NATIONAL ENERGY TECHNOLOGY LABORATORY

DIRECT AIR CAPTURE CASE STUDIES: SOLVENT SYSTEM

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Disclaimer

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This report was peer reviewed by three independent peer reviewers with various affiliations and from relevant scientific disciplines to ensure that information presented is based on sound and credible science and considered technically adequate, competently performed, properly documented, and in compliance with established quality requirements. As a step beyond standard internal quality assurance and quality control procedures, the U.S. Department of Energy (DOE) Office of Fossil Energy and Carbon Management/NETL are committed to rigorous peer review of key work products to meet the quality standards of the research community. The following individuals served on the peer review panel:

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LIST OF EXHIBITS

ACRONYMS AND ABBREVIATIONS

°C	Degrees Celsius
°F	Degrees Fahrenheit
AACE	International (formerly Association for the Advancement of Cost Engineering)
acfm	Actual cubic feet per minute
Ar	Argon
ASU	Air separation unit
BBR4	Bituminous Baseline Revision 4
BFD	Block flow diagram
BOP	Balance of plant
Btu	British thermal unit
C_2H_6	Ethane
C_3H_8	Propane
C_4H_{10}	*n*-Butane
$CaCO_3$	Calcium carbonate
CaO	Calcium oxide
$Ca(OH)_2$	Calcium hydroxide
CDR	Carbon dioxide removal
CE	Carbon Engineering
CF	Capacity factor
CH_4	Methane
CH_4S	Methanethiol
cm	Centimeter
CO	Carbon monoxide
CO_2	Carbon dioxide
COC	Cost of CO_2 capture
COE	Cost of electricity
CPU	CO_2 compression and purification unit
CS	Carbon steel
CT	Combustion turbine
CTG	Combustion turbine generator
CWT	Cold water temperature
d	Day
DAC	Direct air capture
DCS	Distributed control system
DOE	Department of Energy
Eng'g CM H.O.& Fee	Engineering construction management, home office, and fees
FCR	Fixed charge rate
FECM	Office of Fossil Energy and Carbon Management
ft	Foot
ft^3	Cubic foot
gal	Gallon
GJ	Gigajoule
gpm	Gallons per minute
h, hr	Hour
H_2O	Water
H_2S	Hydrogen sulfide
HCl	Hydrogen chloride
Hg	Mercury
HHV	Higher heating value
HRSG	Heat recovery steam generator
HVAC	Heating, ventilation, and air conditioning
HWT	Hot water temperature
Hz	Hertz
in	Inch
IOU	investor-owned utility
IP	Intermediate pressure
ISO	International Organization for Standardization
K	Kelvin
K_2CO_3	Potassium carbonate
K_2SO_3	Potassium sulfite
kg	Kilogram
kJ	Kilojoule
km	Kilometer
KOH	Potassium hydroxide
kV	Kilovolt
kW, kWe	Kilowatt electric
kWh	Kilowatt-hour
kWt	Kilowatt thermal
lb	Pound
lb-S	Pounds of sulfur
LHV	Lower heating value
LNB	Low-NOx burner
LP	Low pressure
lpm	Liters per minute
m	Meter
M, MM	Million
m^3	Cubic meter

min	Minute
MJ	Megajoule
MPa	Megapascal
MVA	Mega-volt ampere
MWe	Megawatt electric
MWh	Megawatt-hour
N/A	Not applicable/available
N_2	Nitrogen
NaOH	Sodium hydroxide
NETL	National Energy Technology Laboratory
NGCC	Natural gas combined cycle
NOx	Oxides of nitrogen
O-H	Overhead
O&M	Operation and maintenance
O_2	Oxygen
p.f.	Power factor
ph	Phase
PM	Particulate matter
ppm	Parts per million
ppmv	Parts per million, volume
ppmvd	Parts per million, volume dry
psi	Pounds per square inch
psia	Pound per square inch absolute
psig	Pound per square inch gauge
QGESS	Quality Guidelines for Energy System Studies
R&D	Research and development
RO	Reverse osmosis
S	Second
scf	Standard cubic feet
scfm	Standard cubic feet per minute
scm	Standard cubic meter
SCR	Selective catalytic reduction
SO_2	Sulfur dioxide
STG	Steam turbine generator
T&S	Transport and storage
TASC	Total as-spent capital
TOC	Total overnight cost
tonne	Metric ton (1,000 kg)
TPC	Total plant cost
tpd	Tons per day
U.S.	United States
V	Volt
WG	Water (gauge)
wt%	Weight percent
y, yr	Year

EXECUTIVE SUMMARY

In recent years, there has been a significant increase in research focused on direct air capture (DAC); however, the technology is considered immature with additional research and development (R&D) required to reduce the cost of carbon dioxide (CO_2) removal from the atmosphere. Cost estimates for CO_2 removal from the atmosphere reported by various sorbent- and solvent-based DAC technology developers span the range of $95–600/tonne, with a stated goal to reduce costs below $100/tonne by 2030. [1] [2] Independent and transparent assessments of DAC technology are required to inform future technology development in this area and focus R&D on the parameters that offer the most potential performance and cost improvement.

The objective of this case study is to provide cost and performance estimates for a solvent-based DAC system. Two cases are considered, based on the net CO_2 removal rates. Case 1 has a net CO_2 removal rate of 903,970 tonnes/yr and Case 1A has a net CO_2 removal rate of 100,000 tonnes/yr. The case study is an example configuration for a solvent-based DAC plant and does not represent an optimized design. Most sub-systems were modeled directly from Carbon Engineering (CE) process data; [3] vendor quotes and engineering, procurement, and construction industrial experience were used to model the balance of plant. Independent operating and capital cost estimates based on commercially available technology from reputable suppliers were developed by Black & Veatch using their in-house cost estimating references. The capital cost estimates represent an AACE International (AACE) Class 5 estimate, with an uncertainty range of +/-50 percent. Sensitivity analysis was conducted on multiple process and cost parameters to determine which parameters offer the most potential performance and cost improvement. Only inside the plant fence line emissions are characterized; a full life cycle analysis was not performed as part of this case study.

The solvent-based DAC system evaluated in this case study comprises air contactors that remove 74.5 percent of the CO_2 from the inlet air using a potassium hydroxide (KOH) solvent, [1] pellet reactors that convert potassium carbonate (K_2CO_3) to calcium carbonate ($CaCO_3$) pellets and KOH using calcium hydroxide ($Ca(OH)_2$), steam slakers that dry the $CaCO_3$ pellets and also regenerate $Ca(OH)_2$ through the reaction of calcium oxide (CaO) with water (H_2O), an oxy-fired calciner in which the $CaCO_3$ pellets are converted to CaO and CO_2, a low-pressure air separation unit (ASU) to provide oxygen to the calciner, and a CO_2 compressor for the CO_2 product. The non-DAC components include a natural gas combined cycle (NGCC) comprising a combustion turbine that produces electricity to support the plant auxiliary load, a heat recovery steam generator (HRSG) to generate steam for power production in a small Rankine bottoming cycle, and an absorber column that captures 90 percent of the CO_2 present in the HRSG flue gas using a slip stream of KOH from the air contactors.

The Case 1 DAC system was sized to achieve a net CO_2 capture rate of 903,970 tonnes/yr from the atmosphere. To achieve a net atmospheric CO_2 reduction of 903,970 tonnes/yr, it is necessary to remove 909,225 tonnes/yr from the air and an additional 187,131 tonnes/yr from the NGCC flue gas. The additional 5,255 tonnes/yr removed in the air contactors accounts for the 2.6 percent of CO_2 in the flue gas not captured by the post combustion CO_2 absorber or the air contactor. All CO_2 produced in the oxy-fired calciner (314,926 tonnes/yr) is captured.

A scaled-down version of Case 1, Case 1A, achieving a net CO_2 removal rate of 100,000 tonnes/yr, was also evaluated. This lower capacity is in line with the capture threshold required by the 2018 45Q tax legislation. [4] Case 1A removes 100,478 tonnes CO_2/yr from the air, and an additional 16,748 tonnes CO_2/yr from the NGCC flue gas. A total of 478 tonnes CO_2/yr are emitted from the NGCC flue gas after capture. All the CO_2 produced in the oxy-fired calciner (34,700 tonnes CO_2/yr) is captured. Detailed reporting is only provided for Case 1.

The cost of CO_2 capture (COC) represents the price of CO_2 on a net CO_2 removal basis. The financial assumptions applied in calculating the COC are the same assumptions applied for NGCC cases in NETL's Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 4. [5] Financial parameters that would be most realistic for the DAC process are presently undefined as there are no large-scale commercial projects to date.

Direct air capture (DAC) systems are an immature technology, lacking a history of commercial deployment at scale. The cost estimate methodology presented in this report is the same as that typically employed by NETL for mature plant designs and does not fully account for the unique cost premiums associated with the initial, complex integrations of established and emerging technologies in a commercial application. Thus, it is anticipated that initial deployments of plants based on the cases found in this report may incur costs higher the presented estimates. Absent demonstrated first-of-a-kind plant costs associated with a specific plant configuration/technology, it is difficult to explicitly project fully mature, nth-of-a-kind values. Consequently, the cost estimates provided herein represent neither first-of-a-kind nor nth-of-a-kind costs. Nevertheless, the application of a consistent methodology, and the presentation of detailed equipment specifications and costs based on contemporary sources, facilitate comparison between cases and improve upon publicly available estimates characterized by more opaque and less detailed methods and sources.

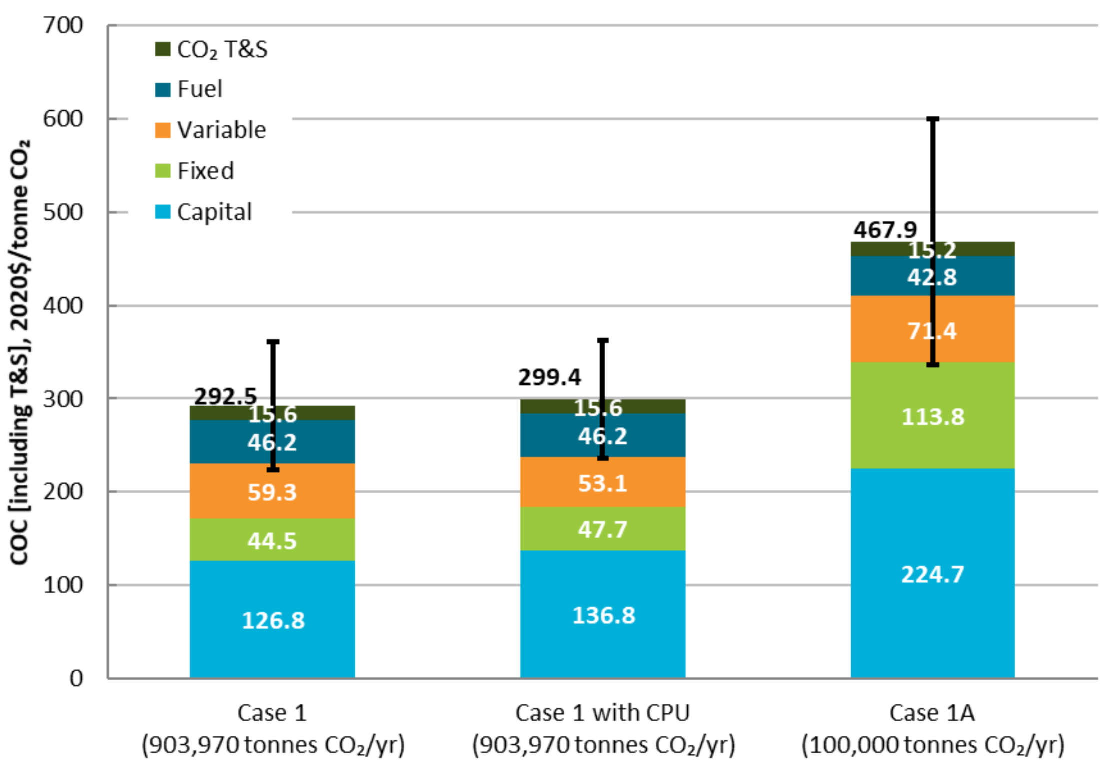
Anticipated actual costs for projects based upon any of the cases presented herein are also expected to deviate from the cost estimates in this report due to project- and site-specific considerations (e.g., contracting strategy, local labor costs and availability, seismic conditions, water quality, financing parameters, local environmental concerns, weather delays) that may make construction more costly. Such variations are not captured by the reported cost uncertainty.

Continuing research, development, and demonstration (RD&D) is expected to result in designs that are more advanced than those assessed by this report, leading to costs that are lower than those estimated here.

Using the methodology described in this study, the cost of CO_2 capture (COC) for the DAC plant is \$293/tonne as shown in Exhibit ES-1. The optimistic end of these results approaches the high end of capture cost ranges reported in the literature for solvent DAC systems; in publicly available literature, CE reported the net COC at \$94–232/tonne CO_2 for various configurations. [1]

At the smaller scale of 100,000 tonnes CO_2/yr, diseconomy of scale raises the COC to $468/tonne. An additional consideration, given the system configuration employing an oxy-fired calciner to produce the CO_2 product stream, is the purity of the CO_2 and its suitability for transport and storage. The CO_2 product purity in Case 1 of this study does not meet CO_2 purity guidelines set forth by the National Energy Technology Laboratory (NETL) Quality Guidelines for Energy System Studies [6]; therefore, the COC of a modified Case 1 incorporating a CO_2 compression and purification unit (CPU) is included in Exhibit ES-1 to ensure that the CO_2 product meets purity specifications. This case also considers the sale of excess electricity generated by the on-site combustion turbine and steam turbine (13.5 MW) at a sale price of $60/MWh (whereas cases 1 and 1A do not).

Exhibit ES-1. Cost of CO_2 capture results for Case 1 (net CO_2 removed basis)



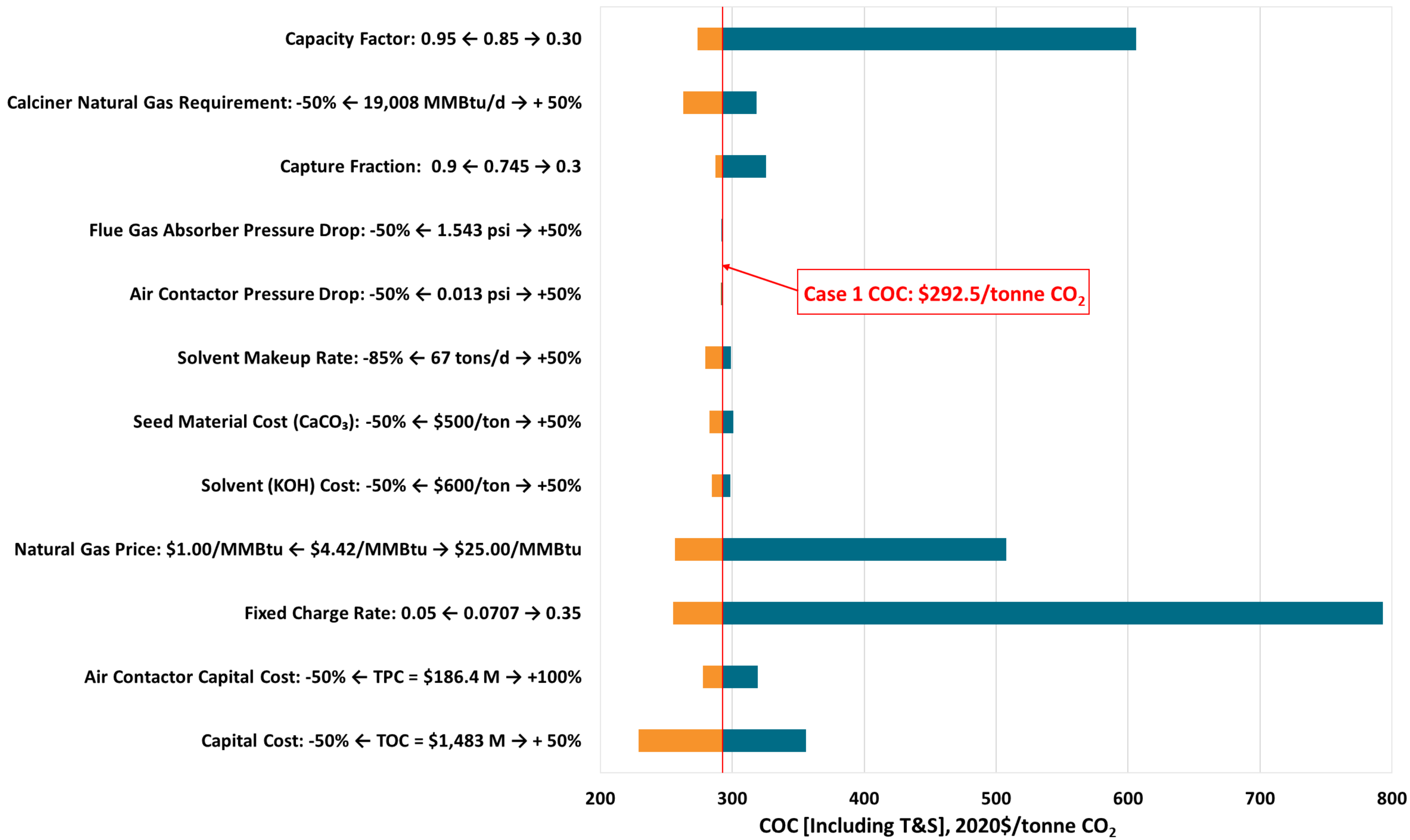
Note: Error bars represent uncertainty range of capital cost estimates (+/-50%)

Each of the COC results shown in Exhibit ES-1 include error bars that reflect the +/-50 percent uncertainty present in the capital cost estimate. The COC uncertainty ranges presented are not reflective of other changes, such as variation in fuel price, labor price, capacity factor, or other factors. Anticipated actual costs for projects based upon any of the cases presented herein are also expected to deviate from the cost estimates in this report due to project- and site-specific considerations (e.g., contracting strategy, local labor costs and availability, seismic conditions, water quality, financing parameters, local environmental concerns, weather delays) that may make construction more costly. Such variations are not captured by the reported cost uncertainty.

To account for such uncertainties, a sensitivity analysis was performed on several critical parameters to gauge their relative impact on the system performance and COC. The additional parameters of interest include capacity factor, system pressure drop, natural gas price, solvent makeup rate, solvent cost, capture fraction, calciner natural gas requirement, financial assumptions (fixed charge rate [FCR]), and a single case addressing revenue from excess electricity generation and CO_2 product purity. A summary of the sensitivity results is shown in Exhibit ES-2.

