NATIONAL ENERGY TECHNOLOGY LABORATORY

COST OF CAPTURING CO_2 FROM INDUSTRIAL SOURCES

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Disclaimer

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ACRONYMS AND ABBREVIATIONS

°C	Degrees Celsius
°F	Degrees Fahrenheit
AACE	AACE International (formerly Association for the Advancement of Cost Engineering)
abs	Absolute
AGR	Acid gas removal
Ar	Argon
Aspen	Aspen Plus®
atm	Atmosphere
B	Billion
BBR4	Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 4
BEC	Bare erected cost
BFD	Block flow diagram
BFS	Blast furnace stove
BOF	Basic oxygen furnace
BPD	Barrels per day
Btu	British thermal unit
C_2H_6	Ethane
C_3H_8	Propane
C_4H_{10}	Butane
CCF	Capital charge factor
CCS	Carbon capture and storage/sequestration
CCSI	Carbon Capture Simulation Initiative
CF	Capacity factor
CH_4	Methane
CH_4S	Methanethiol
CO	Carbon monoxide
COC	Cost of CO_2 capture
CO_2	Carbon dioxide
COG	Coke oven gas
CTL	Coal-to-liquids
DOE	Department of Energy
EAF	Electric arc furnace
Eng'g CM H.O & Fee	Engineering construction management home office and fees
EO	Ethylene oxide
EOR	Enhanced oil recovery
EPC	Engineering/procurement/ construction
EPCC	Engineering, procurement, and construction cost
EPRI	Electric Power Research Institute
FGD	Flue gas desulfurization
ft^3	Cubic foot
FT	Fischer-Tropsch
gal	Gallon
GHG	Greenhouse gas
gpm	Gallons per minute
GTL	Gas-to-liquids
h, hr	Hour
H_2	Hydrogen
H_2O	Water
H_2S	Hydrogen sulfide
He	Helium
HHV	Higher heating value
HX	Heat exchanger
I&C	instrumentation and control
IEAGHG	IEA Greenhouse Gas R&D Programme
kg	Kilogram
kJ	Kilojoule
KO	Knockout
kW, kWe	Kilowatt electric
lb	Pound
LHV	Lower heating value
M	Million
m^3	Cubic meter
MEA	Monoethanethiol
MMBtu	Million British thermal units
MMCFD	Million cubic feet per day
MMSCFD	Million standard cubic feet per day
mol%	Mole percent
MPa	Megapascal
MW, MWe	Megawatt electric
MWh	Megawatt-hour
N/A	Not applicable/available
N_2	Nitrogen

NaOH	Sodium hydroxide
NETL	National Energy Technology Laboratory
NG	Natural gas
NGP	Natural gas processing
NOx	Oxides of nitrogen
O&M	Operation and maintenance
O_2	Oxygen
O-H	Overhead
PC	Portland cement
PPS	Power plant stack
PSA	Pressure swing adsorption
psia	Pound per square inch absolute
psig	Pound per square inch gauge
QGESS	Quality Guidelines for Energy System Studies
R&D	Research and development
scf	Standard cubic feet
SCR	Selective catalytic reduction
SMR	Steam methane reforming
SO_2	Sulfur dioxide
SOx	Oxides of sulfur
T&S	Transport and storage
TAG®	Technical Assessment Guide
TASC	Total as-spent cost
TEG	Triethylene glycol
TOC	Total overnight cost
tonne	Metric ton (1,000 kg)
TPC	Total plant cost
U.S.	United States
USD	U.S. dollar
USDA	U.S. Department of Agriculture
USGS	United States Geological Survey
V-L	Vapor-liquid
yr	Year

EXECUTIVE SUMMARY

The objective of this report is to provide an estimate of the cost to capture carbon dioxide (CO_2) from selected industrial processes. The following nine processes were chosen for analysis due to either the high purity of the CO_2 emission source (99–100 mole percent CO_2) or the large quantity of CO_2 potentially available. The processes considered in this study are summarized in Exhibit ES-1, where "CO_2 Available for Capture" represents the amount of pure CO_2 in the capture stream described in the table for each case, at a 100 percent capacity factor (CF).

Exhibit ES-1. Industrial sources of CO_2 case summary

Case Class	Process	Base Plant Production Capacity	Capture Stream Description	CO_2 Available for Capture (M tonnes CO_2/year)
High Purity	Ammonia	394,000 tonnes/year	Stripping vent: 23.52 psia	0.486
	Ethylene Oxide	364,500 tonnes/year	Acid gas removal CO_2 stream: 43.5 psia	0.122
	Ethanol	50 M gal/year	Fermenter off-gas: 17.40 psia	0.143
	Natural Gas Processing	330 MMSCFD	CO_2 vent: 23.52 psia	0.649
	Coal-to-Liquids	50,000 BPD	AGR CO_2 streams: 160 psia, 265 psia, and 300 psia	8.74
	Gas-to-Liquids	50,000 BPD	AGR CO_2 stream: 265 psia	1.86
Low Purity	Refinery Hydrogen	87,000 tonnes/year	Raw syngas from SMR: 399.9 psia	0.405
	Cement	1.3 M tonnes/year	Kiln off-gas: 14.7 psia	1.21
	Steel/Iron	2.54 M tonnes/year	COG PPS: 14.7 psia COG/BFS: 14.7 psia	3.74 (total of both capture streams)

Note: COG = coke oven gas; PPS = power plant stack; BFS = blast furnace stove

For each industrial process considered, available plant information, such as existing average plant size, projected new development plant size, or existing plant operations data was used to develop a reference plant for this study. Plant size is one factor affecting the amount of CO_2 available for capture from an industrial process. Other factors are specific to each industry. For example, the ammonia industry captures and re-uses CO_2 in urea production, and natural gas processing (NGP) plant CO_2 emissions are dependent upon the raw gas compositions entering the facility. As such, specific assumptions related to CO_2 availability are necessary to establish each representative plant and to suggest the industry's average CO_2 emissions.

For each process, the CO_2 capture cost for a greenfield facility and a retrofit facility was calculated with the latter being calculated by applying a retrofit factor to the greenfield total plant cost (TPC). For the iron/steel process, only a retrofit case is given since the representative plant is a basic oxygen furnace facility, which are no longer being constructed. For the coal-to-liquids (CTL) and gas-to-liquids (GTL) cases, no retrofit case is given, since no plants currently exist domestically, and it is assumed that none will be constructed without CO_2 capture. The cost metric of interest is the cost of CO_2 captured in U.S. dollars per tonne, as calculated in Equation ES-1. In this report, costs are presented in December 2018 real dollars.

$$COC\left(\frac{\$}{tonne\ CO_2}\right) = \frac{TOC * CCF + FOM + VOM + PF + PP}{tonnes\ CO_2\ captured\ per\ year}$$ ***Equation ES-1***

Where:

TOC – Total overnight costs of equipment added for the application of CO_2 capture

CCF – Capital charge factor, based on industry-specific financial assumptions as detailed in Section 3.2

FOM – Annual fixed operating & maintenance (O&M) costs

VOM – Annual variable O&M costs

PF – Purchased fuel

PP – Purchased power

The high purity emissions sources are inherently produced by their base plants at CO_2 concentrations suitable for pipeline transport, requiring only compression, associated intercooling, and, in some cases, glycol dehydration. The low purity sources considered offer emission streams with CO_2 concentrations below that which is acceptable for pipeline use, per guidance in National Energy Technology Laboratory's (NETL) "Quality Guidelines for Energy System Studies (QGESS): CO_2 Impurity Design Parameters" specifications. [1] As such, the refinery hydrogen, cement, and iron/steel cases require CO_2 removal systems along with compression, associated intercooling, and glycol dehydration. For the CO_2 removal systems, two capture rates were evaluated, 90 and 99 percent, to evaluate the cost of capturing the CO_2 from the emissions streams defined in Exhibit ES-1.[a]

Exhibit ES-2 provides the resulting greenfield and retrofit cost of CO_2 capture (COC), where appropriate, for each case considered in this study, along with the capital, variable and fixed O&M, purchased power and/or natural gas (NG) fuel cost components for each case. For each case, other than those of iron/steel, the individual cost components shown (i.e., capital costs, fixed O&M costs, variable O&M costs, and purchased power/natural gas) represent the cost components that add to the total COC in greenfield applications. For iron/steel, those individual cost components represent retrofit costs. In addition, each high purity source shows the total retrofit COC, which is estimated based on methodology described in Section 3.3, except for the CTL and GTL cases. As there are no existing CTL or GTL plants in the domestic industrial fleet, it

[a] This report does not consider capture of the CO_2 produced by the natural gas-fired boiler used for steam generation in the low purity cases (i.e., for solvent regeneration) or other process streams outside of those defined in Exhibit ES-1. If this CO_2 was captured, it would greatly impact the results presented herein. Such an analysis is discussed in the future work considerations detailed in Section 9.

is assumed that future (i.e., greenfield) builds would include carbon capture (i.e., retrofit capture applications at CTL or GTL facilities would not be expected). Further details regarding the estimation of capital, operating, and maintenance costs are provided within the body of the report.

Exhibit ES-2. COC from industrial sources

Case		Capital Costs	Fixed O&M Costs	Variable O&M Costs	Purchased Power/ Natural Gas	Greenfield COC	Retrofit COC
Ammonia		6.1	3.9	2.7	6.3	**19.0**	**19.0**
Ethylene Oxide		9.4	9.8	1.7	5.2	**26.0**	**26.2**
Ethanol		14.1	9.2	1.7	6.8	**31.8**	**32.0**
NGP		6.2	3.4	1.5	5.0	**16.1**	**16.2**
CTL		2.0	0.7	0.3	2.6	**5.6**	**N/A**
GTL		2.9	1.2	0.3	1.9	**6.4**	**N/A**
Refinery Hydrogen	90% Capture	22.8	15.6	5.3	16.2	**59.9**	**61.7**
	99% Capture	21.3	14.4	5.1	16.5	**57.3**	**58.9**
Cement	90% Capture	22.8	11.1	6.1	22.6	**62.7**	**64.3**
	99% Capture	21.8	10.6	5.9	22.6	**60.8**	**62.4**
Iron/Steel	90% Capture	28.0	9.5	5.7	22.6	**N/A**	**65.9**
	99% Capture	27.8	9.3	5.6	22.6	**N/A**	**65.4**

Note: All values expressed in December 2018 U.S. dollars per tonne CO_2.

The results show that CTL has the lowest greenfield COC, followed by GTL, NGP, ammonia, ethylene oxide (EO), ethanol, refinery hydrogen, and finally, cement, which has the highest greenfield COC. Retrofit applications exclude CTL and GTL, but follow the same cost pattern; however, the highest retrofit COC is the iron/steel case.

For the low purity cases, the normalized COC (\$/tonne CO_2) decreases slightly with increasing capture rate (i.e., from 90 to 99 percent capture). The cost of the capture system and associated consumables increases at a lesser rate than that of the amount of CO_2 captured (i.e., a 10 percent increase from 90 to 99 percent capture). This is the effect of accuracy ranges of the capital cost estimates from the capture system vendor (-25/+40 percent) and the cost scaling methodology employed in this study. [2] [3] The margin of error associated with the cost estimate indicate that with increasing capture rate in the low purity cases, the COC is effectively the same. The reported minor increase in capital cost with increased capture rate (up to 99 percent for sources with CO_2 purity greater than 12 percent) based on vendor furnished cost and performance estimates has been validated by independent modeling performed by the carbon capture simulation initiative team at NETL and has been reported independently in literature. [4] Exhibit ES-3 shows the error in the calculated capture system BEC associated with the vendor's quoted uncertainty rate alongside the amount of CO_2 captured in the cement case from 90 to 99 percent capture rate. Similar graphs in for the refinery hydrogen and iron/steel cases can be found in Section 6.1.10 and Section 6.3.10, respectively.

Exhibit ES-3. Capture system BEC and amount of CO_2 captured versus capture rate

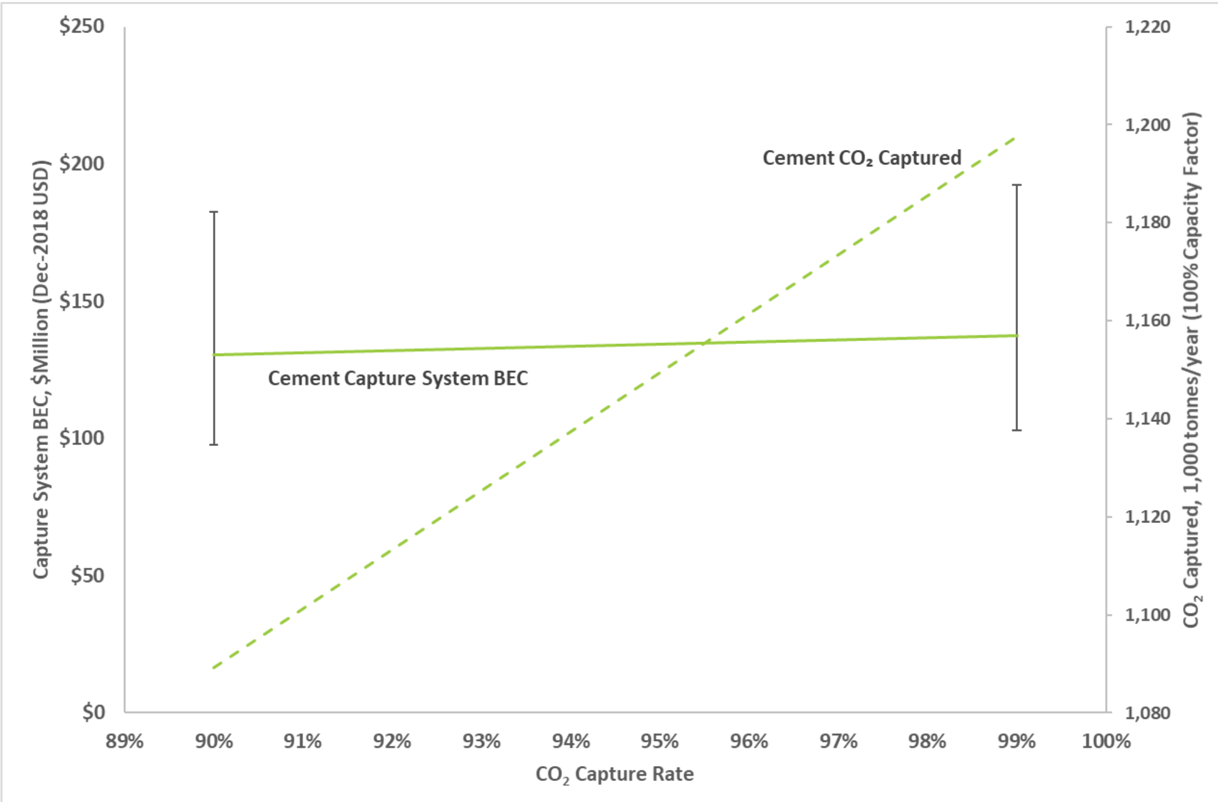
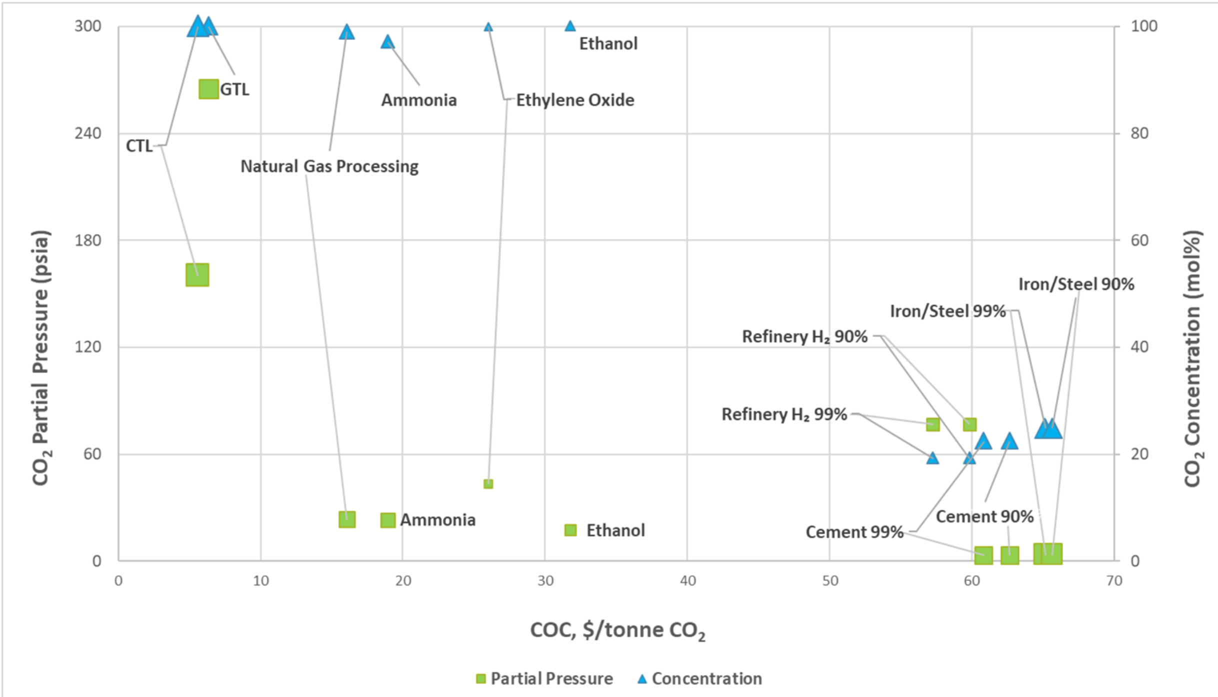


Exhibit ES-4 shows a plot of the COC versus the assumed CO_2 stream partial pressure and the assumed CO_2 concentrations for each of the base cases considered in this report. The general trend shows that as both the CO_2 concentration and the CO_2 partial pressure decrease, the COC of CO_2 increases. The average COC for the six processes with CO_2 concentration greater than 95 percent is $17.5/tonne, while the average COC for the three processes with CO_2 concentration less than 50 percent is $62.0/tonne. The partial pressure in the high purity cases is mainly reflective of the CO_2 concentration.

Exhibit ES-4. COC versus CO_2 partial pressure and CO_2 concentration



Note: Marker size is relatively indicative of CO_2 captured (tonnes/year).

The trends observed in this study may not be universally applicable because the assumptions made for each case in this study may not apply to all real-world examples of a specific industry. Additionally, concentration trends are emphasized due to the potential misleading nature of partial pressure values. In some instances, partial pressure can have directly recognizable effects on the COC; higher pressures will reduce the size of and duty of compression equipment, but this may not always be the case. For example, a stream with a total pressure of 1,000 psia, and a concentration of 10 percent CO_2, would have a partial pressure of 100 psia. For the cases in this study, this partial pressure would be considered high, and might be expected to result in a low COC. However, for this example, capture and/or purification would be required, and therefore the resulting COC would not be expected to follow the partial pressure trend observed in Exhibit ES-4.

There are also exceptions to these trends driven by economies of scale. Such a relationship is demonstrated in Exhibit ES-4 when comparing the results of NGP and ammonia. The CO_2 stream partial pressures are equivalent, and the concentrations are also the same at 99 percent. However, the greenfield COCs were calculated to be \$16.1/tonne CO_2 for NGP and \$19.0/tonne CO_2 for ammonia, about an 18 percent difference. This is a result of the amount of CO_2 available for capture in each case. Based on the assumptions made for each representative plant, NGP has 649,225 tonne/year CO_2 available, while ammonia only has 486,227 tonne/year available. Therefore, while the CO_2 stream partial pressures and concentrations are equivalent, there is 33 percent more CO_2 available for capture and sale at the NGP reference plant, resulting in a lower normalized CO_2 capture cost. The factors noted above in Exhibit ES-4, namely CO_2 partial pressure, concentration, and economies of scale (i.e., CO_2 available at each representative plant), result in a significant range of CO_2 capture costs. The highest greenfield COC, the cement case with 90 percent capture, is more than eleven times the price of the least expensive case (i.e., CTL).

In addition, the assumptions regarding the quality of the CO_2 emissions stream from the base plant in each case may greatly impact the COC. For instance, the base cement case assumes that the kiln off-gas is suitable to be sent directly for CO_2 separation; however, cement industry members suggest that the kiln off-gas may have higher-than-acceptable levels of oxides of sulfur (SOx)/oxides of nitrogen (NOx) and would require the addition of selective catalytic reduction (SCR) and flue gas desulfurization (FGD). A sensitivity to this case was performed to evaluate the effect of adding these unit operations to the cement cases. The amount of SOx/NOx was not directly characterized; instead, the FGD and SCR costs were scaled from Case B12B of Revision 4 of NETL's "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity" based on the quantity of gas to be treated (i.e., the total flow of kiln off-gas). [5] Case B12B presents an SCR with a 78 percent NOx removal efficiency and an FGD that removes 2,000 ppm, by volume, of SOx from the coal boiler flue gas stream. The results of this sensitivity analysis show that the addition of a similar SCR and FGD to the cement plant's CO_2 capture system would increase greenfield COC by 23–25 percent with a COC of $74.8/tonne CO_2 and $78.0/tonne CO_2 for 99 and 90 percent capture, respectively.

While the calculation of a COC demonstrates the capture costs across different industries based on a specific set of plant assumptions, another important consideration is the amount of CO_2 available from each industry. Neglecting CO_2 transportation costs, if two industries demonstrate approximately equivalent normalized COCs, but one has a significantly larger supply, the industry with the larger supply would offer the more effective decarbonization[b] application at the same or similar normalized cost. Exhibit ES-5 shows the CO_2 emissions by industry in the United States, while Exhibit ES-6 presents a plot of COC versus the amount of domestic CO_2 emissions, both based on the Environmental Protection Agency Facility Level Information on Greenhouse Gases Tool as of the 2020 reporting year.[c] [6] The COCs are those calculated in this study for greenfield sites except for iron/steel, which is for a retrofit application. This plot shows the cost of the source relative to the potentially capturable emissions in the United States.

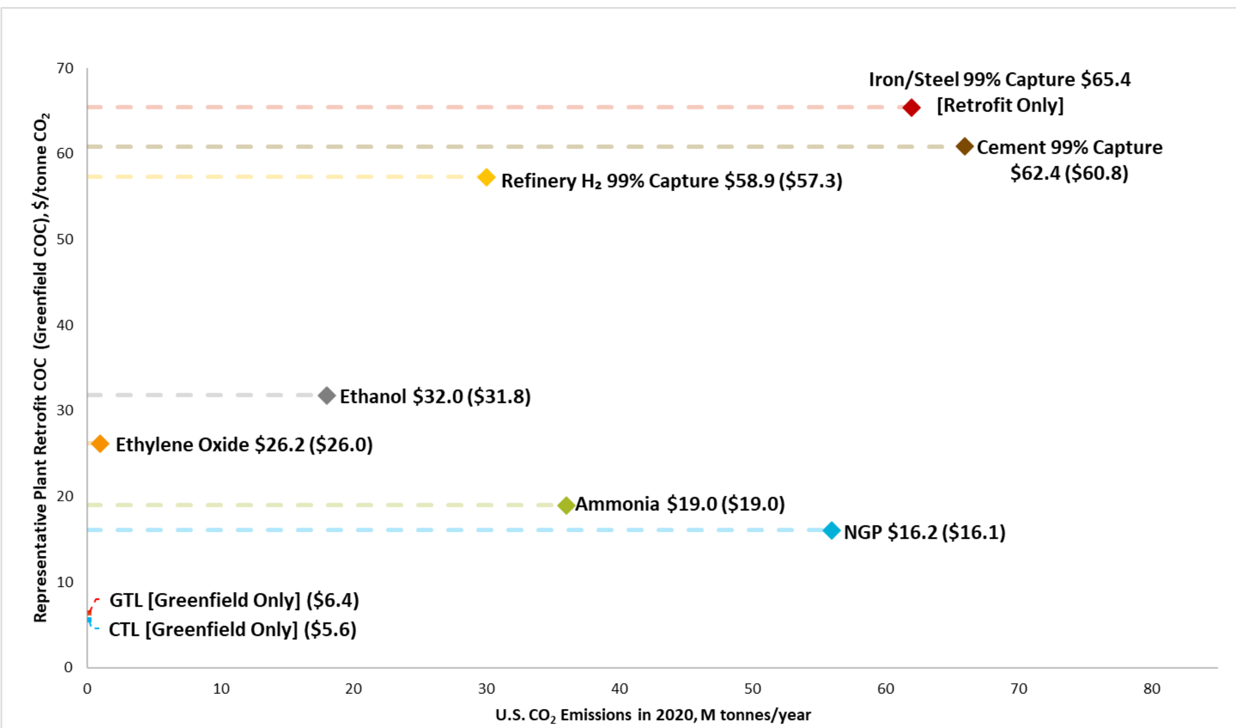
[b] Decarbonization within the context of this report is defined as the reduction of point-source emissions from industrial processes. Lifecycle analysis of decarbonization efforts as it relates to the CO_2 capture operations evaluated in this report is not considered but could be considered in future work opportunities.

[c] CO_2 emissions related to EO production are not reported in Environmental Protection Agency's Facility Level Information on Greenhouse Gases Tool; as such, the total emissions were estimated based on the total EO production as of 2019 [53] and an emissions factor of 1:3 CO_2:EO on a molar basis, according to reaction stoichiometry as detailed in Section 5.2.

Exhibit ES-5. U.S. industrial CO_2 emissions by industry

Industry	U.S. Total CO_2 Emissions in 2020 (M tonnes CO_2/year) [6]
Ammonia	36
Ethylene Oxide	0.95
Ethanol	18
Natural Gas Processing	56
Coal-to-Liquids	0
Gas-to-Liquids	0
Refinery Hydrogen	30
Cement	66
Steel/Iron	62

Exhibit ES-6. Representative plant COC results versus U.S. industrial CO_2 emissions



Note: Only the 99 percent capture cases are shown for low purity sources in Exhibit ES-6.

Based on emissions rates, of the industrial plants with existing operations (i.e., excluding CTL and GTL), EO is the least impactful decarbonization option given the small amount of CO_2 available for capture (0.95 M tonnes/year), and cement manufacturing is the most impactful option with the largest amount of CO_2 available (66 M tonnes/year). Based on normalized COC, NGP is the least expensive industrial source of CO_2 within the existing U.S. fleet with a price of $16.1/tonne, and iron/steel is the most expensive option with a price of $64.8/tonne.

Sensitivities to CF, cost of purchased power, plant size in terms of CO_2 emissions per year, and capital charge factor (CCF) were analyzed for each greenfield case. A sensitivity to natural gas price was also performed for the greenfield low purity cases. In these cases, natural gas is burned in an industrial boiler, described in Section 4.3, to generate steam for solvent regeneration in the CO_2 capture process. Lastly, a sensitivity to the retrofit factor applied to generate retrofit application costs was evaluated for each case, excluding CTL and GTL, which do not have retrofit applications. The plant size sensitivity results for each case, evaluated across the typical plant size ranges specific to each industry, can be found in the corresponding sections, and all other sensitivity analyses can be found in Section 7.2.

