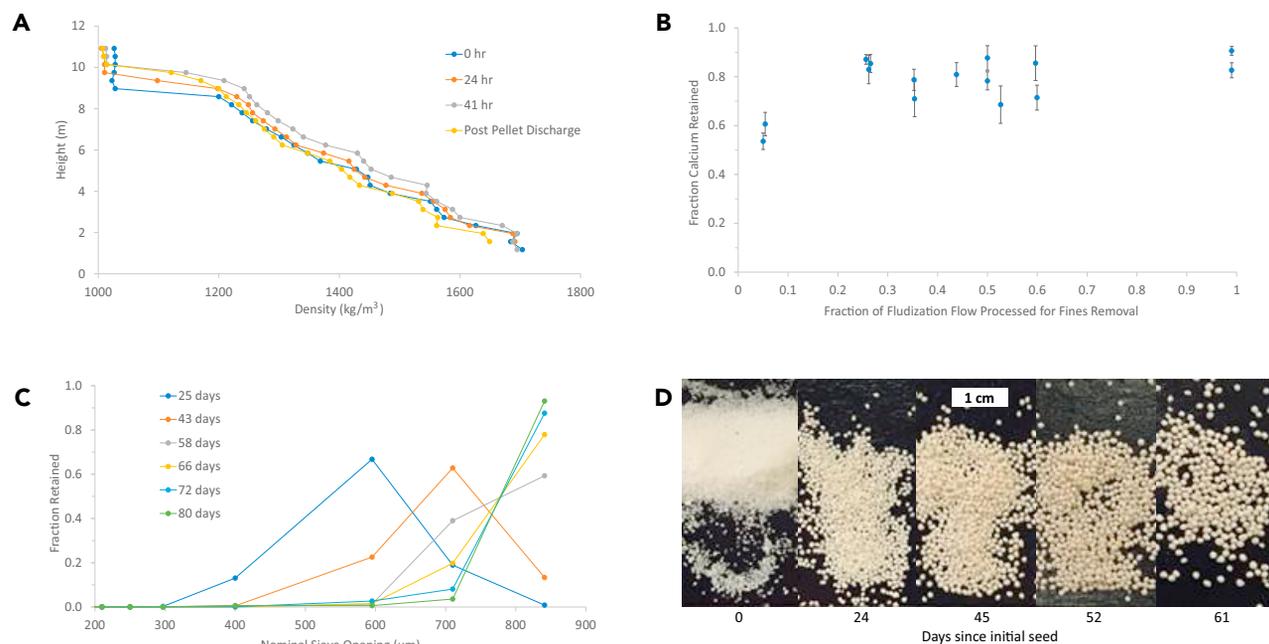


Figure 4. Selected Data from Pilot Pellet Reactors

- (A) Pellet reactor density profile during operation and post pellet discharge. Density of slurry at each point within the reactor determined using the apparent immersed weight of an object of known weight and volume.
- (B) Calculated calcium retention as a function of the fraction of recirculating flow processed to remove inert “fines” (calcium carbonate) in the benchtop pellet reactor.
- (C) Size distribution of sample from bottom of pellet reactor during bed development. Time evolution in days, to show growth of seed material to mature pellet bed measured by passage through calibrated filter screens.
- (D) Photos illustrating time evolution of CaCO₃ pellets sampled from bottom of pellet reactor bed.

improvements in energy trade-off between energy requirement and retention as we continue to adjust process parameters.

Pilot Calciner

The pilot calciner was designed in close collaboration with Technip following their common practice, which uses data from a specially designed 0.15-m interior-diameter calciner to provide accurate performance predictions for a commercial-scale system. Technip’s design uses a steel vessel with an externally heated jacket that effectively eliminates heat loss through the vessel walls and allows the small high surface-to-volume ratio testbed to mimic the bed behavior and in-bed heat transport that would be present in a large refractory lined commercial-scale calciner. Jacket heating is accomplished with a gas-fired external heater and additional electric heaters to manage cold spots.