The process parameters studied in the sensitivity analysis did not have a significant impact on the COC. Given the level of process development by CE on their solvent-based DAC system, bolstered by pilot plant testing, future work using the latest data should be considered. Continuing research, development, and demonstration is expected to result in designs that are more advanced than those assessed by this report, leading to lower costs. Financial parameters, unrelated to process development, were found to have a significant impact on the COC.

Exhibit ES-2. Summary of COC sensitivity results



1 INTRODUCTION

In 2018, the National Academies of Sciences, Engineering, and Medicine released the report "Negative Emissions Technologies and Reliable Sequestration: A Research Agenda." The report focuses on technologies that remove carbon dioxide (CO_2) from the atmosphere, so that it may be stored or utilized. The report assesses five carbon dioxide removal (CDR) technologies, including direct air capture (DAC). The report also makes research and development (R&D) recommendations for advancing these technologies and driving down the cost of deployment.

Technology developers have projected that the cost of CDR will rapidly fall over the next few years. Climeworks, whose technology applies an amine-functionalized solid sorbent, stated in 2019 that the cost to remove CO_2 from the atmosphere using their technology is roughly $600/tonne, and they project that cost to drop to $200/tonne in the next three to four years, with a long-term goal of less than $100/tonne by 2030. [2] Climeworks have demonstrated their technology at a 4,000 tonne/yr scale, and Global Thermostat, who utilize an amine-modified monolith, have demonstrated their technology at a 1,000 tonne/yr pilot. [7] [8] Carbon Engineering (CE), who apply a solvent technology to remove CO_2 from the atmosphere, published a techno-economic analysis in 2018 that projects a cost to remove CO_2 from the atmosphere of $94–232/tonne, depending on financial considerations, regional energy costs, and other factors. [1] The underlying assumptions used to generate public DAC cost results (system performance, capital cost assumptions, financial parameters, inclusion of credits or incentives) are often not presented publicly, and that lack of transparency hampers efforts to validate cost claims.

Without the inclusion of financial incentives, the cost of CDR using these technologies is presently higher than the cost to capture CO_2 from larger point sources, such as industrial processes (e.g., cement and steel manufacture), [9] or flue gas from utility-scale fossil-fueled power generation, [5]; nonetheless, these technologies continue to evolve with R&D and pilot-scale testing. While not currently cost competitive with point source capture, DAC technologies address a different problem; specifically, hard-to-abate emissions (e.g., airplanes) as well as legacy CO_2 emissions already present in the atmosphere.

The objective of this study is to provide cost and performance estimates for a solvent-based DAC system. Two cases with different annual CO_2 removal rates are considered. Case 1 was sized to achieve 903,970 tonnes/year net CO_2 removal. Case 1A achieves a net CO_2 removal rate of 100,000 tonnes/year; this net removal rate meets the minimum threshold for a DAC facility to qualify for 2018 45Q tax credits and may represent a more achievable scale for technologies in earlier phases of development. [4] The case study is an example configuration for a solvent-based DAC plant and does not represent an optimized design. While most process sub-systems are modeled directly from CE's process data, [3] some sub-systems were modeled based on vendor quotes and engineering, procurement, and construction industrial experience. Independent capital and operating cost estimates based on commercially available technology from reputable suppliers were developed by Black & Veatch using their in-house cost estimating references. The capital cost estimates represent an AACE International (AACE) Class 5 estimate, with an uncertainty range of +/-50 percent. Since there is limited public information for industrial-scale DAC systems, sensitivity analysis was conducted on multiple process and cost parameters. A full life cycle analysis was not performed as part of this case study, only inside the plant fence line emissions are characterized.

2 LITERATURE REVIEW

CE was identified as the sole company pursuing solvent-based DAC in the report "Negative Emissions Technologies and Reliable Sequestration: A Research Agenda." [8] A literature review completed in 2019 aimed to identify alternative solvent DAC processes, solvents, and companies pursuing their commercialization, recognizing alternatives were likely at a lower development level than that presented by CE.

Most of the public literature in 2019 regarding solvent-based DAC systems, outside of references published by CE, discussed bench-scale solvent performance and did not provide detailed process configuration information. Exhibit 2-1 shows examples of the solvents discussed in the literature. Strong hydroxide solvents were most often discussed and include sodium hydroxide (NaOH), calcium hydroxide ($Ca(OH)_2$), and potassium hydroxide (KOH). The use of $Ca(OH)_2$ is proposed in earlier literature and has been critiqued for its relatively poor CO_2 mass transfer performance, low solubility in water, and other disadvantages. [10] Additionally, an amino acid-based solution is being investigated by Oak Ridge National Laboratory. [11]

Exhibit 2-1. DAC solvents in literature

Reference	Solvent Type
Lackner et al. 1999 [12]	$Ca(OH)_2$
Herzog 2003 [13]	$Ca(OH)_2$
Stolaroff et al. 2008 [14]	NaOH
Holmes and Keith 2012 [15]	NaOH/KOH
Mazzotti et al. 2013 [16]	NaOH
Brethome et al. 2018 [11]	amino acid solution (glycine and sarcosine)
de Jonge et al. 2019 [17]	NaOH

Alternatives to the solvents mentioned in Exhibit 2-1 were not widely discussed, though one report suggested chilled ammonia and ionic liquids are potential research areas for DAC applications. [18] In addition to scientific journals, a global patent search was included in the literature search. A summary of companies holding DAC-related patents by 2019 is shown in Exhibit 2-2. At the time the literature search was conducted, the only companies holding solvent DAC-related patents appear to be UT-Battelle (associated with Oak Ridge National Laboratory) and CE. Based on publicly available literature, it appears that by 2019 the UT-Battelle technology was in early phases of bench-scale testing. The literature and patent search suggest that CE is the leading company focused on the commercialization of a solvent DAC process. Since 2019, solvent-based DAC R&D has progressed, and novel solvents and concepts that are not included in this literature review have emerged. [19] [20]

Exhibit 2-2. DAC-related patents

Companies Holding DAC-Related Patent(s)	Technology Description
Climeworks	Solid sorbent
Carbon Sink (Parent Company of Infinitree)	Solid sorbent
Global Research Technologies (now Carbon Sink)	Solid sorbent
Kilimanjaro Energy (now Carbon Sink)	Solid sorbent
Carbon Engineering	Liquid solvent
Global Thermostat	Solid sorbent
Skytree	Solid sorbent
UT-Battelle (Oak Ridge National Laboratory)	Liquid solvent

3 Design Basis

3.1 Site and Fuel Characteristics

The cases considered in this study are assumed to be located at a generic plant site in the midwestern United States, with site characteristics and ambient conditions as presented in Exhibit 3-1 and Exhibit 3-2. The ambient conditions are the same as International Organization for Standardization (ISO) conditions. The National Energy Technology Laboratory's (NETL) "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 4" (BBR4) provides a starting point for the ambient conditions for the plant. [5] Adjustments to the air composition were made based on more recent atmospheric data to reflect current concentrations of CO_2 in the atmosphere. An atmospheric CO_2 content of 403.9 ppmv is assumed for the cases in this study.

Exhibit 3-1. Site characteristics

Parameter	Value
Location	Greenfield, Midwestern U.S.
Topography	Level
Size (DAC), acres	260
Transportation	Rail or Highway
Water	50% Municipal and 50% Ground Water

Exhibit 3-2. Site ambient conditions

Parameter	Midwest ISO		
	BBR4 [5]	**DAC**	
Elevation, m (ft)	0 (0)	0 (0)	
Barometric Pressure, MPa (psia)	0.101 (14.696)	0.101 (14.696)	
Average Ambient Dry Bulb Temperature, °C (°F)	15 (59)	15 (59)	
Average Ambient Wet Bulb Temperature, °C (°F)	10.8 (51.5)	10.8 (51.5)	
Design Ambient Relative Humidity, %	60	60	
Cooling Water Temperature, °C (°F)[A]	15.6 (60)	15.6 (60)	
	Air composition, mass %		**Air composition, mole %**
N_2	75.055	74.983	77.243
O_2	22.998	23.050	20.784
Ar	1.280	1.272	0.919
H_2O	0.616	0.633	1.014
CO_2	0.050	0.062	0.040 (403.9 ppmv)
Total	**100.00**	**100.00**	**100.00**

[A]The cooling water temperature is the cooling tower cooling water exit temperature. This is set to 4.8°C (8.5°F) above ambient wet bulb conditions in ISO cases

The land area for the DAC plant is assumed to be 260 acres, based on plant layout drawings provided by CE. [3] The land area requirement includes a plant buffer boundary of 500 feet between the plant proper and the fence line.

Natural gas is utilized as the fuel for on-site electricity generation to satisfy plant auxiliary loads, as well as for process heat requirements of the solvent system; its composition is presented in Exhibit 3-3. The natural gas properties are taken from the 2019 revision of the Quality Guidelines for Energy System Studies (QGESS) document "Specification for Selected Feedstocks." [21]

Exhibit 3-3. Natural gas composition

Component		Volume Percentage	
Methane	CH_4	93.1	
Ethane	C_2H_6	3.2	
Propane	C_3H_8	0.7	
n-Butane	C_4H_{10}	0.4	
Carbon Dioxide	CO_2	1.0	
Nitrogen	N_2	1.6	
Methanethiol[A]	CH_4S	5.75×10^{-6}	
	Total	**100.0**	
	LHV		**HHV**
kJ/kg (Btu/lb)	47,201 (20,293)		52,295 (22,483)
MJ/scm (Btu/scf)	34.52 (927)		38.25 (1,027)

[A]The sulfur content of natural gas is primarily composed of added mercaptan (methanethiol [CH_4S]) with trace levels of hydrogen sulfide (H_2S) [22]
Note: Fuel composition is normalized, and heating values are calculated using Aspen

The levelized natural gas price is \$4.19/GJ (\$4.42/MMBtu) on a higher heating value (HHV) basis, delivered to the Midwest. [23] Fuel costs are levelized over an assumed 30-year plant operational period with an assumed on-line year of 2023.

3.2 Environmental Targets

The environmental targets that would be enforced for a plant of the type presented in this study are presently unclear. However, NETL's BBR4 presents air emission targets for natural gas combined cycle (NGCC) plants, and these targets are reproduced for reference in Exhibit 3-4. [5]

Exhibit 3-4. NGCC emissions targets [5]

Pollutant	NGCC (lb/MWh-gross)
SO_2	0.90
NOx	0.43
PM (Filterable)	N/A
Hg	N/A
HCl	N/A

These air emission targets for NGCC power plants were applied when assessing the air emissions produced by the DAC plant.

3.3 DAC Plant Size

Case 1 was sized to achieve 903,970 tonnes/year net CO_2 removal. Since this is a net removal target, the gross CO_2 removal rate from the air (and, therefore, the actual CO_2 product flow rate from the DAC plant) will be higher. The excess CO_2 required to be removed from the air in order to meet the net removal target is dependent on many factors, including electrical auxiliary load, plant efficiency, system configuration (e.g., electricity generation on-site versus purchased power from the grid), and DAC system performance characteristics.

A scaled-down Case 1A was also developed. The scaled-down version of the process targets a net CO_2 removal rate of 100,000 tonnes/yr; this net removal rate meets the minimum threshold for a DAC facility to qualify for 45Q tax credits and may represent a more achievable scale for technologies in earlier phases of development. [4]

For perspective on the selected DAC plant size, Case B31A from NETL's BBR4 (a 727-MWnet 2017 F-Class combustion turbine-based NGCC, without CO_2 capture equipment, operating at an 85% capacity factor) emits 1.7 M tonnes/yr of CO_2, or 0.35 tonnes/MWh_{net}. The 903,970 tonnes CO_2/yr Case 1 DAC plant size is equivalent to 301 MW_{net} worth of flue gas CO_2 from Case B31A, or 41 percent of the net plant output. The 100,000 tonnes CO_2/yr Case 1A DAC plant size is equivalent to 33.3 MW_{net} worth of flue gas CO_2 from Case B31A, or 4.6 percent of the net plant output.

3.4 Capital Costs

The capital cost estimates documented in this report reflect the uncertainty ranges shown in Exhibit 3-5.

Exhibit 3-5. Capital cost uncertainty ranges

Technology	Uncertainty Range	AACE Classification
Case 1/1A	+/-50	Class 5

The Case 1 and Case 1A uncertainty range of +/-50 percent is consistent with the AACE Class 5 cost estimate (i.e., feasibility study) [24] [25], based on the level of engineering design performed.

Cost estimates for plant components that represent mature technologies, which have been widely deployed at commercial scale (e.g., combustion turbine), reflect nth-of-a-kind on the technology commercialization maturity spectrum. The costs of these components have dropped over time due to "learning by doing" and risk reduction benefits that result from serial deployments as well as continued R&D. Cost estimates for plant components that are not yet fully mature (e.g., air contactors, oxy-fired calciner at the scale required by the system in this study) use the same cost estimation methodology as for mature plant components, which does not fully account for the unique cost premiums associated with the initial, complex integrations of emerging technologies in a commercial application. Thus, it is anticipated that initial deployments of these plants may incur costs higher than those reflected within this report. Without demonstrated first-of-a-kind plant costs associated with a specific plant configuration/technology, it is difficult to explicitly project fully mature, nth-of-a-kind values. Consequently, the cost estimates provided herein represent neither first-of-a-kind nor nth-of-a-kind costs.

Actual project costs for the plant-type considered in this study are expected to deviate from the cost estimates in this report due to project- and site-specific considerations (e.g., contracting strategy, local labor rates, and others) that may make construction more costly. These variations are not captured by the reported cost uncertainty.

Finally, continuing research, development, and demonstration is expected to result in designs that are more advanced than those assessed by this report, leading to lower costs.

3.5 CO_2 Transport and Storage

The cost of CO_2 transport and storage (T&S) in a deep saline formation is estimated using the Department of Energy (DOE) Office of Fossil Energy and Carbon Management (FECM)/NETL CO_2 Transport Cost Model (CO_2 Transport Cost Model) and the FECM/NETL CO_2 Saline Storage Cost Model (CO_2 Storage Cost Model). Additional detail on development of these costs is available in the 2019 revision of the QGESS "Carbon Dioxide Transport and Storage Costs in NETL Studies." [26]

Due to the variances in the geologic formations that make up saline formations across the United States, the cost to store CO_2 will vary depending on location. Storage cost results from the CO_2 Storage Cost Model align with generic plant locations from the NETL studies:

- Midwest plant location – Illinois Basin
- Texas plant location – East Texas Basin
- North Dakota plant location – Williston Basin
- Montana plant location – Powder River Basin

The far-right column of Exhibit 3-6 shows the total T&S costs used in NETL system studies for each plant location rounded to the nearest whole dollar. Only the $10/tonne value is used in this report since all cases are assumed to be located in the Midwest.

Exhibit 3-6. CO_2 transport and storage costs

Plant Location	Basin	Transport (2018 $/tonne)	Storage Cost at 25 Gigatons (2018 $/tonne)	T&S Value for System Studies[A] (2018 $/tonne)
Midwest	Illinois	2.07	8.32	10
Texas	East Texas		8.66	11
North Dakota	Williston		12.98	15

Montana	Powder River		19.84	22

[A]The sum of T&S costs rounded to the nearest whole dollar

3.6 Cost of CO_2 Capture Calculation Methodology

NETL has provided guidance on methods for calculating cost of electricity (COE) for power plants in NETL techno-economic analyses in its QGESS "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." [27] The COE equation used is provided below.

$$COE = CC + OM + FP \qquad (1)$$

Where:

CC = capital charges for the plant

OM = operation and maintenance (O&M) costs for the plant

FP = fuel costs for the plant

The annual O&M and fuel costs for the plant are calculated based on system performance and added to the capital charges.

The capital charge portion of COE is calculated by multiplying the plant total as-spent cost (TASC) with a fixed charge rate (FCR). TASC is calculated by taking the plant total overnight cost (TOC) and multiplying by a TASC/TOC ratio. The determination of the FCR and TASC/TOC ratio is presented in the referenced QGESS and is based on financial parameter assumptions common to the power industry. The product of FCR and TASC/TOC is referred to as the fixed charge factor.

The COE methodology outlined above is applied to calculate the cost of CO_2 capture (COC) by the DAC plant. The FCR used in these calculations is 0.0707 and the TASC/TOC ratio used is 1.093. These values were generated for a plant with a three-year construction period and are applied to NGCC cases in other NETL studies. [5] A sensitivity study on the FCR is presented in Section 5.3.3 to bound the potential impacts of industry-specific financial parameter assumptions on the resulting COC.