The general results of the sensitivities evaluated are as follows:

- As CF varies from 65 to 95 percent, the COC for each case decreases, most notably in the Refinery H_2 90 percent capture case where a $18.0/tonne CO_2 decrease is observed across the sensitivity range. An 85 percent CF was assumed for the cases in this study.
- As purchased power price increases, the COC also increases. This study assumes that all electricity requirements are provided by purchasing power from the grid. In cases requiring additional power beyond just compression, such as power for auxiliary loads in the CO_2 separation processes, the COC increase is more dramatic. The largest increase across the sensitivity range was observed in the iron/steel and cement cases at $16.4/tonne.
- The sensitivity to CCF is important as different industries may have access to different costs of capital. The CCF for each case was developed by NETL's Energy Markets Analysis Team based on market financial data respective to each industrial sector. Details of the financial factors used in this study are given in Section 3.2. As CCF varies from 5 percent to 35 percent, the capture costs can increase by up to $150.2/tonne as observed in the refinery hydrogen case with 90 percent capture.
- The final sensitivity to natural gas price showed that as the natural gas price varied over the range $3–10/MMBtu, the COC may rise as much as $30.6/tonne CO_2 as was observed in the iron/steel 90 percent capture case.[d]

This study uses the COC and CO_2 supply to compare nine potential industrial CO_2 sources. The results are representative of the assumptions regarding the reference plant and its CO_2 emissions stream(s). Scale and location will impact results for actual plants. Methods of CO_2 transport and storage (T&S) and the associated costs are considerations that could ultimately change the economic impact of implementing carbon capture at a specific plant. T&S costs were not considered in this study; however, Section 2 examines the location of individual plants in each industry relative to CO_2 pipelines and current EOR sites to qualitatively identify relative advantages or disadvantages for decarbonization in each industry, as it relates to T&S. To estimate T&S costs, users may refer to NETL's "Quality Guidelines for Energy System Studies (QGESS): Carbon Dioxide Transport and Storage Costs in NETL Studies" for guidance. [7]

[d] This report does not consider capture of the CO_2 produced by the NG-fired boiler. If this CO_2 was captured, it would impact the results presented herein greatly, due to the lower concentration of CO_2 in the flue gas stream compared to that of the low purity industrial sources considered. It would also increase the amount of CO_2 available for capture, as NG consumption increases. Such an analysis is discussed in the future work considerations detailed in Section 9.

1 INTRODUCTION

With a global initiative to reduce greenhouse gas (GHG) emissions, several common industrial processes have been identified as potential opportunities for carbon dioxide (CO_2) capture. Of the 9 processes considered in this report, 7 have existing operations in the United States, contributing just under 270 M tonnes per year of CO_2 emissions in 2020 based on reporting to the Environmental Protection Agency. [6] Industrial plant CO_2 emissions sources offer advantages when considering decarbonization due to their relatively high concentrations of CO_2 in emissions streams, which may lead to lower normalized capture costs. With high CO_2 concentrations, separation equipment costs are minimized, or even eliminated in cases where CO_2 streams are 99–100 percent pure. This study evaluates nine representative plants with CO_2 emissions sources having relatively high concentrations to determine the cost of CO_2 capture.

The cost of CO_2 capture (COC) in each case, as defined by Equation 1-1, considers the equipment required for CO_2 removal, if applicable, and compression, as well as the balance of plant equipment as detailed in Section 4.3 through Section 4.6, and operation and maintenance (O&M), purchased power, and fuel costs, as applicable. Throughout the report, "CO_2 capture" refers to the incremental equipment required to prepare the CO_2 emissions stream for pipeline transport (i.e., compression and intercooling, auxiliary equipment, CO_2 removal systems, etc.).

$$COC\left(\frac{\$}{tonne\ CO_2}\right) = \frac{TOC * CCF + FOM + VOM + PF + PP}{tonnes\ CO_2\ captured\ per\ year} \qquad \textbf{Equation 1-1}$$

Where:

TOC – Total overnight costs of equipment added for the application of CO_2 capture

CCF – Capital charge factor, based on industry-specific financial assumptions as detailed in Section 3.2

FOM – Annual fixed O&M costs

VOM – Annual variable O&M costs

PF – Purchased fuel

PP – Purchased power

Estimates of financing scenarios specific to each industry were applied to the capital costs to account for return on equity and financing costs. Financial methodology and the resulting financial factors for each case are presented in Section 3.

1.1 ASSUMPTIONS

There are many industrial processes that produce CO_2 emissions, and as such, criteria were established to justify the inclusion of an industrial process in this report. First, an industrial plant must be representative of either a relatively large amount of CO_2 emissions (i.e., an emissions source that could benefit from economies of scale) or of a 99–100 percent pure CO_2 stream. The second criterion for inclusion is that an industrial plant is likely to provide a relatively low normalized COC. This condition is highly dependent upon the first criteria, as normalized COC values are a function of CO_2 availability. Power production plants are not considered in this study, as they are evaluated in NETL's collection of baseline studies, such as "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity." [5] Process models were developed for each case based on guidance in NETL's "Quality Guidelines for Energy Systems Studies (QGESS): Process Modeling Design Parameters," and applicable model assumptions are shown in Exhibit 1-1. [8]

Exhibit 1-1. Process design assumptions

Site Characteristics		
Location		Greenfield, Midwestern U.S.
Topography		Level
Size, acres		10
Particulate Matter Disposal		Off-Site
Water Supply		50% Municipal and 50% Ground Water
Site Ambient Conditions		
Elevation, meter (feet)		0 (0)
Barometric Pressure, MPa (psia)		0.101 (14.696)
Average Ambient Dry Bulb Temperature, °C (°F)		15 (59)
Average Ambient Wet Bulb Temperature, °C (°F)		10.8 (51.5)
Design Ambient Relative Humidity, %		60
Cooling Water Temperature, °C (°F)		15.6 (60)
Natural Gas Characteristics		
Component		**Volume %**
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	$5.75x10^{-6}$
	LHV	**HHV**
kJ/kg (Btu/lb)	47,201 (20,293)	52,295 (22,483)
Megajoule/standard cubic meter (Btu/scf)	34.52 (927)	38.25 (1,027)
Air composition based on published psychrometric data, mass %		
Nitrogen	N_2	75.055
Oxygen	O_2	22.998
Argon	Ar	1.280
Water	H_2O	0.616
Carbon Dioxide	CO_2	0.050

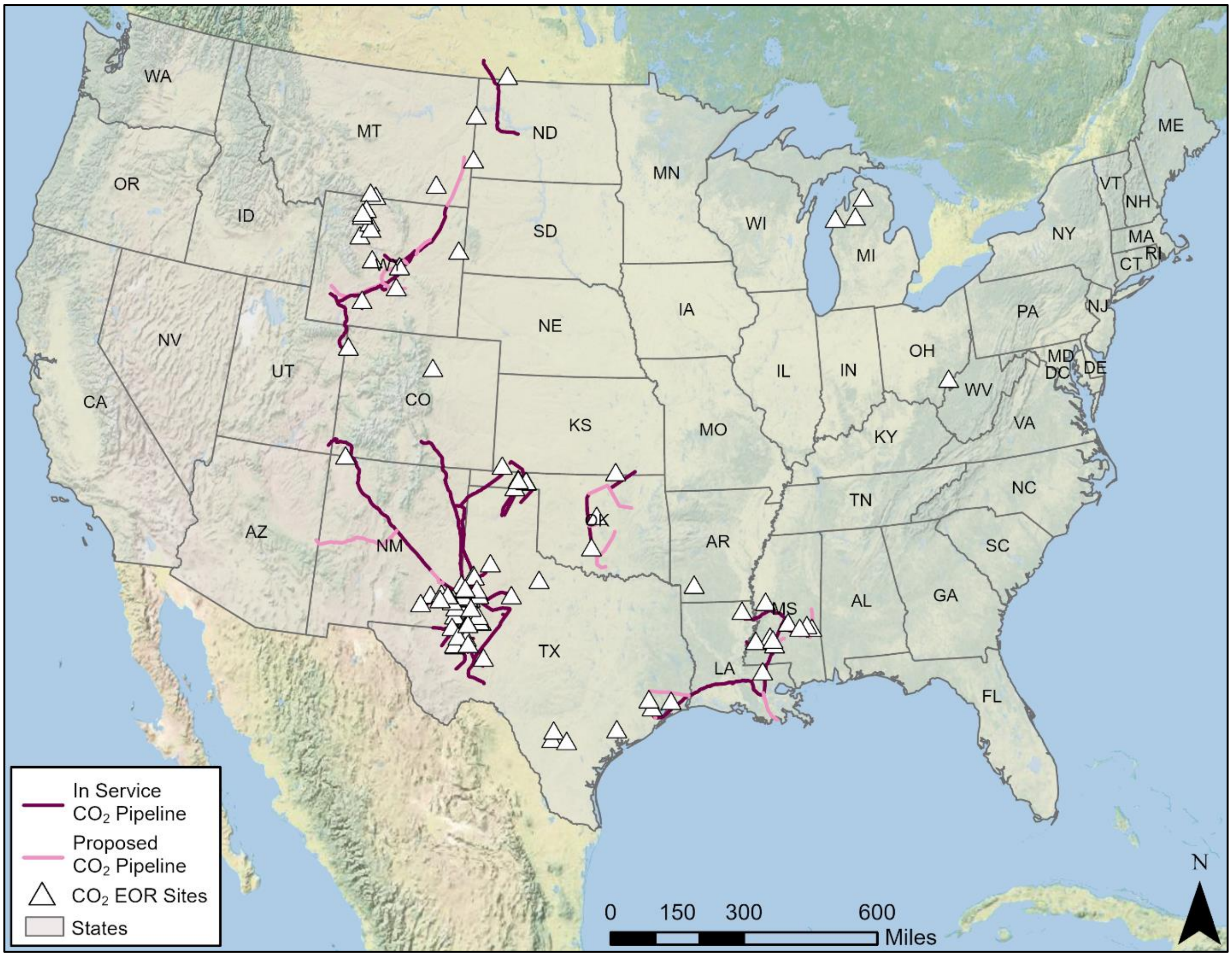
[A] The sulfur content of natural gas is primarily composed of added Mercaptan (methanethiol [CH_4S]) with trace levels of hydrogen sulfide (H_2S)

2 Plant Sites and CO_2 End-Use

The assumption made for this study is that the final CO_2 product is transported via pipeline to be utilized in enhanced oil recovery (EOR) applications, and as such applies the specifications for CO_2 product purity, pressure, and temperature after capture and compression per National Energy Technology Laboratory's (NETL) "QGESS: CO_2 Impurity Design Parameters" specifications. [1] The viability of adding capture to a representative plant would ultimately be dependent upon the costs for transport and storage (T&S) of the CO_2 captured in addition to the COCs evaluated in this report. T&S costs are not considered in the metric of value, COC of CO_2, in this study but should be considered by an owner evaluating capture implementation at an industrial facility. Other uses for the CO_2 may be available to owners, but those alternate possibilities were not considered for the purpose of this report. In addition, analysis of the base plants for each of the nine processes considered falls outside the scope of this study (i.e., cost of cement production before and after CO_2 capture).

Leaving the system boundary of this study is a CO_2 stream that has been purified, where necessary, and compressed to pipeline specifications of 2,200 psig per QGESS specification. [1] While detailed pipeline specifications such as pressure drop, length, and other characteristics, are not considered in this report, and as noted in Exhibit 1-1, the study assumes a generic midwestern plant for the purposes of consistency in process modeling, it is useful to highlight potential industrial CO_2 capture locations and their relative locations to sites/transport mechanisms that could be utilized. Exhibit 2-1 shows existing CO_2 pipelines and EOR injection sites, while the seven maps that follow, Exhibit 2-2 through Exhibit 2-7, illustrate the proximity of plants for each industrial source type to the existing CO_2 pipeline and EOR infrastructure. There are currently no U.S. coal-to-liquids (CTL) or gas-to-liquids (GTL) plants in operation, so no map is given for these cases.

Exhibit 2-1. Existing CO_2 pipelines and active EOR injection sites



A large percentage of ammonia plants are in close proximity to existing CO_2 pipelines and EOR injection sites, as shown in Exhibit 2-2. The bars on the chart represent gross (light blue) and net (dark blue) ammonia production at each plant. As noted in Section 5.1.2, the representative ammonia production in the United States was considered at gross capacity, but in some ammonia plants, portions of gross ammonia and CO_2 produced are further utilized to make ammonia derivatives, such as ammonium nitrate or urea. Alternate use of CO_2 in ammonia plants is outside the scope of this study, but net capacities are shown alongside gross capacities in Exhibit 2-2 for reference or future use.

Exhibit 2-2. Ammonia plant locations and existing CO_2 pipelines and EOR injection sites

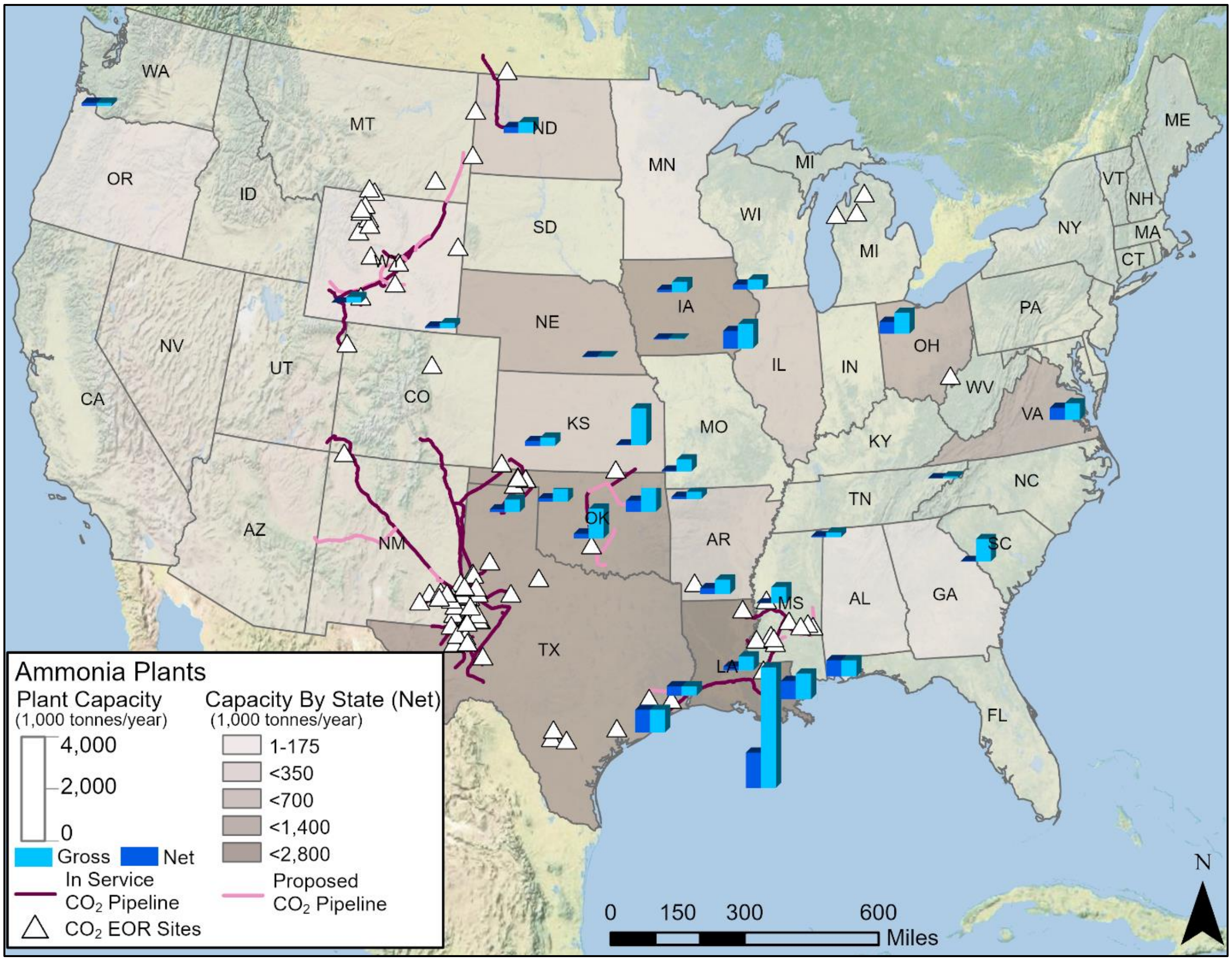
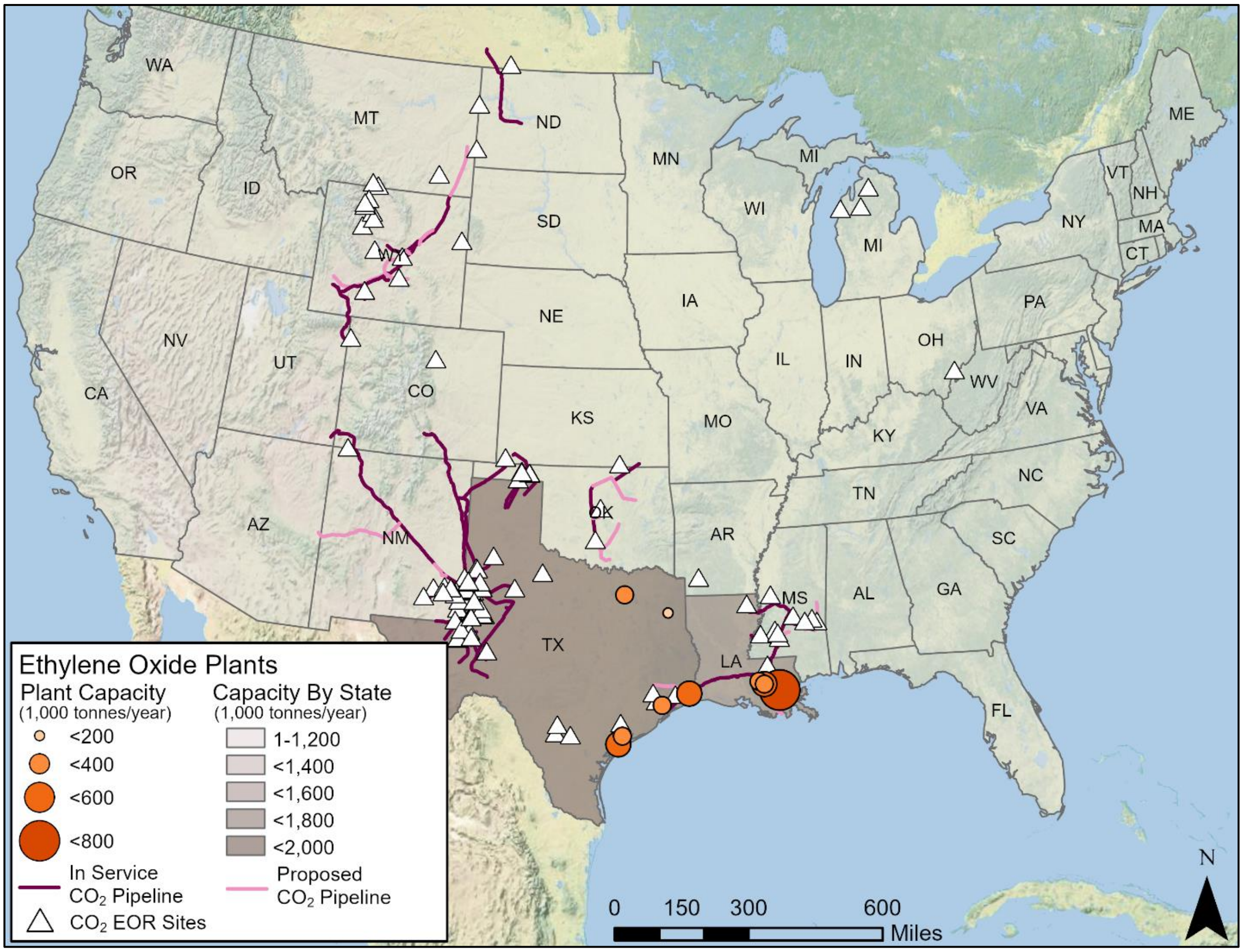


Exhibit 2-3 shows the location of EO plants and their relation to existing CO_2 pipelines and EOR injection sites. U.S. EO production is concentrated in Texas and Louisiana. Of the 15 U.S. EO plants, 6 are located very close to existing EOR pipelines and injection sites. Therefore, from a location standpoint, EO presents a potentially advantageous option for capture integration. However, due to the small scale of the existing EO plants (i.e., the small amount of CO_2 available for capture), diseconomies of scale may deter implementation.

Exhibit 2-3. EO plant locations and existing CO_2 pipelines and EOR injection sites



As shown in Exhibit 2-4, a large percentage of ethanol plant locations are not near existing CO_2 pipelines or EOR injection site locations; however, most of the ethanol processing facilities are grouped in the Midwest and could potentially realize economies of scale collectively to justify the addition of a new CO_2 pipeline for connection to existing infrastructure. This scenario falls outside the scope of this study but could be considered in future work.

Exhibit 2-4. Ethanol plant locations and existing CO_2 pipelines and EOR injection sites

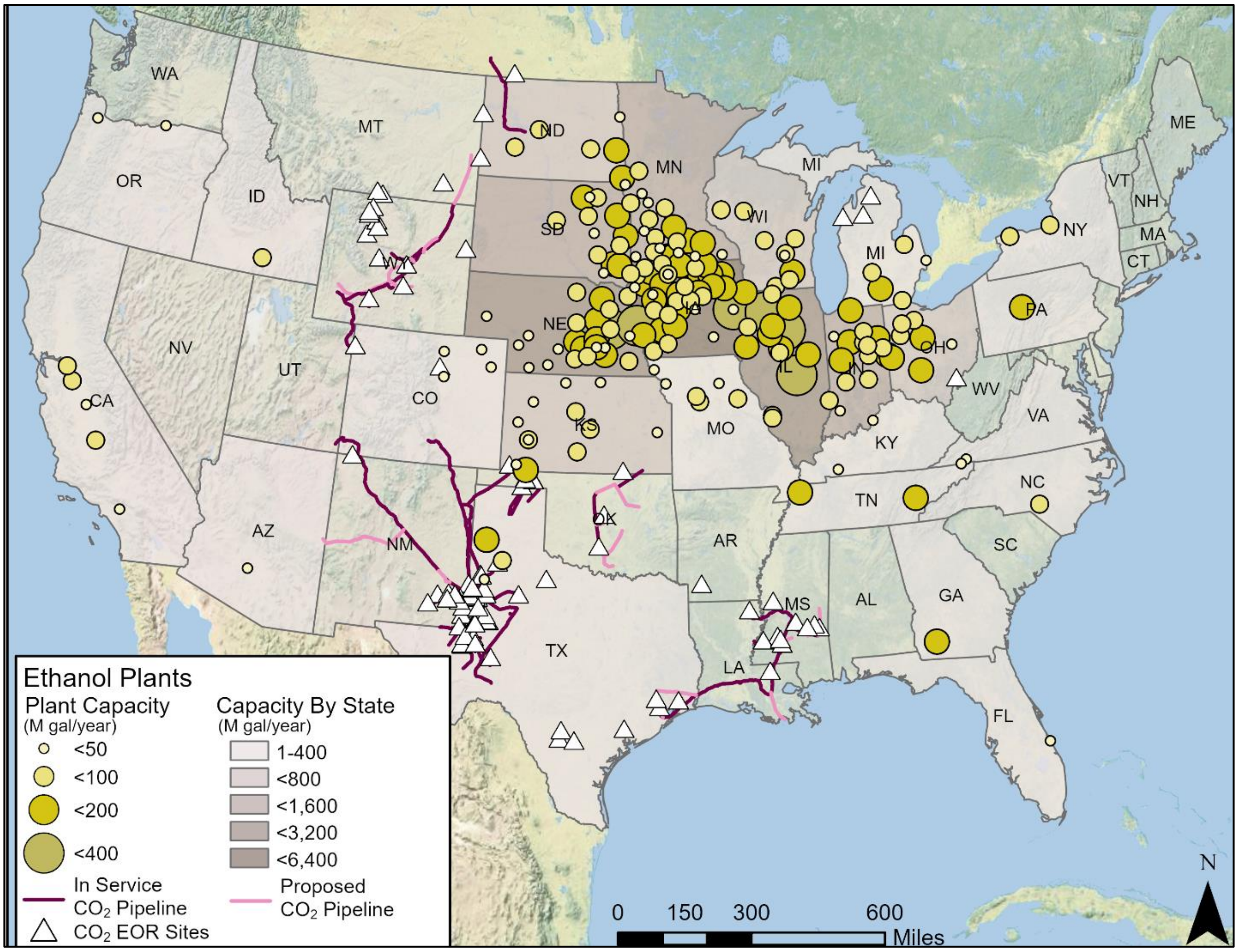


Exhibit 2-5 shows the location of natural gas processing (NGP) facilities and their relations to existing CO_2 pipelines and EOR injection sites. Plant capacities are shown on this map; however, given the 471 NGP facilities, each treating a different amount of natural gas (NG) with widely varying CO_2 concentrations, there may not be a direct correlation between capacity and CO_2 available. This means that a large facility processing NG with low CO_2 concentration may have less CO_2 available than a smaller facility processing NG with a much higher CO_2 concentration.

Exhibit 2-5. NGP plant locations and existing CO_2 pipelines and EOR injection sites

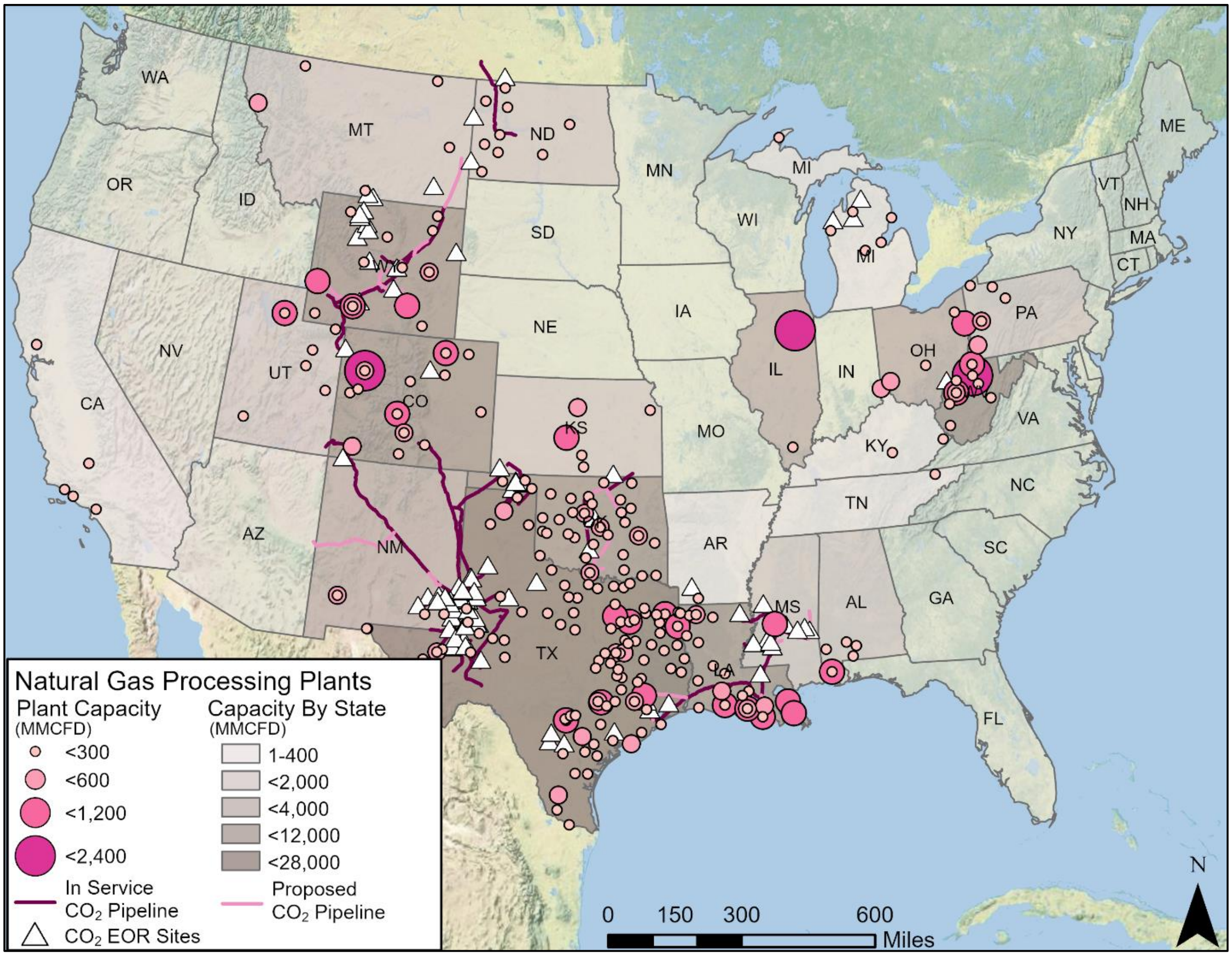


Exhibit 2-6 shows the location of U.S. refineries that produce hydrogen, and their proximity to existing CO_2 pipelines and EOR injection sites. There are many refineries near existing EOR pipelines and injection sites. However, the map is only intended to show the relative crude throughput capacity of the refineries, and not the amount of CO_2 available. There is not necessarily a direct relationship between refinery capacity and CO_2 available for capture.