Hazen Research (Golden, CO) did initial high-temperature testing of CE’s pellets, and then CE worked with Hazen, and BC Research Inc. (Richmond, BC) to design, procure, and commission the calciner to meet Technip’s test specifications. The pilot is an oxy-fired circulating fluid bed calciner with an 8.5-m-tall riser. It differs from the commercial configuration in that it uses a fluoseal for solids recirculation but not for solids discharge, does not include preheat for the pellets or oxygen, and uses an air quench and baghouse prior to venting to manage off-gases.

The important performance goal was to achieve a high material flux and high calcination, in conditions where reaction enthalpy is driven by in-bed combustion using

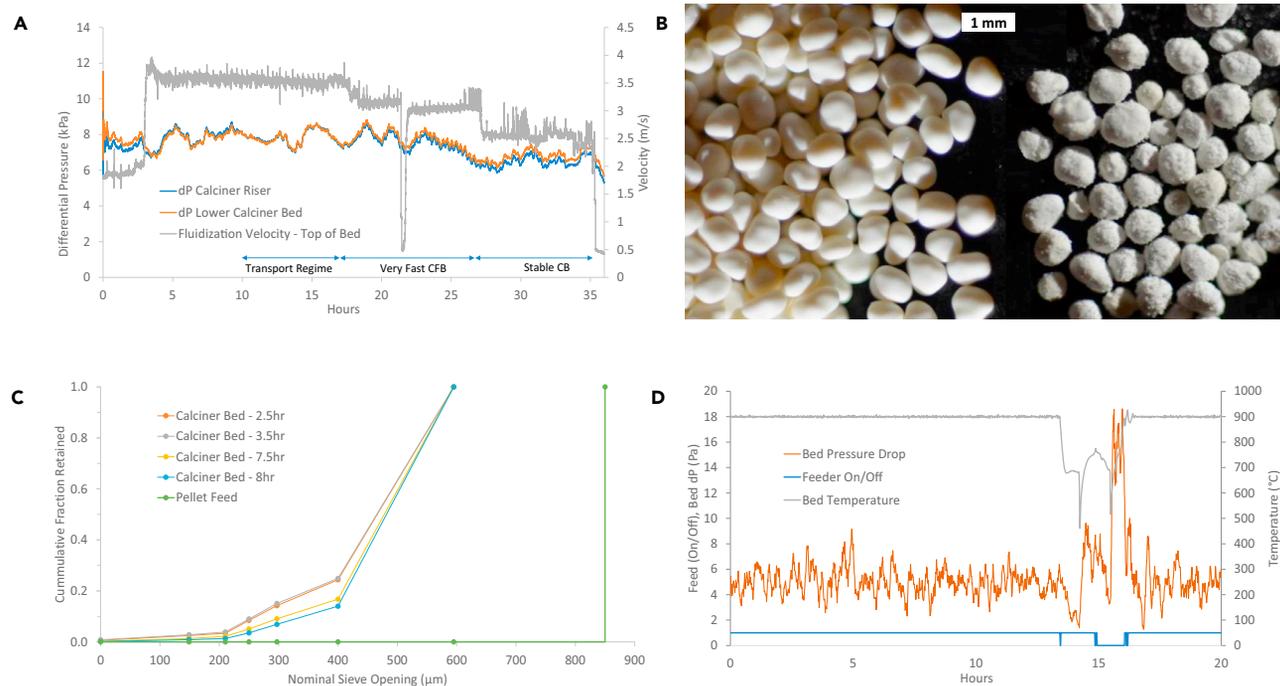


Figure 5. Selected Data from Pilot Calciner

(A) Gas as velocity at bed top computed from flow rates along with pressure drop across two sections of calciner riser. Fluidization regimes as diagnosed from the pressure difference and bed stability at various flow velocities as are indicated.

(B) Photo showing CaCO_3 pellets as they are fed into the calciner and CaO pellets at discharge.

(C) Size distribution of bed sampled bed material and feed pellets are measured by passage through calibrated filter screens.