The equation shown below provides the full calculation for the COC for the DAC plant:

$$Cost\ of\ CO_2\ Capture\ (\frac{\$}{tonne}) = \frac{(FCR)(TASC)+OC_{FIX}+(CF)(OC_{VAR})+(CF)(FP)}{(CF)(F_{CO2-X})} \qquad (2)$$

Where:

FCR = fixed charge rate taken from the referenced QGESS [27]

TASC = total as-spent capital

OC_{FIX} = the sum of all first-year-of-operation fixed annual operating costs

CF = plant capacity factor, assumed to be constant (or levelized) over the operational period; expressed as a fraction

OC_{VAR} = the sum of all first-year-of-operation variable annual operating costs at 100 percent capacity factor (excluding fuel), offset by any byproduct revenues

FP = the sum of annual fuel costs at 100 percent capacity factor; a natural gas price of \$4.42/MMBtu is used, sourced from NETL's QGESS "Fuel Prices for Selected Feedstocks in NETL Studies" [23]

$F_{CO2\text{-}X}$ = annual flow of CO_2 from the plant, which can have differing perspectives; for DAC_{net}, the flow from the plant is the net CO_2 removed from the atmosphere (903,970 tonne/yr), $F_{CO2\text{-}net}$; for DAC_{gross}, the flow from the plant is the gross CO_2 removed across the air contactor, $F_{CO2\text{-}DACGross}$; for $Plant_{gross}$, the flow from the plant is the gross DAC removal, the CO_2 product flow from the NGCC plant, and the CO_2 product from the oxy-fired calciner, $F_{CO2\text{-}PlantGross}$; These three calculation bases refer to only direct emissions from the DAC facility, and do not refer to a life cycle analysis of the facility

4 DAC System Description

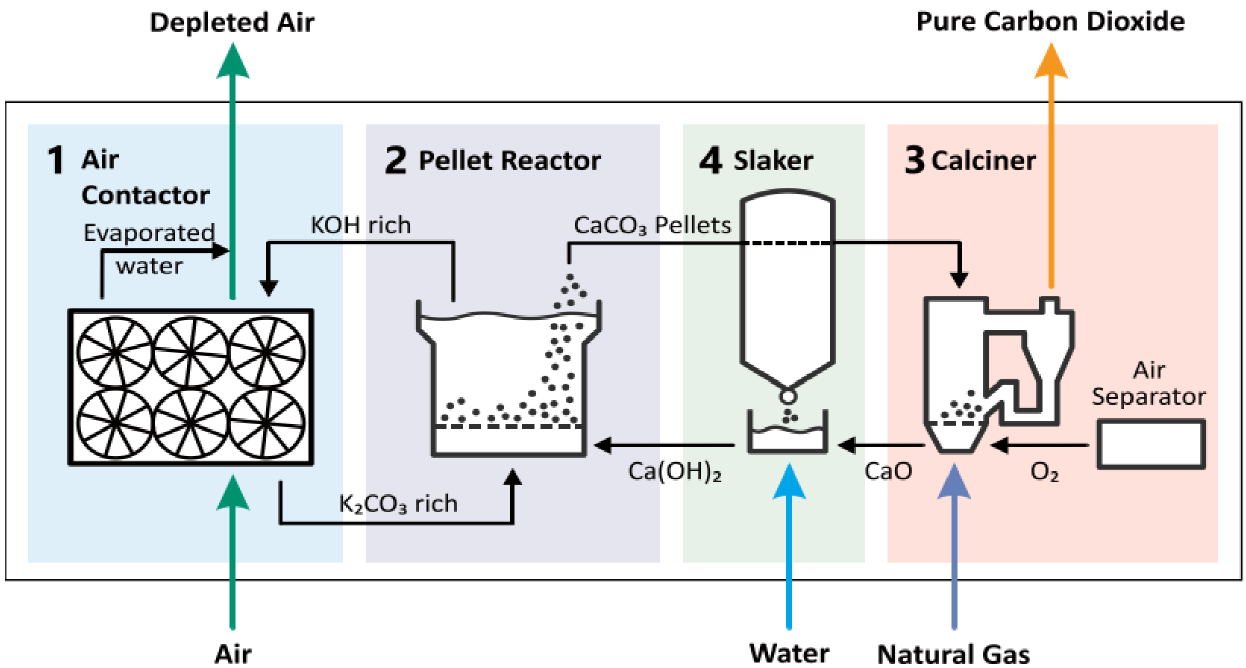
The process configuration considered in Case 1 includes the use of an NGCC plant to provide the electrical auxiliary load of the DAC system. CO_2 present in the flue gas from the NGCC plant is captured at a rate of 90 percent with an absorber column using a portion of the KOH solvent from the air contactors. The purified flue gas leaving the NGCC absorber is sent to a blower, and then co-mingled with inlet air and passed through fans that provide the motive force to deliver the air to the DAC air contactors and overcome the pressure drop of the duct distribution system and air contactor. Purified air and flue gas leave the system at the exit of the air contactors.

The NGCC plant is modeled after Case B31B presented in NETL's BBR4. [5] Sub-system descriptions for the NGCC plant can be found in the reference report and are not replicated here.

There were two notable changes made to Case B31B to better represent the DAC system. Given the size, and electrical demand of the DAC system, a smaller aeroderivative combustion turbine (CT) is considered; the CT considered in this case is an LM6000 aeroderivative CT with a gross electrical output of about 52 MW, as compared to the larger F-class CT considered in Case B31B. The second notable change is that Case B31B considered a triple pressure reheat heat recovery steam generator (HRSG), supplying a triple pressure steam turbine bottoming cycle. The HRSG in Case 1 maximizes heat recovery from the slaker, calciner, and CT flue gas, and produces steam at 4.2 MPa (609.2 psia) and 0.5 MPa (75.3 psia). A small portion of the low-pressure (LP) steam is super-heated to fulfil the triethylene glycol dryer requirements for CO_2 compression. The remainder of the steam generated is sent to a dual-pressure steam turbine bottoming cycle. Excess power is generated in the reference case, but sale of this electricity is not included in base case results.

The sub-systems specific to the DAC portion of the process are modeled directly from CE's process data. [3] The DAC plant layout was assumed to match the layout provided by CE in their public reports. Air contactor rows are spaced 250 meters apart to prevent low-CO_2 concentration air from being drawn into downwind air contactors. Assuming this spacing, with a standard buffer to the fence line of 500 feet, approximately 260 acres of land are required. The following sub-sections provide additional description of sub-systems specific to the DAC portion of the process; a simplified diagram of the DAC process is shown in Exhibit 4-1 for reference.

Exhibit 4-1. DAC simplified process schematic [3]



4.1 Air Contactor

A CO_2-lean solution of KOH is circulated over packing, bringing it into contact with the air in a high efficiency cross-flow configuration air contactor. The CO_2 present in the atmosphere at low concentration is absorbed in the caustic solution. The CO_2-rich solution is pumped to the pellet reactor system for further processing.

A side stream of the CO_2-rich solution is used to capture CO_2 from the NGCC power plant flue gas using a separate absorber unit. The treated NGCC power plant flue gas stream leaves the above absorber and is piped into the air contactor unit to remove 74.5 percent of the leftover CO_2. [1] [3]

The air contactor is based on commercial cooling tower technology. While the geometry and fluid chemistry differ from conventional cooling towers, the design relies on many of the same components, including fans, structured packings, demisters, fluid distribution systems, and fiber-reinforced plastic structural components.

4.2 Pellet Reactor

The pellet reactor is fed with the CO_2-rich solution from the air contactor, along with slaked lime. The pellet reactor hosts the causticization reaction, converting the feed stock into solid calcium carbonate ($CaCO_3$) and rejuvenating the KOH solution. The $CaCO_3$ pellets are separated out, washed, and sent to the steam slaker to be dried and then sent to the calciner.

4.3 STEAM SLAKER

The steam slaker receives the washed $CaCO_3$ pellets from the pellet reactor system and hot quicklime from the calciner. Steam drives the slaking reaction where the quicklime (CaO) particles are hydrated into slaked lime. Heat generated during the slaking is used to heat and dry the $CaCO_3$ pellets. The slaked lime is re-used in the pellet reactor and the dried $CaCO_3$ pellets are sent to the calciner.

The slaker is a refractory lined bubbling/turbulent fluid bed that is fluidized by recirculating steam flow.

4.4 CALCINER

The dried $CaCO_3$ pellets are calcined, producing quicklime and high purity CO_2. The CO_2 is cooled down and residual solids removed through heat recovery cyclones. The CO_2 stream is then dehydrated and compressed. The quicklime is sent to the steam slaker.

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 and natural gas is injected directly into the fluidized bed.

4.5 CO_2 ABSORBER

The gas turbine exhaust stream is stripped of CO_2 at a rate of 90 percent using a conventional counterflow gas liquid column, using a portion of the fluid stream leaving the air contactor. The absorber outlet is ducted to the main air contactor where the remaining CO_2 is captured at a rate of 74.5 percent. [1] [3]

5 CASE 1 PERFORMANCE AND COST ESTIMATES

This section provides the process description and performance and cost results for Case 1. The system description follows the block flow diagram (BFD) in Exhibit 5-1 and stream numbers reference the same exhibit.

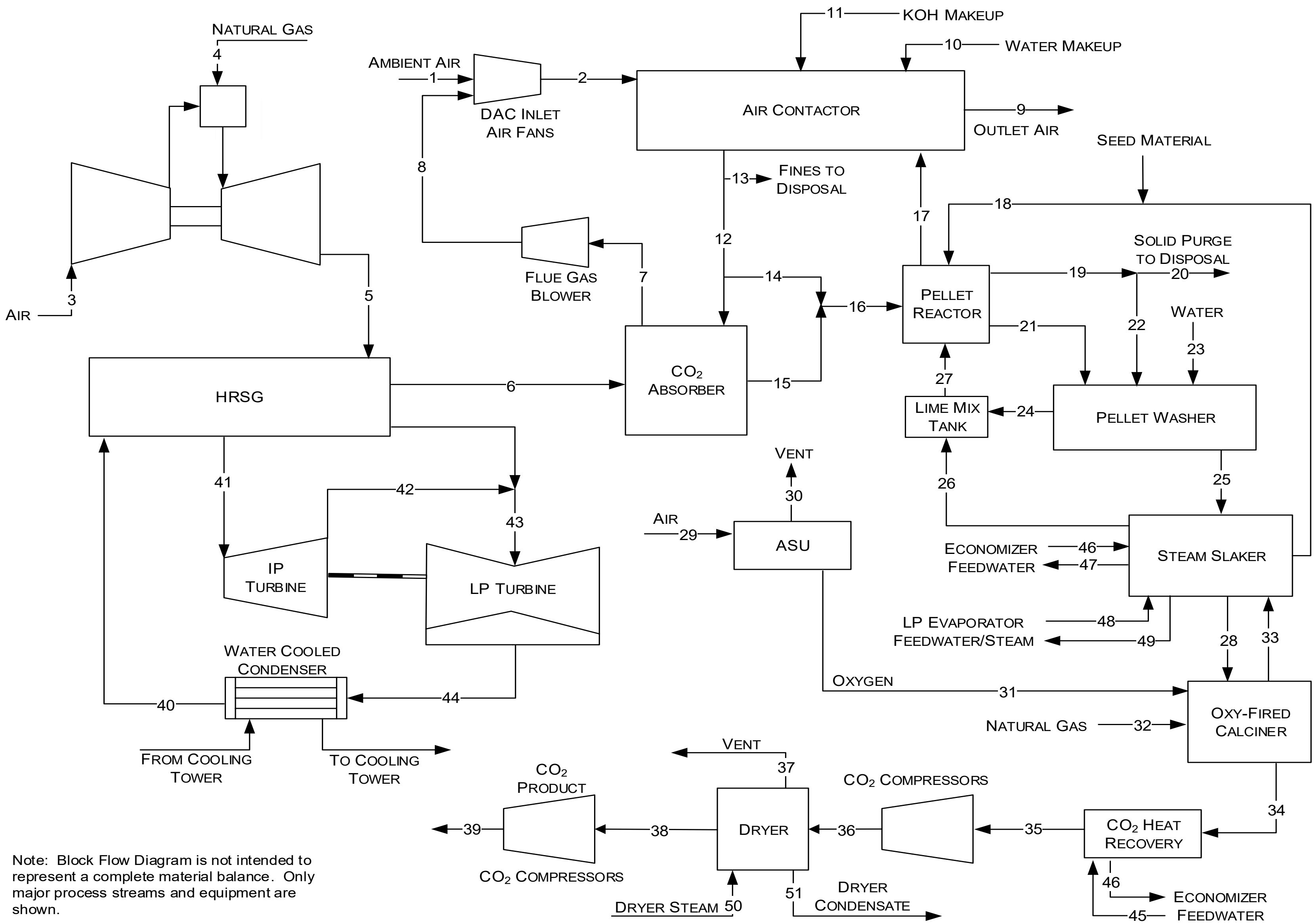
5.1 CASE 1 – PROCESS DESCRIPTION AND PERFORMANCE RESULTS

Case 1 captures a net 903,970 tonnes CO_2/yr from the atmosphere. The plant produces 13.2 MW of excess electricity; however, Case 1 does not account for revenue generated by selling this excess electricity to the grid.

Ambient air (stream 3) is supplied to an inlet filter and compressed before being combined with natural gas (stream 4) in the dry low-oxides of nitrogen (NOx) burners (LNBs), which are operated to control the rotor inlet temperature at 1,347°C (2,457°F). The flue gas exits the turbine at 505°C (940°F) (stream 5) and passes into the HRSG. The dual-pressure HRSG generates 4.2 MPa (609.2 psia) steam and 0.5 MPa (75.3 psia) steam. A small portion of the LP steam is used for the CO_2 triethylene glycol dryer (stream 50), but the balance of steam is sent to the steam turbine (streams 41 and 43). Flue gas exits the HRSG at 173°C (344°F) and passes to the CO_2 absorber, where the flue gas is contacted with a portion of the KOH solvent leaving the air contactor. The absorber captures 90 percent of the CO_2 in the flue gas. The CO_2 reacts with KOH to form potassium carbonate (K_2CO_3); the K_2CO_3-enriched stream (stream 15) is sent to the pellet reactor for further processing. The purified flue gas is sent through a fan to the air contactor for further CO_2 removal.

Along with the purified flue gas, ambient air (stream 1) is sent through fans and a duct system to distribute air to the 277 DAC air contactors (stream 2). The air contactors remove 74.5 percent of the inlet CO_2 via the reaction of KOH with CO_2 to produce K_2CO_3. The purified air and flue gas exit the process through the air contactors (stream 9). The K_2CO_3 rich stream is sent to the pellet reactor (stream 16), where K_2CO_3 reacts with $Ca(OH)_2$ to produce $CaCO_3$ and KOH. KOH (stream 17) is recycled back to the air contactor, and the $CaCO_3$ pellets are washed and sent to the steam slaker for drying (stream 25). In the steam slaker, calcium oxide (CaO) reacts with water to produce $Ca(OH)_2$, which is recycled back to the pellet reactor (stream 27). Heat generated via the exothermic reaction between CaO and water is recovered to generate steam. The dried $CaCO_3$ pellets (stream 28) are sent to the oxy-fired calciner, where they undergo thermal decomposition to form CaO and CO_2. An LP air separation unit (ASU) produces 138,072 lb/hr of oxygen (O_2) at 17.4 psia (stream 31) for use in the oxy-fired calciner. The calciner product CO_2 includes the CO_2 removed in the flue gas absorber, the air contactors and high purity CO_2 produced through the oxy-combustion of natural gas in the calciner. Heat is recovered from the CO_2 product before it is dried and compressed to 15.2 MPa (2,200 psig) (stream 39).

Exhibit 5-1. Case 1 BFD, solvent-based DAC system



Note: Block Flow Diagram is not intended to represent a complete material balance. Only major process streams and equipment are shown.

Overall plant performance is summarized in Exhibit 5-2; Exhibit 5-3 provides a detailed breakdown of the auxiliary power requirements.

Exhibit 5-2. Case 1 plant performance summary

Performance Summary	
Combustion Turbine Power, MWe	52
Steam Turbine Power, MWe	27
Total Gross Power, MWe	**78**
ASU, kWe	14,220
CO_2 Compression, kWe	23,360
Air Fans, kWe	7,210
Balance of Plant, kWe	20,331
Total Auxiliaries, MWe	**65**
Net Power, MWe	**13**
HHV Combustion Turbine Efficiency, %	37.0%
LHV Combustion Turbine Efficiency, %	41.0%
Steam Turbine Cycle Efficiency, %	53.1%
Steam Turbine Heat Rate, kJ/kWh (Btu/kWh)	6,785 (6,431)
Condenser Duty, GJ/hr (MMBtu/hr)	268 (254)
Calciner Natural Gas Feed Flow, kg/hr (lb/hr)	15,981 (35,233)
NGCC Natural Gas Feed Flow, kg/hr (lb/hr)	9,632 (21,234)
Total Natural Gas Feed Flow, kg/hr (lb/hr)	**25,613 (56,467)**
Total HHV Thermal Input, kWt	**372,063**
Total LHV Thermal Input, kWt	**335,823**
Capacity Factor	85%
NGCC Flue Gas CO_2 Captured[A], tonne/yr	187,131
DAC CO_2 Removed from Air (Gross), tonne/yr	909,225
Calciner CO_2 Captured, tonne/yr	314,926
NGCC Flue Gas CO_2 Emitted to Air, tonne/yr	5,255
Net CO_2 Removed from Air, tonnes/yr	**903,970**
Total Plant CO_2 Flow to Storage, tonnes/yr	**1,411,282**

[A]Includes CO_2 captured in the flue gas absorber and air contactors

Exhibit 5-3. Case 1 plant power summary

Power Summary	
Combustion Turbine Power, MWe	52
Steam Turbine Power, MWe	27
Total Gross Power, MWe	**78**
Auxiliary Load Summary	
Circulating Water Pumps, kWe	2,310
Combustion Turbine Auxiliaries, kWe	110
Condensate Pumps, kWe	130
Cooling Tower Fans, kWe	1,190
CO_2 Compression, kWe	23,360
Ground Water Pumps, kWe	790
Miscellaneous Balance of Plant[A], kWe	60
SCR, kWe	1
Steam Turbine Auxiliaries, kWe	20
Transformer Losses, kWe	370
Air Fans, kWe	7,210
Absorber Flue Gas Blower, kWe	1,490
KOH Pumps, kWe	8,970
Lime Pumps, kWe	40
Pellet Reactor Fluidization Pumps, kWe[B]	3,130
Pellet Reactor Filtration Pumps, kWe[B]	1,720
Air Separation Unit Auxiliaries, kWe	210
Air Separation Unit Main Air Compressor, kWe	14,010
Total Auxiliaries, MWe	**65.1**
Net Power, MWe	**13.2**

[A]Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads
[B]These units are modeled separately from the rest of the DAC process

5.1.1 Environmental Performance

It is assumed that all emissions in this process originate in the NGCC portion of the plant (other potential sources of air emissions are not considered). For example, KOH slip from the air contactors is a possibility, but not quantified in this section due to a lack of available data. A summary of the plant air emissions is shown in Exhibit 5-4.

Exhibit 5-4. Case 1 air emissions

	kg/GJ (lb/MMBtu)	tonne/yr (ton/yr)[A]	kg/MWh (lb/MWh)[B]	lb/lb CO_2 net captured
SO_2	0.0004 (0.0009)	1.5 (1.7)	0.003 (0.006)	0.0000017
NOx	0.002 (0.004)	7 (8)	0.012 (0.026)	0.0000077
Particulate	0.000 (0.000)	0 (0)	0.000 (0.000)	0.0000
Hg	0.000 (0.000)	0 (0)	0.000 (0.000)	0.0000
CO	0.000 (0.000)	0 (0)	0.000 (0.000)	0.0000
CO_2	1.4 (3.3)	5,255 (5,792)	9 (20)	0.00581

[A]Calculations based on an 85 percent CF
[B]Emissions based on gross power

As discussed previously in Section 3.2, it is presently unclear to what environmental targets the DAC plant would be subject. However, based on the air emission targets laid out for reference NGCC power plants, Case 1 would comply with air emission regulations for NGCC plants for sulfur dioxide (SO_2) and NOx.

The natural gas was assumed to contain the domestic average value of total sulfur of 0.34 grains/100 scf (4.71×10^{-4} lb-S/MMBtu). [22] It was also assumed that the added methanethiol (CH_4S) was the sole contributor of sulfur to the natural gas. No sulfur capture systems were implemented. Due to the presence of SO_2 in the flue gas (0.18 ppm) it is possible that K_2SO_3 would be produced in the absorber and/or air contactor; this is not accounted for in Case 1. If produced, K_2SO_3 would need to be purged from the system and would potentially cause issues such as fouling. Additionally, the formation of K_2SO_3 may result in lower sulfur emissions than reported.