Exhibit 2-6. Refinery hydrogen (U.S. refineries) plant locations and existing CO_2 pipelines and EOR injection sites

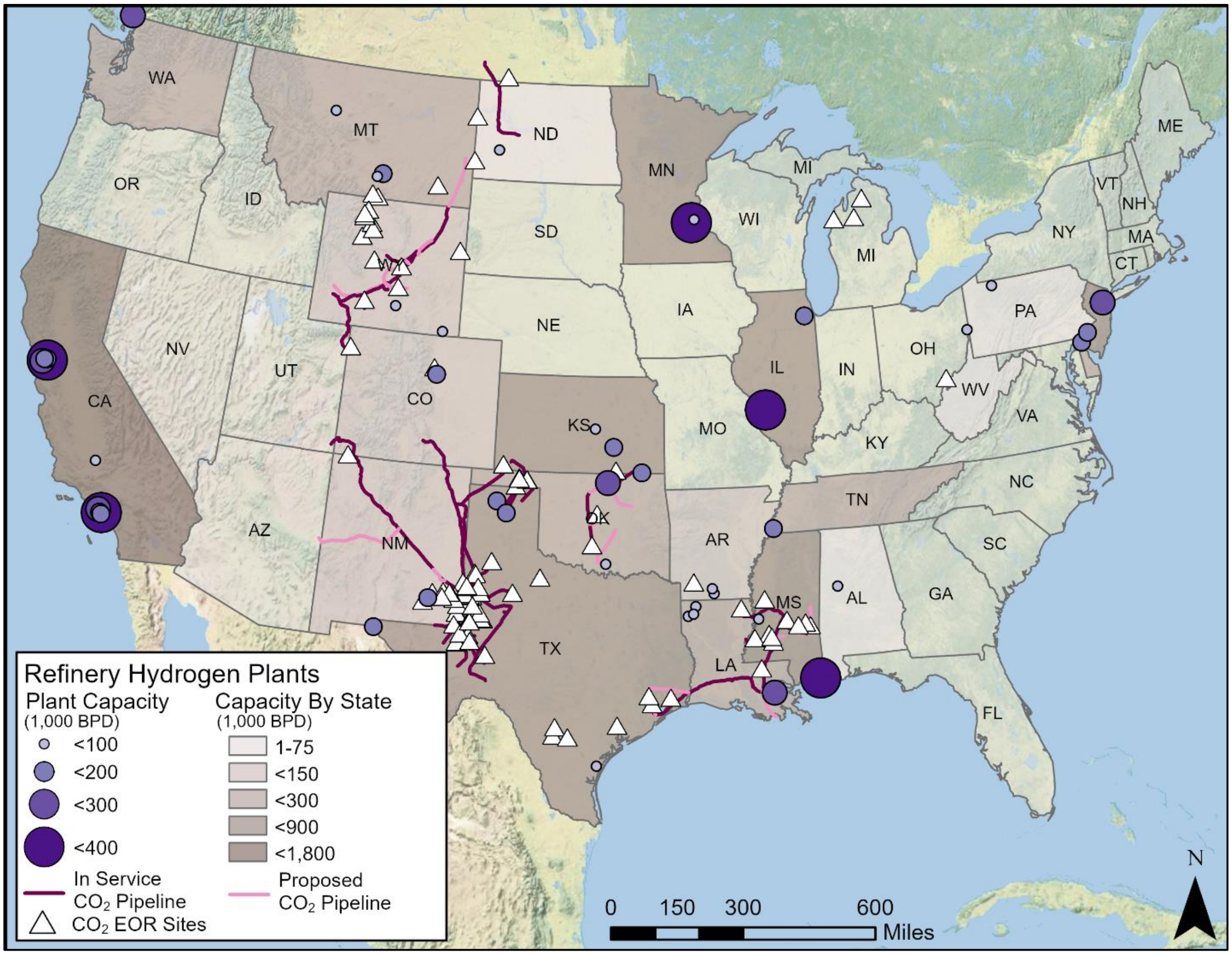


Exhibit 2-7 shows the location of cement plants and their relation to existing CO_2 pipelines and EOR injection sites. Some cement plants are located relatively close to existing infrastructure and given the typically larger scale of cement production capacity, and consequently larger amount of CO_2 emissions available, construction of a connecting pipeline for other cement facilities may be a viable means of decarbonization in the cement industry. This is scenario is not evaluated within the context of this study.

Exhibit 2-7. Cement plant locations and existing CO_2 pipelines and EOR injection sites

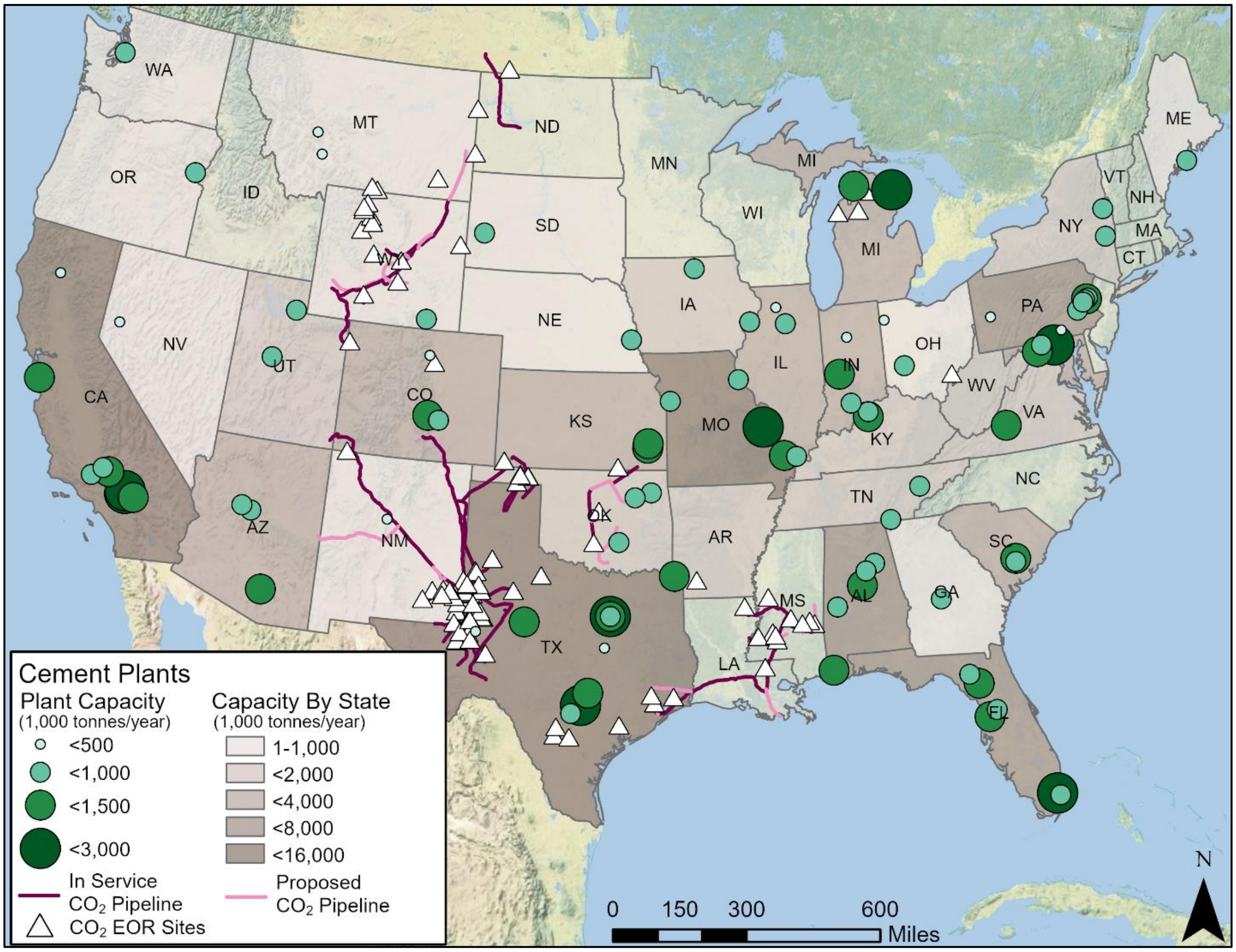
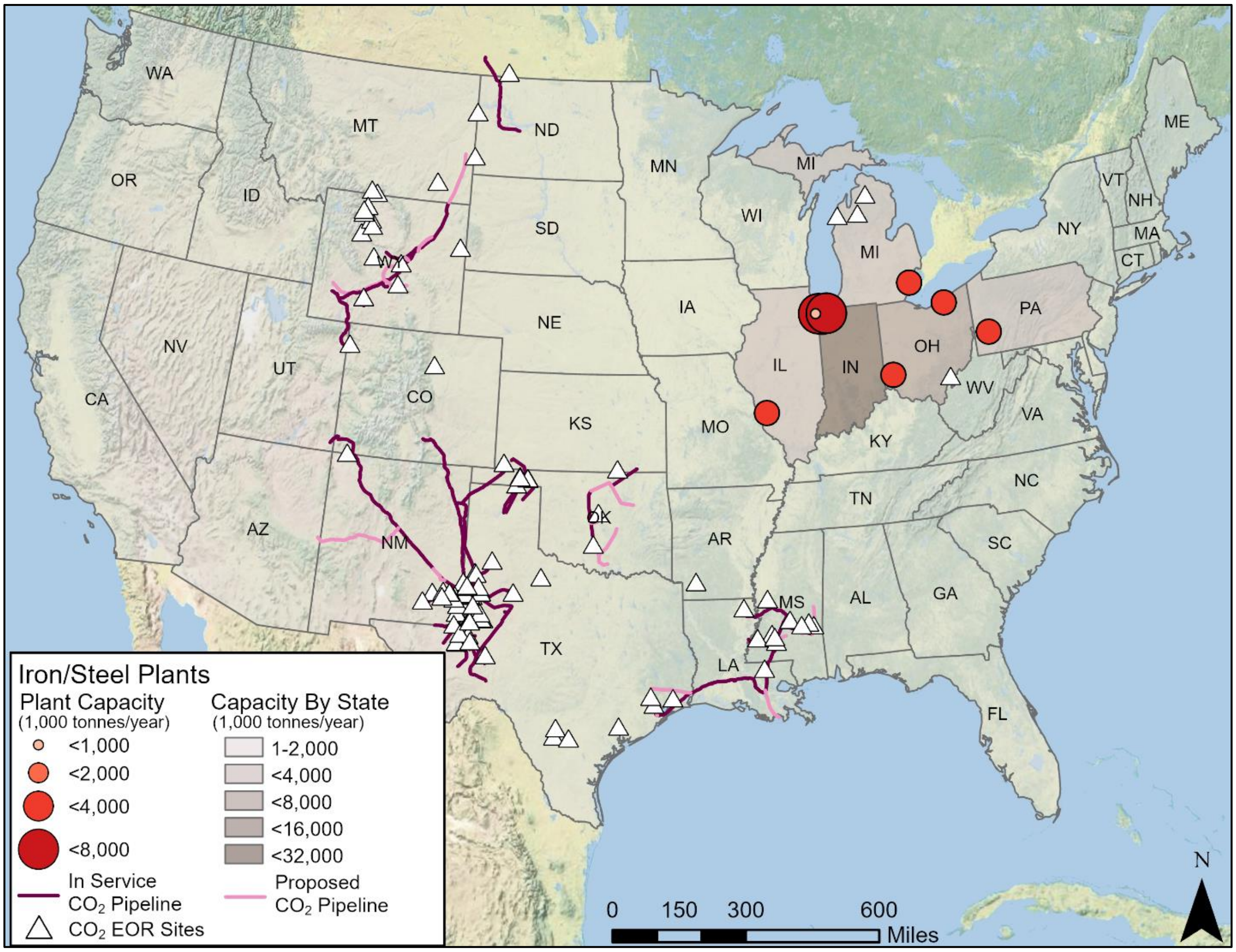


Exhibit 2-8 shows currently operating steel basic oxygen furnace (BOF) plants and their relation to existing CO_2 pipelines and EOR injection sites. Steel does not appear to provide ease of implementation for EOR end-use because many facilities would not be able to utilize any of the existing EOR infrastructure. However, based on this study's assumptions, steel plants represent the largest amount of CO_2 available among the industries considered that are currently operating plants in the United States; therefore, construction of connecting pipelines may be a viable means of decarbonization in the steel industry. This scenario is not evaluated in the context of this study.

Exhibit 2-8. Steel (BOF) plant locations and existing CO_2 pipelines and EOR injection sites



3 Economic Analysis Overview

The industrial sources considered in this study are grouped into "High Purity" and "Low Purity" groups, based on the concentration of CO_2 in the stream to be captured. The prior iteration of this report applied global financial assumptions based on the simple delineation between high and low purity sources. This approach relied on the fact that high purity sources would only require compression, whereas low purity sources would require CO_2 removal and compression, and each would have distinct construction, and thus capital expenditure, periods. For this revision update, capital expenditure assumptions have been maintained, but additional detail regarding each specific industry's financial assumptions have been added based on market data analysis performed by NETL's Energy Markets Analysis Team in October 2021.

3.1 Cost Estimating Methodology

Detailed information pertaining to topics such as contracting strategy; engineering, procurement, and construction (EPC) contractor services; estimation of capital cost contingencies; owner's costs; cost estimate scope; economic assumptions; and finance structures are available in the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." [9] Select portions are repeated in this report for completeness.

Costs of Mature Technologies and Designs:

The cost estimates for cases that only contain fully mature technologies, which have been widely deployed at commercial scale (e.g., high purity cases, which only require compression) reflect nth-of-a-kind on the technology commercialization maturity spectrum. The costs of such technologies have dropped over time due to "learning by doing" and risk reduction benefits that result from serial deployments as well as from continuing research and development (R&D). All process equipment in the estimates found herein is commercially available, so no process contingencies were added to those cases, except for those which require purification (i.e., low purity cases) via acid gas removal as detailed in Section 4.2.

Costs of Emerging Technologies and Designs:

The cost estimates for cases that include technologies that are not yet fully mature (e.g., capture systems for low purity cases) use the same cost estimating methodology as for mature technologies, 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 expected that addition of capture equipment in low purity cases may incur costs higher than those estimated for a mature technology. As such, process contingency of 17 percent is applied to the CO_2 removal system for low purity cases based on engineering judgment and for consistency of process contingencies applied for similar technologies in other NETL studies. [5]

Other Factors:

Actual reported project costs for all the plant types 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, 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.

3.1.1 Capital Costs

As illustrated in Exhibit 3-1, this report defines capital cost at five levels: BEC, EPCC, TPC, TOC, and TASC. BEC, EPCC, TPC, and TOC are "overnight" costs and are expressed in "base-year" dollars. The base year is the first year of capital expenditure. TASC is expressed in mixed, current-year dollars over the entire capital expenditure period, which is assumed to last one year in high purity cases and three years in low purity cases. The cost estimates presented in this study are considered Class 4 estimates, as defined by AACE International (AACE) 16R-90. [10]

The Bare Erected Cost (BEC) comprises the cost of process equipment, on-site facilities and infrastructure that support the plant (e.g., shops, offices, labs, road), and the direct and indirect labor required for its construction and/or installation. The cost of EPC services and contingencies are not included in BEC.

The Engineering, Procurement and Construction Cost (EPCC) comprises the BEC plus the cost of services provided by the EPC contractor. EPC services include detailed design, contractor permitting (i.e., those permits that individual contractors must obtain to perform their scopes of work, as opposed to project permitting, which is not included here), and project/construction management costs.

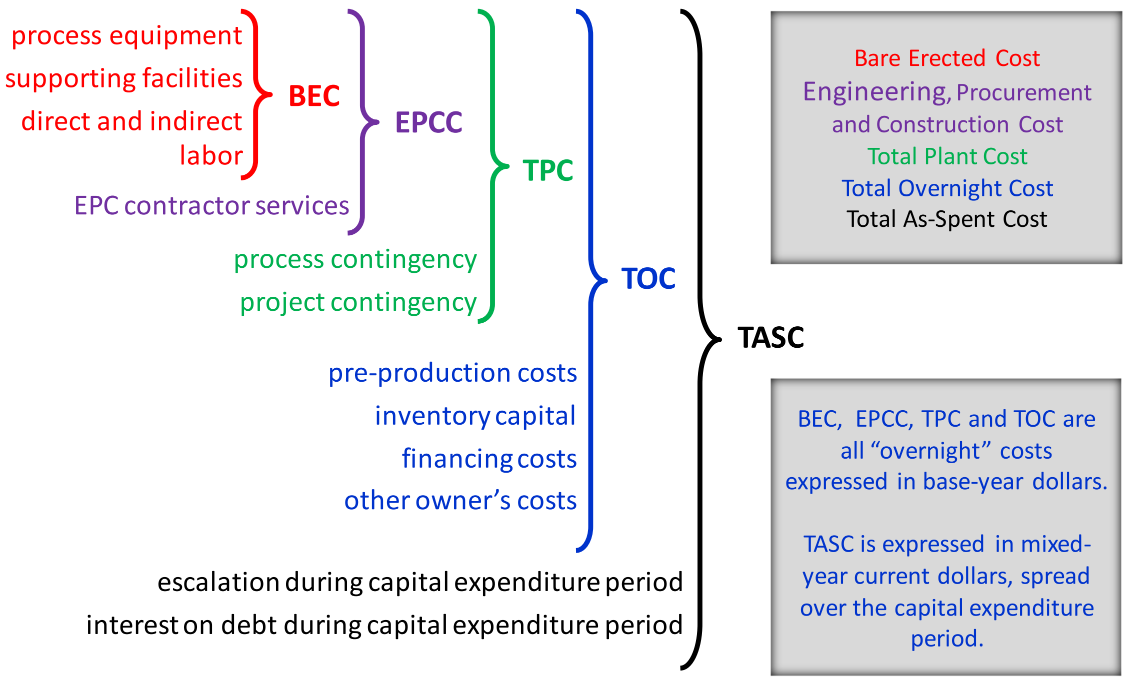
The Total Plant Cost (TPC) comprises the EPCC plus project and process contingencies.

The AACE 16R-90 states that project contingency for a "budget-type" estimate (AACE Class 4 or 5) should be 15–30 percent of the sum of BEC, EPC fees, and process contingency. [10] Therefore, a 20 percent project contingency was added to each cost account across all cases.

The Total Overnight Cost (TOC) comprises the TPC plus all other overnight costs, including owner's costs. TOC does not include escalation during construction or interest during construction.

The Total As-Spent Cost (TASC) is the sum of all capital expenditures as they are incurred during the capital expenditure period including their escalation. TASC also includes interest during construction, comprising interest on debt and a return on equity.

Exhibit 3-1. Capital cost levels and their elements



3.1.1.1 Cost Estimate Basis and Classification

The TPC and O&M costs for each of the cases in the report were estimated based on adjusted vendor-furnished data and scaled estimates from previous NETL studies. Reference costs are scaled based on direction from NETL's QGESS "Capital Cost Scaling Methodology: Revision 4 Report." [3] An underlying assumption of this cost scaling methodology is that capital equipment is available and scalable at any size/capacity. In real applications, equipment may only be manufactured in discrete sizes, which would potentially differ from the costs presented herein. This is particularly applicable for the "Plant Capacity Sensitivity Analysis" found in the analysis subsections for each of the industrial plant types. Those sensitivity analyses are generated assuming continuous equipment capacities and costs and using generic scaling of cost components, rather than by following the QGESS capital cost scaling methodology for every capacity across the plant size range. For the purposes of this analysis, it is assumed that margins of error associated with discrete versus continuous costs and equipment capacities would be within the scope of an AACE Class 4 estimate.

3.1.1.2 System Code-of-Accounts

The costs are grouped according to a process/system-oriented code of accounts. This type of code-of-account structure has the advantage of grouping all reasonably allocable components of a system or process, so they are included in the specific system account.[e]

[e] This would not be the case had a facility, area, or commodity account structure been chosen instead.

3.1.1.3 Price Fluctuations

During the writing of this report, the prices of equipment and bulk materials used as reference costs fluctuated because of various market forces. All vendor quotes used to develop these estimates were adjusted to December 2018 dollars accounting for the price fluctuations. The Chemical Engineering Plant Cost Index [11] was used as needed for these adjustments. While such overall indices are nearly constant, it should be noted that the cost of individual equipment types may still deviate from the December 2018 reference point.

In addition to year dollar effects on the costs presented in this study, the location of the actual installation can influence pricing due to transport and shipping constraints, workforce availability, etc. It is assumed that these contingencies are covered within the range of accuracy of the report (AACE Class 4).

3.1.1.4 Owner's Costs

Owner's costs were estimated based on the 2019 revision of the QGESS document "Cost Estimation Methodology for NETL Assessment of Power Plant Performance." [9] Owner's costs are split into three categories: pre-production costs, inventory capital, and other costs.

Pre-production allocations are expected to carry the specific plants through substantial completion, and to commercial operation. Substantial completion is intended to represent the transfer point of the facility from the EPC contractor (development entity) to the end user or owner, and is typically contingent on mutually acceptable equipment closeout, successful completion of facility-wide performance testing, and full closeout of commercial items. Exhibit 3-2 presents descriptions of the owner's costs estimated for the cases in this report.

Exhibit 3-2. Estimated amounts for owner's costs

Owner's Cost	Estimated Amount
Prepaid Royalties	Any technology royalties are assumed to be included in the associated equipment cost, and thus are not included as an owner's cost
Production (Start-up) Costs	• 6 months operating labor • 1 month maintenance materials at full capacity • 1 month non-fuel consumables at full capacity • 1 month waste disposal • 25% of one month's fuel cost at full capacity • 2% of TPC Compared to AACE 16R-90, this includes additional costs for operating labor (6 months versus 1 month) to cover the cost of training the plant operators, including their participation in startup, and involving them occasionally during the design and construction. AACE 16R-90 [10] and Electric Power Research Institute (EPRI) Technical Assessment Guide (TAG®) [12] differ on the amount of fuel cost to include; this estimate follows EPRI
Inventory Capital	• 0.5% of TPC for spare parts • 60-day supply (at full capacity) of fuel. Not applicable for NG

Owner's Cost	Estimated Amount
	• 60-day supply (at full capacity) of non-fuel consumables (e.g., chemicals and catalysts) that are stored on site. Does not include catalysts and adsorbents that are batch replacements such as water gas shift, carbonyl sulfide, and selective catalytic reduction catalysts and activated carbon AACE 16R-90 [10] does not include an inventory cost for fuel, but EPRI TAG® [12] does
Land	• \$3,000/acre, 10 acres • Note: This land cost is based on a site in a rural location
Financing Costs	• 2.7% of TPC This financing cost (not included by AACE 16R-90 [10]) covers the cost of securing financing, including fees and closing costs but not including interest during construction. The "rule of thumb" estimate (2.7% of TPC) is based on a 2019 professional communication with Black & Veatch
Other Owner's Costs	• 15% of TPC This additional lumped cost is not included by AACE 16R-90 [10] or EPRI TAG® [12]. The "rule of thumb" estimate (15% of TPC) is based on a 2019 professional communication with Black & Veatch

3.1.2 Operation and Maintenance Costs

The production costs or operating costs and related maintenance expenses pertain to those charges associated with operating and maintaining equipment over its expected life. The O&M costs calculated in this study are incremental costs related to the capture, compression, and ancillary equipment evaluated and thus are not indicative of the O&M costs of the base plant. These O&M costs include the following:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal
- Co-product or by-product credit (that is, a negative cost for any by-products sold)

There are two components of O&M costs: fixed O&M, which is independent of production, and variable O&M, which is proportional to production. Taxes and insurance are included as fixed O&M costs, totaling two percent of the TPC.

3.1.2.1 Operating Labor

Operating labor cost was determined based on the number of operators required for the addition of capture and compression where applicable for each case. For high purity cases, which require only the addition of compression and associated utilities, one additional operator was considered. Low purity cases require acid gas removal (AGR) units and an industrial boiler alongside compression and the utilities associated with each additional process unit. As such, 2.3 additional operators were considered for low purity cases, which is the difference in operating labor required for a supercritical pulverized coal power plant with and without capture, per NETL's "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity" results. [5] The average base labor rate used to determine annual cost is $38.50/hour. The associated labor burden is estimated at 30 percent of the base labor rate.

3.1.2.2 Maintenance Material and Labor

Maintenance cost was evaluated based on relationships of maintenance cost to initial capital cost. This represents a weighted analysis in which the individual cost relationships were considered for each major plant component or section.

3.1.2.3 Administrative Support and Labor

Labor administration and overhead charges are assessed at a rate of 25 percent of the burdened O&M labor.

3.1.2.4 Consumables

The cost of consumables, including fuel, was determined based on individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours.

Quantities for major consumables such as NG for fuel and purchased power were taken from technology-specific energy and mass balance diagrams developed for each plant application. Fuel cost is $4.42/MMBtu, and power is purchased at a cost of $60/MWh. Sensitivity analyses relating COC to purchased power price and NG price are detailed in Section 7.2.3 and Section respectively. Other consumables were evaluated based on the quantity required using reference data.

The quantities for initial fills and daily consumables were calculated on a 100 percent operating capacity basis. The annual cost for the daily consumables was then adjusted to incorporate the annual plant operating basis, or capacity factor (CF). An 85 percent CF was assumed for all cases. Initial fills of the consumables, fuels, and chemicals may be accounted for directly in the O&M tables or included with the equipment pricing in the capital cost.

3.1.2.5 Waste Disposal

Waste quantities and disposal costs were determined/evaluated similarly to the consumables. Waste streams are individually reported, and disposal costs are reported for each waste stream, where applicable.

3.2 Capital Charge Factors

The financial assumptions for each case were developed by NETL's Energy Markets Analysis Team in October 2021 based on market data respective to each industrial sector. These factors are summarized in Exhibit 3-3 and Exhibit 3-4. All values are expressed in real dollar terms.

Exhibit 3-3. Financial assumptions for high purity sources

Financial Parameter	Ammonia	EO	Ethanol	NGP	CTL/GTL
Fixed Charge Rate	5.33%	4.63%	6.64%	5.82%	7.32%
TASC/TOC Ratio	1.035	1.025	1.047	1.039	1.054
Capital Charge Factor	5.51%	4.74%	6.96%	6.05%	7.71%
Debt/Equity Ratio	54/46	48/52	36/64	43/57	32/68
Payback Period	30 years				
Interest on Debt	5.15%				
Levered Return on Equity (Asset Weighted)	1.50%	0.04%	4.51%	2.96%	5.54%
Capital Expenditure Period	1 year				
Capital Distribution	1st year – 100%				

Exhibit 3-4. Financial assumptions for low purity sources

Financial Parameter	Refinery Hydrogen	Cement	Iron/Steel
Fixed Charge Rate	4.39%	5.08%	6.90%
TASC/TOC Ratio	1.036	1.054	1.091
Capital Charge Factor	4.55%	5.35%	7.53%
Debt/Equity Ratio	33/67	42/58	39/61
Payback Period	30 years		
Interest on Debt	5.15%		
Levered Return on Equity (Asset Weighted)	0.41%	1.42%	5.02%
Capital Expenditure Period	3 years		
Capital Distribution	1st year – 10%; 2nd year – 60%; 3rd year – 30 %		

The result of the economic analysis is a calculated COC of CO_2, which represents the cost to the owner, per tonne of CO_2 captured. This cost includes the capital expenditures, escalated at the assumed nominal general inflation rate of two percent per year, providing the stipulated rate of return on equity over the entire economic analysis period. Assuming all annual costs also escalate at the same inflation rate, the COC is essentially the sum of the O&M costs and the annualized capital cost charges, all normalized to the annual plant CO_2 flow rate.