(D) Measured bed temperature and pressure drop for a selected 20-hr duration, showing an upset, and recovery to equilibrium operation, at hr 14 due to an operator error, which led to auxiliary equipment shutting down.

oxygen and natural gas as the sole fluidizing gas. Reaction conversion $>98\%$ was measured using chemical analysis and X-ray diffraction analysis on discharged pellets.

Figure 5 shows selected pilot plant calciner data, and Figure S4 shows images of CE's pilot calciner and industrial analogs. We attained the specified 90 kg- CaCO_3 /hr feed rate over hundreds of hours of run time with stable bed performance, and with stable combustion with excess oxygen concentration of 20%. Note that given the small bed diameter, we did not expect to obtain stable combustion at lower excess oxygen. Based on performance of their commercial oxy-fired calciners, Technip does not expect difficulty achieving the design ratio in a commercial-scale calciner. Throughput is determined by fluidization velocity in the circulating bed regime. Figure 5A demonstrates the reactor's transition from the transport regime (minimal pressure drop) to a stable circulating bed with a 6.5 kPa pressure drop as fluidization velocity is reduced. Feed and product pellets are shown Figure 5B. The fluidization characteristics are, in turn, determined by the particle size distributions shown in Figure 5C.

Another important goal was to determine that rates of fouling—the deposition of materials on interior surfaces of the calciner—were acceptably low for commercial operation. Two kinds of fouling are relevant. Re-carbonization fouling occurs if temperature drops below the re-carbonation temperature allowing CaO and CO_2 to form CaCO_3 , which forms a hard deposit on surfaces. Alkali fouling is driven by

the sticky alkali species, particularly influenced by carryover of K from the aqueous process. Alkali fouling was judged to pose significant process risk and was an important objective for the pilot. While re-carbonation fouling posed challenges to achieving stable operation of the pilot, Technip anticipates minimal re-carbonation in a larger refractory lined commercial system for which precise temperature control is easier than in our small high-aspect ratio pilot. Once sufficient temperature control was achieved to minimize re-carbonation fouling, the pilot was operated for 90 hr of near-continuous high-temperature operation and minimal alkali fouling was observed, providing confidence to proceed to a commercial-scale calciner.

Process Economics

The cost of new technologies is inherently uncertain. While technology developers may have the most relevant knowledge, they may also have incentives to underestimate costs. In considering the cost of DAC, it is useful to distinguish between adsorption-based technologies^{20,42} that typically require manufacture of hardware not yet available in a competitive market at a relevant price, and technologies such as the process described here that requires construction of an industrial facility that will perform a novel process, but that is constructed using commodity equipment and methods. Uncertainties in the first case arise from scaling up a manufacturing process for a new product (the absorber) system, while the uncertainties in the second arise from estimating project construction costs for a new facility. In both cases, additional uncertainty comes in estimating the performance (e.g., capture rate) and from the cost of energy inputs.

Capital Cost

CE uses the structured front end loading (FEL) process⁴³ for project management. CE has begun development of FEL-3 engineering for a commercial validation plant with a CO₂ capture capacity of order 2 kt-CO₂/year. The project engineering and costing for the 1 Mt-CO₂/year plant described here is FEL-1.

CE's cost estimating process starts with vendors of the major nonstandard unit operations: air contactor, pellet reactor, and calciner/steam-slaker. SPX, RHDHV, and Technip have each worked with CE through years of development, and each have provided budgetary estimates for commercial equipment. The character of these estimates varies with the business model of the vendor. SPX, for example, generally provides firm quotes for the erection of a complete cooling tower on a site prepared by the customer, whereas Technip and RHDHV provide detailed engineering and a limited amount of unique equipment and then work with the customer's EPC firm to oversee construction and commissioning.

All other components are common industrial process equipment available from multiple vendors. Cost estimates for these components start with rough estimates using consultants, equipment vendors, and standard engineering reference sources. CE's engineering group then uses simple multiplicative cost estimating factors to go from equipment costs to estimates of total plant cost.