The CT considered was based on a vendor quote for a LM6000 aeroderivative CT. The reference CT is designed to achieve approximately 1.8 ppmvd NOx emissions (at 15 percent O_2) using a dry LNB in the combustion turbine generator (CTG)—the dry LNBs reduce the emissions to about 9 ppmvd (at 15 percent O_2) [28]—and selective catalytic reduction (SCR) equipment—the SCR system is designed for 82.7 percent NOx reduction. [29]

The pipeline natural gas was assumed to contain no mercury (Hg) or hydrogen chloride (HCl), resulting in zero emissions.

The CO_2 absorber removes 90 percent of the CO_2 in the NGCC flue gas and an additional 74.5 percent of the remaining CO_2 in the purified flue gas is removed in the air contactors. The KOH in the air contactors also removes 74.5 percent of the CO_2 in the DAC inlet air. All of the CO_2 produced from the oxy-combustion of natural gas in the calciner is recovered. The total annual CO_2 emission rate in Case 1 is 5,255 tonnes/yr.

The carbon balance for the plant is shown in Exhibit 5-5. The carbon input to the plant consists of carbon in the natural gas fed to both the CT and calciner, and carbon as CO_2 in the air fed to the CT, the DAC air contactors, and the ASU. Carbon leaves the plant as CO_2 through the air contactors, solid disposal streams, the CO_2 product stream, and other vents.

Exhibit 5-5. Case 1 carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Natural Gas (NGCC)	6,957 (15,337)	Air Contactor Outlet Air	11,598 (25,570)
Natural Gas (Calciner)	11,543 (25,448)	Fines Disposal	81 (178)
NGCC Air	82 (180)	Pellet Reactor Solids Disposal	652 (1,437)
Air Contactor Inlet Air	44,744 (98,644)	Air Separation Unit Vent	47 (103)
Air Separation Unit Inlet Air	47 (103)	CO_2 Product	51,447 (113,422)
Seed Material ($CaCO_3$)	470 (1,036)	CO_2 Dryer Vent	13 (29)
		CO_2 Knockout	0.6 (1.4)
Total	**63,842 (140,748)**	**Total**	**63,842 (140,748)**

As shown in Exhibit 5-6, the sulfur content of the natural gas is insignificant. The sulfur oxides produced by the CT is assumed to leave the process through the exit of the air contactors. The SO_2 produced in the calciner is assumed to exit the process through the CO_2 dryer vent.

Exhibit 5-6. Case 1 sulfur balance

Sulfur In		Sulfur Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Natural Gas (NGCC)	0.1 (0.2)	Stack Gas	0.1 (0.2)
Natural Gas (Calciner)	0.2 (0.4)	CO_2 Dryer Vent	0.2 (0.4)
Total	**0.3 (0.6)**	**Total**	**0.3 (0.6)**

Exhibit 5-7 shows the overall water balance for Case 1. The summary includes the water balance based upon calculations for CO_2 drying, CO_2 compression recovery, air contactor solvent makeup, pellet washer, and cooling tower.

Exhibit 5-7. Case 1 water balance

Water Use	Water Demand	Internal Recycle	Raw Water Withdrawal	Process Water Discharge	Raw Water Consumption
	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)	m^3/min (gpm)
CO_2 Drying	–	–	–	0.007 (1.8)	-0.007 (-1.8)
CO_2 Compression Recovery	–	0.5 (145)	-0.5 (-145)	–	-0.5 (-145)
Air Contactor Solvent Makeup	14 (3,804)	–	14 (3,804)	–	14 (3,804)
Pellet Washer	10 (2,693)	–	10 (2,693)	–	10 (2,693)
Cooling Tower	9.0 (2,375)	–	9.0 (2,375)	2.0 (534)	7.0 (1,840)
Total	**34 (8,871)**	**0.5 (145)**	**33 (8,726)**	**2.0 (536)**	**31 (8,190)**

5.1.2 Energy Balance

An overall plant energy balance is provided in tabular form in Exhibit 5-8. The cooling tower load includes the condenser, flue gas absorber pre-cooler, ASU, pellet reactor and CO_2 compressor intercooler, and other miscellaneous cooling loads.

Exhibit 5-8. Case 1 overall energy balance (0 °C [32 °F] reference)

	HHV	Sensible + Latent	Power	Total
Heat In GJ/hr (MMBtu/hr)				
Natural Gas (CT)	504 (477)	0.3 (0.3)	–	504 (478)
Air (CT)	–	15 (14)	–	15 (14)
Natural Gas (Calciner)	836 (792)	0.6 (0.5)	–	836 (793)
Air Contactor Inlet Air	–	8,104 (7,681)	–	8,104 (7,681)
Potassium Hydroxide Makeup	–	0.0 (0.0)	–	0.0 (0.0)
Air Separation Unit Inlet Air	–	8.4 (8.0)	–	8.4 (8.0)
Pellet Reactor Seed Material	–	-118 (-112)	–	-118 (-112)
Raw Water Makeup	–	124 (118)	–	124 (118)
Auxiliary Power	–	–	234 (222)	234 (222)
TOTAL	1,339 (1,270)	8,134 (7,710)	234 (222)	9,708 (9,201)
Heat Out GJ/hr (MMBtu/hr)				
Air Contactor Outlet Air	–	8,302 (7,869)	–	8,302 (7,869)
Fines Disposal	–	-7.4 (-7.0)	–	-7.4 (-7.0)
Pellet Reactor Solids Disposal	–	-36 (-34)	–	-36 (-34)
Slaker Solids Disposal	–	-70 (-66)	–	-70 (-66)
Air Separation Unit Vent Air	–	6.3 (5.9)	–	6.3 (5.9)
Cooling Tower Load[A]	–	1,174 (1,113)	–	1,174 (1,113)
CO_2 Product Stream	–	-42 (-39)	–	-42 (-39)
Deaerator Vent	–	0.0 (0.0)	–	0.0 (0.0)
Ambient Losses[B]	–	28 (27)	–	28 (27)
Power	–	–	282 (267)	282 (267)
TOTAL	–	9,356 (8,868)	282 (267)	9,638 (9,135)
Unaccounted Energy[C]	–	70 (66)	–	70 (66)

[A]Includes the condenser, flue gas absorber pre-cooler, ASU, pellet reactor and CO_2 compressor intercooler, and other miscellaneous cooling loads.
[B]Ambient losses include all losses to the environment through radiation, convection, etc. Sources of these losses include the combustor, superheater, and transformers
[C]By difference

5.2 Case 1 – Equipment List

Major equipment items for the total plant (NGCC plant and DAC system) are shown in the following tables. The accounts used in the equipment list correspond to the account numbers used in the cost estimates in Section 5.3. In general, the design conditions include a 10 percent contingency for flows and heat duties and a 21 percent contingency for heads on pumps and fans.

Case 1 – Account 3: Feedwater and Miscellaneous Balance of Plant Systems

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Condensate Pumps	Vertical canned	1,120 lpm @ 1410 m H_2O (290 gpm @ 4610 ft H_2O) 870 lpm @ 160 m H_2O (230 gpm @ 540 ft H_2O) 290 lpm @ 160 m H_2O (80 gpm @ 520 ft H_2O)	3	1
2	Auxiliary Boiler	Shop fabricated, water tube	18,000 kg/hr, 2.8 MPa, 343°C (40,000 lb/hr, 400 psig, 650°F)	1	0
3	Service Air Compressors	Flooded screw	13 m^3/min @ 0.7 MPa (450 scfm @ 100 psig)	2	1
4	Instrument Air Dryers	Duplex, regenerative	13 m^3/min (450 scfm)	2	1
5	Closed Cycle Cooling Heat Exchangers	Plate and frame	13 GJ/hr (13 MMBtu/hr)	2	0
6	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	5,200 lpm @ 20 m H_2O (1,400 gpm @ 70 ft H_2O)	2	1
7	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 110 m H_2O (1,000 gpm @ 350 ft H_2O)	1	1
8	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 80 m H_2O (700 gpm @ 250 ft H_2O)	1	1
9	Raw Water Pumps	Stainless steel, single suction	5,100 lpm @ 20 m H_2O (1,300 gpm @ 60 ft H_2O)	2	1
10	Filtered Water Pumps	Stainless steel, single suction	150 lpm @ 50 m H_2O (40 gpm @ 160 ft H_2O)	2	1
11	Filtered Water Tank	Vertical, cylindrical	145,000 liter (38,000 gal)	1	0
12	Makeup Water Demineralizer	Multi-media filter, cartridge filter, RO membrane assembly and electro-deionization unit	330 lpm (90 gpm)	1	0
13	Liquid Waste Treatment System	–	10 years, 24-hour storm	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
14	Gas Pipeline	Underground, coated carbon steel, wrapped cathodic protection	20 m^3/min @ 3.2 MPa (698 acfm @ 460 psia) 39 cm (16 in) standard wall pipe	16 km (10 mile)	0
15	Gas Metering Station	–	20 m^3/min (698 acfm)	1	0

Case 1 – Account 5: CO_2 Compression

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CO_2 Dryer	Triethylene glycol	Inlet: 81 m^3/min @ 2.1 MPa (2,847 acfm @ 305 psia) Outlet: 2.0 MPa (291 psia) Water Recovered: 418 kg/hr (921 lb/hr)	1	0
2	CO_2 Compressor	Reciprocating compressor	4.0 m^3/min @ 15.3 MPa, 150°C (157 acfm @ 2,218 psia, 302°F)	1	0
3	CO_2 Aftercooler	Shell and tube heat exchanger	Outlet: 15.3 MPa, 30°C (2,215 psia, 86°F) Duty: 54 GJ/hr (51 MMBtu/hr)	1	0

Case 1– Account 6: Combustion Turbine and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Combustion Turbine	LM6000 Aeroderivative	50 MW	1	0
2	Combustion Turbine Generator	Hydrogen Cooled	60 MVA @ 0.9 p.f., 18 kV, 60 Hz, 3-phase	1	0

Case 1 – Account 7: HRSG, Ductwork, and Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	CS plate, type 409SS liner	46 m (150 ft) high x 4.1 m (14 ft) diameter	1	0
2	Heat Recovery Steam Generator	Drum, multi-pressure with economizer section	IP steam – 66,539 kg/hr, 4.1 MPa/420°C (146,694 lb/hr, 595 psig/788°F) LP steam – 64,433 kg/hr, 0.4 MPa/174°C (142,050 lb/hr, 60 psig/346°F)	1	0
3	SCR Reactor	Space for spare layer	540,000 kg/hr (1,190,000 lb/hr)	1	0
4	SCR Catalyst	–	Space available for an additional catalyst layer	1 layer	0
5	Dilution Air Blowers	Centrifugal	1.3 m^3/min @ 108 cm WG (40 scfm @ 42 in WG)	1	1
6	Ammonia Feed Pump	Centrifugal	0.3 lpm @ 90 m H_2O (0.1 gpm @ 300 ft H_2O)	1	1
7	Ammonia Storage Tank	Horizontal tank	4,000 liter (1,000 gal)	1	0

Case 1 – Account 8: Steam Turbine and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	28 MW 4.1 MPa/420°C (595 psig/788°F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	30 MVA @ 0.9 p.f., 18 kV, 60 Hz, 3-phase	1	0
3	Surface Condenser	Two pass, divided waterbox including vacuum pumps and integrated deaerator	290 GJ/hr (280 MMBtu/hr), Inlet water temperature 16°C (60°F), Water temperature rise 11°C (20°F)	1	0
4	Steam Bypass	One per HRSG	50% steam flow @ design steam conditions	1	0

Case 1 – Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	232,000 lpm @ 30 m (61,000 gpm @ 100 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11°C (51.5°F) wet bulb/16°C (60°F) CWT/ 27°C (80°F) HWT/ 1,290 GJ/hr (1,220 MMBtu/hr) heat duty	1	0

Case 1 – Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CTG Transformer	Oil-filled	18 kV/345 kV, 60 MVA, 3-ph, 60 Hz	2	0
2	STG Transformer	Oil-filled	18 kV/345 kV, 30 MVA, 3-ph, 60 Hz	1	0
3	High Voltage Auxiliary Transformer	Oil-filled	345 kV/13.8 kV, 13 MVA, 3-ph, 60 Hz	2	0
4	Medium Voltage Transformer	Oil-filled	18 kV/4.16 MVA, 35 MVA, 3-ph, 60 Hz	1	1
5	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 11 MVA, 3-ph, 60 Hz	1	1
6	CTG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	18 kV, 3-ph, 60 Hz	1	0
7	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	18 kV, 3-ph, 60 Hz	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
8	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
9	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
10	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Case 1 – Account 12: Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS – Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser black and white)	Operator stations/printers and engineering stations/printers	1	0
2	DCS – Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS – Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

Case 1 – Account 15: Direct Air Capture

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	CO_2 Absorber	Conventional counterflow gas-liquid column	541,000 kg/hr (1,194,000 lb/hr) 5.2 wt% CO_2 concentration	1	0
2	CO_2 Absorber Blower	-	499,859 kg/hr @ 0.1 MPa (1,102,000 lb/hr @ 16 psia)	1	0
3	Recirculation Pump (P-1011)	Centrifugal	370,954 lpm @ 10 m H_2O (97,995 gpm @ 20 ft H_2O)	2	1
4	Booster Pump to Absorber (P-1012)	Centrifugal	111,253 lpm @ 20 m H_2O (29,390 gpm @ 70 ft H_2O)	3	0
5	Flue Gas Rich Solution Pump (P-6030)	Centrifugal	111,501 lpm @ 20 m H_2O (29,455 gpm @ 60 ft H_2O)	3	0
6	Air Contactor	Cross-flow configuration over plastic packing	156 m^3/hr of KOH	277	0
7	Air Fans	-	526,167 kg/hr @ 0.1 MPa, (1,160,000 lb/hr @ 15 psia)	277	0
8	KOH Sump	Steel basin	15,750 m^3 1.6 hr Residence Time	4	0
9	KOH Recycle Pump	Centrifugal	375,951 lpm @ 70 m H_2O (99,315 gpm @ 220 ft H_2O)	2	1
10	Pellet Reactor	Fluidized bed reactive crystallizer	22 tonnes/day CO_2 processed	178	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
11	Pellet Washer	-	19,656 kg/hr @ 386,994 kg/hr, (43,334 lb/hr @ 853,176 lb/hr)	20	0
12	Lime Distribution Tank	-	54 m^3/hr 0.5 hr Residence Time	15	0
13	Lime Mix Tank	-	323 m^3/hr 1 hr Residence Time	2	0
14	Lean Solution Tank	-	4,000 m^3 0.5 hr Residence Time	5	
15	Lime Distribution Pump (P-1510 and P-1535)	Centrifugal	8,585 lpm @ 20 m H_2O (2,268 gpm @ 70 ft H_2O)	2	0
16	Fluidization Pump (P1210)	Centrifugal	1,465,194 lpm @ 4 m H_2O (2,988,106 gpm @ 10 ft H_2O)	4	0
17	Primary Filtration	Pinned bed clarifier	12 m diameter	45	0
18	Primary Filtration Pump (P1700)	Centrifugal	968,810 lpm @ 7 m H_2O (1,975,784 gpm @ 20 ft H_2O)	2	0
19	Secondary Filtration	Filter Press	1,321 m^2 of filtering surface	2	0
20	Steam Slaker	Refractory lined bubbling/turbulent fluid bed (fluidized by steam)	850 m^3	2	0
21	ASU Main Air Compressor	Centrifugal, multi-stage	4,000 m^3/min @ 0.5 MPa (146,000 scfm @ 70 psia)	1	0
22	Cold Box	Vendor design	1,700 tonne/day (1,900 tpd) of 95% purity oxygen	1	0
23	Calciner	Oxygen-fired circulating fluidized bed	340,027 lb/hr $CaCO_3$	2	0

5.3 Case 1 – Cost Estimate Results

Exhibit 5-9 shows a detailed breakdown of the capital costs; Exhibit 5-10 shows the owner's costs, TOC, and TASC; Exhibit 5-11 shows the initial and annual O&M costs; Exhibit 5-12 provides guidance for scaling DAC-specific equipment costs; Exhibit 5-13 shows the COC breakdown. The capital cost estimate presented represents an AACE Class 5 estimate, with an uncertainty range of +/-50 percent. Estimates were developed on a basis for emerging technologies and designs similar to that used in the past for NETL reports. That is, the cost estimates for plant designs that include technologies that are not yet fully mature (e.g., DAC) use the same cost estimating methodology as for mature plant designs, which does not fully account for the unique cost premiums associated with the initial, complex integrations of emerging technologies in a commercial application. Thus, it is anticipated that initial deployments of these plants may incur costs higher than those reflected within this report. All major equipment components and features are based on commercially proven technology from reputable suppliers; no non-standard designs are required.

The capital costs reported for accounts 3 through 14 were scaled based on prior NETL reference cases, as these accounts represent generic balance of plant equipment. Costs reported in accounts 1, 2, and 15 were independently developed by Black & Veatch using a combination of in-house cost data, vendor references, and analogous references for novel equipment. O&M consumable unit costs were also provided by Black & Veatch.