For a CO_2 source with a higher flow rate (same CO_2 purity and pressure), a corresponding increase in the flow rate of the captured CO_2, requirement for consumables, size of capture equipment, etc., occurs; however, the COC is expected to be roughly equivalent or, in some cases, lower due to the economies of scale associated with the cost of the larger equipment. This is especially apparent when comparing the costs of each low purity case at two different capture rates (e.g., cement at 90 percent and 99 percent capture). Ultimately, the CCF, which is the product of the fixed charge rate and the TASC/TOC ratio, applied in each case can have a dramatic effect on the COC calculated. A sensitivity analysis evaluating this relationship is presented in Section 7.2.1.

3.3 Retrofit Factors

Retrofit factors for power plants retrofitting amine solvent-based CO_2 capture technologies were developed in the NETL study "Retrofit Cost Analysis for Post-combustion CO_2 Capture" (Retrofit Study). [13] The retrofit factors, as presented in the Retrofit Study, are technology- and size-specific, and significant factors would be ignored when applying them to other configurations, such as the ones in this study. Examples of assumptions that would affect the implementation of the retrofit factors from the Retrofit Study include:

The high purity sources do not require a CO_2 separation system
CO_2 separation is performed using Shell Cansolv post-combustion amine-based capture process in the steel and cement cases, a process that differs from that of the monoethanethiol (MEA) systems that were used to develop the retrofit factors in the Retrofit Study [13]
Shell's ADIP-Ultra amine-based pre-combustion capture process is the basis for purification of the CO_2 stream in the refinery hydrogen case, which differs greatly from the post-combustion MEA systems within the Retrofit Study [13]
These industrial sources are significantly smaller than the utility scale power plants for which the retrofit factors in the Retrofit Study were developed [13]

The areas where these retrofit factors would be more directly applicable are the 'Ductwork & Stack' accounts, which can have a retrofit factor as high as 1.6. The BEC of the 'Ductwork & Stack' account in the cement case with 99 percent capture, for example, is $15,274,000. Application of a 1.6 retrofit factor would add an additional $9,164,400 for the 'Ductwork & Stack' line item. With the cement plant case having a greenfield TOC of $424,897,000 application of this 1.6 retrofit factor would represent a 2.2 percent increase in the TOC for 'Ductwork & Stack' alone.

Engineering judgment was used to determine a more generic factor to be applied to the cases in this report, in lieu of those presented in the Retrofit Study. As an alternative, for high purity cases a retrofit factor of 1.01 was applied to the TPC as a blanket retrofit cost increase, and a retrofit factor of 1.05 was applied to the TPC of low purity cases. Without a formalized procedure for applying the retrofit factors, it is best to consider the retrofit factor as a single capital cost sensitivity, from which the true cost of a retrofit (which has overriding project and site-specific considerations) can be refined as more information is available for a specific design case. A sensitivity analysis examining the effect on COC related to the retrofit factor applied is discussed in Section 7.2.2.

4 Equipment

4.1 Compression

Two different types of compressors are used for the cases in this study, an integrally geared centrifugal compressor and a reciprocating compressor. The type of compressor selected for each case is chosen based on the mass flow of CO_2 to the first compression stage as well as the suction conditions at stage one.

4.1.1 Reciprocating Compressor

A quote for a five-stage reciprocating compressor was used to represent compression for cases listed in Exhibit 4-1. The referenced compression quoted a suction pressure of 17.4 psia, suction temperature of 80°F, and an inlet flow to stage one of 35,991 lb/hr. The discharge pressure was quoted as 2,200 psia with a total power requirement of 1.72 MW. The reciprocating compressor was modeled with alterations as applicable, resulting in the specifications shown in Exhibit 4-1.

Exhibit 4-1. Reciprocating compressor cases specifications

Case	Number of Compression Stages	Inlet Flow to Compression Stage 1 (lb/hr)	Suction Pressure (psia)	Suction Temperature (°F)	Discharge Pressure (psia)
Ammonia	5	122,946	23.5	69	2,214.7
EO	4	30,578	43.5	96	2,214.7
Ethanol	5	36,000	16.4	80	2,214.7

4.1.2 Centrifugal Compressor

Quotes for integrally geared centrifugal compressors were used to represent compression in the cases listed in Exhibit 4-2. Two separate quotes were used, the first of which was provided for the development of NETL's "Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity," Revision 4 (BBR4). [5] The second quote for a centrifugal compressor was obtained as part of the development of this study, specifically for application in the refinery hydrogen case.

Given that the CTL and GTL cases are taken from previous NETL reports, they implement the same compression train performance and cost used in their respective reports, converted to current year dollar. Those reports employ integrally geared centrifugal compressors specifically designed for their respective CO_2 flowrates and conditions. This type of compressor is particularly advantageous for CTL and refinery hydrogen cases, where CO_2 is available at multiple pressures, and requires a special compression train that can accommodate multiple suction pressures. Exhibit 4-2 shows the cases using integrally geared centrifugal compression and their case specifications.

Exhibit 4-2. Integrally geared centrifugal compressor cases specifications

Case	Number of Compression Stages	Inlet Flow to Compression Stage 1 (lb/hr)	Suction Pressure (psia)	Suction Temperature (°F)	Discharge Pressure (psia)
NGP	8	164,059	23.5	69	2,214.7
Steel/Iron COG/BFS 90% Capture	8	424,424	28.9	87.8	2,214.7
Steel/Iron COG/BFS 99% Capture	8	466,701	28.9	87.8	2,214.7
Steel/Iron COG PPS 90% Capture	8	426,791	28.9	87.8	2,214.7
Steel/Iron COG PPS 99% Capture	8	469,304	28.9	87.8	2,214.7
Cement 90% Capture	8	275,388	28.9	87.8	2,214.7
Cement 99% Capture	8	302,818	28.9	87.8	2,214.7
Refinery Hydrogen 90% Capture	7	93,136[B, C]	28.3/90.8[D]	104.0/215.6	2,214.7
Refinery Hydrogen 99% Capture	7	104,553[B, C]	28.3/90.8[D]	104.0/215.6	2,214.7
CTL	N/A[A]	2,200,423[B]	160/265/300[E]	N/A	2,214.7
GTL	N/A[A]	467,794	265	100	2,214.7

[A] Both CTL and GTL are assumed to use eight total compression stages, but this is not explicitly stated in the respective reports.
[B] Flow reported is total. The individual flows at each of the multiple suction pressures sum to the total flow.
[C] These flowrates fall below the lower operating limit detailed in Section 4.1.1, but a specific performance and cost quote was obtained for application in the refinery hydrogen cases. The quote data is proprietary; thus, details are not included within this report.
[D] A second inlet to compression was considered as part of the compressor design (proprietary) for refinery hydrogen cases due to AGR specifications and process flow.
[E] The CTL process produces three high purity CO_2 streams at three pressures. Details related to the compressor for the CTL case are provided in Section 5.5.

As mentioned, all compressors discharge at a pressure of 2,214.7 psia (2,200 psig). This is the pipeline pressure specification assumed in this study, which is given in the QGESS for CO_2 for use in EOR applications. [1] However, it should be noted that EOR field pressure requirements can vary from location to location, and pressures as low as 1,200 psig could be acceptable. [14]

4.2 CO_2 Capture and Purification[f]

For cases requiring CO_2 separation and purification prior to compression, an AGR unit was used. The AGR unit also provides polishing of residual sulfur components in the CO_2 capture stream.

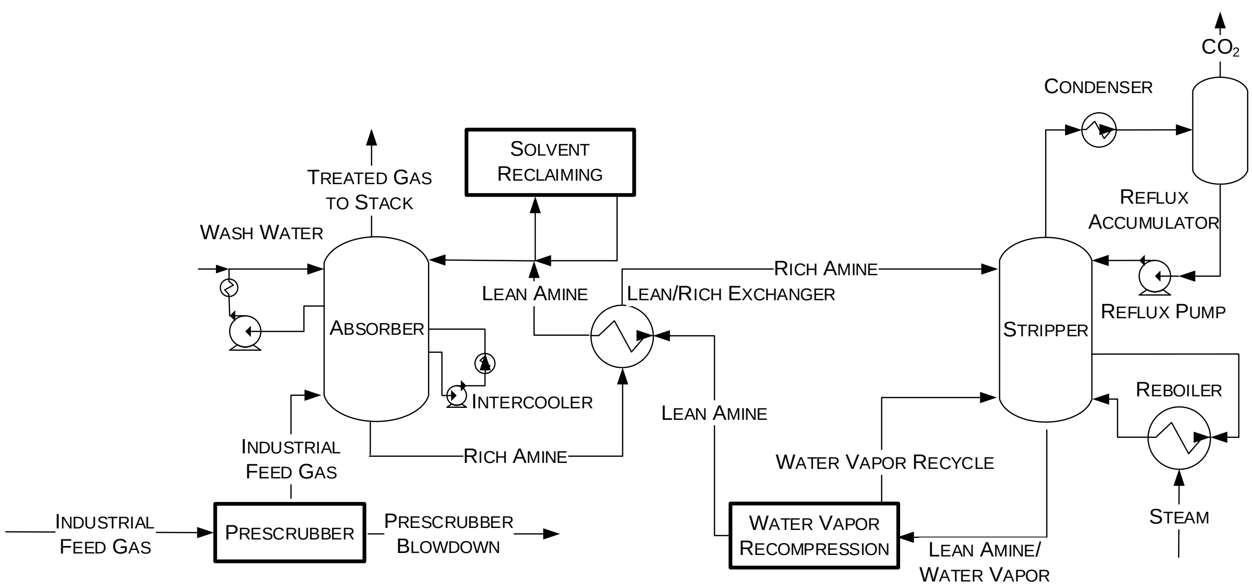
[f] Much of the text and descriptions within this section were sourced, with permission, from data provided by Shell to NETL, unless otherwise noted. The information relates to a CO_2 removal system designed by Shell.

The performance and cost information for the AGR units employed in this study are based on data provided by Shell in 2021. The quote provided specific cost and performance metrics at individual capture rates (i.e., 90, 95, and 99 percent) for each representative industrial plant. The unit cost is scaled based on CO_2 product mass flow (60 percent) and inlet flow to the adsorber (40 percent), per specifications in “QGESS Capital Cost Scaling Methodology: Revision 4 Report.” [3] Cases where an AGR is used include refinery hydrogen, iron/steel, and cement. The CO_2 removal efficiency of the AGR unit is represented at two rates, 90 percent and 99 percent, for each case. For the purposes of this study, performance and cost data for the AGR units was obtained from Shell for the specific flue gas streams representative of the low purity industrial sources, not scaled or applied from quotes provided for power-related capture systems.

4.2.1 Cansolv Post-Combustion Capture

The AGR system utilized in the iron/steel and cement cases is the Cansolv CO_2 Capture technology commercially offered by Shell. This amine-based, post-combustion process is designed to recover high purity CO_2 from dilute streams that contain O_2, such as flue gas from coal-fired power plants, combustion turbine exhaust gas, and other industrial waste gas streams, such as those evaluated in this report. A typical flowsheet for the process is shown in Exhibit 4-3.

Exhibit 4-3. Shell’s Cansolv CO_2 capture typical process flow diagram



4.2.1.1 Pre-scrubber

The CO_2-laden gas from the industrial source (cement or iron/steel plant) is sent through a booster fan to drive the gas through downstream equipment starting with the pre-scrubber inlet cooling section. The cooler is operated as a direct contact cooler that saturates and sub-cools the feed gas stream. Saturation and sub-cooling are beneficial to the system as they improve the amine absorption capacity, thus reducing amine circulation rate. In cement or steel applications, in or after the cooling section the feed gas is also scrubbed with caustic to capture residual acid compounds (SO_2, hydrogen chloride, etc.).

4.2.1.2 CO_2 Absorber

The Cansolv absorber is a single, rectangular, acid resistant, steel- or resin-lined concrete structure containing stainless-steel packing, a typical design for large-scale units. There is a packed section used for CO_2 absorption, and another packed section used for water-wash. This specific absorber geometry and design provides several cost advantages over more traditional column configurations while maintaining equivalent or elevated performance. The feed gas enters the absorber and flows counter-current to the Cansolv solvent.

The lean solvent absorbs 90–99 percent of the inlet CO_2, depending on the design capture rate, and the remaining CO_2 exits the main absorber section and enters the water-wash section of the absorber. Prior to entering the bottom packing section, hot amine is collected, removed, and pumped through a heat exchanger (HX) to provide intercooling and maintain a low temperature favorable to absorption. The cooled amine is then sent back to the absorber just above the final packed section.

The water-wash section at the top of the absorber is used to remove volatiles or entrained amine from the treated gas, as well as to condense and retain water in the system. The wash water is removed from the bottom of the wash section, pumped through a HX, and is then re-introduced at the top of the wash section. This wash water is made up of recirculated wash water as well as water condensed from the treated gas; excess water resulting from condensation overflows to the lower absorption section through a chimney tray. The CO_2-lean gas treated in the water-wash section is then released to the atmosphere.

4.2.1.3 Amine Regeneration

The rich amine is collected at the bottom of the absorber and pumped through multiple parallel rich/lean HXs where heat from the lean amine is exchanged with the rich amine. The Cansolv rich/lean solvent HXs are a stainless-steel plate and frame type with a typical 5°C (9°F) approach temperature. The rich amine continues and enters the stripper near the top of the column.

The stripper is a stainless-steel vessel using structured stainless-steel packing. The regenerator reboiler uses low pressure steam to boil water vapor from the solvent; this vapor flows upwards, counter-current to the rich amine flowing downwards, and removes CO_2 from the amine. Steam is provided by the NG-fired boiler described in Section 4.3. The Cansolv regenerator reboiler is a stainless-steel plate and frame type with a 3°C (5°F) approach temperature. Lean amine is collected in the stripper bottoms and flows to a flash vessel where water vapor is released. This lean solvent is then pumped through the same rich/lean HX to exchange heat from the lean amine to the rich amine and continues to the lean amine tank.

The water vapor and stripped CO_2 flow up the stripper where they are contacted with recycled reflux to condense a portion of the vapor and collect entrained solvent droplets. The remaining gas continues to the condenser where it is partially condensed. The two-phase mixture then flows to a reflux accumulator where the CO_2 product gas is separated and sent to the CO_2

compressor at approximately 0.2 MPa (29 psia), and the remaining water is collected and returned to the stripper as reflux.

The flow of steam to the regenerator reboiler is proportional to the rich amine flow to the stripper; however, the flow of low-pressure steam is also dependent on the stripper top temperature.

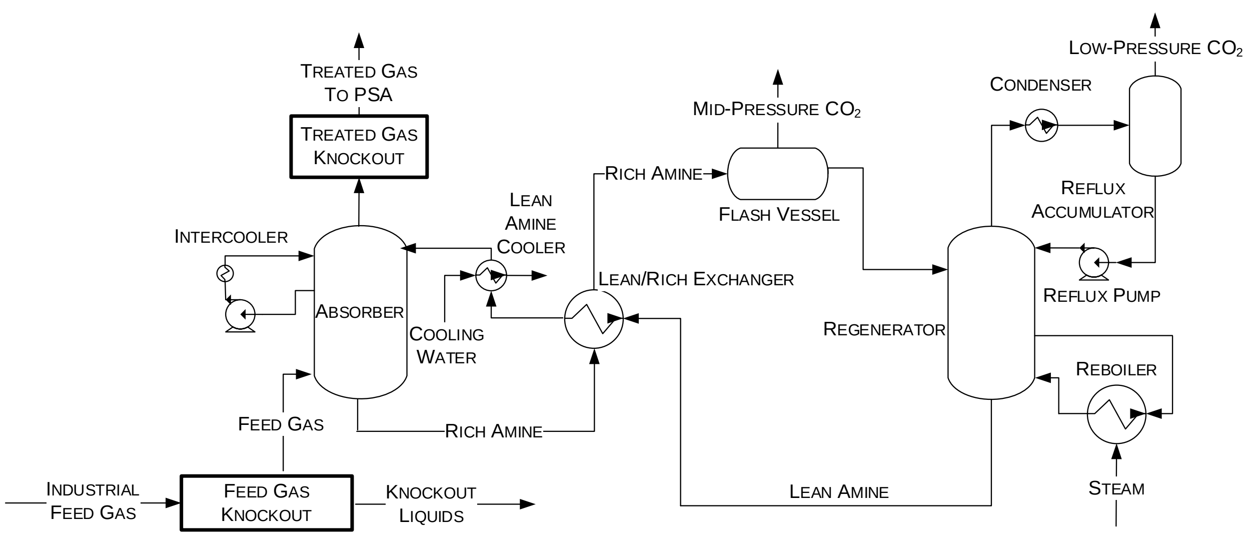
4.2.1.4 Amine Purification

The purpose of the amine purification, or amine reclaiming, section is to remove a portion of the heat-stable salts as well as ionic and non-ionic amine degradation products. The Cansolv amine purification (reclaiming) is essentially a distillation operation, in which the usable amine is boiled off the degraded solvent, which is recovered at the bottom of the column for disposal.

4.2.2 ADIP-Ultra Pre-Combustion Capture

The AGR utilized in the refinery hydrogen case is the ADIP-Ultra CO_2 capture technology developed by Shell. This pre-combustion process, the latest evolution of the ADIP-Ultra process, uses a proprietary amine-based solvent capable of bulk removal of CO_2 from high pressure gas streams. This technology has been deployed and is currently in operation at Shell's Quest facility in Alberta, Canada. [15] A typical flowsheet is shown in Exhibit 4-4.

Exhibit 4-4. ADIP-Ultra CO_2 capture typical process flow diagram



4.2.2.1 CO_2 Absorber

The feed gas is sent through a knockout vessel to remove water and liquid hydrocarbons if any are present. The knockout vessel produces a saturated vapor stream that is sent to the CO_2 absorber. A lean solvent stream enters the top of the absorber and flows down over trays to absorb CO_2 from the feed gas stream. The feed gas stream flows countercurrent to the solvent stream, which absorbs 90–99 percent of inlet CO_2, depending on the design capture rate.

Treated gas exits through the top of the absorber and is sent through a second knockout vessel to remove entrained amine droplets using a mist pad before being routed to the pressure-swing adsorption unit for the production of high purity hydrogen. A rich solvent stream exits through the sump of the absorber and is routed towards the amine regeneration section.

4.2.2.2 Amine Regeneration

The rich solvent stream flows through a rich/lean HX, where rich solvent is heated by lean solvent moving to the absorber. To minimize reboiler duty and compression power, part of the CO_2 (mid-pressure) in the rich amine is then flashed off in a hot flash vessel and routed towards compression and dehydration.

The remaining rich amine liquid continues to the stripper, entering near the top of the column. The regenerator reboiler indirectly uses low pressure steam to produce water vapor that flows upwards, counter-current to the rich amine flowing downwards, and removes CO_2 from the amine. Steam is provided by the NG-fired boiler described in Section 4.3. The lean solvent flows from the bottom of the regenerator tower and is pumped through the same rich/lean HX to exchange heat from the lean amine to the rich amine and continues to the absorber.

The acid gas from the stripping section is washed in the water wash section of the regenerator to remove entrained amine. The gas is then cooled in an overhead condenser and sent to a reflux vessel where CO_2 and water are separated. Low-pressure CO_2 is sent to compression and dehydration, while water is returned to the stripper via regenerator reflux pumps.

4.3 Industrial Boiler

AGR unit configurations detailed in the prior two sections require low pressure steam at 71 psia for solvent regeneration. Since no assumptions regarding available steam are made about the base plants, cases requiring CO_2 separation and purification also require the addition of a boiler for steam production.

A quote for an industrial steam boiler was obtained from CleaverBrooks in March 2021. [16] The boiler produces superheated steam at 100 psig. For each case requiring an AGR unit, the total heat required from 71 psia steam for solvent regeneration was calculated, and that amount of heat delivered from the referenced boiler was modeled as part of the Aspen Plus® (Aspen) simulation. Boiler auxiliary power requirements for pumps and compressors were scaled based on the quoted information. Consumables include NG fuel usage, as predicted by the Aspen model for each case, and feedwater makeup, calculated by methods consistent with those used to estimate feedwater makeup in BBR4 cases.

4.4 Cooling Water Unit

As previously stated, no characterization of the base plant for each process was assessed; as such, no assumptions were made regarding the existing plant's cooling water system. Therefore, it is assumed for the purpose of this report that any cooling required by the compression train, and in some cases the AGR unit, must be supplied by a stand-alone cooling water unit.

Power consumption estimates for the cooling water system (i.e., circulating water pumps and cooling tower fans) were calculated based on methodology consistent with that of BBR4 cases. Cost estimates for the cooling water system were scaled from Case B11A-BR of NETL's "Eliminating the Derate of Carbon Capture Retrofits" (Derate Study) based on QGESS guidance for capital cost scaling. [17] [3] This account was scaled from the Derate Study because Case B11A-BR is more representative of the size range for the cooling water system associated with the cases in this report.

4.5 Heat Exchangers

Cooling of the product CO_2 is required for all cases following compression to meet the pipeline temperature specification of 86°F, and in some cases, cooling is also required preceding compression. For cases using a reciprocating compressor, post-cooling of the compressed product CO_2 is included in the compressor quote. The quoted discharge temperature of the centrifugal compressors referenced are higher than the pipeline specification temperature of 86°F and require cooling. For those cases, after-cooler costs were scaled from BBR4 Case B12B based on HX duty as predicted by Aspen, consistent with QGESS cost scaling methodology. Cases with reciprocating compression do not depict an aftercooler HX in the block flow diagrams (BFDs) throughout Section 5. For the cases with centrifugal compression, the HX is depicted downstream of the compressor in the BFDs throughout Section 5 and Section 6.

Cooling of the CO_2 at the inlet of the compression train is dependent on the quoted compression train suction temperature and the base plant assumptions regarding the temperature at which the CO_2 is available. A pre-cooler HX is required only for the Ethanol case, where fermentation produces a CO_2 stream with a temperature of 320°F, which far exceeds the suction temperature of the reciprocating compressor employed. The cost of this exchange was developed from heuristics in Analysis, Synthesis, and Design of Chemical Processes, assuming a floating head shell-and-tube HX with a heat transfer coefficient equal to 6.2 Btu/hour-square foot-°F. [18]

4.6 Ancillary Equipment, Buildings, and Structures

Ancillary equipment associated with implementing the capture and compression systems in this report include an accessory electrical plant and instrumentation and control (I&C) equipment. In addition, some site improvements, such as ground preparation and additional facilities, would be required for the construction and ongoing operation of the equipment considered. Estimates for these costs were scaled per QGESS guidance based on Case B11A-BR of the Derate Study, as the costs of this reference case are approximately comparable to those that would be incurred with the addition of the equipment detailed throughout the prior sub-sections. [17]

5 COST AND PERFORMANCE: HIGH PURITY SOURCES

The sources discussed in this section are considered high purity sources, meaning the available CO_2 does not require AGR to meet EOR pipeline specifications. In some high purity cases, dehydration of the CO_2 stream using a triethylene glycol (TEG) system may be required.

5.1 AMMONIA

It is estimated that the U.S. gross ammonia production in 2019 was over 19.2 M tonnes. [19] In all but one plant in the United States, the ammonia production process first reforms a NG feedstock to produce hydrogen (H_2), carbon monoxide (CO), and CO_2. The unconverted CO from reforming is then shifted to produce more H_2 and CO_2. The optimum ratio of H:N for ammonia synthesis is 3:1; therefore, the amount of CO_2 removed from the post-shift stream must be high to optimize the H:N ratio. A portion of the CO_2 removed from the post-shift stream is often captured and reused to produce urea, by reacting ammonia with CO_2. The amount of CO_2 captured and reused for ammonia derivatives will vary from plant to plant based on production capacities and market opportunities for each product. With CO_2 removal inherent to the ammonia process, coupled with the need for CO_2 to convert ammonia into ammonia derivatives, ammonia processing is a potentially low-cost option for industrial CO_2 capture.

5.1.1 Size Range

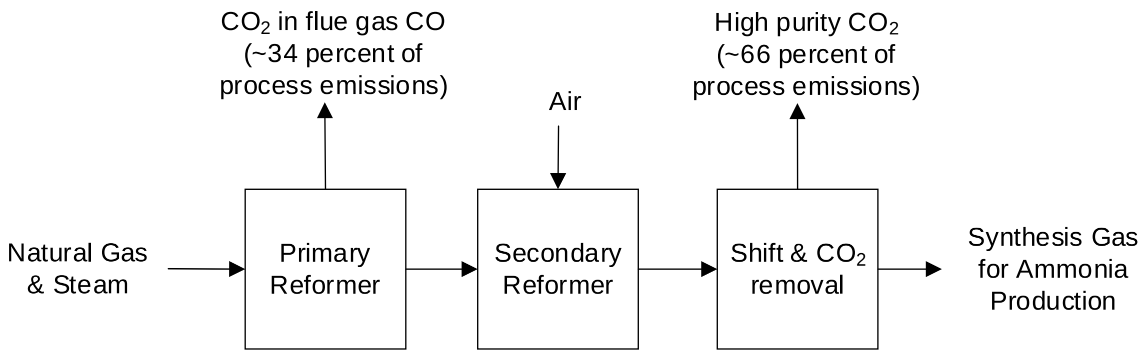
As of 2019, there were 32 ammonia plants in the United States, 19 of which fell in the range of 0.1–0.6 M tonnes/year (0.11–0.66 M tons/year) production capacity, and nine had a capacity of 600,000 tonnes/year or greater. The largest U.S. ammonia plant has a capacity of 4.3 M tonnes/year. [19] For the purposes of this study, the ammonia case is represented with a production capacity of 394,000 tonnes ammonia/year.

5.1.2 CO_2 Point Sources

The main point sources of CO_2 emissions in an ammonia plant comes from the flue gas from the primary reformer and the vent from the CO_2 stripper that separates CO_2 from the ammonia syngas. Of these two, only the CO_2 stripper vent is considered a high purity source of CO_2. The primary reformer flue gas has a CO_2 concentration of approximately 18 mol% and would be considered a low purity source of CO_2. [20] [21] As such, it is not considered in this study case but may be evaluated as part of future work, as discussed in Section 9.1.

An article published by KBR Technology [22] concerning CO_2 capture in the ammonia industry stated that for an average ammonia plant producing 660,000 tonnes/year ammonia, approximately 34 percent of CO_2 emissions come from the primary reformer flue gas and 66 percent are emitted by the CO_2 stripper vent. The total CO_2 produced in ammonia production (i.e., that of both the primary reformer and the CO_2 stripper) is 1.87 tonnes CO_2/tonne ammonia. [22] Applying this emissions factor and the fact that 66 percent of the CO_2 emissions would be captured from the stripper vent as a high purity source, the representative 394,000 tonnes ammonia/year plant produces 486,227 tonnes CO_2 vented from the CO_2 stripper. It is assumed that the stripper vent CO_2 concentration is 99 percent by volume. [23] The ammonia production process, using NG as a feedstock, is depicted in a basic BFD (Exhibit 5-1) to further illustrate the point-sources of CO_2 described in this section.

Exhibit 5-1. Ammonia production via NG reforming



In some ammonia production facilities, portions of the ammonia and the CO_2 emissions are further processed to create ammonia derivatives. For this study, it is assumed that the ammonia produced by the representative plant is not used for derivative production, and as such, the CO_2 emitted is not needed for reprocessing within the plant. In practical applications, the amount of CO_2 available would be affected by derivative manufacturing, as well as by process configurations and operating parameters affecting the ratio of CO_2 emitted from the stripper and the primary reformer. This would have to be evaluated on a case-by-case basis, and the assumptions in this study are employed to present an illustrative COC in a representative ammonia production plant.