As a complement to these bottom-up in-house estimates, CE engaged Solaris MCI (Surrey, BC, Canada), a midsized EPC firm, to provide a substantially independent project cost estimate. Solaris worked with major vendors and used their proprietary cost database and factorial project cost estimating methods. The Solaris scope included civil works and utilities. The resulting capital cost and summary justifications are provided in [Table 3](#). In our estimates, we define equipment cost (EC) as major purchased equipment, Materials cost (MC) as materials such as cement and

Table 3. Capital Costs, Given in \$M USD (in 2016 Dollars)

Module	Early Plant			Nth		Justification
	EC	MC	LC	TDFC	TDFC	
Air contactor	\$114.2	\$48.0	\$50.0	\$212.2	\$132.8	SPX provided an estimate for DAC air contactors designed in conjunction with CE. Scope includes contactor structure, top and bottom basins, all liquid piping, fans, gear-reducer, and fan motors. SPX estimate includes field installation. This estimate includes civil work, concrete foundations, liquid pumps and supporting utilities, piping, and instrumentation and controls. Solaris then used estimating tools to provide final EC, MC, and LC roll ups
Pellet reactor	\$76.9	\$28.4	\$25.5	\$130.7	\$94.8	quote from DHV was for major equipment, material, and labor but did not include design, engineering management, and permitting. Solaris then estimated equipment costs at 60% of the RHDHV estimate and broke down the equipment costs into components in DHV quote. Solaris then used ACCE to build up TDFC. Working backward from the ACCE output, 60% of the TDFC is within 1.5% of the original 60% estimate in RHDHV quote
Calciner-slaker	\$43.8	\$18.1	\$15.8	\$77.7	\$63.6	Technip provided a budgetary cost estimate for the steam slaker and calciner equipment which totaled \$16-20 M. Using typical scaling factors based on their project experience installing this type of equipment, they estimated that the TDFC should be between \$70 and \$90 M. Solaris used this information as the basis of their costing analysis in ACCE. The number of calciner/steam slaker units (40 and 12, respectively) was adjusted so the TDFC output from ACCE landed within the \$70-90 M estimate from Technip
Air separation unit	\$38.0	\$0.0	\$16.3	\$54.3	\$46.7	based on quoted data to Solaris for similar ASU projects. Cost source from vendor budget quote for a 1,500 t/day O ₂ unit (95% purity). Labor estimated at 250,000 hr from Solaris in-house experience. Nth plant has a 20% reduction in material cost with the same labor hours
CO ₂ compressor	\$17.2	\$1.4	\$1.4	\$19.9	\$15.5	assumed to be a multistage centrifugal compressor; four stages, inter-stage coolers and scrubbers included. Sizing data estimated based on HYSYS simulation for compression of CO ₂ , and Solaris internal database
Steam turbine	\$6.7	\$0.4	\$0.4	\$7.51	\$5.8	cost of equipment was generated by ACCE based on CE-required specs and compared against available vendor budgetary quotes and in-house data from similar facilities
Power plant	\$32.7	\$0.9	\$1.4	\$35.03	\$26.7	cost of equipment was generated by ACCE based on CE-required specifications and compared against available vendor budgetary quotes and in-house data from similar facilities
Fines filter	\$17.6	\$7.1	\$6.2	\$30.9	\$24.8	CE commissioned Procorp to perform a filtration costing study based on CE's process conditions, which produced the estimated equipment cost. The equipment cost used similar ratios of EC, MC, and LC to TDFC to develop a TDFC
Other equipment	\$96.9	\$3.4	\$2.5	\$102.9	\$77.0	assumed to be 20% of the cost for the rest of the process equipment, including the CO ₂ absorber for the power cycle. Auxiliary equipment was input as a quoted item where only the cost of material could be specified; 20% assumption was made to include the ancillary equipment for which no data on design basis was available to Solaris. Pipe rack cost was generated by ACCE
Buildings	\$2.5	\$0.0	\$4.2	\$6.7	\$5.8	cost of equipment/area generated by ACCE. Building dimensions estimated from plot plan
Transformer	\$ -	\$18.6	\$1.2	\$19.8	\$16.7	assuming 75,000 kVA rated load at 25 kV. Cost from ACCE based on CE-required specs and validated by Solaris against vendor budgetary quotes
Total direct field costs				\$697.7	\$510.1	
Indirect field costs				\$88.5	\$68.8	Solaris estimate including field construction supervision, start-up, and commissioning
Total field costs				\$786.2	\$578.9	
Engineering				\$135.2	\$78.0	assumed to be 12% of TPC less engineering for early plants, 10% of TPC less engineering for N th plant
Other project costs				\$48.2	\$35.8	Solaris estimated charges for home office, G&A, and contract fee
Contingency				\$157.2	\$86.8	assumed to be 20% of total field costs for early plant, and 15% for N th
Total non-field costs				\$340.6	\$200.6	
Total project costs				\$1,126.8	\$779.5	