Exhibit 5-9. Case 1 total plant cost details

Case:		DAC-1					Estimate Type:			Conceptual	
Plant Size (net tonnes CO_2/yr):		903,970					Cost Base:			March 2020	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/tonne (net)
1	Material Handling										
1.1	Pneumatic Conveyor (Slaker to Calciner)	$1,694	$368	$410	$0	$2,472	$494	$0	$445	$3,411	$4
1.2	KOH Makeup Loading to Air Contactor	$183	$0	$55	$0	$238	$48	$0	$43	$328	$0
1.3	Fines/Grids Disposal from Air Contactor	$89	$21	$46	$0	$156	$31	$0	$28	$215	$0
1.4	Seed Material to Pellet Reactor (incl. Solid Purge equip)	$1,353	$0	$244	$0	$1,597	$319	$0	$287	$2,204	$2
1.9	Foundations	$0	$748	$986	$0	$1,734	$347	$0	$312	$2,392	$3
	Subtotal	$3,318	$1,137	$1,740	$0	$6,196	$1,239	$0	$1,115	$8,550	$9
2	Material Preparation & Feed										
2.5	Solvent Preparation Equipment	$521	$22	$107	$0	$650	$130	$0	$117	$897	$1
2.6	Solvent Storage & Feed	$873	$0	$329	$0	$1,202	$240	$0	$216	$1,659	$2
2.9	Solvent Feed Foundation	$0	$350	$307	$0	$656	$131	$0	$118	$906	$1
	Subtotal	$1,394	$372	$743	$0	$2,508	$502	$0	$452	$3,462	$4
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$529	$907	$453	$0	$1,889	$378	$0	$340	$2,607	$3
3.2	Water Makeup & Pretreating	$8,725	$873	$4,944	$0	$14,542	$2,908	$0	$3,490	$20,941	$23
3.3	Other Feedwater Subsystems	$213	$70	$66	$0	$350	$70	$0	$63	$482	$1
3.4	Service Water Systems	$2,608	$4,979	$16,123	$0	$23,710	$4,742	$0	$5,690	$34,143	$38
3.5	Other Boiler Plant Systems	$230	$84	$209	$0	$523	$105	$0	$94	$722	$1
3.6	Natural Gas Pipeline and Start-Up System	$6,278	$270	$203	$0	$6,750	$1,350	$0	$1,215	$9,316	$10
3.7	Wastewater Treatment Equipment	$4,696	$0	$2,878	$0	$7,575	$1,515	$0	$1,818	$10,907	$12
3.9	Miscellaneous Plant Equipment	$6,018	$789	$3,059	$0	$9,866	$1,973	$0	$2,368	$14,208	$16
	Subtotal	$29,298	$7,971	$27,936	$0	$65,205	$13,041	$0	$15,079	$93,325	$103
5	CO_2 Compression										
5.4	CO_2 Compression & Drying	$15,449	$5,353	$5,575	$0	$26,376	$5,275	$0	$6,330	$37,982	$42
5.5	CO_2 Compressor Aftercooler	$476	$182	$179	$0	$837	$167	$0	$201	$1,205	$1
5.12	Gas Cleanup Foundations	$0	$387	$419	$0	$806	$161	$0	$193	$1,161	$1
	Subtotal	$15,925	$5,922	$6,172	$0	$28,019	$5,604	$0	$6,725	$40,348	$45

Case:		DAC-1						Estimate Type:			Conceptual	
Plant Size (net tonnes CO_2/yr):		903,970						Cost Base:			March 2020	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/tonne (net)
6	Combustion Turbine & Accessories										
6.1	Combustion Turbine Generator	$22,112	$0	$1,346	$0	$23,458	$4,692	$0	$4,222	$32,371	$36
6.3	Combustion Turbine Accessories	$804	$0	$49	$0	$853	$171	$0	$154	$1,177	$1
6.4	Compressed Air Piping	$0	$265	$60	$0	$326	$65	$0	$59	$449	$0
6.5	Combustion Turbine Foundations	$0	$277	$300	$0	$577	$115	$0	$138	$831	$1
	Subtotal	$22,916	$543	$1,754	$0	$25,213	$5,043	$0	$4,573	$34,829	$39
7	HRSG, Ductwork, & Stack										
7.1	Heat Recovery Steam Generator	$5,657	$0	$1,414	$0	$7,071	$1,414	$0	$1,273	$9,758	$11
7.2	Heat Recovery Steam Generator Accessories	$333	$0	$62	$0	$395	$79	$0	$71	$545	$1
7.3	Ductwork	$0	$206	$143	$0	$350	$70	$0	$63	$482	$1
7.4	Stack	$1,972	$0	$366	$0	$2,338	$468	$0	$421	$3,226	$4
7.5	Heat Recovery Steam Generator, Ductwork & Stack Foundations	$0	$154	$144	$0	$298	$60	$0	$72	$429	$0
7.6	Selective Catalytic Reduction System	$589	$247	$345	$0	$1,181	$236	$0	$213	$1,630	$2
	Subtotal	$8,551	$607	$2,475	$0	$11,633	$2,327	$0	$2,112	$16,071	$18
8	Steam Turbine & Accessories										
8.1	Steam Turbine Generator & Accessories	$4,768	$0	$698	$0	$5,466	$1,093	$0	$984	$7,543	$8
8.2	Steam Turbine Plant Auxiliaries	$28	$0	$62	$0	$90	$18	$0	$16	$124	$0
8.3	Condenser & Auxiliaries	$1,788	$0	$957	$0	$2,745	$549	$0	$494	$3,789	$4
8.4	Steam Piping	$1,067	$0	$432	$0	$1,499	$300	$0	$270	$2,069	$2
8.5	Turbine Generator Foundations	$0	$235	$388	$0	$622	$124	$0	$149	$896	$1
	Subtotal	$7,651	$235	$2,537	$0	$10,423	$2,085	$0	$1,913	$14,421	$16
9	Cooling Water System										
9.1	Cooling Towers	$7,092	$0	$2,148	$0	$9,240	$1,848	$0	$1,663	$12,751	$14
9.2	Circulating Water Pumps	$940	$0	$58	$0	$998	$200	$0	$180	$1,377	$2
9.3	Circulating Water System Auxiliaries	$7,475	$0	$986	$0	$8,461	$1,692	$0	$1,523	$11,676	$13
9.4	Circulating Water Piping	$0	$2,071	$1,875	$0	$3,946	$789	$0	$710	$5,445	$6
9.5	Make-up Water System	$471	$0	$605	$0	$1,075	$215	$0	$194	$1,484	$2
9.6	Component Cooling Water System	$305	$0	$234	$0	$539	$108	$0	$97	$744	$1
9.7	Circulating Water System Foundations	$0	$479	$795	$0	$1,274	$255	$0	$306	$1,835	$2
	Subtotal	$16,284	$2,550	$6,701	$0	$25,534	$5,107	$0	$4,673	$35,314	$39

Case:		DAC-1						Estimate Type:			Conceptual	
Plant Size (net tonnes CO_2/yr):		903,970						Cost Base:			March 2020	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies			Total Plant Cost	
				Direct	Indirect			Process	Project		$/1,000	$/tonne (net)
11						Accessory Electric Plant						
11.1	Generator Equipment	$685	$0	$517	$0	$1,202	$240	$0	$216		$1,659	$2
11.2	Station Service Equipment	$8,043	$0	$690	$0	$8,733	$1,747	$0	$1,572		$12,051	$13
11.3	Switchgear & Motor Control	$11,485	$0	$1,993	$0	$13,478	$2,696	$0	$2,426		$18,600	$21
11.4	Conduit & Cable Tray	$0	$2,777	$8,002	$0	$10,779	$2,156	$0	$1,940		$14,875	$16
11.5	Wire & Cable	$0	$4,145	$7,409	$0	$11,554	$2,311	$0	$2,080		$15,945	$18
11.6	Protective Equipment	$584	$0	$2,029	$0	$2,613	$523	$0	$470		$3,606	$4
11.7	Standby Equipment	$222	$0	$205	$0	$426	$85	$0	$77		$588	$1
11.8	Main Power Transformers	$374	$0	$8	$0	$382	$76	$0	$69		$527	$1
11.9	Electrical Foundations	$0	$19	$50	$0	$69	$14	$0	$17		$99	$0
	Subtotal	$21,394	$6,941	$20,902	$0	$49,237	$9,847	$0	$8,867		$67,951	$75
12						Instrumentation & Control						
12.1	Natural Gas Combined Cycle Control Equipment	$252	$0	$161	$0	$413	$83	$21	$77		$594	$1
12.2	Combustion Turbine Control Equipment	$430	$0	$274	$0	$705	$141	$35	$132		$1,013	$1
12.3	Steam Turbine Control Equipment	$404	$0	$258	$0	$662	$132	$33	$124		$952	$1
12.4	Other Major Component Control Equipment	$686	$0	$437	$0	$1,124	$225	$56	$211		$1,616	$2
12.5	Signal Processing Equipment	$566	$0	$17	$0	$583	$117	$29	$109		$838	$1
12.6	Control Boards, Panels & Racks	$151	$0	$92	$0	$243	$49	$12	$46		$349	$0
12.7	Distributed Control System Equipment	$8,413	$0	$257	$0	$8,670	$1,734	$434	$1,626		$12,464	$14
12.8	Instrument Wiring & Tubing	$695	$556	$2,223	$0	$3,473	$695	$174	$651		$4,993	$6
12.9	Other Instrumentation & Controls Equipment	$481	$0	$1,114	$0	$1,595	$319	$80	$299		$2,293	$3
	Subtotal	$12,079	$556	$4,834	$0	$17,469	$3,494	$873	$3,275		$25,111	$28
13						Improvements to Site						
13.1	Site Preparation	$0	$204	$4,327	$0	$4,531	$906	$0	$1,087		$6,525	$7
13.2	Site Improvements	$0	$655	$866	$0	$1,522	$304	$0	$365		$2,191	$2
13.3	Site Facilities	$629	$0	$660	$0	$1,289	$258	$0	$309		$1,856	$2
	Subtotal	$629	$859	$5,854	$0	$7,342	$1,468	$0	$1,762		$10,573	$12
14						Buildings & Structures						
14.1	Combustion Turbine Area	$0	$145	$77	$0	$222	$44	$0	$40		$306	$0

Case:		DAC-1					Estimate Type:			Conceptual	
Plant Size (net tonnes CO_2/yr):		903,970					Cost Base:			March 2020	
Item No.	Description	Equipment Cost	Material Cost	Labor		Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies		Total Plant Cost	
				Direct	Indirect			Process	Project	$/1,000	$/tonne (net)
14.3	Steam Turbine Building	$0	$902	$1,200	$0	$2,102	$420	$0	$378	$2,901	$3
14.4	Administration Building	$0	$198	$134	$0	$331	$66	$0	$60	$457	$1
14.5	Circulation Water Pumphouse	$0	$48	$24	$0	$71	$14	$0	$13	$98	$0
14.6	Water Treatment Buildings	$0	$674	$614	$0	$1,287	$257	$0	$232	$1,776	$2
14.7	Machine Shop	$0	$276	$177	$0	$452	$90	$0	$81	$624	$1
14.8	Warehouse	$0	$231	$139	$0	$370	$74	$0	$67	$510	$1
14.9	Other Buildings & Structures	$0	$208	$150	$0	$358	$72	$0	$64	$494	$1
14.10	Waste Treating Building & Structures	$0	$354	$632	$0	$986	$197	$0	$178	$1,361	$2
	Subtotal	$0	$3,034	$3,146	$0	$6,180	$1,236	$0	$1,112	$8,528	$9
15						Direct Air Capture System					
15.9	CO_2 Absorber System	$10,496	$11,996	$12,487	$0	$34,979	$6,996	$3,498	$6,821	$52,293	$58
15.10	Air Contactor System	$78,751	$22,500	$23,422	$0	$124,673	$24,935	$12,467	$24,311	$186,386	$206
15.11	Pellet Reactor System	$89,273	$51,013	$53,102	$0	$193,388	$38,678	$19,339	$37,711	$289,115	$320
15.12	Steam Slaker	$18,902	$6,481	$6,747	$0	$32,129	$6,426	$3,213	$6,265	$48,033	$53
15.13	ASU	$73,507	$16,802	$17,490	$0	$107,799	$21,560	$10,780	$21,021	$161,159	$178
15.14	Calciner	$35,704	$14,281	$14,867	$0	$64,852	$12,970	$6,485	$12,646	$96,953	$107
15.15	Foundations	$0	$12,305	$4,395	$0	$16,701	$3,340	$0	$3,006	$23,047	$25
	Subtotal	$306,633	$135,378	$132,509	$0	$574,520	$114,904	$55,782	$111,781	$856,987	$948
	Total	$446,071	$166,106	$217,303	$0	$829,480	$165,896	$56,655	$163,438	$1,215,469	$1,345

Exhibit 5-10. Case 1 owner's costs

Description	2020$/1,000	2020$/tonne
Pre-Production Costs		
6 Months All Labor	$7,966	$9
1 Month Maintenance Materials	$1,358	$2
1 Month Non-Fuel Consumables	$3,628	$4
1 Month Waste Disposal	$267	$0
25% of 1 Month's Fuel Cost at 100% CF	$1,024	$1
2% of TPC	$24,309	$27
Total	$38,552	$43
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$6,529	$7
0.5% of TPC (spare parts)	$6,077	$7
Total	$12,606	$14
Other Costs		
Initial Cost for Catalyst and Chemicals	$932	$1
Land	$778	$1
Other Owner's Costs	$182,320	$202
Financing Costs	$32,818	$36
TOC	**$1,483,477**	**$1,641**
TASC Multiplier (IOU, 33 year)	1.093	
TASC	**$1,620,916**	**$1,793**

Exhibit 5-11. Case 1 initial and annual operating and maintenance costs

Case:	DAC-1	Solvent DAC w/1x1 CT NGCC			Cost Base:	Mar-20
Plant Size:	**903,970**	**tonnes of CO_2 captured (net)**			**Capacity Factor (%):**	**85.0**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		1.0
Operating Labor Burden:		30.00	**% of base**	Operator:		3.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		2.0
				Lab Techs, etc.:		2.0
				Total:		**8.0**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonne-net)**
Annual Operating Labor:					$3,507,504	$4
Maintenance Labor:					$9,237,568	$10
Administrative & Support Labor:					$3,186,268	$4
Property Taxes and Insurance:					$24,309,390	$27
Total:					$40,240,730	$45
Variable Operating Costs						
					($)	**($/tonne-net)**
Maintenance Material:					$13,856,352	$15
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (gal/1000):	-	6,283	$1.90	$0	$3,703,643	$4
Makeup and Waste Water Treatment Chemicals (ton):	-	18.7	$550	$0	$3,193,656	$4
Ammonia (19 wt%, ton):	-	0.35	$300	$0	$32,513	$0
SCR Catalyst (ft^3):	706	0.387	$150	$105,882	$18,000	$0
Triethylene Glycol (gal):	w/equip.	466	$6.80	$0	$983,395	$1
Potassium Hydroxide (100 wt% pure; ton)	1,377	69.8	$600	$826,236	$13,001,025	$14
Calcium Carbonate (100 wt% pure; ton)	w/equip.	104	$500	$0	$16,068,945	$18
Subtotal:				$932,117	$37,001,177	**$41**
Waste Disposal						
SCR Catalyst (ft^3):	0	0.387	$2.50	$0	$300	$0.0
Triethylene Glycol (gal):	0	466	$0.35	$0	$50,616	$0.1
Solids to Disposal[A] (ton):	0	227	$38.00	$0	$2,671,191	$3.0
Subtotal:				$0	$2,722,107	$3.0
Variable Operating Costs Total:				$932,117	$53,579,636	$59
Fuel Cost						
Natural Gas (MMBtu):	0	30,469	$4.42	$0	$41,783,811	$46
Total:				**$0**	**$41,783,811**	**$46**

[A]Disposal Streams Contain: KOH, H_2O, $CaCO_3$, K_2CO_3

5.3.1 Cost Estimate Scaling

The majority of Case 1's cost estimate was scaled using Case B31B from NETL's BBR4 as a reference. [5] Guidance on scaling the balance of sub-accounts that relate to Case B31B can be found in NETL's QGESS: "Capital Cost Scaling Methodology: Revision 4 Report." [30] Exceptions to this approach include costs for the CT, HRSG (Case 1 considers a dual pressure HRSG, whereas Case B31B considers a triple pressure reheat HRSG), and steam turbine (similar to the HRSG, Case 1 only considers a lower pressure steam turbine). The remainder of the capital cost estimates for these accounts were developed by Black & Veatch using their in-house cost estimating references.

Black & Veatch also developed capital cost estimates for all sub-accounts in accounts 1, 2, and 15 using their in-house cost estimating references, which include vendor-supplied data. Scaling of these costs are not covered in the previously referenced QGESS report [30]; therefore, this section provides additional perspective. Exhibit 5-12 presents guidance on the scaling parameters of accounts 1, 2, and 15.

For certain equipment items, such as 15.9 and 15.12–15, scaling of cost follows the typical economy of scale approach. For equipment items 15.10 and 15.11, the number of individual units included in the plant process is large and it is expected that the unit cost would be maintained, but the total cost would change based on changing number of units required. Depending on the new size of the plant considered, the number of units required may necessitate slight adjustments in the size of a repeating unit. In this instance, it would be appropriate to scale the individual unit cost for size, but that new unit size cost should then be replicated for the new total number of units required.

Exhibit 5-12. Scaling parameters for DAC-specific equipment

Account Number	Item Description	Parameter	Range	Exponent
1	**SOLVENT HANDLING**			
1.1	Pneumatic Conveyor (Slaker to Calciner)	Calcium Carbonate Dry Pellets Flow Rate, lb/hr	571,000 – 952,000	0.6
1.2	KOH Makeup Loading to Air Contactor	KOH Makeup Rate (pure KOH), tpd	50 – 84	0.6
1.3	Fines/Grids Disposal from Air Contactor	Solids to Disposal, tpd	170 – 280	0.5
1.4	Seed Material to Pellet Reactor (incl. Solid Purge equip)	Calcium Carbonate Seed Material Required, tpd	78 – 130	0.6
1.9	Foundations	KOH Makeup Rate (tpd) + $CaCO_3$ Makeup Rate (tpd)	130 – 210	0.6
2	**SOLVENT PREPARATION & FEED**			
2.5	Solvent Preparation Equipment	KOH Makeup Rate (tpd) + $CaCO_3$ Makeup Rate (tpd)	130 – 210	0.6
2.6	Solvent Storage & Feed	KOH Makeup Rate (tpd) + $CaCO_3$ Makeup Rate (tpd)	130 – 210	0.6
2.9	Solvent Feed Foundation	KOH Makeup Rate (tpd) + $CaCO_3$ Makeup Rate (tpd)	130 – 210	0.6
15	**DAC SYSTEM – SOLVENT**			
15.9	CO_2 Absorber System	Gas Flow to CO_2 Absorber, acfm	181,000 – 302,000	0.7
15.10	Air Contactor System	Gas Flow to Air Contactor, acfm	95,700,000 – 159,500,000	1.0
15.11	Pellet Reactor System	Pellet Reactor CO_2 Processed, tpd	2,900 – 4,900	1.0
15.12	Steam Slaker	Calcium Carbonate Dry Pellets, lb/hr	571,000 – 952,000	0.7
15.13	ASU	Oxygen Production, tpd	1,400 – 2,200	0.7
15.14	Calciner	Calcium Carbonate Dry Pellets, lb/hr	571,000 – 952,000	0.7
15.15	Foundations	Gross DAC Capture, tonne CO_2/yr	1,058,000 – 1,763,000	0.8

5.3.2 Cost of CO_2 Capture Results

Using the methodology presented in Section 3.6, Exhibit 5-13 presents the results for the COC for Case 1.