5.1.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the ammonia process for the purpose of this study:

- The representative ammonia plant has a capacity of 394,000 tonnes ammonia per year
- The ammonia process feedstock is NG
- The gas from the stripper vent is assumed 99 volume percent CO_2 and the balance of the stream (1 volume percent) is assumed to be water
- The total high purity CO_2 amount produced by the plant is 736,750 tonnes CO_2/year (at 100 percent CF); the amount generated from the stripper vent is 486,227 tonnes CO_2/year at 100 percent CF and neglecting process losses or CO_2 reuse in ammonia derivative production
- The temperature of the CO_2 at the stripper vent outlet is 69°F
- The pressure of the CO_2 at the stripper vent outlet is 23.52 psia
- The end product CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

5.1.4 CO_2 Capture System

Only cooling and compression is required for the ammonia case. Reciprocating compression discussed previously in Section 4.1.1 is modeled and the costs for the compressor and ancillary equipment is estimated as outlined in Section 3 and Section 4. Based on mass flow rate, this represents a large scale with up to 3.39 times the quoted flow rate.

5.1.5 BFD, Stream Table, and Performance Summary

There is no cooling of the high purity CO_2 stream from the ammonia plant since it is assumed that the overhead condenser of the stripping column discharges at a temperature of 69°F. A water knockout step is considered to avoid water condensation within the compression train. The costs for the water knockout were estimated using methods in Analysis, Synthesis, and Design of Chemical Processes. [18] After compression, the CO_2 product stream is cooled and sent directly for EOR or other usage. Exhibit 5-2 gives the BFD for this process. Exhibit 5-3 provides the stream table.

Exhibit 5-2. Ammonia CO_2 capture BFD

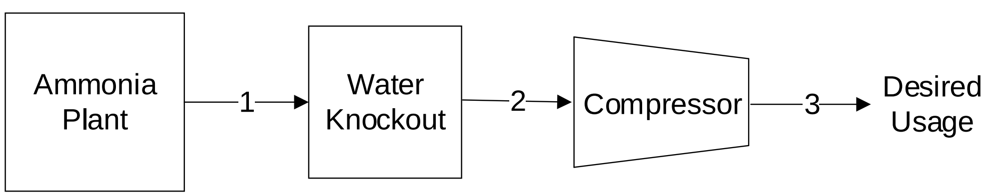


Exhibit 5-3. Ammonia stream table

	1	2	3
V-L Mole Fraction			
Ar	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000
CO_2	0.9709	0.9887	0.9995
SO_2	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000
H_2O	0.0291	0.0113	0.0005
H_2S	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	1,299	1,276	1,261
V-L Flowrate (kg/hr)	56,189	55,767	55,488
Temperature (°C)	21	21	30
Pressure (MPa, abs)	0.16	0.2	15.3

	1	2	3
Steam Table Enthalpy (kJ/kg)[A]	8,841	8,791	8,755
Aspen Plus Enthalpy (kJ/kg)[B]	-9,021	-8,968	-9,195
Density (kg/m^3)	3.0	2.9	630.1
V-L Molecular Weight	43.3	43.7	44.0
V-L Flowrate (lb$_{mol}$/hr)	2,864	2,812	2,780
V-L Flowrate (lb/hr)	123,876	122,946	122,330
Temperature (°F)	69	69	86
Pressure (psia)	23.5	23.5	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	3,801	3,779	3,764
Aspen Plus Enthalpy (Btu/lb)[B]	-3,878	-3,855	-3,953
Density (lb/ft^3)	0.184	0.183	39.3

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance results are based on the reciprocating compressor quote and are provided in Exhibit 5-4.

Exhibit 5-4. Performance summary

Performance Summary	
Item	**394,000 tonnes ammonia/year (kWe)**
CO_2 Compressor	5,770
Circulating Water Pumps	60
Cooling Tower Fans	30
Total Auxiliary Load	**5,860**

5.1.6 Capture Integration

In an existing ammonia plant, a cooling water system that could accommodate the additional cooling needs of the compressor intercoolers modeled in this case may be in place to satisfy the condenser cooling duty for the CO_2 removal system. This is especially true if an ammonia plant is designed to produce ammonia derivatives. However, for this study, a stand-alone cooling system is required to provide for the compressor's intercooling needs. In real applications, the inclusion of an additional cooling water system would be evaluated on a case-by-case basis.

5.1.7 Power Source

Given the relatively small amount of CO_2, the compression power consumption is 5.77 MW. Power consumption estimates for the cooling system were scaled as described in Section 4.4. The total power requirement was calculated to be 5.86 MW, which includes all power required by the compression train and the cooling water system. Purchased power cost is estimated at a rate of $60/MWh as discussed in Section 3.1.2.4.

5.1.8 Economic Analysis Results

The economic results for CO_2 capture application in an ammonia plant are presented in this section. Owner's costs (Exhibit 5-5), capital costs (Exhibit 5-6), and O&M costs are calculated as discussed in Section 3.1. Retrofit costs were determined by applying a retrofit factor to TPC as discussed in Section 3.3. The greenfield TOC for the ammonia case is $45.6 M. The corresponding greenfield COC is $19.0/tonne CO_2, and the COC is $19.0/tonne CO_2 in retrofit applications. The small difference between greenfield and retrofit COC in this case is not apparent due to rounding.

Exhibit 5-5. Owner's costs for ammonia greenfield site

Description	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs		
6 Months All Labor	$423	$1
1-Month Maintenance Materials	$35	$0
1-Month Non-Fuel Consumables	$70	$0
1-Month Waste Disposal	$3	$0
25% of 1-Month Fuel Cost at 100% CF	$0	$0
2% of TPC	$747	$2
Total	**$1,278**	**$3**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$134	$0
0.5% of TPC (spare parts)	$187	$0
Total	**$321**	**$1**
Other Costs		
Initial Cost for Catalyst and Chemicals	$0	$0
Land	$30	$0
Other Owner's Costs	$5,602	$12
Financing Costs	$1,008	$2
TOC	**$45,587**	**$94**
TASC Multiplier (Ammonia, 31 year)	1.035	
TASC	**$47,162**	**$97**

Exhibit 5-6. Capital costs for ammonia greenfield site

	Case:	Ammonia						Estimate Type:		Conceptual	
	Representative Plant Size:	394,000 tonnes ammonia/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
	5	Flue Gas Cleanup									
5.1	Inlet Water Knockout for Compression	$11	$0	$2	$0	$14	$2	$0	$3	$19	$0
5.4	CO_2 Compression & Drying	$6,192	$929	$2,070	$0	$9,192	$1,609	$0	$2,160	$12,960	$27
5.5	CO_2 Compressor Aftercooler	w/5.4	w/5.4	w/5.4	w/5.4	$0	$0	$0	$0	$0	$0
5.7	TEG Dryer (within compression train)	$1,900	$285	$635	$0	$2,821	$494	$0	$663	$3,977	$8
	Subtotal	**$8,104**	**$1,214**	**$2,708**	**$0**	**$12,026**	**$2,105**	**$0**	**$2,826**	**$16,957**	**$35**
	7	Ductwork & Stack									
7.3	Ductwork	$0	$164	$114	$0	$277	$49	$0	$65	$391	$1
	Subtotal	**$0**	**$164**	**$114**	**$0**	**$277**	**$49**	**$0**	**$65**	**$391**	**$1**
	9	Cooling Water System									
9.1	Cooling Towers	$163	$0	$50	$0	$213	$37	$0	$50	$301	$1
9.2	Circulating Water Pumps	$13	$0	$1	$0	$14	$2	$0	$3	$20	$0
9.3	Circulating Water System Aux.	$321	$0	$42	$0	$364	$64	$0	$85	$513	$1
9.4	Circulating Water Piping	$0	$149	$135	$0	$283	$50	$0	$67	$399	$1
9.5	Make-up Water System	$52	$0	$67	$0	$119	$21	$0	$28	$167	$0
9.6	Component Cooling Water System	$23	$0	$18	$0	$41	$7	$0	$10	$58	$0
9.7	Circulating Water System Foundations	$0	$19	$31	$0	$50	$9	$0	$12	$71	$0
	Subtotal	**$572**	**$167**	**$344**	**$0**	**$1,084**	**$190**	**$0**	**$255**	**$1,528**	**$3**
	11	Accessory Electric Plant									
11.2	Station Service Equipment	$1,725	$0	$148	$0	$1,873	$328	$0	$440	$2,642	$5
11.3	Switchgear & Motor Control	$2,679	$0	$465	$0	$3,143	$550	$0	$739	$4,432	$9
11.4	Conduit & Cable Tray	$0	$348	$1,003	$0	$1,352	$237	$0	$318	$1,906	$4
11.5	Wire & Cable	$0	$922	$1,648	$0	$2,570	$450	$0	$604	$3,624	$7
	Subtotal	**$4,404**	**$1,270**	**$3,265**	**$0**	**$8,939**	**$1,564**	**$0**	**$2,101**	**$12,604**	**$26**
	12	Instrumentation & Control									
12.8	Instrument Wiring & Tubing	$353	$282	$1,130	$0	$1,765	$309	$0	$415	$2,489	$5
12.9	Other I&C Equipment	$434	$0	$1,005	$0	$1,439	$252	$0	$338	$2,029	$4
	Subtotal	**$787**	**$282**	**$2,135**	**$0**	**$3,204**	**$561**	**$0**	**$753**	**$4,518**	**$9**
	13	Improvements to Site									
13.1	Site Preparation	$0	$22	$440	$0	$461	$81	$0	$108	$651	$1
13.2	Site Improvements	$0	$102	$136	$0	$238	$42	$0	$56	$336	$1
13.3	Site Facilities	$117	$0	$123	$0	$240	$42	$0	$56	$339	$1
	Subtotal	**$117**	**$124**	**$698**	**$0**	**$940**	**$164**	**$0**	**$221**	**$1,325**	**$3**
	14	Buildings & Structures									
14.5	Circulation Water Pumphouse	$0	$9	$7	$0	$17	$3	$0	$4	$23	$0
	Subtotal	**$0**	**$9**	**$7**	**$0**	**$17**	**$3**	**$0**	**$4**	**$23**	**$0**
	Total	**$13,985**	**$3,231**	**$9,271**	**$0**	**$26,487**	**$4,635**	**$0**	**$6,225**	**$37,347**	**$77**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 5-7 while Exhibit 5-8 shows the COC for greenfield and retrofit sites for the representative ammonia plant.

Exhibit 5-7. Initial and annual O&M costs for ammonia greenfield site

Case:	**Ammonia**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**394,000 tonnes ammonia/year**				**Capacity Factor (%):**	**85**
			O&M Labor			
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		1.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**1.0**
			Fixed Operating Costs			
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$438,438	$0.90
Maintenance Labor:					$239,021	$0.49
Administrative & Support Labor:					$169,365	$0.35
Property Taxes and Insurance:					$746,941	$1.54
Total:					**$1,593,765**	**$3.28**
			Variable Operating Costs			
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$358,532	$0.87
			Consumables			
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	46	$1.90	$0	$27,119	$0.07
Makeup and Waste Water Treatment Chemicals (ton):	0	0.1	$550.00	$0	$24,747	$0.06
Triethylene Glycol (gal):	w/equip.	312	$6.80	$0	$658,287	$1.59
Subtotal:				**$0**	**$710,152**	**$1.72**
			Waste Disposal			
Triethylene Glycol (gal):		312	$0.35	$0	$33,882	$0.08
Subtotal:				**$0**	**$33,882**	**$0.08**
Variable Operating Costs Total:				**$0**	**$1,102,566**	**$2.67**

Exhibit 5-8. COC for 394,000 tonnes/year ammonia greenfield and retrofit[A]

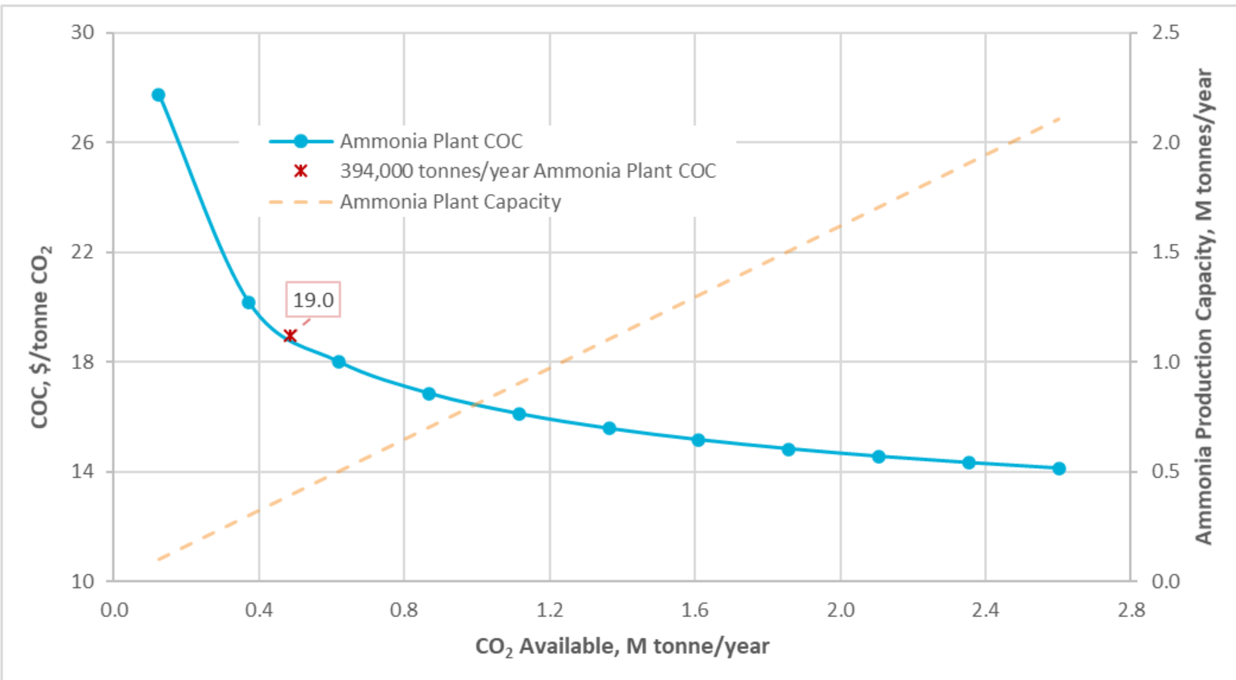
Component	Greenfield Value, \$/tonne CO_2	Retrofit Value, \$/tonne CO_2
Capital	6.1	6.1
Fixed	3.9	3.9
Variable	2.7	2.7
Purchased Power	6.3	6.3
Total COC	**19.0**	**19.0**

[A]Differences in COC for greenfield and retrofit applications of this case are not apparent due to rounding.

5.1.9 Plant Capacity Sensitivity Analysis

An analysis of the sensitivity of greenfield COC to ammonia plant capacity is shown in Exhibit 5-9. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 5-9. Ammonia plant capacity sensitivity



Note: The data point for the COC at a 394,000 tonnes/year ammonia plant does not fall on the COC line due to data point increments and plot formatting.

5.1.10 Ammonia Conclusion

The high purity CO_2 stream produced from ammonia plants makes them a relatively low-cost industrial process for CO_2 capture since the plant itself acts as the separation medium. Economic analysis of the additional CO_2 compression system required for capture resulted in a COC of CO_2 equal to \$19.0/tonne CO_2 for a greenfield site and \$19.0/tonne CO_2 for a retrofit application. The small disparities (not visible due to rounding[g]) between greenfield and retrofit cases are the result of unknown difficulties required for a retrofit installation versus a greenfield application, assuming adequate plot plan space for the retrofit case exists. The sensitivity analysis for plant capacity, when varied from 0.1 M tonnes/year to 2.1 M tonnes/year ammonia production, showed a change in COC of \$13.6/tonne CO_2.

[g] For instance, the TASC for the retrofit ammonia case is \$47.5 million, which is higher in comparison to the TASC for the greenfield ammonia case (i.e., \$47.2 million) as presented in Exhibit 5-5.

It should be noted that for existing U.S. ammonia plants producing excess high purity CO_2, this CO_2 may already be processed and sold for other uses. For example, in addition to urea and other ammonia derivative production, some ammonia plants also produce food-grade liquid CO_2 as a sellable product. This would reduce or eliminate the amount of high purity CO_2 potentially available for capture as evaluated in this study. This scenario was not considered in this study as it would need to be evaluated on a case-by-case basis.

5.2 Ethylene Oxide

Ethylene oxide (EO) is a colorless flammable gas that is mainly used as a raw material for production of several industrial chemical intermediates. When assessed by region, 73 percent of North American EO production goes directly to synthesis of ethylene glycol, which is used in antifreeze, polyester, liquid solvents, and plastics production. [24]

EO is produced by direct oxidation of ethylene in the presence of a silver catalyst. The reaction conditions range 200–300°C and 10–30 bar. [24] Literature suggests that with the catalyst driving the competing reactions (Equation 5-1) towards more EO production, CO_2 is produced during the oxidation reaction in a ratio of 6:2 EO:CO_2 on a molar basis. As a result of the competing steam and CO_2 producing, CO_2 concentration of the emissions stream can range 30–100 percent CO_2 [25] with the balance of the emissions stream being water, but most references give a range of 95–100 percent CO_2 concentration, indicating that a purification step (i.e., water removal from the emissions stream) is inherent to the EO production plant. [26]

$$C_2H_4 + {}^1/_2\,O_2 \xrightarrow{Silver\ Catalyst} C_2H_4O$$
$$C_2H_4 + 5O_2 \xrightarrow{yields} 2CO_2 + 2H_2O$$

Equation 5-1

5.2.1 Size Range

Current EO U.S. plant sizes range 105,000–770,000 tonnes. [27] Exhibit 5-10 shows the ten U.S. EO production facilities and their associated capacity as of 2007.

Exhibit 5-10. 2007 U.S. EO production facility capacities

Company	Location	Capacity (1,000 tonnes EO/year)
BASF	Geismar, Louisiana	220
Dow Chemical	Plaquemine, Louisiana	275
Dow Chemical	Seadrift, Texas	430
Dow Chemical	Taft, Louisiana	770
Eastman Chemical	Longview, Texas	105
Formosa Plastics	Point Comfort, Texas	250
Huntsman	Port Neches, Texas	460
LyondellBasell	Bayport, Texas	360
Old World Industries	Clear Lake, Texas	355
Shell Chemicals	Geismar, Louisiana	420

The U.S. contains 10 major producers totaling an EO production of 3.6 M tonnes. The average 2007 U.S. plant capacity is 364,500 tonnes EO, which is representative of the majority of EO plants and, thus, is the production capacity basis for the EO case in this study. With a 6:2 ratio of EO:CO_2, a plant with a 3.6 M tonnes annual EO production capacity would produce 121,500 tonnes CO_2/year at 100 percent CF. The International Energy Agency Greenhouse Gas R&D Programme (IEAGHG) database gives an average annual emission for the 52 worldwide EO production sites of 150,000 tonnes CO_2 per plant [24], which is within range of the assumed emissions rate for the representative EO plant evaluated.

5.2.2 CO_2 Point Sources

EO is considered a high purity source of CO_2. The process has a single CO_2 source: the CO_2 removal system that is assumed an inherent part of the EO production process. The removal system may be one of several types—physical sorbents such as Rectisol or Selexol, chemical sorbents such as aqueous amines, or cryogenic separation systems. This study assumes that the base plant employs a physical sorbent Rectisol unit, with the CO_2 stream to be captured available at a pressure of 43.5 psia and a temperature of 96°F. For this study, the concentration of the CO_2 emissions stream is assumed to be 100 percent CO_2.

5.2.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the EO process for the purpose of this study:

- The representative plant has a production capacity of 364,500 tonnes of EO/year
- The CO_2 generated at 100 percent CF is 121,500 tonnes CO_2/year.
- The CO_2 stream is 100 percent CO_2
- Due to 100 percent purity, only compression and cooling are required
- The CO_2 stream temperature is 96°F
- The CO_2 stream pressure is 43.5 psia
- The end product CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

5.2.4 CO_2 Capture System

For the EO case considered in this report, CO_2 separation is an inherent part of base plant operations, and only the addition of compression and associated intercooling are required. Given the low CO_2 flowrate, reciprocating compression is employed and scaled for this case. Based on mass flow rate, this represents a scale down of 15 percent versus the quoted flow rate as given previously in Section 4.1.2.

The suction pressure to the first stage of the reciprocating compressor is quoted as 17.43 psia, which is below the assumed stream pressure for this case of 43.5 psia. However, the assumed CO_2 stream pressure nearly matches the quoted 44.04 psia suction pressure to the second stage of the compressor. Therefore, when implementing this quote, the first stage is bypassed, and the CO_2 stream is introduced into the second stage. This reduces the overall power consumption of the compression train. The cost was adjusted to account for the removal of the first stage by scaling on power requirement, resulting in a 21.4 percent reduction in cost, as compared to the quoted value.

5.2.5 BFD, Stream Table, and Performance Summary

Since the EO absorption/separation process releases 100 percent pure CO_2, only cooling and compression is required for the CO_2 stream to be sent directly for EOR or other usage. As shown in Exhibit 5-11, the vent, which is at a lower temperature than required by the compressor, is sent directly to the compression train. Since the compression train includes a post-cooler, after-cooling is not represented here. Exhibit 5-12 provides the stream table.

Exhibit 5-11. EO CO_2 capture BFD

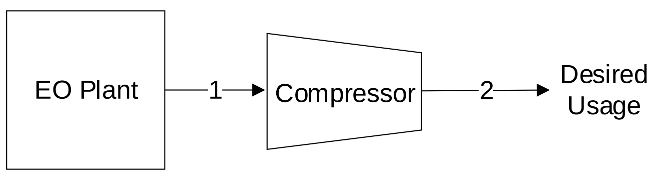


Exhibit 5-12. EO stream table

	1	2
V-L Mole Fraction		
AR	0.0000	0.0000
CH_4	0.0000	0.0000
CO	0.0000	0.0000
CO_2	1.0000	1.0000
SO_2	0.0000	0.0000
H_2	0.0000	0.0000
H_2O	0.0000	0.0000
H_2S	0.0000	0.0000
N_2	0.0000	0.0000
Total	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	315	315
V-L Flowrate (kg/hr)	13,870	13,870
Temperature (°C)	36	30
Pressure (MPa, abs)	0.30	15.3
Steam Table Enthalpy (kJ/kg)[A]	8,759	8,753
Aspen Plus Enthalpy (kJ/kg)[B]	-8,935	-9,193
Density (kg/m^3)	5.2	629
V-L Molecular Weight	44.0	44.0
V-L Flowrate (lb_{mol}/hr)	695	695

	1	2
V-L Mole Fraction		
V-L Flowrate (lb/hr)	30,578	30,578
Temperature (°F)	96	86
Pressure (psia)	43.5	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	3,765	3,763
Aspen Plus Enthalpy (Btu/lb)[B]	-3,841	-3,952
Density (lb/ft^3)	0.325	39.3

[A]Steam table reference conditions are 32.02°F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance summary is provided in Exhibit 5-13.

Exhibit 5-13. Performance summary

Performance Summary	
Item	364,500 tonnes/year (kW$_e$)
CO_2 Compressor	1,180
Circulating Water Pumps	10
Cooling Tower Fans	10
Total Auxiliary Load	**1,200**

5.2.6 Capture Integration

The reactor effluent is received by the AGR absorber at a temperature of 410°F [28] and requires cooling, indicating an existing cooling water system. A cooling water system from the retrofit could potentially be integrated into the existing plant's cooling water system; however, depending on the size of the existing cooling water system and the design cooling temperature range, it might be more economical to install a stand-alone cooling system rather than increase the existing cooling system. This would have to be evaluated on a case-by-case basis. If a power plant using a steam cycle is present within the EO facility, an efficient HX could capture this energy to heat condensate make-up.

For the purposes of this study, it is assumed that an additional, stand-alone cooling water unit will perform the necessary cooling for compression intercooling. However, there is a potential for integration of make-up water to be used to feed or partially feed the cooler thereby reducing the unit's size or replacing it with a simple heat exchanger depending on the size of the plant. These options are not evaluated within the scope of this study.

5.2.7 Power Source

Given the relatively small amount of CO_2, the compressor power consumption is 1.18 MW. Power consumption estimates for the cooling water system were scaled as described in Section 4.4. The total power requirement was approximated to be 1.2 MW, which includes all power required by the compression train and the cooling water system. Purchased power cost is estimated at a rate of $60/MWh as discussed in Section 3.1.2.4. Given that the EO reaction is exothermic, and this additional heat is possibly used to generate steam, an EO plant may already generate power on-site for other usage, and this power may be available as an alternative to purchasing power from the grid. The availability of on-site power would need to be evaluated on a case-by-case basis and is not considered within the scope of this report.

5.2.8 Economic Analysis Results

The economic results for CO_2 capture application in an EO plant are presented in this section. Owner's costs (Exhibit 5-14), capital costs (Exhibit 5-15), and O&M costs are calculated as discussed in Section 3.1. Retrofit costs were determined by applying a retrofit factor to TPC as discussed in Section 3.3. The greenfield TOC for the EO case is $20.4 M. The corresponding greenfield COC is $26.0/tonne CO_2, and the COC is $26.2/tonne CO_2 in retrofit applications.

Exhibit 5-14. Owner's costs for EO greenfield site

Description	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs		
6 Months All Labor	$341	$3
1-Month Maintenance Materials	$16	$0
1-Month Non-Fuel Consumables	$1	$0
1-Month Waste Disposal	$0	$0
25% of 1-Month Fuel Cost at 100% CF	$0	$0
2% of TPC	$333	$3
Total	**$690**	**$6**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$1	$0
0.5% of TPC (spare parts)	$83	$1
Total	**$84**	**$1**
Other Costs		
Initial Cost for Catalyst and Chemicals	$0	$0
Land	$30	$0
Other Owner's Costs	$2,495	$21
Financing Costs	$449	$4
TOC	**$20,385**	**$168**
TASC Multiplier (EO, 31 year)	1.025	
TASC	**$20,892**	**$172**

Exhibit 5-15. Capital costs for EO greenfield site

	Case:	Ethylene Oxide						Estimate Type:		Conceptual	
	Representative Plant Size:	364,500 tonnes EO/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
	5	Flue Gas Cleanup									
5.4	CO_2 Compression & Drying	$2,352	$353	$786	$0	$3,491	$611	$0	$820	$4,922	$41
5.5	CO_2 Compressor Aftercooler	w/5.4	w/5.4	w/5.4	w/5.4	$0	$0	$0	$0	$0	$0
	Subtotal	**$2,352**	**$353**	**$786**	**$0**	**$3,491**	**$611**	**$0**	**$820**	**$4,922**	**$41**
	7	Ductwork & Stack									
7.3	Ductwork	$0	$41	$29	$0	$70	$12	$0	$16	$99	$1
	Subtotal	**$0**	**$41**	**$29**	**$0**	**$70**	**$12**	**$0**	**$16**	**$99**	**$1**
	9	Cooling Water System									
9.1	Cooling Towers	$52	$0	$16	$0	$68	$12	$0	$16	$95	$1
9.2	Circulating Water Pumps	$4	$0	$0	$0	$4	$1	$0	$1	$5	$0
9.3	Circulating Water System Aux.	$125	$0	$17	$0	$142	$25	$0	$33	$200	$2
9.4	Circulating Water Piping	$0	$58	$53	$0	$111	$19	$0	$26	$156	$1
9.5	Make-up Water System	$25	$0	$32	$0	$57	$10	$0	$13	$81	$1
9.6	Component Cooling Water System	$9	$0	$7	$0	$16	$3	$0	$4	$23	$0
9.7	Circulating Water System Foundations	$0	$8	$13	$0	$21	$4	$0	$5	$30	$0
	Subtotal	**$215**	**$66**	**$138**	**$0**	**$418**	**$73**	**$0**	**$98**	**$590**	**$5**
	11	Accessory Electric Plant									
11.2	Station Service Equipment	$873	$0	$75	$0	$947	$166	$0	$223	$1,336	$11
11.3	Switchgear & Motor Control	$1,355	$0	$235	$0	$1,590	$278	$0	$374	$2,241	$18
11.4	Conduit & Cable Tray	$0	$176	$507	$0	$684	$120	$0	$161	$964	$8
11.5	Wire & Cable	$0	$466	$834	$0	$1,300	$227	$0	$305	$1,833	$15
	Subtotal	**$2,227**	**$642**	**$1,651**	**$0**	**$4,521**	**$791**	**$0**	**$1,062**	**$6,374**	**$52**
	12	Instrumentation & Control									
12.8	Instrument Wiring & Tubing	$287	$230	$919	$0	$1,437	$251	$0	$338	$2,026	$17
12.9	Other I&C Equipment	$353	$0	$818	$0	$1,171	$205	$0	$275	$1,651	$14
	Subtotal	**$640**	**$230**	**$1,737**	**$0**	**$2,607**	**$456**	**$0**	**$613**	**$3,677**	**$30**
	13	Improvements to Site									
13.1	Site Preparation	$0	$16	$320	$0	$336	$59	$0	$79	$474	$4
13.2	Site Improvements	$0	$75	$99	$0	$174	$30	$0	$41	$245	$2
13.3	Site Facilities	$85	$0	$90	$0	$175	$31	$0	$41	$247	$2
	Subtotal	**$85**	**$90**	**$509**	**$0**	**$685**	**$120**	**$0**	**$161**	**$965**	**$8**
	14	Buildings & Structures									
14.5	Circulation Water Pumphouse	$0	$4	$3	$0	$7	$1	$0	$2	$10	$0
	Subtotal	**$0**	**$4**	**$3**	**$0**	**$7**	**$1**	**$0**	**$2**	**$10**	**$0**
	Total	**$5,520**	**$1,427**	**$4,852**	**$0**	**$11,799**	**$2,065**	**$0**	**$2,773**	**$16,636**	**$137**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 5-16, while Exhibit 5-17 shows the COC for greenfield and retrofit sites for the representative EO plant.