Estimated capital costs are for a plant design shown in Figure 2 that has a capacity of 0.98 Mt-CO₂/year from the air. Annual quantity captured is lower given utilization. ACCE, Aspen Capital Cost Estimator.

piping not included in EC, and labor cost (LC) as on-site construction labor. Total direct field costs (TDFC) are the sum of EC, MC, and LC. Indirect field costs (IFC) comprise material and labor for construction along with benefits, burdens, consumables, insurance, and other miscellaneous costs. Non-field costs (NFC) comprise engineering, contingency, other project costs such as home office, general and administrative, contract fees, and taxes. Total project cost (TPC) is the sum of TDFC, IFC, and NFC.

Contingency allowances on large projects account for three risk categories: project, strategic, and contextual. Project risks comprise equipment and supply chain risks along with all site-related risks. Strategic risks account for business issues such as joint venture negotiations, changes to the project objectives, or resource management. Contextual risks are those resulting from dependence on current laws, geopolitics, and economic conditions. We account for project risks by adding 20% of TDFC and IFC for the early and plant configuration and 15% for Nth, but we excluded strategic and contextual allowances. Note that the 20% project contingency used here is too low for a first of a kind plant.⁴⁴ CE anticipates that its first plants will be smaller than that the 1 Mt-CO₂/year analyzed here. Our "early plant" value assumes risks have been reduced by construction of a sub-scale commercial plant.

Non-fuel Operating Costs

Solaris and CE estimated non-fuel-operations and maintenance (O&M) unit by unit, relying on industrial experience with similar equipment, and with separate estimates

made for first and Nth plant designs (Table 2). Variable O&M costs include water, labor, and make-up chemicals. We use a water cost of 0.1 \$/m³, the average water acquisition cost for manufacturing in Canada. Water costs are highly variable, but as an upper bound, seawater desalination at a cost of 1 \$/m³ would only add roughly 5 \$/tCO₂ to variable O&M. Fixed O&M is dominated by labor costs, and fixed O&M dominates total O&M so, for simplicity, we report the sum of fixed and variable O&M in Table 2 using a utilization, *U*, of 90%.

Levelized Costs

The levelized costs per ton CO₂ captured from the atmosphere are the sum of levelized capital cost, non-fuel-O&M, and energy costs. The levelized capital cost is $CI \times CRF/U$, where capital intensity (*CI*) is the capital cost per unit capacity, e.g., \$/t-CO₂-year.

The capital recovery factor (*CRF*) is a levelized annual charge on capital divided by the overnight capital cost. If interest on debt is 5%, for example, and return on equity capital is 12.5% then the weighted average cost of capital is 8% for a project with 60% debt financing, yielding a *CRF* of 9.4% if the project is amortized over 25 years. Actual project finance uses a more complex financial structure and includes taxes, but for a rough engineering-economic estimate we reduce all financial uncertainties to variation in the *CRF*. Table 2 presents results for *CRFs* of 7.5% and 12.5%

The final factor is energy costs. These vary sharply around the world. The results in Table 2 assume a natural gas cost of 3.5 \$/GJ and use show results for electricity costs of 30 and 60 \$/MWhr.