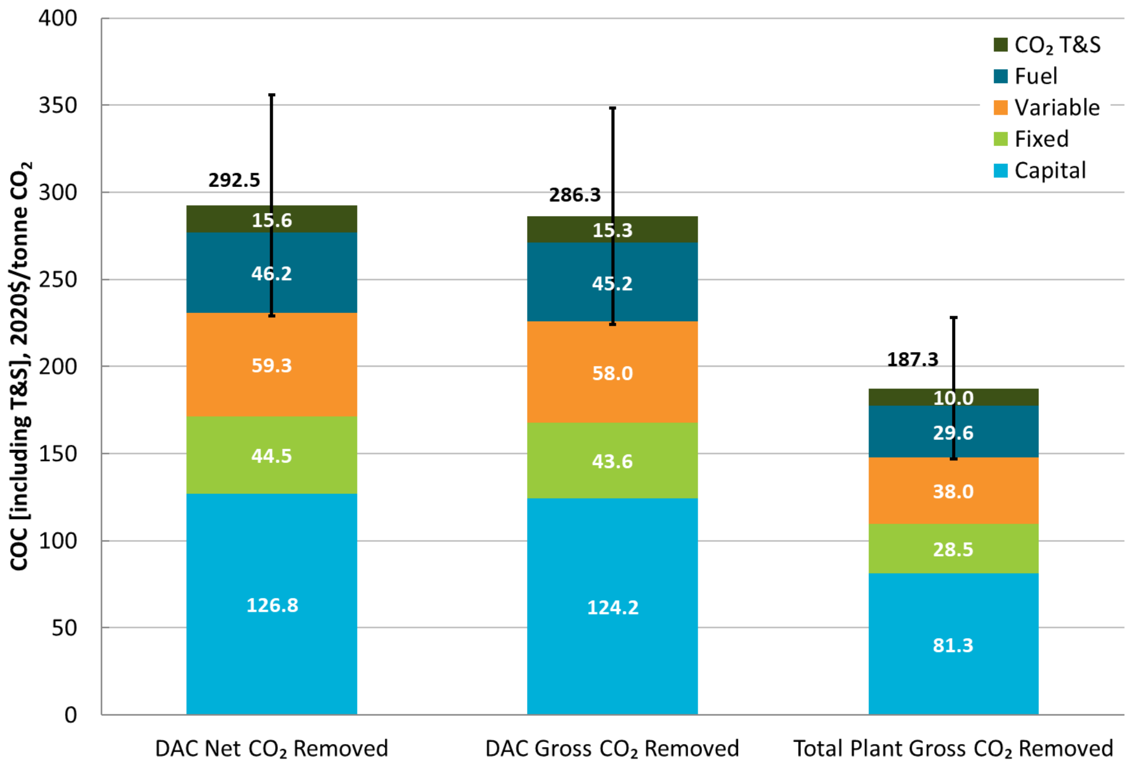
Exhibit 5-13. Case 1 COC

Component	COC DAC_{net}, 2020$/tonne	COC DAC_{gross}, 2020$/tonne	COC $Plant_{gross}$, 2020$/tonne
Capital	126.8	124.2	81.3
Fixed	44.5	43.6	28.5
Variable	59.3	58.0	38.0
Fuel	46.2	45.2	29.6
Total (Excluding T&S)	**276.9**	**271.0**	**177.3**
CO_2 T&S	15.6	15.3	10.0
Total (Including T&S)	**292.5**	**286.3**	**187.3**

For the COC DAC_{net} result of $293/tonne CO_2 (including T&S), a total CO_2 flow of 903,970 tonnes/yr is used. For the COC DAC_{gross} result of $286/tonne CO_2 (including T&S), a total CO_2 flow of 923,527 tonnes/yr is used; this represents the gross CO_2 captured from the atmosphere and the flue gas CO_2 that is captured in the air contactor. For the COC $Plant_{gross}$ result of $187/tonne CO_2 (including T&S), a total CO_2 flow of 1,411,282 tonne/yr is used, which represents the gross CO_2 captured by the DAC system from the atmosphere, the CO_2 captured from the CT exhaust gas, plus the CO_2 captured from the oxy-fired calciner. It is important to note that the financial assumptions used in Case 1 can be considered overly optimistic, and applying financial parameters that reflect the risk associated with a new technology like DAC would result in a higher COC.

Exhibit 5-14 presents the COC results graphically and includes error bars relating to the uncertainty in the capital cost estimate. As highlighted previously, the capital estimate represents an AACE Class 5 estimate, with an uncertainty range of +/-50 percent. The COC ranges presented are not reflective of other uncertainty, such as variation in fuel price, labor price, capacity factor, or other factors.

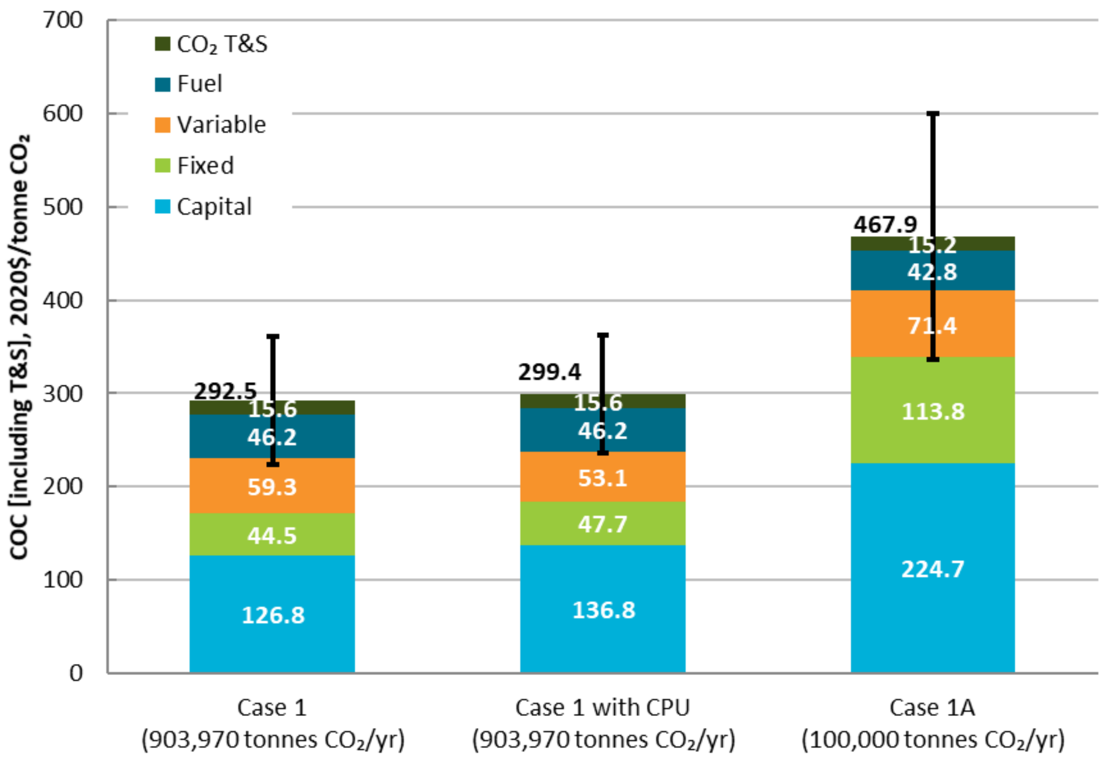
Exhibit 5-14. Case 1 COC plot and uncertainty ranges



Note: Error bars represent uncertainty range of capital cost estimates (+/- 50%)

Exhibit 5-15 illustrates the net COC for Case 1 compared to other cases evaluated. At the smaller scale of 100,000 tonnes/yr CO_2 (Case 1A), diseconomy of scale raises the COC result to $467/tonne. Because the CO_2 product purity in Case 1 does not meet CO_2 purity guidelines, an additional case is included that evaluates the addition of a CO_2 compression and purification unit (CPU); this case also sells excess electricity generated (13.5 MW) at $60/MWh. Including a CPU and selling excess electricity to the grid results in a net COC of $299/tonne.

Exhibit 5-15. Cost of CO_2 capture results for Case 1 (net CO_2 removed basis) compared to modified cases



Note: Error bars represent uncertainty range of capital cost estimates (+/- 50%)

5.3.3 Sensitivity Analysis

Given data gaps in the literature and the early technology readiness level of DAC technology overall, a sensitivity analysis was conducted on multiple process and cost parameters to gauge their impact on the final system performance and COC. The parameters of interest include total plant capital cost, air contactor capital cost, capacity factor, calciner natural gas requirement, capture fraction, system pressure drop, solvent cost, solvent lifetime, natural gas price, and financing assumptions (FCR). Exhibit 5-16 summarizes the sensitivity study results described in this section and plots the potential impacts such that the importance of different parameters can be weighed against each other. The sensitivity analysis did not identify any key process parameters that could have significant impact on the results, if improved. Conversely, financial parameters, like FCR, have a significant impact on the results. Due to the risk associated with this technology coupled with an uncertain future market, it is difficult to identify the appropriate financial structure.

Exhibit 5-16. Summary of COC sensitivity results

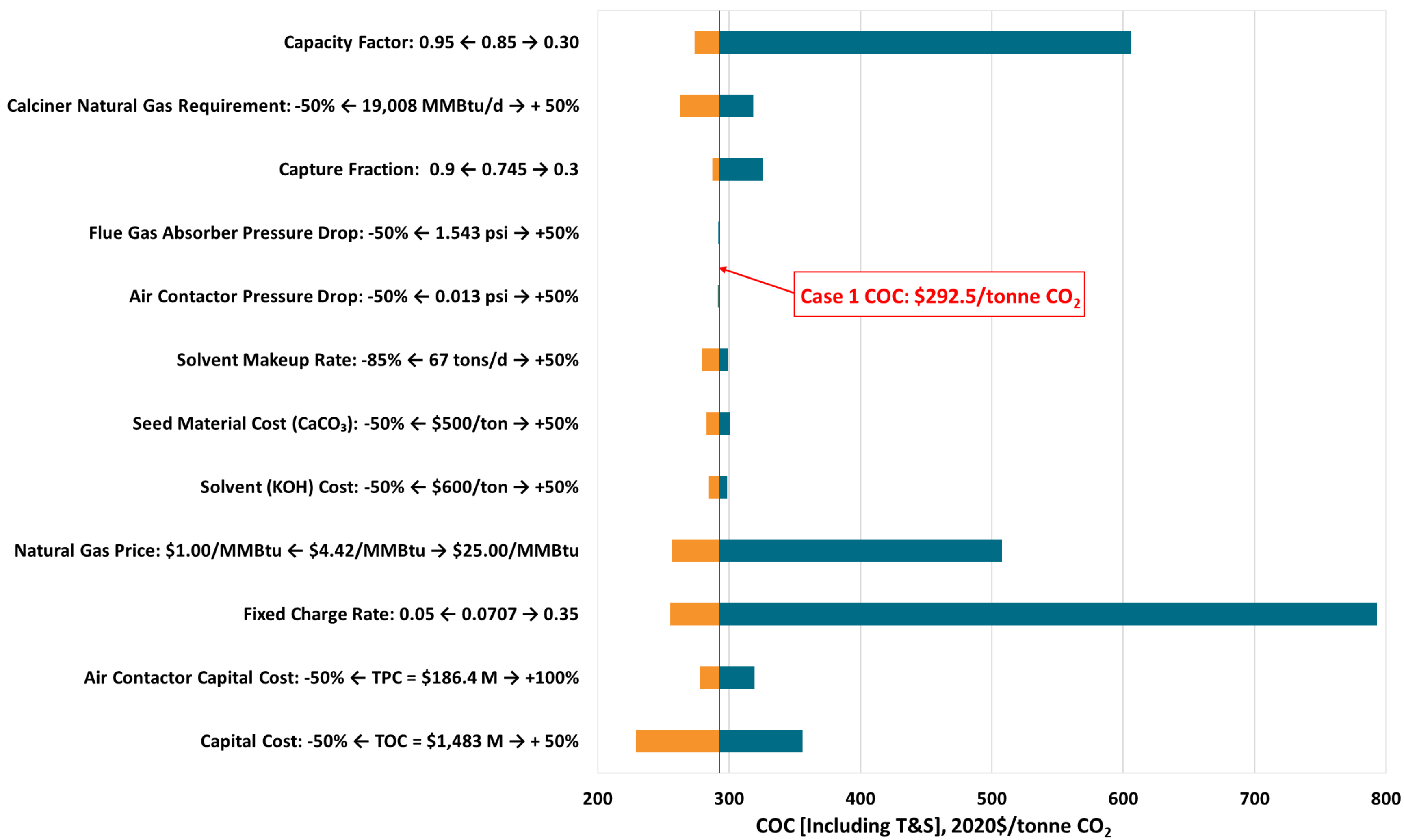


Exhibit 5-17 shows the COC sensitivity to natural gas price for the three different bases of calculation (DAC_{net}, DAC_{gross}, and Total $Plant_{Gross}$). The natural gas price is varied over the range of \$0.95/GJ (\$1/MMBtu–77 percent reduction from the reference) to \$23.75/GJ (\$25/MMBtu–466 percent increase from the reference). This range encompasses Henry Hub spot market prices over the past 20 years. The results show that at the low natural gas price point, COC is reduced by 12 percent versus the reference; whereas at the high natural gas price point, COC increases by 74 percent versus the reference. Fuel price accounts for approximately 16 percent of the COC (including T&S); therefore, COC is not overly sensitive to small changes to the price of natural gas, but large increases in natural gas price will have a significant impact on COC.

Exhibit 5-17. COC sensitivity to natural gas price

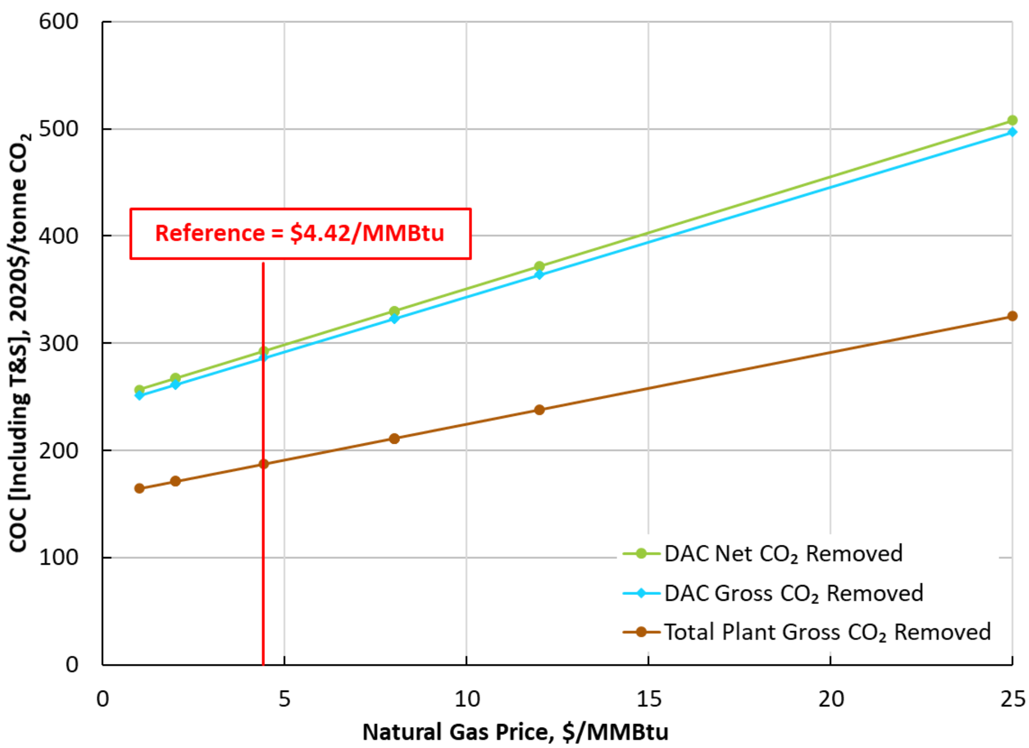


Exhibit 5-18 shows the COC sensitivity to solvent cost for the three different bases of calculation. The solvent cost range considered is +/-50 percent around the reference solvent cost, quoted by Black & Veatch. The maximum and minimum parameter ranges result in only a 2 percent increase or decrease in the COC. For Case 1, variable O&M accounts for only 20 percent of the COC. Of this 20 percent, the annual solvent make-up cost accounts for 24 percent of the variable O&M; therefore, the solvent cost is found to be a relatively non-impactful parameter in terms of COC.

Exhibit 5-18. COC sensitivity to solvent cost

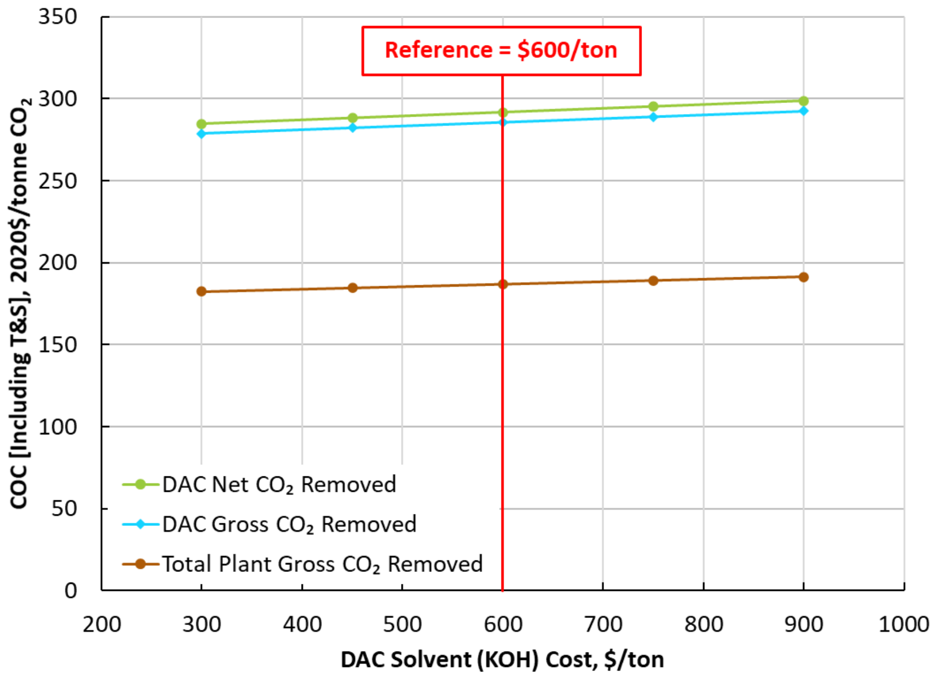


Exhibit 5-19 shows the COC sensitivity to solvent makeup rate for the three different bases of calculation. The reference case, based on engineering judgement and CE-reported values, requires 67 tons/day of solvent makeup. A wide sensitivity range spanning 10 tons/day (-85 percent) to 100 tons/day (+50 percent) was selected; the COC shows about a 4 percent decrease at the low end and a 2 percent increase at the high end of this range.