Exhibit 5-16. Initial and annual O&M costs for EO greenfield site

Case:	Ethylene Oxide				Cost Base:	Dec 2018
Representative Plant Size:	364,500 tonnes EO/year				Capacity Factor (%):	85
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		1.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**1.0**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$438,438	$3.61
Maintenance Labor:					$106,470	$0.88
Administrative & Support Labor:					$136,227	$1.12
Property Taxes and Insurance:					$332,718	$2.74
Total:					**$1,013,852**	**$8.34**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$159,705	$1.55
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	10	$1.90	$0	$6,099	$0.06
Makeup and Waste Water Treatment Chemicals (ton):	0	0.03	$550.00	$0	$5,260	$0.05
Subtotal:				**$0**	**$11,359**	**$0.11**
Variable Operating Costs Total:				**$0**	**$171,063**	**$1.66**

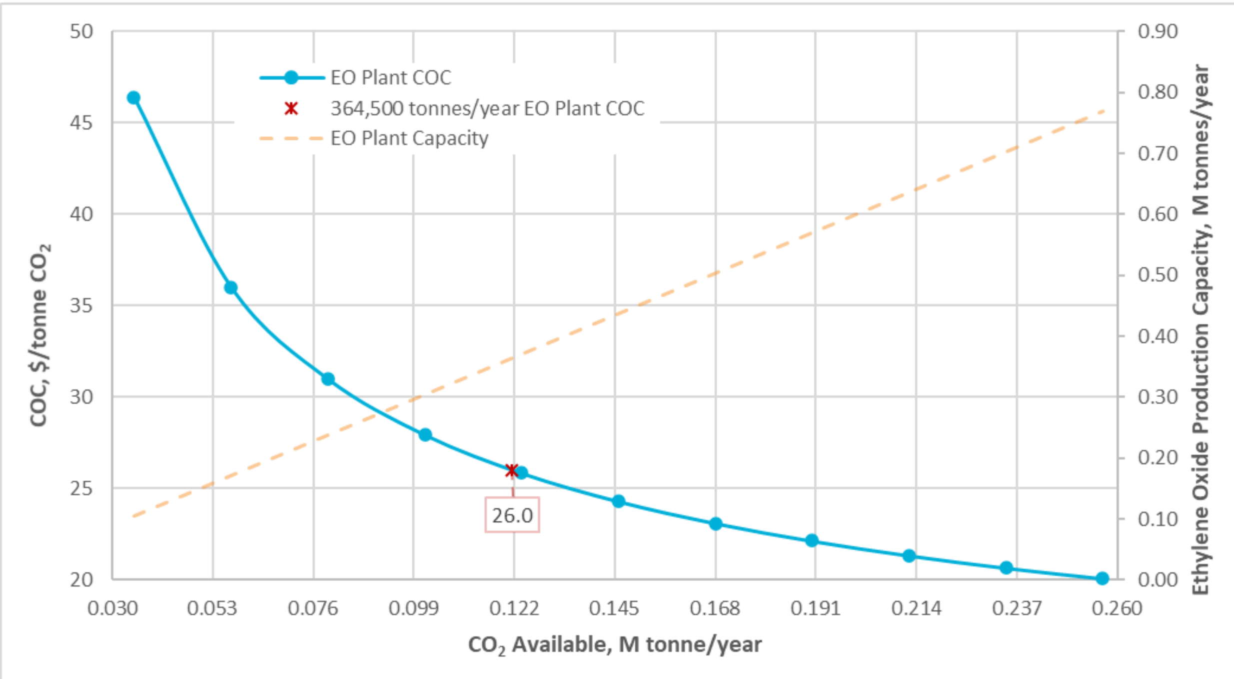
Exhibit 5-17. COC for 364,500 tonnes/year EO greenfield and retrofit

Component	Greenfield Value, $/tonne CO_2	Retrofit Value, $/tonne CO_2
Capital	9.4	9.4
Fixed	9.8	9.9
Variable	1.7	1.7
Purchased Power	5.2	5.2
Total COC	**26.0**	**26.2**

5.2.9 Plant Capacity Sensitivity Analysis

An analysis of the sensitivity of greenfield COC to EO plant capacity is shown in Exhibit 5-18. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 5-18. EO plant capacity sensitivity



5.2.10 Ethylene Oxide Conclusion

The high purity CO_2 stream produced from EO plants makes them a relatively low-cost industrial process for CO_2 capture, as the plant itself performs the separation of CO_2 under normal operating conditions. A CO_2 compression system for a 364,500 tonnes/year EO plant was modeled to estimate the COC of capturing CO_2 from the AGR system. The results showed the COC of CO_2 to be $26.0/tonne CO_2 for a greenfield site and $26.2/tonne CO_2 for a retrofit site. The small disparities between greenfield and retrofit cases are the result of unknown difficulties required for a retrofit installation versus a greenfield application, assuming adequate plot plan space for the retrofit case exists.

The plant size sensitivity showed that as plant size decreased from 770,000 tonnes EO/year to 105,000 tonnes EO/year, the COC increased by $26.3/tonne CO_2. As the plant size is decreased, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC.

5.3 ETHANOL

Ethanol production generates as a byproduct a high purity CO_2 stream greater than 85 percent by volume. [29] Though not a large-scale CO_2 producer, the COC is assumed to be relatively low. One project where CO_2 is being captured from ethanol refining is the DOE-funded Archer Daniel's Midlands project in Decatur, IL. The purpose of the project is to demonstrate how the next generation of technologies capture and store or reuse industrial CO_2 emissions. [30] The project design states a goal to capture approximately 1 M tons of CO_2/year using dehydration and compression and store the captured CO_2 in the Mt. Simon Sandstone Formation saline reservoir. [30]

5.3.1 Size Range

There are 208 ethanol refineries in the United States demonstrating a wide range of production, with 90 percent of these refineries using the dry-mill process. [31]. Of the 208 ethanol refineries in the United States, 66 of the plants (approximately 32 percent) fall between 40 and 60 M gallons/year. [32] Exhibit 5-19 shows the quantities of ethanol production ranges and the number of plants in each designated range. It is important to note that plant capacities would affect the COC presented, and a sensitivity analysis evaluating the effect of plant size on COC is included in Section 5.3.9. However, the effects would be noted at the equipment selection level. For instance, CO_2 produced from a 50 M gallons/year plant versus a 215+ M gallons/year plant requires a different type of compression (reciprocating versus centrifugal). This is due to the quantity of CO_2 produced at each plant. Discussion of the different types of compression can be found in Section 4.1.

Exhibit 5-19 U.S. Ethanol plant capacities and quantities

Capacity Range (M gallons/year)	Number of Plants
0–50	59
40–60	**66**
51–100	81
101–150	57
>150	11

Since a large portion of existing ethanol plants, 66 have smaller production capacities of 40–60 M gallons/year, the plant size chosen was 50 M gallons/year, and utilized reciprocating compression. It was also assumed that the plant uses the dry mill process with corn as the feedstock of choice.

5.3.2 CO_2 Point Sources

The major point sources of CO_2 emissions at an ethanol plant result from the fermentation process and fuel burning to provide required process heat. Of these two sources, only the fermentation off-gas stream is considered high purity and is the basis for the ethanol case in this report. The fuel burning stream may be considered as future work, as detailed in Section 9.1.

A study by the Illinois State Geological Survey [33] investigated the inventory of stationary CO_2 emissions in the Illinois Basin in 2007. The study reviewed a wide range of industrial processes, including ethanol plants. They used the relationship given in Equation 5-2 to calculate the amount of CO_2 emissions from the fermentation point source.

$$CO_{2\,Fermentation}\left(\frac{tonne}{year}\right) = \frac{\left[ethanol\ production\ \left(\frac{gal}{year}\right) * EF\ \left(\frac{lbCO_2}{gal}\right)\right]}{2{,}000\,\frac{lb}{ton} * 1.01231\ \frac{ton}{tonne}}$$

Equation 5-2

Where

EF = emission factor, feedstock dependent

The generic plant assumed in the Illinois Stage Geological Survey study utilizes corn as the feedstock, giving an EF equal to 6.31 lb CO_2/gallons ethanol. The EF was formulated in the Illinois Stage Geological Survey study through communication with representatives from existing ethanol plants in the Illinois Basin. [33] Using this relationship, the representative ethanol plant will generate approximately 143,042 tonnes CO_2/year from fermentation (at 100 percent CF), with a production capacity of 50 M gallons of ethanol/year.

A report published by the Global Carbon Capture and Sequestration (CCS) Institute in 2010 states that “the emission in ethanol plants arise from fermentation of biomass such as sugar cane or corn. Fermentation results in a pure stream of CO_2, which significantly reduces the cost for applying CCS.” [34] The fermentation process occurs at a temperature of 140–180 °C (284–356 °F). [35] Therefore, the fermentation stream is assumed to be 100 percent CO_2 and may be sent directly for cooling and compression. Other sources [30] have referenced the presence of water in the fermentation CO_2 stream. This is a possibility; however, water knockout drums would be present in the CO_2 compression train and, thus, further purification before processing would be unnecessary.

5.3.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the ethanol process for the purpose of this study:

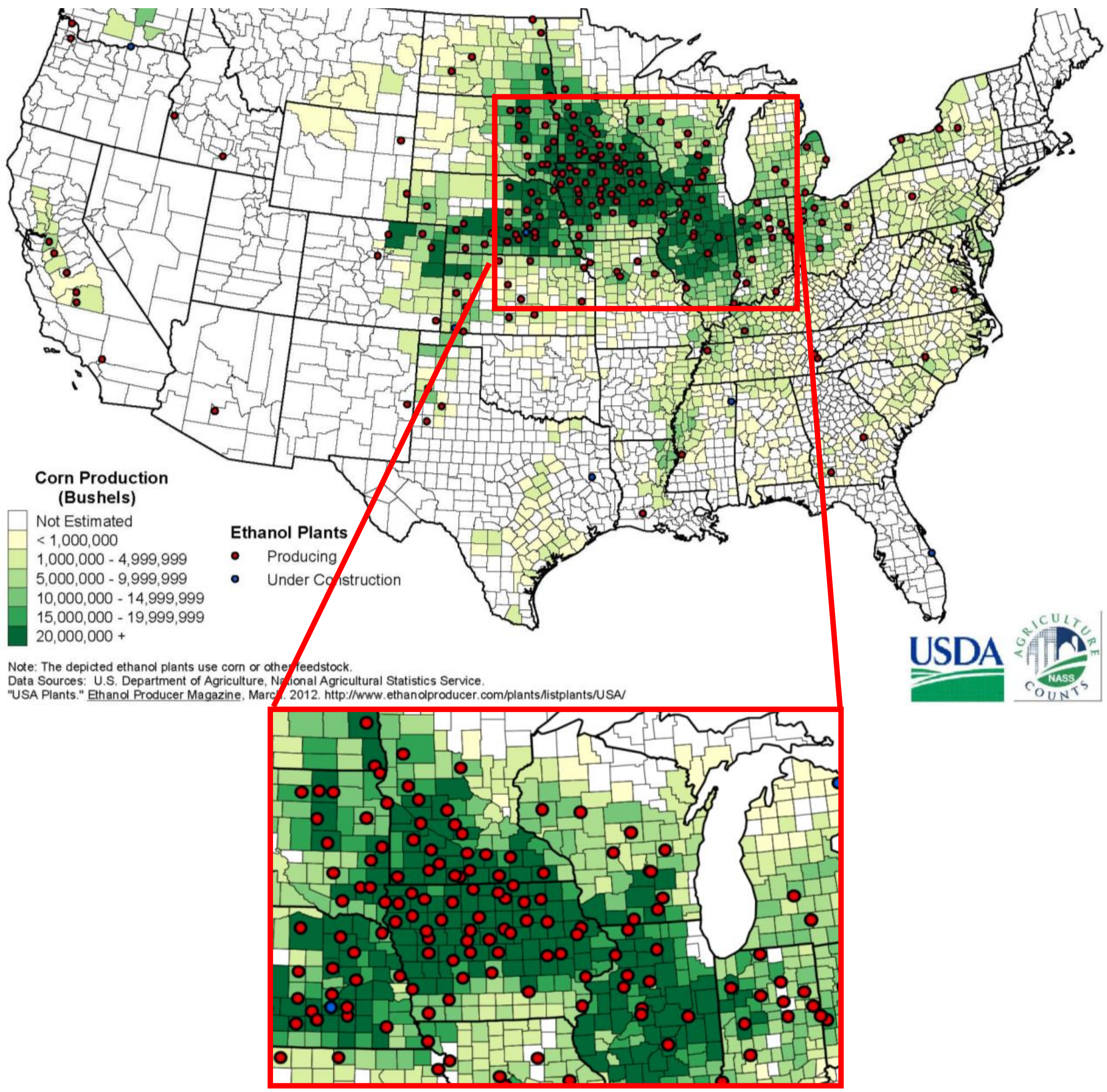
- The base plant is representative of an ethanol plant producing 50 M gallons of ethanol/year
- The plant uses the dry-mill process with corn as the feedstock
- The fermentation off-gas, assumed to be 100 percent CO_2, is the only high purity point source considered
- The CO_2 generated at 100 percent CF is 143,042 tonnes CO_2/year
- The CO_2 temperature is 320°F
- The CO_2 pressure is 17.4 psia

- The end product CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

5.3.4 CO_2 Capture System

Exhibit 5-20 [36] is a map provided by the U.S. Department of Agriculture (USDA) showing the production of corn by county in comparison to the location of U.S. ethanol plants, as of March 2012. As expected, the ethanol plants are mostly located near areas of high corn production, namely the Midwest states. The highest density of ethanol plants occurs in Illinois, Iowa, Minnesota, and Nebraska.

Exhibit 5-20. U.S. ethanol plant locations



Source: USDA [36]

The trend for the ethanol industry is smaller plants, which in turn produce smaller CO_2 streams and require compression equipment capable of handling smaller flows. This requirement is satisfied by using reciprocating compression discussed in Section 4.1.1; however, an alternative to smaller equipment could be to combine the emissions streams from multiple nearby plants for a single, larger compressor to compress the aggregate CO_2 for EOR use. Such a scenario is not considered in the scope of this study but could be evaluated in future work as described in Section 9.2.

5.3.5 BFD, Stream Table, and Performance Summary

Since the fermentation process releases 100 percent pure CO_2, only cooling and compression is required for the CO_2 stream to be sent directly for EOR or other usage. As shown in Exhibit 5-21, the fermentation vent is cooled through a HX, compressed (with interstage cooling and after-cooling) to meet EOR pipeline specifications for pressure and temperature. Exhibit 5-22 provides the stream table.

Exhibit 5-21. Ethanol CO_2 capture BFD

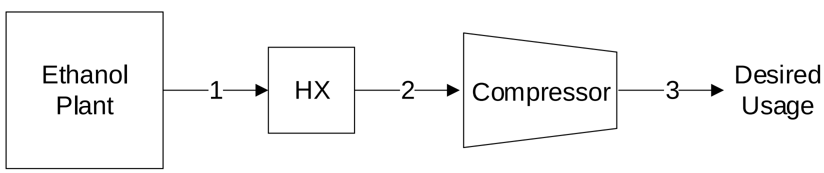


Exhibit 5-22. Ethanol stream table

	1	2	3
V-L Mole Fraction			
AR	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000
CO_2	1.0000	1.0000	1.0000
SO_2	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000
H_2O	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	371	371	371
V-L Flowrate (kg/hr)	16,329	16,329	16,329
Temperature (°C)	160	27	30
Pressure (MPa, abs)	0.12	0.1	15.3
Steam Table Enthalpy (kJ/kg)[A]	8,762	8,759	8,753
Aspen Plus Enthalpy (kJ/kg)[B]	-8,819	-8,941	-9,193

	1	2	3
V-L Mole Fraction			
Density (kg/m^3)	1.5	2.0	629
V-L Molecular Weight	44.0	44.0	44.0
V-L Flowrate (lb$_{mol}$/hr)	818	818	818
V-L Flowrate (lb/hr)	36,000	36,000	36,000
Temperature (°F)	320	80	86
Pressure (psia)	17.4	16.4	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	3,767	3,766	3,763
Aspen Plus Enthalpy (Btu/lb)[B]	-3,791	-3,844	-3,953
Density (lb/ft^3)	0.092	0.125	39.3

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance results are based on compressor quotes discussed in Section 4.1.1 and scaled auxiliary loads for the cooling water system as discussed in Section 4.4. The performance summary is provided in Exhibit 5-23.

Exhibit 5-23. Performance summary

Performance Summary	
Item	**50 M Gal Ethanol/year (kWe)**
CO_2 Compressor	1,810
Circulating Water Pumps	20
Cooling Tower Fans	10
Total Auxiliary Load	**1,840**

5.3.6 Capture Integration

The fermentation process occurs at a temperature of 140–180°C (284–356°F). Any cooling water system from the retrofit could be integrated into the existing plant's cooling water system; however, depending on the size of the existing cooling water system and the design cooling temperature range, it might be more economical to install a stand-alone cooling system rather than increase the existing cooling system. For the purposes of this study, it is assumed that an additional, stand-alone cooling water unit will perform the necessary cooling for capture and compression since integration with the base plant is outside the scope of this report. However, there is a potential for integration of make-up water to be used to feed or partially feed the cooling unit, thereby reducing the unit's size; there is also the potential that the heat removed from compression could be recycled within the plant to produce dried distiller grain solids. This product is produced by drying the solids that remain after fermentation. Heat for dried distiller grain solids drying is generally provided by NG.

5.3.7 Power Source

Given the relatively small amount of CO_2, the compression power consumption is 1.81 MW. Power consumption estimates for the cooling system were scaled as described in Section 4.4. The total power requirement was calculated to be 1.85 MW, which includes all power required by the compression train and the cooling water system. Purchased power cost is estimated at a rate of $60/MWh as discussed in Section 3.1.2.4.

5.3.8 Economic Analysis Results

The economic results for CO_2 capture application in an ethanol plant are presented in this section. Owner's costs (Exhibit 5-24), capital costs (Exhibit 5-25), and O&M costs are calculated as discussed in Section 3.1. Retrofit costs were determined by applying a retrofit factor to TPC as discussed in Section 3.3. The greenfield TOC for the ethanol case is $24.7 M. The corresponding greenfield COC is $31.8/tonne CO_2, and the COC is $32.0/tonne CO_2 in retrofit applications.

Exhibit 5-24. Owner's costs for ethanol greenfield site

Description	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs		
6 Months All Labor	$355	$2
1-Month Maintenance Materials	$19	$0
1-Month Non-Fuel Consumables	$2	$0
1-Month Waste Disposal	$0	$0
25% of 1-Month Fuel Cost at 100% CF	$0	$0
2% of TPC	$404	$3
Total	**$779**	**$5**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$2	$0
0.5% of TPC (spare parts)	$101	$1
Total	**$103**	**$1**
Other Costs		
Initial Cost for Catalyst and Chemicals	$0	$0
Land	$30	$0
Other Owner's Costs	$3,028	$21
Financing Costs	$545	$4
TOC	**$24,672**	**$172**
TASC Multiplier (Ethanol, 31 year)	1.047	
TASC	**$25,840**	**$181**

Exhibit 5-25. Capital costs for ethanol greenfield site

Case:		Ethanol						Estimate Type:		Conceptual	
Representative Plant Size:		50 M gallons ethanol/year						Cost Base:		Dec 2018	
Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/tonnes/yr ($CO_2$)
	5					Flue Gas Cleanup					
5.1	Inlet Cooler for Compression Train	$63	$0	$13	$0	$76	$13	$0	$18	$107	$1
5.4	CO_2 Compression & Drying	$3,053	$458	$1,021	$0	$4,532	$793	$0	$1,065	$6,390	$45
5.5	CO_2 Compressor Aftercooler	w/5.4	w/5.4	w/5.4	w/5.4	$0	$0	$0	$0	$0	$0
	Subtotal	**$3,116**	**$458**	**$1,034**	**$0**	**$4,608**	**$806**	**$0**	**$1,083**	**$6,497**	**$45**
	7					Ductwork & Stack					
7.3	Ductwork	$0	$112	$78	$0	$190	$33	$0	$45	$268	$2
	Subtotal	**$0**	**$112**	**$78**	**$0**	**$190**	**$33**	**$0**	**$45**	**$268**	**$2**
	9					Cooling Water System					
9.1	Cooling Towers	$75	$0	$23	$0	$99	$17	$0	$23	$139	$1
9.2	Circulating Water Pumps	$5	$0	$0	$0	$6	$1	$0	$1	$8	$0
9.3	Circulating Water System Aux.	$171	$0	$23	$0	$193	$34	$0	$45	$273	$2
9.4	Circulating Water Piping	$0	$79	$71	$0	$150	$26	$0	$35	$212	$1
9.5	Make-up Water System	$32	$0	$41	$0	$73	$13	$0	$17	$102	$1
9.6	Component Cooling Water System	$12	$0	$9	$0	$22	$4	$0	$5	$31	$0
9.7	Circulating Water System Foundations	$0	$11	$18	$0	$28	$5	$0	$7	$40	$0
	Subtotal	**$296**	**$90**	**$186**	**$0**	**$571**	**$100**	**$0**	**$134**	**$805**	**$6**
	11					Accessory Electric Plant					
11.2	Station Service Equipment	$1,049	$0	$90	$0	$1,139	$199	$0	$268	$1,606	$11
11.3	Switchgear & Motor Control	$1,629	$0	$283	$0	$1,912	$335	$0	$449	$2,695	$19
11.4	Conduit & Cable Tray	$0	$212	$610	$0	$822	$144	$0	$193	$1,159	$8
11.5	Wire & Cable	$0	$561	$1,002	$0	$1,563	$274	$0	$367	$2,204	$15
	Subtotal	**$2,678**	**$773**	**$1,985**	**$0**	**$5,436**	**$951**	**$0**	**$1,277**	**$7,665**	**$54**
	12					Instrumentation & Control					
12.8	Instrument Wiring & Tubing	$304	$243	$972	$0	$1,519	$266	$0	$357	$2,142	$15
12.9	Other I&C Equipment	$373	$0	$865	$0	$1,238	$217	$0	$291	$1,746	$12
	Subtotal	**$677**	**$243**	**$1,837**	**$0**	**$2,757**	**$482**	**$0**	**$648**	**$3,887**	**$27**
	13					Improvements to Site					
13.1	Site Preparation	$0	$17	$349	$0	$366	$64	$0	$86	$516	$4
13.2	Site Improvements	$0	$81	$108	$0	$189	$33	$0	$44	$267	$2
13.3	Site Facilities	$93	$0	$98	$0	$191	$33	$0	$45	$269	$2
	Subtotal	**$93**	**$99**	**$554**	**$0**	**$746**	**$131**	**$0**	**$175**	**$1,052**	**$7**
	14					Buildings & Structures					
14.5	Circulation Water Pumphouse	$0	$5	$4	$0	$9	$2	$0	$2	$13	$0
	Subtotal	**$0**	**$5**	**$4**	**$0**	**$9**	**$2**	**$0**	**$2**	**$13**	**$0**
	Total	**$6,860**	**$1,779**	**$5,678**	**$0**	**$14,317**	**$2,505**	**$0**	**$3,364**	**$20,187**	**$141**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 5-26, while Exhibit 5-27 shows the COC for greenfield and retrofit sites for the representative ethanol plant.

Exhibit 5-26. Initial and annual O&M costs for ethanol greenfield site

Case:	**Ethanol**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**50 M gallons ethanol/year**				**Capacity Factor (%):**	**85**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		1.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**1.0**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$438,438	$3.07
Maintenance Labor:					$129,194	$0.90
Administrative & Support Labor:					$141,908	$0.99
Property Taxes and Insurance:					$403,732	$2.82
Total:					**$1,113,272**	**$7.78**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$193,791	$1.59
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	17	$1.90	$0	$9,946	$0.08
Makeup and Waste Water Treatment Chemicals (ton):	0	0.1	$550.00	$0	$8,577	$0.07
Subtotal:				**$0**	**$18,523**	**$0.15**
Variable Operating Costs Total:				**$0**	**$212,314**	**$1.75**

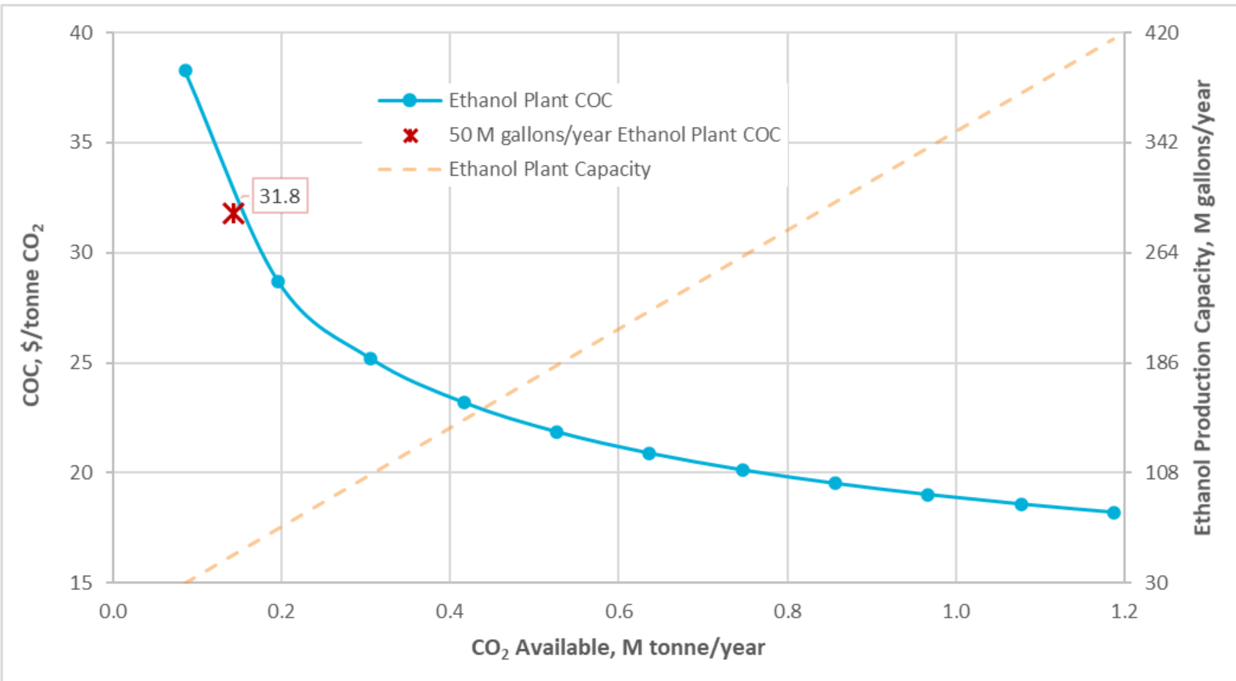
Exhibit 5-27. COC for 50 M gallons/year ethanol greenfield and retrofit

Component	Greenfield Value, $/tonne CO_2	Retrofit Value, $/tonne CO_2
Capital	14.1	14.2
Fixed	9.2	9.2
Variable	1.7	1.8
Purchased Power	6.8	6.8
Total COC	**31.8**	**32.0**

5.3.9 Plant Capacity Sensitivity Analysis

An analysis of the sensitivity of greenfield COC to ethanol plant capacity is shown in Exhibit 5-28. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 5-28. Ethanol plant capacity sensitivity



Note: The data point for the COC at a 50 M gallon/year ethanol plant does not fall directly on the COC line due to data point increments and plot formatting.