DISCUSSION

Given the potential importance of estimates of DAC cost to climate policy,^{3,12,45} we compare our work with prior estimates, discuss life cycle emission, and speculate about the prospects for reducing costs.

Comparison with Prior Estimates

The most influential prior estimate of DAC costs was provided by a 2011 American Physical Society (APS) study.⁴ The study estimated the cost of an aqueous Ca-looping technology like that presented here. The APS “realistic” case had costs of 780 \$/t-CO₂-avoided and 550 \$/t-CO₂-captured, where the “avoided” value includes emission from electricity supply outside the plant boundary. Our cost range is 94–232 \$/t-CO₂ captured, and if we use the financial and gas price assumptions of the APS (*CRF* = 12% and 6 \$/GJ), then our costs would be 107–249 \$/t-CO₂ for the A and B variants in Table 2.

What accounts for this difference?

The cost discrepancy is primarily driven by divergent design choices rather than by differences in methods for estimating performance and cost of a given design. Our own estimates of energy and capital cost for the APS design roughly match the APS values. The most important design choices involved the contactor including (1) use of vertically oriented counterflow packed towers, (2) use of Na⁺ rather than K⁺ as the cation which reduces mass transfer rates by about one-third,^{28,46} and (3) use of steel packings which have larger pressure drop per unit surface area than the packing we chose²⁸ and which cost 1,700 \$/m³, whereas the PVC tower packings we use cost less than 250 \$/m³.

The APS study and a subsequent paper¹⁴ assumed use of hundreds of vertically oriented counterflow packed towers using a design common to post-combustion CCS. This design requires 190 kWhr/t-CO₂ of energy and has a capital intensity of \$1,304 \$/t-CO₂-year. Our comparable contactor cost is 285 \$/t-CO₂-year computed by multiplying the TDFC value in Table 3 by our average ratio of TPC/TDFL. In rough summary, the APS contactor packed tower design yielded a roughly 4-fold higher capital cost per unit inlet area, and also used packing with 6-fold higher cost, and 2-fold larger pressure drop.

The APS contactor design was motivated by concern about the environmental hazard posed by drift from a contactor using cooling-tower components: "If a caustic solution is used to capture CO₂ and the contactor is unenclosed, then the treated air at a large air capture complex can entrain a mist that releases tonnes of caustic solution per day into the environment... the costs to control these mists may be significant, and in some locations permitting may not even be possible."¹⁴ This assumption appears to be incorrect. The measured drift at the outflow of our contactor is well below indoor air quality limits (see section on the Pilot Contactor). Moreover, it is not clear what an "unenclosed" contactor means, since any DAC contactor must exchange air with the atmosphere. Vertical towers specified by the APS commonly use mist eliminators that operate on the same physical principle with very similar specifications to the mist eliminators used in cooling towers and in our design. But vertical towers—especially an array of hundreds—have much higher capital costs per unit volume enclosed than our design based on large-scale cooling towers.

Another important difference in the calciner thermal energy demand, where APS estimated 8.1 GJ/t-CO₂ processed by the calciner compared with our 5.25 GJ/t-CO₂ estimate. Our lower energy demand arises in part from design choices regarding heat integration along with the use of steam slaking. The APS applied a 75% thermal efficiency to results from prior study,²⁵ but it is unclear what this efficiency derating means as the direct radiative and conductive heat loss from large calciners vessels is minimal.

Finally, the APS estimates of avoided cost assumed that all electricity is supplied from a grid with an emissions intensity of 610 kg-CO₂/MWhr. We agree with the APS assessment that it would make little sense to build a DAC plant driven by electricity from a high-carbon grid. The primary technology variant analyzed here uses no net grid electricity, but even for a system with electricity inputs, the APS estimate is high. Global average emissions intensity⁴⁷ was 520 kg-CO₂/MWhr in 2013 and it has likely declined since then, with current US emission intensity⁴⁸ falling to 450 kg-CO₂/MWhr. Global intensities would only be relevant in a world with perfect universal electric transmission. In practice, there are many locations where low-carbon renewables are transmission constrained. These are the locations where DAC with electricity imports makes sense. At a 300 kg-CO₂/MWhr grid intensity and with the 0.366 MWhr/t-CO₂ demand of our "C" variant DAC plant, for example, the grid emissions would be 117 kg-CO₂ per ton captured from the atmosphere. This raises the avoided cost only 13% above the CO₂ captured cost. This is not an unreasonable constraint to siting of DAC plants, given that roughly 20% of world's electricity⁴⁷ now has an average intensity below 300 kg-CO₂/MWhr.