Exhibit 5-19. COC sensitivity to solvent makeup rate

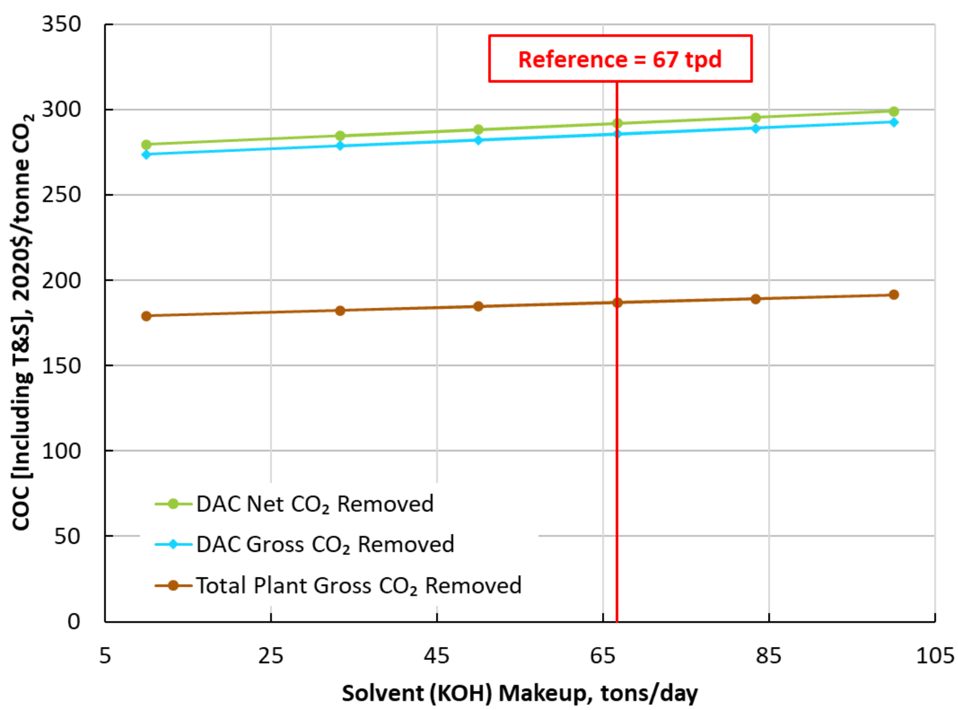


Exhibit 5-20 shows the COC sensitivity to FCR for the three different bases of calculation. The reference case assumes an FCR of 0.0707, which is the value used for NGCC plant LCOE calculations as discussed previously in Section 3.6. This value was selected based on an assumed three-year construction period. Financial parameters that would be most realistic for the DAC process are presently undefined as there are no large-scale commercial projects to date. Considering the high-risk associated with the lack of maturity of DAC technology, the reference case FCR is likely favorably low for Case 1. Alternate financial parameter assumptions accounting for the high-risk nature of this technology would likely result in a higher FCR. However, when special financial considerations or programs (e.g., loan guarantee programs) are accounted for, the reference case FCR value could be considered high. For this reason, a wide FCR sensitivity range was selected to span all possible scenarios.

Given the slope of the lines in Exhibit 5-20, the resulting COC is significantly impacted by the FCR. Doubling the FCR would result in approximately a 43 percent increase in the COC, and reducing the FCR to 0.05 would result in approximately a 13 percent decrease in the COC.

As will be outlined in Section 7, a deeper analysis of possible financial scenarios should be performed.

Exhibit 5-20. COC sensitivity to fixed charge rate

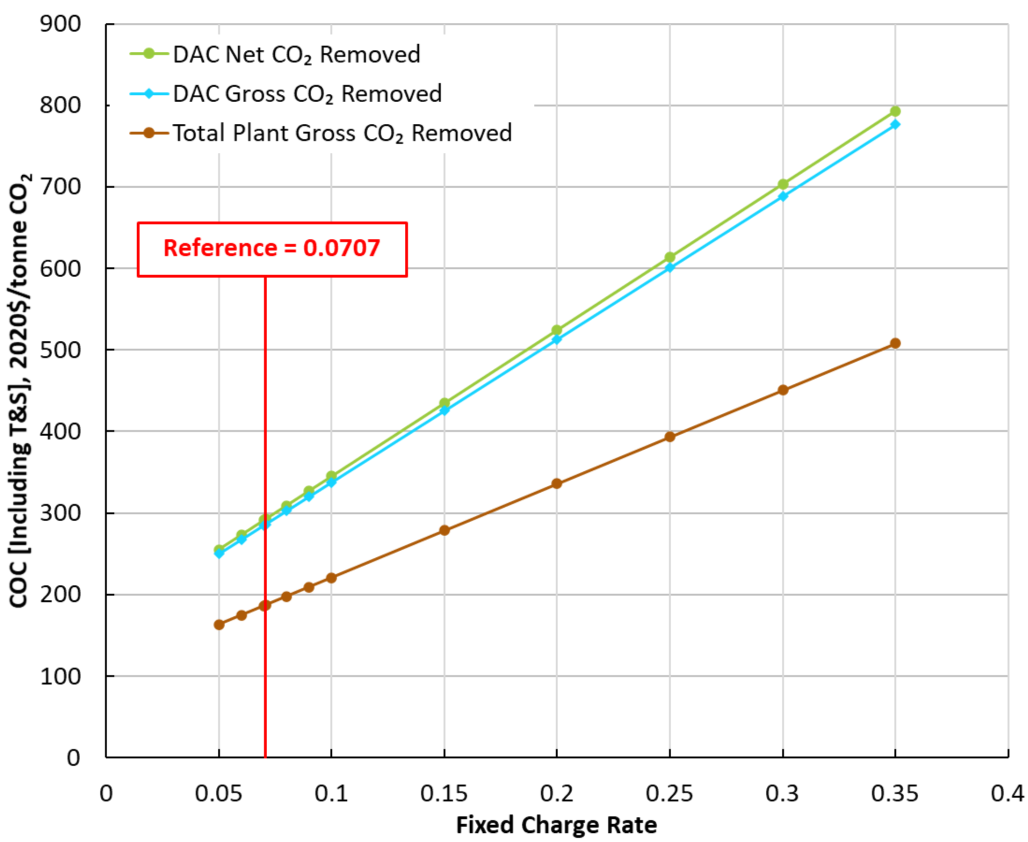
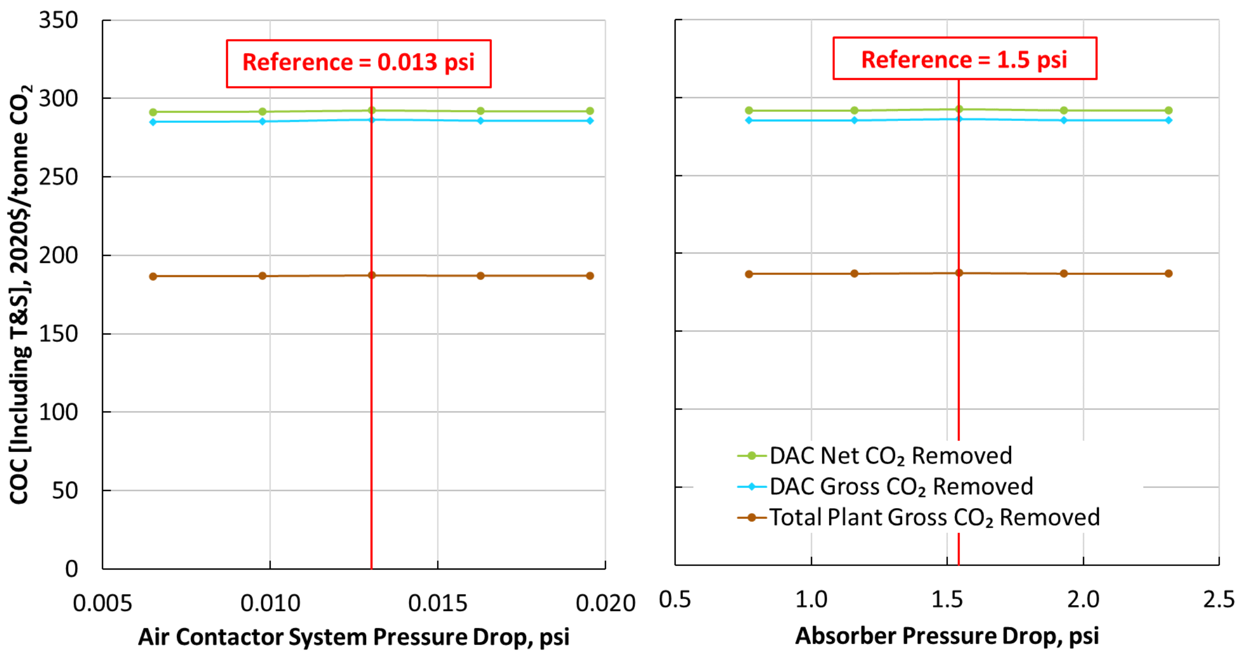


Exhibit 5-21 shows the COC sensitivity to system pressure drop across the air contactor and the flue gas absorber. Based on engineering judgement and CE-reported values, the pressure drop across the air contactor and the absorber in Case 1 are 0.013 psi and 1.5 psi, respectively. Over the range considered by the sensitivity to these variables (+/-50 percent), the impact on COC is negligible; this is because the capital cost estimate and scaling methodology are not granular enough to show a response to these changing process variables. The compressor capital costs are not broken out from the air contactor capital costs, and therefore, changes in the compressor parameters are not reflected in the cost estimate. The primary sub-accounts affected by the pressure drop are those that have costs scaled on the auxiliary load.

Exhibit 5-21. COC sensitivity to system pressure drop



In order to compare Case 1 with sorbent DAC technologies (that may experience larger pressure drops across the air contactor) an additional sensitivity that extends the range of pressure drops considered is shown in Exhibit 5-22. A pressure drop of 0.7 psi across the air contactor (increased from 0.013 psi in the reference case) results in a 71 percent increase in the COC. The sensitivity case with a 0.7 psi pressure drop requires more electricity than is generated onsite; it is assumed that the additional electricity required (approximately 361 MW) is purchased for $60/MWh. Since no life cycle analysis has been conducted as part of this study, and only inside the plant fence line emissions are characterized, no carbon footprint for this electricity purchase is assumed. A full analysis building in carbon footprint for the purchased electricity would increase the costs shown in the plot. It is worth noting that pressure drop is only one of the design parameters for an air contactor. These parameters would have to be considered in conjunction to evaluate the tradeoffs and for cost optimization, which is beyond the scope of this analysis.

Exhibit 5-22. COC sensitivity to air contactor pressure drop

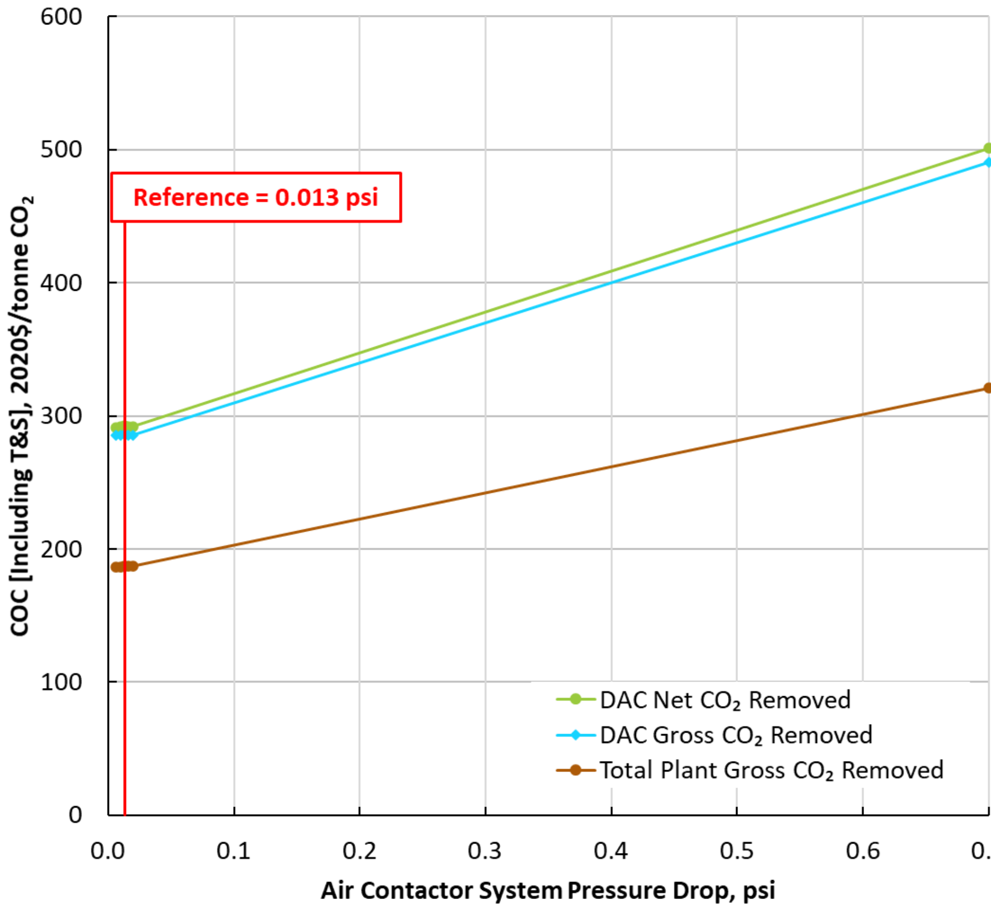


Exhibit 5-23 shows the COC sensitivity to DAC system capture fraction for the three different bases of calculation. Based on CE-reported performance, the reference Case 1 assumes that the air contactors remove 74.5 percent of the inlet CO_2 present in the combined air and flue gas stream. A wide sensitivity range was selected to evaluate the impact of capture fraction on the COC. As the capture fraction increases, less ambient air is required to meet the target net capture rate, reducing auxiliary load and air contactor costs. At a capture rate of 30 percent, 1.46 billion lb/hr of ambient air must be treated in the air contactors; this is 151 percent more air than is required in Case 1 (0.58 billion lb/hr). For comparison, at a capture rate of 90 percent, 0.48 billion lb/hr of ambient air must be treated in the air contactors, 18 percent less than Case 1. Because the cost of the air contactor is only a small portion of the capital cost (15 percent of the TPC), and the cost estimate is not granular enough to account for differences in compressor costs (as discussed in the pressure drop sensitivities), the impact of this parameter is minor. In the sensitivity case considering a capture fraction of 30 percent, more electricity is required than is produced onsite; in this case, it is assumed that the required additional electricity (approximately 4.5 MW) is purchased for $60/MWh. Since no life cycle analysis has been conducted as part of this study, and only inside the plant fence line emissions are characterized, no carbon footprint for this electricity purchase is assumed. A full analysis building in carbon footprint for the purchased electricity would increase the cost.

Exhibit 5-23. COC sensitivity to DAC system capture fraction

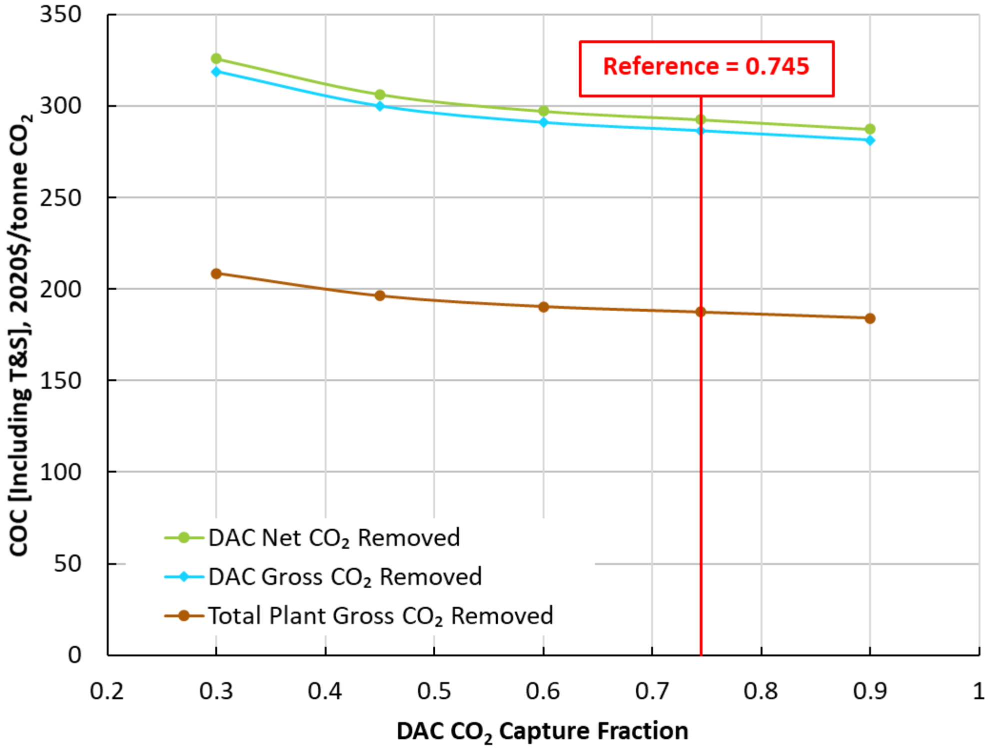


Exhibit 5-24 shows the COC sensitivity to the calciner natural gas requirement for the three different bases of calculation. In the modeled solvent DAC process, the CO_2 captured from the air and flue gas is released through the thermal decomposition of the $CaCO_3$ pellets. Based on results from the process model, which leverages CE process data, to achieve this thermal decomposition, the oxy-fired calciner in Case 1 requires 19,008 MMBtu/day of natural gas. A sensitivity range of +/-50 for the calciner natural gas requirement results in a 9 percent increase or 10 percent decrease in the net COC. As the calciner natural gas requirement increases, the total plant gross COC decreases due to the increased CO_2 product flow from the calciner.