5.3.10 Ethanol Conclusion

The high purity CO_2 stream produced in an ethanol plant makes them relatively low-cost industrial processes for CO_2 capture since they require no costly separation equipment. A CO_2 compression system for a 50 M gallons/year ethanol plant was modeled to estimate the COC of capturing CO_2 from the fermenter. The results showed the COC of CO_2 to be $31.8/tonne CO_2 for a greenfield site and $32.0/tonne CO_2 for a retrofit site. The small disparities between greenfield and retrofit cases are the result of unknown difficulties required for a retrofit installation versus a greenfield application, assuming adequate plot plan space for the retrofit case exists.

The plant size sensitivity showed that as plant size decreased from 415 M gallons/year to 30 M gal/year, the COC increased by $20.1/tonne CO_2. As the plant size is decreased, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC. Though outside of this study's scope, literature discusses food-grade CO_2 capture for potential use instead of EOR. This might be a more economical option, but further evaluation would be required to determine an applicable COC for this alternate CO_2 end-use.

5.4 Natural Gas Processing

Natural gas processing is considered a high purity industrial process with a CO_2 discharge stream composition of 96–99 percent. Since in many applications CO_2 removal is inherently necessary to the processing of natural gas, NGP presents a potentially low-cost source of industrial CO_2 capture.

5.4.1 Size Range

For the purposes of this study, it is assumed that the reference plant has a capacity of 330 MMSCFD at 100 percent capacity. The composition of the raw gas processed is represented by that of a formation in the Michigan Basin producing formation with 10.2 percent CO_2. [37] The full raw gas characteristics are given in Exhibit 5-29, and represent average concentrations of the gas produced in the referenced formation. Given this plant capacity and the raw natural gas CO_2 composition, this plant would generate approximately 649,255 tonnes CO_2/year at 100 percent CF.[h]

Exhibit 5-29. Michigan basin producing formation raw gas characteristics

Michigan Basin Raw Gas Characteristics	
Component	**Average Mole %**
CH_4	82.4
C_2H_6	2.48
C_3H_8	0.37
n-Butane	0.00
i-Butane	0.00
n-Pentane	0.00
i-Pentane	0.00
c-Pentane	0.00
Hexanes	0.00
H_2S	0.00
CO_2	10.2
N_2	2.23
He	0.00
Other	2.32

[h] The assumptions for this study's reference plant are not limited to the Michigan Basin. High CO_2 content coupled with large capacity processing plants may also be found in the Gulf Coast region, the Williston Basin, and the Midwest region, referred to as the Foreland Province, according to the Gas Technology Institute database. [37]

5.4.2 CO_2 Point Sources

Natural gas processing (or gas sweetening) takes raw NG containing 2–70 percent CO_2 by volume and removes CO_2 and other impurities to meet the required pipeline or liquefaction specifications. The single point source is the CO_2 stream from the AGR system, which is generally vented to the atmosphere. The variation in raw natural gas CO_2 content would affect the amount of CO_2 available for capture; however, the concentration of the CO_2 stream to be captured is high, 96–99 percent.

5.4.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the natural gas processing plant for the purpose of this study:

- The representative NGP plant has a capacity of 330 MMSCFD of raw gas processed
- The raw gas CO_2 content is 10.2 mole percent
- The CO_2 generated at 100 percent CF is 649,255 tonnes CO_2/year
- The CO_2 stream temperature is 69°F
- The CO_2 stream pressure is 23.52 psia
- The CO_2 stream is 99 percent CO_2 by volume, balanced with water
- The CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

5.4.4 CO_2 Capture System

Only compression, glycol dehydration, and associated cooling is required for this NGP case. Given the amount of CO_2 available for capture, a centrifugal compressor, discussed in Section 4.1.2, is used to attain 2,200 psig EOR pipeline pressure per QGESS specifications. [1]

5.4.5 BFD, Stream Table, and Performance Summary

Since the stripping column releases 99 volume percent CO_2, only compression with glycol dehydration and cooling is required. Water knockout is used in the compression train to avoid liquid entering the compressors. There is no cooling of the inlet stream required, as it is assumed that the overhead condenser of the stripping column in the base plant discharges at a temperature of 69°F. After compression, the CO_2 product stream is cooled to 120°F and sent directly for EOR or other usage. Exhibit 5-30 gives the BFD for this process. Exhibit 5-31 provides the stream table.

Exhibit 5-30. NGP CO_2 capture BFD

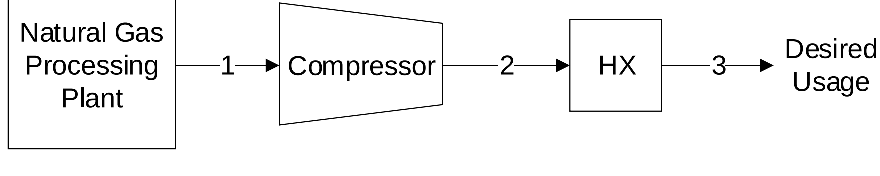


Exhibit 5-31. NGP stream table

	1	2	3
V-L Mole Fraction			
AR	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000
CO_2	0.9900	0.9995	0.9995
SO_2	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000
H_2O	0.0100	0.0005	0.0005
H_2S	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	1,701	1,684	1,684
V-L Flowrate (kg/hr)	74,416	74,109	74,109
Temperature (°C)	21	83	30
Pressure (MPa, abs)	0.16	15.3	15.3
Steam Table Enthalpy (kJ/kg)[A]	8,787	8,758	8,755
Aspen Plus Enthalpy (kJ/kg)[B]	-8,965	-9,034	-9,195
Density (kg/m^3)	2.9	416	630
V-L Molecular Weight	43.8	44.0	44.0
V-L Flowrate (lb_{mol}/hr)	3,750	3,713	3,713
V-L Flowrate (lb/hr)	164,059	163,382	163,382
Temperature (°F)	69	182	86
Pressure (psia)	23.5	2,216.9	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	3,778	3,765	3,764
Aspen Plus Enthalpy (Btu/lb)[B]	-3,854	-3,884	-3,953
Density (lb/ft^3)	0.183	25.9	39.3

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance results are based on the centrifugal compressor discussed in Section 4.1.2 and scaled auxiliary loads for the cooling water system as discussed in Section 4.4. The performance summary is provided in Exhibit 5-32.

Exhibit 5-32. Performance summary

Performance Summary	
Item	**330 MMSCFD (kWe)**
CO_2 Compressor	6,010
Circulating Water Pumps	70
Cooling Tower Fans	40
Total Auxiliary Load	**6,120**

5.4.6 Capture Integration

In this instance, the capture system is inherent to the base plant design, under the assumption that the raw gas CO_2 content is above that of pipeline specifications. Therefore, there is little opportunity for capture integration other than the necessary cooling for compression. Since the base plant is considered outside the scope of this study, a standalone cooling water system is assumed to provide the necessary intercooling for the compression process. However, in real applications, the necessity for a standalone cooling water system would need to be evaluated on a case-by-case basis. There could be potential to integrate make-up water to feed or partially feed the cooling system, thereby reducing the unit's size, or replacing it completely with a simple HX.

5.4.7 Power Source

The compressor power consumption for this case is 6.01 MW. Power consumption estimates for the cooling water system were scaled as described in Section 4.4. The total power requirement was calculated to be 6.12 MW, which includes all power required by the compression train and the cooling water system. Purchased power cost is estimated at a rate of $60/MWh as discussed in Section 3.1.2.4. For practical applications for this type of facility with NG readily available, the power required to operate the cooling water system as well as the compression system could be generated on site. Depending on the size and location of the facility there could be other co-beneficial reasons to produce the required power on-site. This scenario would need to be evaluated on a case-by-case basis and is outside of the scope of this study.

5.4.8 Economic Analysis Results

The economic results for CO_2 capture application in an NGP plant are presented in this section. Owner's costs (Exhibit 5-33), capital costs (Exhibit 5-34), and O&M costs are calculated as discussed in Section 3.1. Retrofit costs were determined by applying a retrofit factor to TPC as discussed in Section 3.3. The greenfield TOC for the NGP case is $56.8 M. The corresponding greenfield COC is $16.1/tonne CO_2, and the COC is $16.2/tonne CO_2 in retrofit applications.

Exhibit 5-33. Owner's costs for NGP greenfield site

Description	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs		
6 Months All Labor	$461	$1
1-Month Maintenance Materials	$44	$0
1-Month Non-Fuel Consumables	$37	$0
1-Month Waste Disposal	$2	$0
25% of 1 Months Fuel Cost at 100% CF	$0	$0
2% of TPC	$934	$1
Total	**$1,477**	**$2**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$68	$0
0.5% of TPC (spare parts)	$233	$0
Total	**$302**	**$0**
Other Costs		
Initial Cost for Catalyst and Chemicals	$0	$0
Land	$30	$0
Other Owner's Costs	$7,004	$11
Financing Costs	$1,261	$2
TOC	**$56,764**	**$87**
TASC Multiplier (NGP, 31 year)	1.039	
TASC	**$58,977**	**$91**

Exhibit 5-34. Capital and costs for NGP greenfield site

	Case:	Natural Gas Processing							Estimate Type:	Conceptual	
	Representative Plant Size:	330 MMSCFD natural gas							Cost Base:	Dec 2018	
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	$/tonnes/yr ($CO_2$)
5	Flue Gas Cleanup										
5.4	CO_2 Compression & Drying	$12,229	$1,834	$4,089	$0	$18,152	$3,177	$0	$4,266	$25,594	$39
5.5	CO_2 Compressor Aftercooler	$86	$14	$37	$0	$136	$24	$0	$32	$192	$0
	Subtotal	**$12,315**	**$1,848**	**$4,126**	**$0**	**$18,288**	**$3,200**	**$0**	**$4,298**	**$25,787**	**$40**
7	Ductwork & Stack										
7.3	Ductwork	$0	$200	$139	$0	$339	$59	$0	$80	$478	$1
	Subtotal	**$0**	**$200**	**$139**	**$0**	**$339**	**$59**	**$0**	**$80**	**$478**	**$1**
9	Cooling Water System										
9.1	Cooling Towers	$183	$0	$57	$0	$239	$42	$0	$56	$338	$1
9.2	Circulating Water Pumps	$15	$0	$1	$0	$16	$3	$0	$4	$22	$0
9.3	Circulating Water System Aux.	$353	$0	$47	$0	$400	$70	$0	$94	$564	$1
9.4	Circulating Water Piping	$0	$163	$148	$0	$311	$54	$0	$73	$439	$1
9.5	Make-up Water System	$56	$0	$72	$0	$128	$22	$0	$30	$180	$0
9.6	Component Cooling Water System	$25	$0	$20	$0	$45	$8	$0	$11	$63	$0
9.7	Circulating Water System Foundations	$0	$21	$34	$0	$55	$10	$0	$13	$77	$0
	Subtotal	**$632**	**$184**	**$378**	**$0**	**$1,194**	**$209**	**$0**	**$281**	**$1,683**	**$3**
11	Accessory Electric Plant										
11.2	Station Service Equipment	$1,757	$0	$151	$0	$1,908	$334	$0	$448	$2,690	$4
11.3	Switchgear & Motor Control	$2,728	$0	$473	$0	$3,201	$560	$0	$752	$4,514	$7
11.4	Conduit & Cable Tray	$0	$355	$1,022	$0	$1,377	$241	$0	$324	$1,941	$3
11.5	Wire & Cable	$0	$939	$1,679	$0	$2,618	$458	$0	$615	$3,691	$6
	Subtotal	**$4,485**	**$1,294**	**$3,325**	**$0**	**$9,104**	**$1,593**	**$0**	**$2,139**	**$12,837**	**$20**
12	Instrumentation & Control										
12.8	Instrument Wiring & Tubing	$355	$284	$1,136	$0	$1,775	$311	$0	$417	$2,503	$4
12.9	Other I&C Equipment	$436	$0	$1,011	$0	$1,447	$253	$0	$340	$2,040	$3
	Subtotal	**$791**	**$284**	**$2,147**	**$0**	**$3,222**	**$564**	**$0**	**$757**	**$4,543**	**$7**
13	Improvements to Site										
13.1	Site Preparation	$0	$22	$443	$0	$465	$81	$0	$109	$656	$1
13.2	Site Improvements	$0	$103	$137	$0	$240	$42	$0	$56	$339	$1
13.3	Site Facilities	$118	$0	$124	$0	$242	$42	$0	$57	$342	$1
	Subtotal	**$118**	**$125**	**$704**	**$0**	**$948**	**$166**	**$0**	**$223**	**$1,337**	**$2**
14	Buildings & Structures										
14.5	Circulation Water Pumphouse	$0	$10	$8	$0	$18	$3	$0	$4	$26	$0
	Subtotal	**$0**	**$10**	**$8**	**$0**	**$18**	**$3**	**$0**	**$4**	**$26**	**$0**
	Total	**$18,342**	**$3,945**	**$10,826**	**$0**	**$33,114**	**$5,795**	**$0**	**$7,782**	**$46,690**	**$72**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 5-35, while Exhibit 5-36 shows the COC for greenfield and retrofit sites for the representative NGP plant.

Exhibit 5-35. Initial and annual O&M costs for NGP greenfield site

Case:	Natural Gas Processing				Cost Base:	Dec 2018
Representative Plant Size:	**330 MMSCFD natural gas**				**Capacity Factor (%):**	**85**
			Operating & Maintenance Labor			
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		1.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**1.0**
			Fixed Operating Costs			
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$438,438	$0.68
Maintenance Labor:					$298,819	$0.46
Administrative & Support Labor:					$184,314	$0.28
Property Taxes and Insurance:					$933,808	$1.44
Total:					**$1,855,379**	**$2.86**
			Variable Operating Costs			
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$448,228	$0.81
			Consumables			
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	53	$1.90	$0	$31,518	$0.06
Makeup and Waste Water Treatment Chemicals (ton):	0	0.2	$550.00	$0	$27,728	$0.05
Triethylene Glycol (gal):	w/equip.	152	$6.80	$0	$321,580	$0.58
Subtotal:				**$0**	**$380,826**	**$0.69**
			Waste Disposal			
Triethylene Glycol (gal):		152	$0.35	$0	$16,552	$0.03
Subtotal:				**$0**	**$16,552**	**$0.03**
Variable Operating Costs Total:				**$0**	**$845,606**	**$1.53**

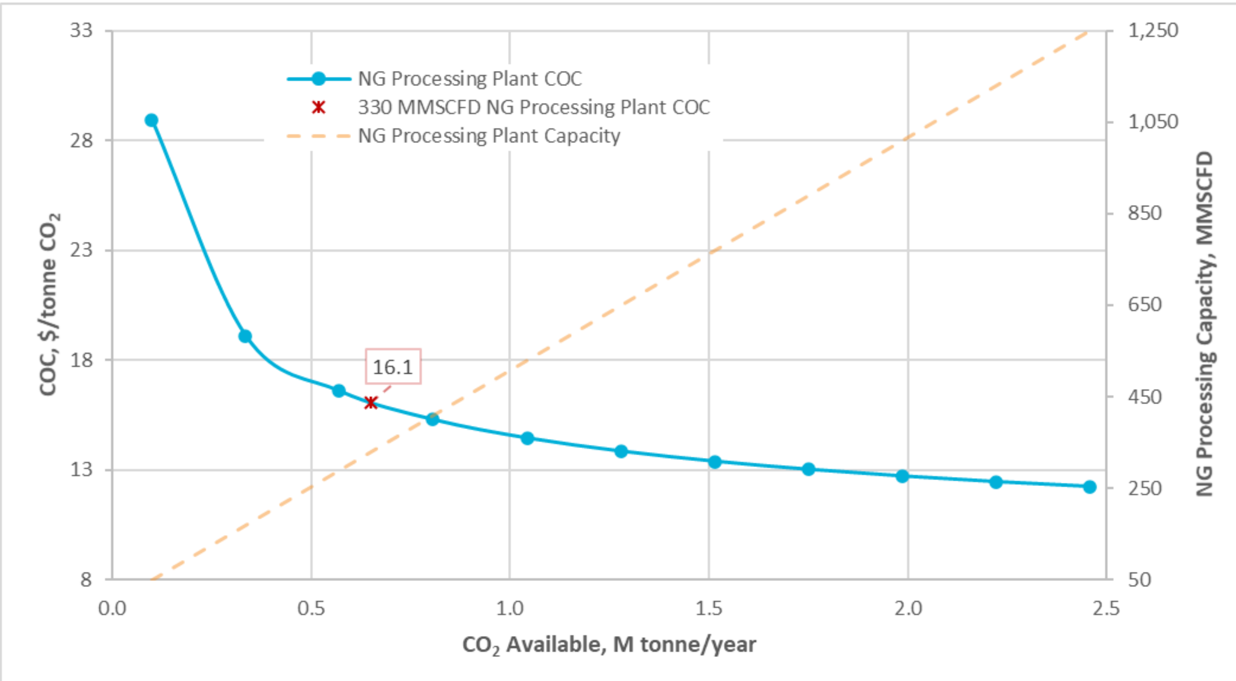
Exhibit 5-36. COC for 330 MMSCFD NGP greenfield and retrofit

Component	Greenfield Value, \$/tonne CO_2	Retrofit Value, \$/tonne CO_2
Capital	6.2	6.3
Fixed	3.4	3.4
Variable	1.5	1.5
Purchased Power	5.0	5.0
Total COC of CO_2	**16.1**	**16.2**

5.4.9 Plant Capacity Sensitivity Analysis

An analysis of the sensitivity of greenfield COC to NGP plant capacity is shown in Exhibit 5-37. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 5-37. NGP plant capacity sensitivity



5.4.10 Natural Gas Processing Conclusion

The high purity CO_2 stream produced from NGP plants makes them a relatively low-cost industrial process since CO_2 separation is inherent to normal operations. A CO_2 compression system for a 330 MMSCFD NGP plant was modeled to estimate the COC of capturing CO_2 from the plant's existing AGR. The results showed the COC of CO_2 to be $16.1/tonne CO_2 for a greenfield site and $16.2/tonne CO_2 for a retrofit site. The small disparities between greenfield and retrofit cases are the result of unknown difficulties required for a retrofit installation versus a greenfield application, assuming adequate plot plan space for the retrofit case exists.

The plant size sensitivity, based on the design basis assumptions in this study, showed that as plant size decreased from 1,250 MMSCFD to 50 MMSCFD, the COC increased by $16.7/tonne CO_2. With decreasing plant size, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC.

5.5 COAL-TO-LIQUIDS

Economic and national security concerns related to liquid fuels have revived national interest in alternative liquid fuel sources. Coal-to-Fischer-Tropsch fuels production emerged as a major technology option for many states and the DOE. The 2014 NETL report "Baseline Technical and Economic Assessment of a Commercial Scale Fischer-Tropsch Liquids Facility" ("CTL Study") [38] examined the technical and economic feasibility of a commercial 50,000 barrels per day (BPD) CTL facility. The facility employs gasification and Fischer-Tropsch (FT) technology to produce commercial-grade diesel and naptha liquids from medium-sulfur bituminous coal. The basis for the CTL case in this report is that of the CO_2 sequestration case evaluated in the CTL Study.

5.5.1 Size Range

The CTL Study focuses on a 50,000 BPD CTL production facility, and this is the plant capacity assumed for this study to allow for comparisons across NETL reports. With the given capacity, the CTL facility will produce 8,743,312 tonnes/year of CO_2 at 100 percent CF. The CTL study also considers power production, where the gas turbine and steam turbine produce power in excess of what base plant operations would require, and this excess 4.7 MW was exported to the grid. This reported excess power is on a net basis and does include auxiliary loads for CO_2 compressors. For the purposes of this study, all power requirements are met with power purchased from the grid; however, in some cases the base plant will have excess power available to meet compression and cooling power requirements, as is the case in the CTL study.

5.5.2 CO_2 Point Sources

Within the CTL facility there are two main point sources of CO_2 emissions; the AGR unit in the gasification section and the FT amine AGR unit in the FT section. The gasification section AGR generates CO_2 at two pressures: 160 psia and 300 psia. The FT amine AGR generates CO_2 at 265 psia. These three streams are compressed in one compression train, with the higher-pressure streams added to the train between the appropriate compression stages. The CO_2 product stream has a purity of 100 percent CO_2.

5.5.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the CTL process for the purpose of this study:

- The representative CTL facility has a production capacity of 50,000 BPD
- The CO_2 generated is 8,743,312 tonnes CO_2/year at 100 percent CF
- The CO_2 stream is 100 percent CO_2
- Due to 100 percent purity, only compression and cooling are required
- The CO_2 stream pressures are 160, 265, and 300 psia
- The CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

5.5.4 CO_2 Capture System

The CTL Study considers cases with CO_2 compression for EOR export and, therefore, the base plant acts as the separation process. The specific AGR units used in the CTL Study discharge CO_2 at multiple pressures and, therefore, the compression trains used are configured specifically to handle these compression requirements. Of the vendor quotes discussed in Section 4.1, there is not a compression train quote that accounts for multiple inlet CO_2 streams at multiple pressures. Therefore, the cost and performance specified in the NETL CTL Study are used here. This requires approximation of the amount of cooling water necessary for interstage cooling. [38]

It should be noted that in the CTL Study, after the CO_2 streams are combined, a portion is removed and sent back to the gasifier. For the purposes of this study, this stream is not considered, and all calculations are based on the reported mass flow of the product CO_2 stream (at 2,200 psig) given in the CTL Study. [38]

5.5.5 BFD, Stream Table, and Performance Summary

Since the CTL process releases 100 percent pure CO_2, only cooling and compression is required for the CO_2 stream to be sent directly for EOR or other usage. The compression train used discharges the product CO_2 at 2,200 psig and 121°F and, therefore, after-cooling is required. Exhibit 5-38 gives the BFD for this process, and Exhibit 5-39 provides the stream table.

Exhibit 5-38. CTL CO_2 capture BFD

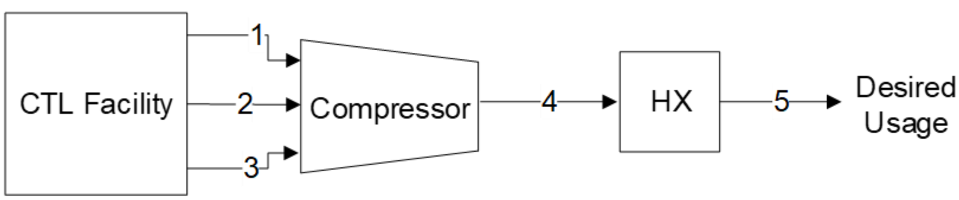


Exhibit 5-39. CTL stream table

	1	2	3	4	5
V-L Mole Fraction					
AR	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	1.0000	1.0000	1.0000	1.0000	1.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0000	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000

	1	2	3	4	5
V-L Mole Fraction					
V-L Flowrate (kg_{mol}/hr)	13,449	7,384	1,846	22,679	22,679
V-L Flowrate (kg/hr)	91,870	324,980	81,245	498,095	498,095
Temperature (°C)	38	16	16	49	30
Pressure (MPa, abs)	1.8	1.1	2.1	15.3	15.3
Steam Table Enthalpy (kJ/kg)[A]	8,759	8,759	8,758	8,755	8,753
Aspen Plus Enthalpy (kJ/kg)[B]	-8,948	-8,961	-8,972	-9,132	-9,188
Density (kg/m^3)	34.2	21.7	43.8	668.2	628.8
V-L Molecular Weight	44.01	44.01	44.01	44.01	44.01
V-L Flowrate (lbmol/hr)	29,649	16,280	4,070	49,998	49,998
V-L Flowrate (lb/hr)	1,304,851	716,458	179,114	2,200,423	2,200,423
Temperature (°F)	100	60	60	121	86
Pressure (psia)	265.0	160.0	300.0	2,214.7	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	3,766	3,766	3,765	3,764	3,763
Aspen Plus Enthalpy (Btu/lb)[B]	-3,847	-3,852	-3,857	-3,926	-3,950
Density (lb/ft^3)	2.14	1.36	2.74	41.7	39.3

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance results are taken from the CTL Study sequestration case that considered CO_2 capture. The performance summary is provided in Exhibit 5-40.

Exhibit 5-40. Performance summary

Performance Summary	
Item	**50,000 BPD (kWe)**
CO_2 Compressor	43,480
Circulating Water Pumps	100
Cooling Tower Fans	50
Total Auxiliary Load	**43,630**

5.5.6 Capture Integration

For the purposes of this study, it is assumed that an additional, stand-alone cooling water system will perform the necessary cooling for capture and compression. No retrofit case is considered for CTL as any new builds would most likely include cooling. However, to make this case comparable to the other cases considered in this study, the cost for cooling must be included in the greenfield COC. Therefore, a stand-alone cooling system is included.

5.5.7 Power Source

The auxiliary power required for this case is 43.6 MW. The total power requirement was approximated to include all power required by the compression train and the cooling water system. Power requirement estimates for the cooling water system were scaled as described in Section 4.4. Purchased power costs are estimated at a rate of $60/MWh as discussed in Section 3.1.2.4. However, for practical applications for this type of facility with power produced on-site and excess power sent to the grid, the power requirements may be met with power generated on-site. For instance, while the CTL Study sequestration case has excess power that would be able to satisfy a portion of this study's power requirement, this scenario should be evaluated on a case-by-case basis, which is not included in the scope of this report.

5.5.8 Economic Analysis Results

The economic results for CO_2 capture application in a CTL plant are presented in this section. Owner's costs (Exhibit 5-41), capital costs (Exhibit 5-42), and O&M costs are calculated as discussed in Section 3.1. The greenfield TOC for the CTL case is $196.9 M. The corresponding greenfield COC is $5.6/tonne CO_2.