Avoided Emissions and Life Cycle Accounting

A full assessment of DAC needs to address emissions on a life cycle basis, accounting for emissions from construction, indirect emissions from production of inputs used during operations, disposal of wastes, fugitive emissions, and decommissioning.

Full life cycle assessment (LCA) is far beyond the scope of this paper, which aims to provide a quantitative process description that may serve as an input to LCA. Here we provide some preliminary observations.

In configurations that use electricity, the grid emission intensity will be a major determinant of LCA emissions. Although, as discussed above, the impact of electricity emission on avoided cost can still be comparatively small.

CE has performed an LCA of CO₂ emissions for a pure sequestration configuration like that in Scenario B. We used the Economic Input Output LCA tool⁴⁹ to estimate construction phase emissions given construction cost estimates slightly lower than those in Table 3. We used a natural gas emission factor of 63.8 kg-CO_{2e}/GJ-NG for upstream and direct combustion emissions, which is relatively high, and published emissions factors for make-up chemicals and disposal. Fugitive leakage from the DAC facility was combined with estimates for leakage during transport and injection and geological storage. The net result was that ~0.9 tons of CO₂ were permanently sequestered for each ton captured from air.

The results of this LCA are subject to considerable uncertainty, but they are broadly consistent with LCAs for similar large energy projects, which generally find that use-phase emissions are much larger than emissions from construction and decommissioning.

Prospects for Technology Development and Cost Reduction

It is difficult to estimate the cost of a technology prior to its widespread deployment. CE has spent several tens of millions of dollars developing DAC technology, yet our performance and cost estimates still carry substantial uncertainty. Our process design choices were substantially driven by a goal of reducing development risk and reducing the capital cost of early plants, rather than by minimizing energy use or ultimate levelized cost. CE adopted a conservative approach to cost and performance estimation, driven, in part, by controversy around the feasibility and cost of DAC. The process described here should therefore be seen as a low-risk starting point rather than a fully optimized least-cost design.

Many small process changes will incrementally improve performance relative to the baseline described here. Some improvements will be implemented as “de-bottlenecking” measures on early plants, while others can only be incorporated in the design of new plants. Beyond minor changes to the process described here, CE is developing alternate DAC processes. One process involves an all-electric variant of the calcium cycle that eliminates natural gas input. Another is an all liquid-phase regeneration system developed from CE’s membrane-enhanced thermal swing DAC process.⁵⁰ These processes are all built on CE’s air contactor as a common platform technology, since the cooling-tower derived aqueous contactor provides a low-risk, low-energy, and low-capital cost front end for DAC.

SUPPLEMENTAL INFORMATION

Supplemental Information includes four figures and can be found with this article online at <https://doi.org/10.1016/j.joule.2018.05.006>.

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AUTHOR CONTRIBUTIONS

Conceptualization: D.W.K., G.H., D.S.A., and K.H. Methodology: D.W.K., G.H., and K.H. Investigation: D.W.K., G.H., and K.H. Resources: G.H. and K.H. Writing – Original Draft: D.K. and G.H. Visualization: G.H. and K.H. Supervision: D.W.K. and D.S.A. Project Administration: D.S.A. Funding Acquisition: D.W.K., G.H., and K.H.

DECLARATION OF INTERESTS

All authors are employees of Carbon Engineering. All authors have an ownership stake in the form of shares and/or options. David W. Keith is a founder of Carbon Engineering and serves on its Board. The authors are among the inventors of the following US Patents: US9095813B2, US8119091B2, US8871008B2, US8728428B1, and US9637393B2.

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