Exhibit 5-24. COC sensitivity to calciner natural gas requirement

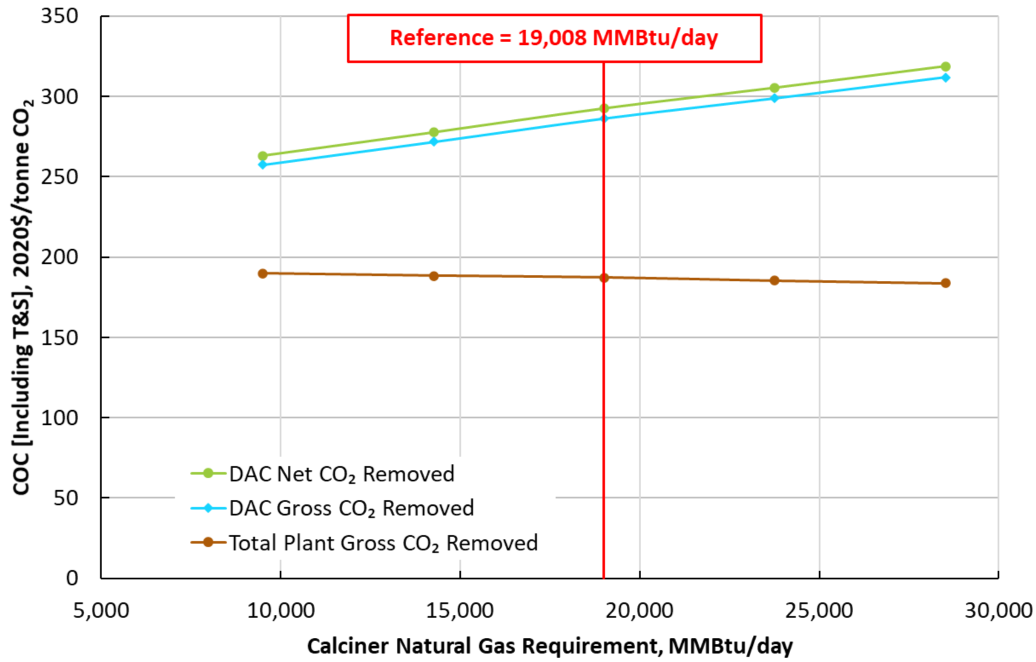
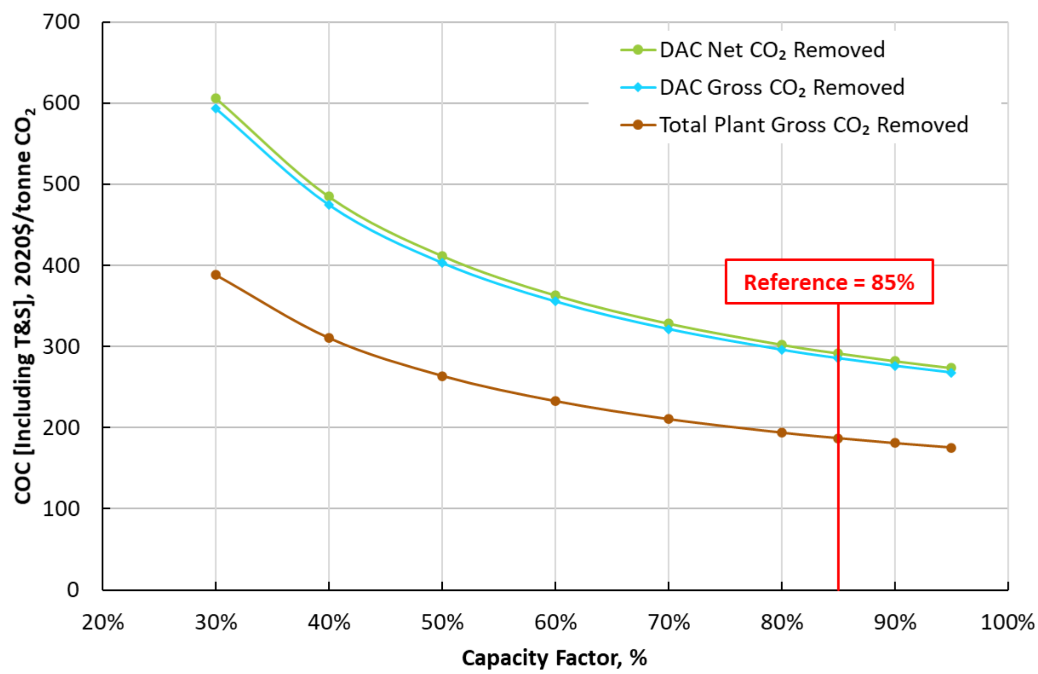


Exhibit 5-25 shows the COC sensitivity to DAC system capacity factor for the three different bases of calculation. The reference Case 1 assumes a capacity factor of 85 percent. This value is generally in line with NETL carbon capture studies and was selected to allow comparison across studies. A wide sensitivity was selected to evaluate the impact of the capacity factor. As expected, as the capacity factor of the DAC plant reduces, the COC increases rapidly, indicating that high capacity factors will be required for a DAC plant to be economically competitive.

Exhibit 5-25. COC sensitivity to capacity factor



The reference Case 1 CO_2 product purity does not meet CO_2 product purity specifications. Specifically, the O_2 concentration is greater than allowed in the specifications (10–100 ppmv depending on use). [6] Therefore, a single sensitivity case was considered where the CO_2 compressor was removed and replaced with a cryogenic CPU.

The CPU data was sourced from a prior NETL report that examined advanced oxy-combustion technologies for coal-fired power plants. [31] Salient data for the CPU as presented in the reference is shown in Exhibit 5-26. For perspective, the relevant Case 1 parameter values are also provided.

As highlighted in Exhibit 5-26, the reference CPU system purifies a stream with an inlet CO_2 concentration of 71.6 mole percent; this difference in inlet purity may introduce minor inconsistencies in the cost estimate results.

Exhibit 5-26. Reference CPU data

Parameter	CPU Reference Value	DAC Case 1 Value
Inlet Flow Rate, lb/hr	1,221,161	502,031
Inlet CO_2 Purity, mole %	71.58	95.5
Inlet Pressure, psia	14.8	14.7
Inlet Temperature, °F	135	371
Outlet CO_2 Product Purity, mole %	99.99	-
Outlet Product Pressure, psig	2,200	-
Total Plant Cost, 2018$/1000	242,814	-

Exhibit 5-27 presents the relevant cost comparison data for Case 1 and the sensitivity case that includes a CPU, as well as the final COC result. Application of the CPU capital cost maintained the same process and project contingencies that were assumed in the reference report, and the same engineering home office and fee percentage that has been applied to the DAC system in this study.

Exhibit 5-27. COC result for Case 1 versus Case 1 with a CPU

Component	Case 1	Case 1 with CPU
CO_2 Product Purity	95.54% CO_2 1.8% O_2	99.99% CO_2 10 ppmv O_2
CO_2 Compressors and Aftercooler TPC, 2020$/1000	40,348	-
Scaled CPU TPC, 2020$/1000	-	137,069
TPC, 2020$/1000	1,215,469	1,311,761
TPC, 2020$/tonne CO_2 net	1,345	1,456
COC, 2020$/tonne CO_2net	292.5	306.1
Percent Increase in COC, %	-	4.6

Replacement of the DAC CO_2 compressor with the CPU adds an additional $96 M TPC to the sensitivity case capital cost. This value represents approximately an 8 percent increase in the TPC, and results in a 4.6 percent increase in the COC.

The CPU cost applied in this sensitivity study inherently assumes a fixed inlet CO_2 purity, and if the DAC process were to provide a CO_2 product stream below this purity, the CPU capital cost, and COC result, would increase.

Case 1 and related sensitivity studies do not consider potential revenue from excess power generated in the process. Exhibit 5-28 shows the results of a single sensitivity case that includes CPU and the sale of excess power to the grid at $60/MWh. Selling the excess electricity generated in Case 1 with CPU (13.5 MW) at $60/MWh results in a 2.4 percent increase in the net COC relative to Case 1.

Exhibit 5-28. COC result for Case 1 with excess electricity sold to the grid

Component	Case 1	Case 1 with CPU and Excess Electricity Sold to Grid
Excess Electricity Generated, MW	13.2	13.5
Assumed Selling Price of Excess Electricity, $/MWh	0	60
COC, 2020$/tonne CO_{2net}	292.5	299.4
Percent Increase in COC, %	-	2.4

5.4 Case 1A – Performance and Cost Results

For evaluation at the 2018 45Q required scale, Case 1 was scaled-down from a net capture rate of 903,970 tonnes CO_2/yr to 100,000 tonnes CO_2/yr. Additional changes to the reference case include a change from the aeroderivative CT to the F-class CT and an adjustment in the CT power output to achieve a net power of approximately 0 MW. This approach, often referred to as a "rubber turbine" approach, allows for the elimination of excess power for sale, which could impact the overall economics of the DAC plant. It is acknowledged that this approach does not reflect today's off-the-shelf CT technology and, therefore, lessens the direct applicability of the case results to future systems that may be constructed.

The following sub-sections provide performance and cost results for Case 1A.

5.4.1 Performance Results

Exhibit 5-29 shows the performance results for Case 1A.

Exhibit 5-29. Performance results for Case 1A

	Case 1A (Solvent System)
Combustion Turbine Power, MWe	4
Steam Turbine Power, MWe	3
Total Gross Power, MWe	**7**
Auxiliary Loads	
Air Fans, kWe	800
ASU, kWe	1,560

	Case 1A (Solvent System)
CO_2 Compression, kWe	2,590
Balance of Plant, kWe	2,247
Total Auxiliaries, MWe	**7**
Net Power, MWe	**0**
Combustion Turbine Natural Gas Feed Flow, kg/hr (lb/hr)	865 (1,908)
Calciner Natural Gas Feed Flow, kg/hr (lb/hr)	1,761 (3,882)
Total Natural Gas Feed Flow, kg/hr (lb/hr)	**2,626 (5,790)**
CO_2 Balance	
CO_2 Captured from Air, tonne/yr	100,478
Combustion Turbine Flue Gas Captured CO_2, tonne/yr*	16,748
Calciner Captured CO_2, tonne/yr	34,700
CO_2 Emitted, tonne/yr	478
Net Capture Rate, tonne/yr	**100,000**
Total Plant CO_2 Flow to Storage, tonnes/yr	**151,926**

*Includes CO_2 captured in the flue gas absorber and air contactors

5.4.2 Cost Estimate Results

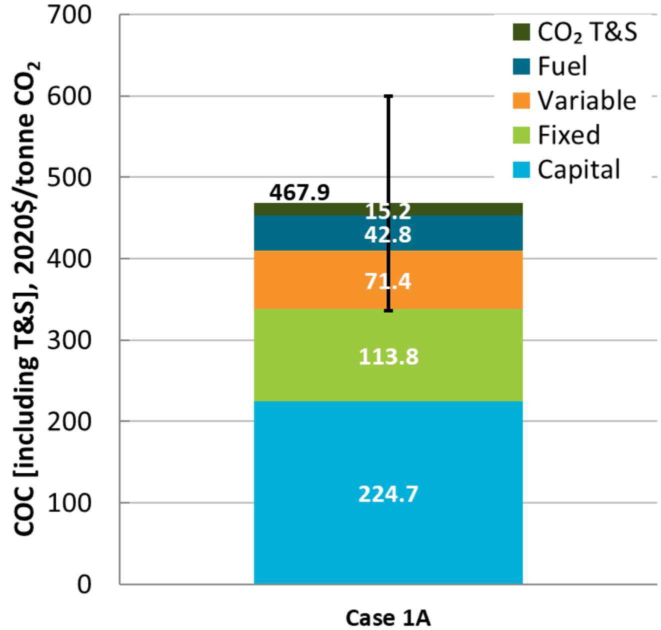
Exhibit 5-30 presents the capital cost estimate results for Case 1A. Costs for DAC-specific equipment are shown in account 15, while capital costs of the remainder of the plant are covered in accounts 1–14.

Exhibit 5-30. Cost results summary for Case 1A

	Case 1A (Solvent System)	
	2020$/1,000	**2020$/tonne (net)**
Accounts 1-14 Total Plant Cost	$107,893	$1,079
Account 15 (DAC) Total Plant Cost	$129,212	$1,292
TPC	$237,105	$2,371
TOC	$290,718	$2,907

Exhibit 5-31 shows the net COC results for Case 1A. Case 1A has a COC of $468/net tonne CO_2.

Exhibit 5-31. COC results for Case 1A



Note: Error bars represent uncertainty range of capital cost estimates (+/- 50%)

5.4.3 CO_2 Product Purity

Similar to Case 1, Case 1A DAC CO_2 product purity does not meet CO_2 product purity O_2 concentration specifications (10–100 ppmv depending on use). [6] Therefore, a single sensitivity case was considered where the CO_2 compressor was removed and replaced with a cryogenic CPU.

The CPU data was sourced from a prior NETL report that examined advanced oxy-combustion technologies for coal-fired power plants. [31] Salient data for the CPU as presented in the reference is shown in Exhibit 5-32. For perspective, the relevant Case 1A parameter values are also provided.

Exhibit 5-32. Reference CPU data

Parameter	CPU Reference Value	DAC Case 1A Value
Inlet Flow Rate, lb/hr	1,221,161	55,324
Inlet CO_2 Purity, mole%	71.58	95.5
Inlet Pressure, psia	14.8	14.7
Inlet Temperature, °F	135	371
Outlet CO_2 Product Purity, mole%	99.99	-
Outlet Product Pressure, psig	2,200	-
Total Plant Cost, 2018$/1000	242,814	-

As highlighted in Exhibit 5-32, the reference CPU system processes about 22 times more inlet gas than the DAC Case 1A system requires to be treated. This difference in scale may introduce minor inconsistencies in the cost estimate results. The reference CPU system also purifies a stream with an inlet CO_2 concentration of 71.6 mole percent. Deviations from this value for the DAC system may also introduce uncertainty in the sensitivity results presented.

Exhibit 5-33 presents the relevant cost comparison data for Case 1A and the sensitivity case that includes a CPU, as well as the final COC result. Application of the CPU capital cost maintained the same process and project contingencies that were assumed in the reference report, and the same engineering home office and fee percentage that has been applied to the DAC system in this study.

Exhibit 5-33. COC result for Case 1A versus Case 1A with a CPU

Component	Case 1A	Case 1A with CPU
CO_2 Product Purity	95.54% CO_2 1.8% O_2	99.99% CO_2 10 ppmv O_2
CO_2 Compressors and Aftercooler TPC, 2020$/1000	15,812	-
Scaled CPU TPC, 2020$/1000	-	29,273
TPC, 2020$/1000	237,105	250,566
TPC, 2020$/tonne CO_2net	2,371	2,514
COC, 2020$/tonne CO_2net	467.9	486.0
Percent Increase in COC, %	-	3.9

Replacement of the DAC CO_2 compressor with the CPU adds an additional $13.5 M TPC to the sensitivity case capital cost. This value represents approximately 6 percent increase in the TPC, and results in a 4 percent increase in the COC.

The CPU cost applied in this sensitivity study inherently assumes a fixed inlet CO_2 purity, and if the DAC process were to provide a CO_2 product stream below this purity, the CPU capital cost, and COC result, would increase.

6 CONCLUSIONS

In recent years, there has been a significant increase in research focused on DAC, but to date, the technology is not fully mature. There have been developers that have advanced to small pilot-scale testing of their processes and published projected cost estimates, [2] [1] but these technologies require further R&D to continue to reduce the COC.

The objective of this study is to provide cost and performance estimates for a solvent-based DAC system. The solvent case evaluated removes a net of 903,970 tonnes/yr of CO_2 from the atmosphere. Accounting for uncertainty in the capital cost estimates, the COC of CO_2 for this case is $230–355/net tonne CO_2, including costs for T&S of the captured CO_2. The optimistic end of these results approaches the high end of COC ranges reported in the literature for solvent DAC systems; in publicly available literature, CE has reported the net COC at $93–232/tonne CO_2. [1] In this study, Case 1 results in a net COC of $293/ton. A scaled-down version of Case 1, Case 1A described in Section 5.4, removes a net of 100,000 tonnes/yr CO_2 from the atmosphere. Due to diseconomies of scale, Case 1A results in a net COC of $468/tonne CO_2 net.

Several parameters were considered for sensitivity analysis, as detailed in Section 5.3.3. As noted, the combination of publicly available data (or lack thereof) and the resulting level of modeling detail plausible based on the available data limits the ability to quantify the impact of uncertainty in the reference case design. The sensitivity studies presented should be viewed as framing the downstream impacts on the overall system, rather than a direct analysis of the selected parameter. The sensitivity analysis did not identify any key process parameters, of those considered, that could have significant impact on the results if improved. For example, parameters such as the calciner natural gas requirement, the CO_2 removal efficiency of the air contactor, the pressure drop of the flue gas absorber or air contactor, and the solvent make-up rate, over the sensitivity range considered would impact the final COC, but the most significant COC reduction was shown to be only a 10 percent reduction when reducing the calciner natural gas requirement by 50 percent. The COC was shown to be relatively unaffected by certain material prices, as demonstrated by the +/-50 percent sensitivity to $CaCO_3$ and KOH purchase prices. The natural gas price sensitivity did demonstrate that fluctuations in the natural gas price can impact the final COC, with a natural gas price of $1.00/MMBtu resulting in a 12 percent reduction in COC, and a natural gas price of $25.00/MMBtu resulting in a 74 percent increase in COC. The base natural gas price was assumed to be $4.42/MMBtu. The two most impactful parameters considered were capacity factor and FCR. Case 1 was assumed to operate at a capacity factor of 85 percent, which leaves little room for improvement. However, if the capacity factor were to drop to the lower bound considered of 30 percent, the COC result was shown to more than double. As discussed in Section 5.3.3, the FCR selected for Case 1 reflects that expected for an NGCC plant, and was chosen due to the underlying three-year construction period assumed. It is unclear at present how future DAC projects will be financed, and how technology demonstration and maturation will aid in de-risking future deployments. Therefore, financial parameters represent a large unknown in this work, and the sensitivity analysis considered a broad range of scenarios, demonstrating that future DAC financing structures could have a significant impact on the COC of the technology.

7 Future Work

Some limitations to this case study were identified, and this section provides several suggestions, but not a complete list, for future work that would aid in refining the DAC solvent system results presented.

7.1 Alternate Configurations

The process configuration used as the basis for this study was a 2016 version of CE's process. [3] Since then, CE has made progress in their design, and several aspects of their process have changed. Their current internal design includes a different design for the air contactors, pellet reactor, and slaker. In future revisions of this case study, these alternate designs should be considered.

Novel solvent-based DAC concepts have been developed since this work was initiated, and additional analysis could explore the cost and performance of these alternate designs.

Additional analysis could also be developed to evaluate solvent DAC technology under alternative atmospheric conditions (e.g., pressure, temperature, and relative humidity).

Another alternate configuration to be considered includes the use of low-carbon or renewable electricity to provide the electricity demand of the process; this would provide useful perspective on configurations suggested in the DAC literature, but is not considered in this case study.

7.2 Financial Parameters

As described in Section 3.6, the financial parameter assumptions used to calculate the COC were sourced from NETL's BBR4, and represent the financial assumptions used to calculate the COE for NGCC power plants. Future work could look to develop DAC-specific financial assumptions: a reference case set of assumptions for today's markets, possibly reflective of the chemical industry, and incorporating high-risk aspects given the lack of maturity of the DAC technology; a future set of assumptions building in the de-risking of DAC as the technology deploys and matures; sensitivity assumptions building in options for special financial considerations or programs (e.g., loan guarantee programs); and others as determined to be appropriate. This future work would improve upon the FCR sensitivity examined in this study, as the scenarios would be more closely tied to real-world financial scenarios.

7.3 Potential Contaminants

Case 1 does not consider the possible presence of unwanted reaction products like potassium sulfite. The presence of potassium sulfite could result in issues like fouling if not purged from the system. The impact and fate of such contaminants should be evaluated in future work.

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