Exhibit 5-41. Owner's costs for CTL greenfield site

Description	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs		
6 Months All Labor	$925	$0
1-Month Maintenance Materials	$153	$0
1-Month Non-Fuel Consumables	$42	$0
1-Month Waste Disposal	$0	$0
25% of 1-Month Fuel Cost at 100% CF	$0	$0
2% of TPC	$3,257	$0
Total	**$4,377**	**$1**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$39	$0
0.5% of TPC (spare parts)	$814	$0
Total	**$853**	**$0**
Other Costs		
Initial Cost for Catalyst and Chemicals	$0	$0
Land	$30	$0
Other Owner's Costs	$24,426	$3
Financing Costs	$4,397	$1
TOC	**$196,924**	**$23**
TASC Multiplier (CTL, 31 year)	1.054	
TASC	**$207,583**	**$24**

Exhibit 5-42. Capital costs for CTL greenfield site

	Case:	CTL						Estimate Type:		Conceptual	
	Representative Plant Size:	50,000 BPD Fischer-Tropsch liquids						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
5						Flue Gas Cleanup					
5.4	CO_2 Compression & Drying	$59,197	$0	$20,070	$0	$79,267	$13,872	$0	$18,628	$111,766	$13
5.5	CO_2 Compressor Aftercooler	$310	$49	$133	$0	$492	$86	$0	$116	$694	$0
	Subtotal	**$59,507**	**$49**	**$20,203**	**$0**	**$79,759**	**$13,958**	**$0**	**$18,743**	**$112,461**	**$13**
7						Ductwork & Stack					
7.3	Ductwork	$0	$246	$171	$0	$417	$73	$0	$98	$588	$0
	Subtotal	**$0**	**$246**	**$171**	**$0**	**$417**	**$73**	**$0**	**$98**	**$588**	**$0**
9						Cooling Water System					
9.1	Cooling Towers	$839	$0	$661	$0	$1,501	$263	$0	$353	$2,116	$0
9.2	Circulating Water Pumps	$233	$0	$17	$0	$250	$44	$0	$59	$352	$0
9.3	Circulating Water System Aux.	$2,663	$0	$351	$0	$3,014	$527	$0	$708	$4,250	$0
9.4	Circulating Water Piping	$0	$1,231	$1,115	$0	$2,346	$410	$0	$551	$3,307	$0
9.5	Make-up Water System	$307	$0	$394	$0	$701	$123	$0	$165	$988	$0
9.6	Component Cooling Water System	$192	$0	$147	$0	$339	$59	$0	$80	$478	$0
9.7	Circulating Water System Foundations	$0	$132	$220	$0	$352	$62	$0	$83	$497	$0
	Subtotal	**$4,233**	**$1,363**	**$2,905**	**$0**	**$8,502**	**$1,488**	**$0**	**$1,998**	**$11,988**	**$1**
11						Accessory Electric Plant					
11.2	Station Service Equipment	$4,090	$0	$351	$0	$4,441	$777	$0	$1,044	$6,261	$1
11.3	Switchgear & Motor Control	$6,349	$0	$1,102	$0	$7,450	$1,304	$0	$1,751	$10,505	$1
11.4	Conduit & Cable Tray	$0	$825	$2,378	$0	$3,204	$561	$0	$753	$4,517	$1
11.5	Wire & Cable	$0	$2,186	$3,907	$0	$6,093	$1,066	$0	$1,432	$8,591	$1
	Subtotal	**$10,439**	**$3,011**	**$7,738**	**$0**	**$21,188**	**$3,708**	**$0**	**$4,979**	**$29,874**	**$3**
12						Instrumentation & Control					
12.8	Instrument Wiring & Tubing	$458	$367	$1,467	$0	$2,292	$401	$0	$539	$3,231	$0
12.9	Other I&C Equipment	$563	$0	$1,305	$0	$1,868	$327	$0	$439	$2,634	$0
	Subtotal	**$1,022**	**$367**	**$2,771**	**$0**	**$4,160**	**$728**	**$0**	**$977**	**$5,865**	**$1**
13						Improvements to Site					
13.1	Site Preparation	$0	$33	$657	$0	$689	$121	$0	$162	$972	$0
13.2	Site Improvements	$0	$153	$203	$0	$356	$62	$0	$84	$502	$0
13.3	Site Facilities	$175	$0	$184	$0	$359	$63	$0	$84	$506	$0
	Subtotal	**$175**	**$186**	**$1,043**	**$0**	**$1,404**	**$246**	**$0**	**$330**	**$1,980**	**$0**
14						Buildings & Structures					
14.5	Circulation Water Pumphouse	$0	$33	$26	$0	$60	$10	$0	$14	$84	$0
	Subtotal	**$0**	**$33**	**$26**	**$0**	**$60**	**$10**	**$0**	**$14**	**$84**	**$0**
	Total	**$75,376**	**$5,255**	**$34,858**	**$0**	**$115,490**	**$20,211**	**$0**	**$27,140**	**$162,840**	**$19**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 5-43, while Exhibit 5-44 shows the COC for a greenfield site for the representative CTL plant.

Exhibit 5-43. Initial and annual O&M costs for CTL greenfield site

Case:	**Coal-to-Liquids**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**50,000 BPD Fischer-Tropsch liquids**				**Capacity Factor (%):**	**85**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		1.0
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**1.0**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$438,438	$0.05
Maintenance Labor:					$1,042,178	$0.12
Administrative & Support Labor:					$370,154	$0.04
Property Taxes and Insurance:					$3,256,808	$0.37
Total:					**$5,107,578**	**$0.58**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$1,563,268	$0.21
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	387	$1.90	$0	$228,019	$0.03
Makeup and Waste Water Treatment Chemicals (ton):	0	1.2	$550.00	$0	$200,179	$0.03
Subtotal:				**$0**	**$428,198**	**$0.06**
Variable Operating Costs Total:				**$0**	**$1,991,465**	**$0.27**

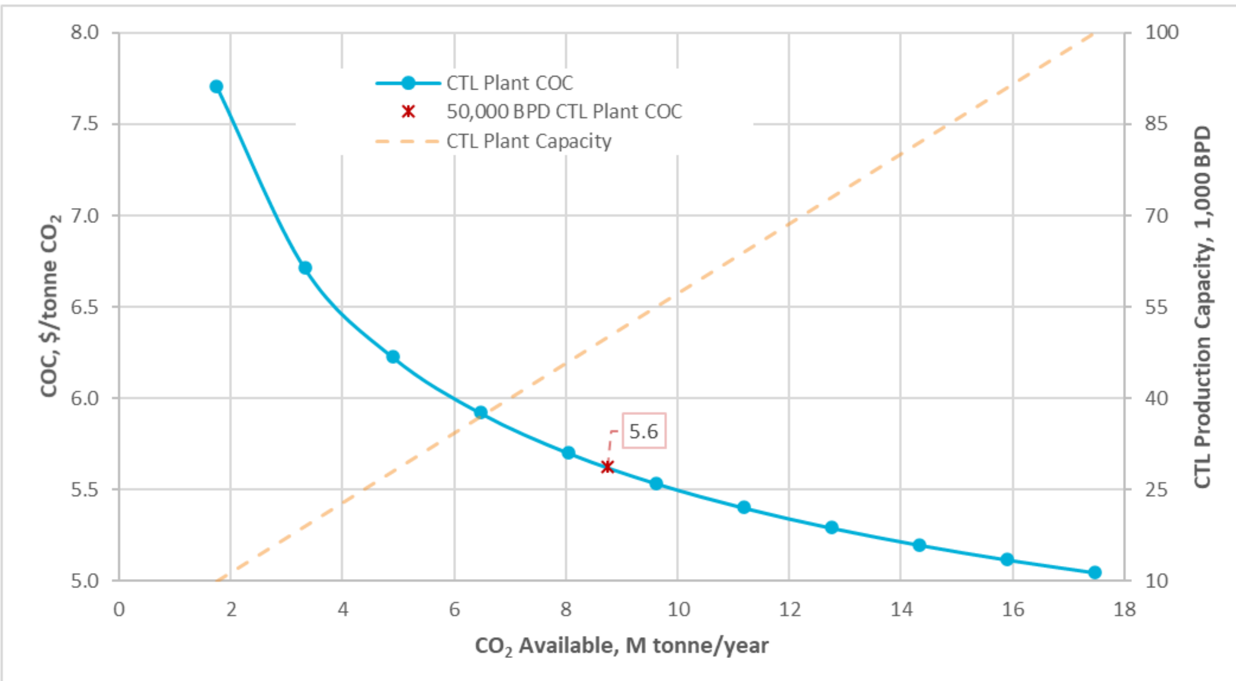
Exhibit 5-44. COC for 50,000 BPD CTL greenfield

Component	Greenfield Value, $/tonne CO_2
Capital	2.0
Fixed	0.7
Variable	0.3
Purchased Power	2.6
Total COC	**5.6**

5.5.9 Plant Capacity Sensitivity Analysis

An analysis of the sensitivity of greenfield COC to CTL plant capacity is shown in Exhibit 5-45. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 5-45. CTL plant capacity sensitivity



5.5.10 Coal-to-Liquids Conclusion

The high purity CO_2 streams produced from CTL plants makes them a relatively low-cost industrial process since the plant performs the CO_2 separation as a part of normal operations. A CO_2 compression system for a 50,000 BPD plant was modeled to estimate the COC of capturing CO_2 from the process. The results showed the COC of CO_2 to be $5.6/tonne CO_2 for a greenfield site. The plant size sensitivity showed that as plant size decreased from 100,000 to 10,000 BPD, the COC increased by $2.7/tonne CO_2. As the plant size is decreased, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC.

5.6 Gas-to-Liquids

Domestic FT GTL technology provides an alternative option for use of U.S. increasing supply of domestic NG. As with CTL, GTL can create a significant economic value while increasing the country's energy security. In their report "Analysis of Natural Gas-to Liquid Transportation Fuels via Fischer-Tropsch" [39] ("GTL Study") published in 2013, NETL evaluated the cost and performance of a 50,000 BPD FT liquids GTL facility. Of the total liquids production, 30 percent is allocated for finished motor gasoline, and 70 percent results in low-density diesel fuel. The system is calibrated to produce predominately liquid fuels; however, electrical power for export is also a co-product after satisfying internal plant power consumption. In its current configuration, the GTL plant exports 40.8 MWe to the grid. This study also considers CO_2 capture and compression with associated performance and cost. The case for this report is that of the GTL Study.

5.6.1 Size Range

The GTL Study plant size is a 50,000 BPD GTL production facility and, therefore, the plant size assumed for this study is 50,000 BPD to allow for comparisons across NETL reports. The 50,000 BPD GTL facility produces 1,858,628 tonnes/year of CO_2 at 100 percent CF. The NETL study also considered power production where the steam turbine produced power in excess of what base plant operations would require, and this excess power is exported to the grid. The GTL plant in the GTL Study has a net of 40.8 MWe available for export. While this study assumes that all power requirements are met with power purchased from the grid, in some cases, such as that of the GTL Study, the base plant will have excess power available to meet or partially meet compression and cooling power requirements.

5.6.2 CO_2 Point Sources

Within the GTL facility, there is one main point source of CO_2 emissions; the AGR unit in the FT section. The FT AGR generates CO_2 at 265 psia and 100°F, with a purity of 100 percent CO_2.

5.6.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the GTL process for the purpose of this study:

- The representative plant has a production capacity of 50,000 BPD
- The CO_2 generated is 1,858,628 tonnes CO_2/year at 100 percent CF
- The CO_2 stream is 100 percent CO_2
- Due to 100 percent purity, only compression and cooling are required
- The CO_2 stream pressure is 265 psia
- The CO_2 stream temperature is 100 °F
- The CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

5.6.4 CO_2 Capture System

NETL's GTL Study considers CO_2 removal and compression for EOR export and, therefore, the base plant separates CO_2 as part of its inherent process. The FT AGR unit used discharges CO_2 at 265 psia and, therefore, the compression train used is configured specifically to handle this higher inlet suction pressure. Of the vendors quotes discussed in Section 4.1, there is not a compression train quote that accounts for higher inlet CO_2 stream pressures. Therefore, the cost and performance specified in the current GTL Study is replicated here, with its cost being adjusted to December 2018 dollars. This will require that the amount of cooling water necessary for interstage cooling be approximated, similar to the CTL case in this study.

5.6.5 BFD, Stream Table, and Performance Summary

Since the GTL process releases 100 percent pure CO_2, only cooling and compression is required for the CO_2 stream to be sent directly for EOR or other usage. The compression train used discharges the product CO_2 at 2,200 psig and 117°F and, therefore, after-cooling is required. Exhibit 5-46 gives the BFD for this process. Exhibit 5-47 provides the stream table.

Exhibit 5-46. GTL CO_2 capture BFD

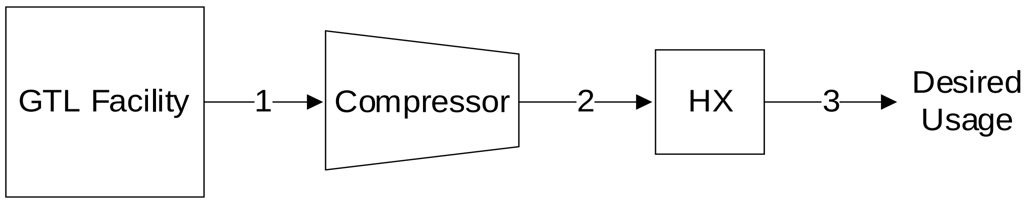


Exhibit 5-47. GTL stream table

	1	2	3
V-L Mole Fraction			
AR	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000
CO_2	1.0000	1.0000	1.0000
COS	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000
H_2O	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	4,821	4,821	4,821
V-L Flowrate (kg/hr)	212,188	212,188	212,188
Temperature (°C)	38	47	30
Pressure (MPa, abs)	1.827	15.270	15.270
Steam Table Enthalpy (kJ/kg)[A]	8,758	8,754	8,753
Aspen Plus Enthalpy (kJ/kg)[B]	-8,948	-9,139	-9,188
Density (kg/m^3)	34.2	688.6	628.8
V-L Molecular Weight	44.0	44.0	44.0
V-L Flowrate (lb_{mol}/hr)	10,629	10,629	10,629
V-L Flowrate (lb/hr)	467,794	467,794	467,794
Temperature (°F)	100	117	86
Pressure (psia)	265.0	2,214.7	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	3,766	3,764	3,763

	1	2	3
V-L Mole Fraction			
Aspen Plus Enthalpy (Btu/lb)[B]	-3,847	-3,929	-3,950
Density (lb/ft^3)	2.14	43.0	39.3

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance results given are taken from the current GTL Study case that considered CO_2 capture. The performance summary is provided in Exhibit 5-48.

Exhibit 5-48. Performance summary

Performance Summary	
Item	**50,000 BPD (kWe)**
CO_2 Compressor	6,700
Circulating Water Pumps	20
Cooling Tower Fans	10
Total Auxiliary Load	**6,730**

5.6.6 Capture Integration

For the purposes of this study, it is assumed that a stand-alone cooling water unit will perform the necessary cooling for capture and compression. No retrofit case is considered for GTL as any new builds would most likely include compression. However, to make this case comparable to the other cases considered in this study, the cost for cooling is included in the greenfield COC.

5.6.7 Power Source

The power consumption is approximated as 6.73 MW, which includes all power required by the compression train and the cooling water system. Power requirement estimates for the cooling water unit were scaled as described in Section 4.4. For practical applications for this type of facility with power produced on-site and excess power sent to the grid, the power requirements may be met with power generated on-site. For instance, while the GTL Study has excess power that would be able to satisfy a portion of this study's power requirement, this scenario should be evaluated on a case-by-case basis, which is not included in the scope of this report.

5.6.8 Economic Analysis Results

The economic results for CO_2 capture application in a GTL plant are presented in this section. Owner's costs (Exhibit 5-49), capital costs (Exhibit 5-50), and O&M costs are calculated as discussed in Section 3.1. The greenfield TOC for the GTL case is \$59.7 M. The corresponding greenfield COC is \$6.4/tonne CO_2.

Exhibit 5-49. Owners' costs for GTL greenfield site

Description	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs		
6 Months All Labor	$471	$0
1-Month Maintenance Materials	$46	$0
1-Month Non-Fuel Consumables	$6	$0
1-Month Waste Disposal	$0	$0
25% of 1-Month Fuel Cost at 100% CF	$0	$0
2% of TPC	$983	$1
Total	**$1,507**	**$1**
Inventory Capital		
60-day supply of fuel and consumables at 100% CF	$6	$0
0.5% of TPC (spare parts)	$246	$0
Total	**$252**	**$0**
Other Costs		
Initial Cost for Catalyst and Chemicals	$0	$0
Land	$30	$0
Other Owner's Costs	$7,375	$4
Financing Costs	$1,328	$1
TOC	**$59,661**	**$32**
TASC Multiplier (GTL, 31 year)	1.054	
TASC	**$62,890**	**$34**

Exhibit 5-50. Capital costs for GTL greenfield site

	Case:	GTL					Estimate Type:			Conceptual	
	Representative Plant Size:	50,000 BPD Fischer-Tropsch liquids					Cost Base:			Dec 2018	
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	$/tonnes/yr ($CO_2$)
5	Flue Gas Cleanup										
5.4	CO_2 Compression & Drying	$14,192	$0	$5,432	$0	$19,624	$3,434	$0	$4,612	$27,670	$15
5.5	CO_2 Compressor Aftercooler	$77	$12	$33	$0	$122	$21	$0	$29	$172	$0
	Subtotal	**$14,269**	**$12**	**$5,465**	**$0**	**$19,746**	**$3,456**	**$0**	**$4,640**	**$27,842**	**$15**
7	Ductwork & Stack										
7.3	Ductwork	$0	$75	$52	$0	$126	$22	$0	$30	$178	$0
	Subtotal	**$0**	**$75**	**$52**	**$0**	**$126**	**$22**	**$0**	**$30**	**$178**	**$0**
9	Cooling Water System										
9.1	Cooling Towers	$197	$0	$61	$0	$257	$45	$0	$61	$363	$0
9.2	Circulating Water Pumps	$16	$0	$1	$0	$17	$3	$0	$4	$24	$0
9.3	Circulating Water System Aux.	$375	$0	$50	$0	$424	$74	$0	$100	$598	$0
9.4	Circulating Water Piping	$0	$173	$157	$0	$330	$58	$0	$78	$466	$0
9.5	Make-up Water System	$59	$0	$75	$0	$134	$23	$0	$31	$189	$0
9.6	Component Cooling Water System	$27	$0	$21	$0	$48	$8	$0	$11	$67	$0
9.7	Circulating Water System Foundations	$0	$22	$36	$0	$58	$10	$0	$14	$82	$0
	Subtotal	**$673**	**$195**	**$401**	**$0**	**$1,269**	**$222**	**$0**	**$298**	**$1,789**	**$1**
11	Accessory Electric Plant										
11.2	Station Service Equipment	$1,831	$0	$157	$0	$1,988	$348	$0	$467	$2,803	$2
11.3	Switchgear & Motor Control	$2,842	$0	$493	$0	$3,335	$584	$0	$784	$4,702	$3
11.4	Conduit & Cable Tray	$0	$369	$1,065	$0	$1,434	$251	$0	$337	$2,022	$1
11.5	Wire & Cable	$0	$978	$1,749	$0	$2,727	$477	$0	$641	$3,845	$2
	Subtotal	**$4,672**	**$1,348**	**$3,463**	**$0**	**$9,484**	**$1,660**	**$0**	**$2,229**	**$13,372**	**$7**
12	Instrumentation & Control										
12.8	Instrument Wiring & Tubing	$359	$288	$1,150	$0	$1,797	$315	$0	$422	$2,534	$1
12.9	Other I&C Equipment	$442	$0	$1,023	$0	$1,465	$256	$0	$344	$2,066	$1
	Subtotal	**$801**	**$288**	**$2,173**	**$0**	**$3,262**	**$571**	**$0**	**$767**	**$4,600**	**$2**
13	Improvements to Site										
13.1	Site Preparation	$0	$22	$452	$0	$474	$83	$0	$111	$669	$0
13.2	Site Improvements	$0	$105	$140	$0	$245	$43	$0	$58	$345	$0
13.3	Site Facilities	$120	$0	$126	$0	$247	$43	$0	$58	$348	$0
	Subtotal	**$120**	**$128**	**$718**	**$0**	**$966**	**$169**	**$0**	**$227**	**$1,362**	**$1**
14	Buildings & Structures										
14.5	Circulation Water Pumphouse	$0	$11	$9	$0	$19	$3	$0	$5	$27	$0
	Subtotal	**$0**	**$11**	**$9**	**$0**	**$19**	**$3**	**$0**	**$5**	**$27**	**$0**
	Total	**$20,536**	**$2,056**	**$12,280**	**$0**	**$34,872**	**$6,103**	**$0**	**$8,195**	**$49,170**	**$26**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 5-51, while Exhibit 5-52 shows the COC for a greenfield site for the representative GTL plant.

Exhibit 5-51. Initial and annual O&M costs for GTL greenfield site

Case:	Gas-to-Liquids				Cost Base:	Dec 2018
Representative Plant Size:	50,000 BPD Fischer-Tropsch liquids				Capacity Factor (%):	85
Operating & Maintenance Labor						
Operating Labor				Operating Labor Requirements per Shift		
Operating Labor Rate (base):		38.50	$/hour	Skilled Operator:		0.0
Operating Labor Burden:		30.00	% of base	Operator:		1.0
Labor O-H Charge Rate:		25.00	% of labor	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**1.0**
Fixed Operating Costs						
					Annual Cost	
					($)	($/tonnes/yr CO_2)
Annual Operating Labor:					$438,438	$0.24
Maintenance Labor:					$314,687	$0.17
Administrative & Support Labor:					$188,281	$0.10
Property Taxes and Insurance:					$983,396	$0.53
Total:					**$1,924,802**	**$1.04**
Variable Operating Costs						
					($)	($/tonnes/yr CO_2)
Maintenance Material:					$472,030	$0.30
Consumables						
	Initial Fill	Per Day	Per Unit	Initial Fill		
Water (/1000 gallons):	0	59	$1.90	$0	$34,632	$0.02
Makeup and Waste Water Treatment Chemicals (ton):	0	0.2	$550.00	$0	$29,863	$0.02
Subtotal:				**$0**	**$64,495**	**$0.04**
Variable Operating Costs Total:				**$0**	**$536,526**	**$0.34**

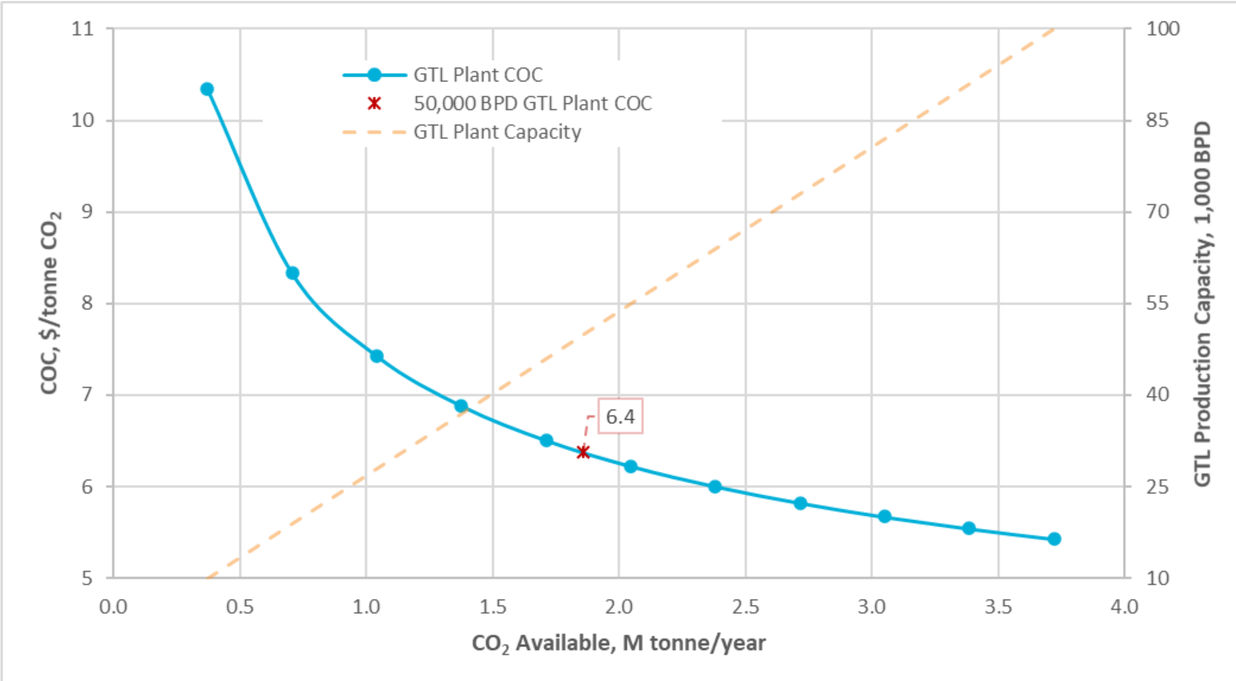
Exhibit 5-52. COC for 50,000 BPD GTL greenfield

Component	Greenfield Value, $/tonne CO_2
Capital	2.9
Fixed	1.2
Variable	0.3
Purchased Power	1.9
Total COC	**6.4**

5.6.9 Plant Capacity Sensitivity Analysis

An analysis of the sensitivity of greenfield COC to GTL plant capacity is shown in Exhibit 5-53. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 5-53. GTL plant capacity sensitivity



5.6.10 Gas-to-Liquids Conclusion

The high purity CO_2 stream produced from GTL plants makes them a relatively low-cost industrial process since the plant performs the CO_2 separation as a part of normal operations. A CO_2 compression system for a 50,000 BPD plant was modeled to estimate the COC of capturing CO_2 from the process. The results showed the COC of CO_2 to be \$6.4/tonne CO_2 for a greenfield site. The plant size sensitivity showed that as plant size decreased from 100,000 to 10,000 BPD, the COC increased by \$4.9/tonne CO_2. As the plant size is decreased, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC.

6 COST AND PERFORMANCE: LOW PURITY SOURCES

The sources discussed in this section are considered low purity sources, meaning the available CO_2 requires purification to meet EOR pipeline specifications. The CO_2 removal systems described in Section 4.2 are employed to purify the CO_2 streams to meet QGESS specifications for EOR pipeline end-use. In all low purity cases, compression, cooling, and TEG dehydration of the CO_2 stream is required following capture and purification.

6.1 REFINERY HYDROGEN

Refineries are an example of an industrial source that currently deploys gas separation technology to produce hydrogen. Like other gas processing, hydrogen production emits CO_2 not only from the process gas, but from the SMR in the form of flue gas, like that of a power plant. NETL has studied hydrogen production with post-combustion CO_2 capture as part of their "Comparison of Commercial, State-of-the-Art, Fossil-Based Hydrogen Production Technologies" [40], evaluating H_2 production via SMR and coal gasification.

6.1.1 Size Range

Size range for hydrogen production varies widely depending on the industry. Ninety-five percent of hydrogen produced in the United States is done so by way of NG reforming in refineries. [41] The Shell Quest CCS facility in Alberta, Canada has successfully captured and stored over 5 million tonnes of CO_2 from a refinery hydrogen production process since its startup in 2015. [42] The Scotford Upgrader near Edmonton, Alberta, Canada includes three hydrogen manufacturing units and produces a total of 367 MMSCFD (322,461 tonnes/year) of hydrogen. As a result, approximately 1.5 M tonnes/year CO_2 is available at the facility. The information provided by Shell regarding their ADIP-Ultra pre-combustion CO_2 capture process detailed in Section 4.2.2 provided cost and performance data for an 87,000 tonnes/year hydrogen production facility, with 404,700 tonnes/year CO_2 available for capture (at 100 percent CF). [2] As such, the representative plant for the refinery hydrogen case will mirror that of the quote provided by Shell. [2]

6.1.2 CO_2 Point Sources

When producing hydrogen via SMR, Shell indicates that advanced capture systems (i.e., 99 percent CO_2 capture rate or greater) are most economically implemented in the raw syngas stream from the SMR. At lower capture rates, a post-combustion CO_2 unit would likely be more economically viable, but for the purpose of comparison of like technologies between cases, the ADIP-Ultra pre-combustion system is employed in both the 90 and 99 percent capture scenarios for the refinery hydrogen case. The pre-combustion AGR captures CO_2 upstream of the pressure-swing adsorption (PSA) unit, which separates the high purity hydrogen from the syngas stream for further processing and end-use. The pre-PSA stream to be purified is characterized in Exhibit 6-1.

Exhibit 6-1. Stream characteristics of raw syngas from SMR

Component	Vapor Mole Fraction
CO_2	0.1918
H_2O	0.0032
CH_4	0.0272
C_2H_6	0.0074
C_3H_8	0.0017
C_4H_{10}	0.0009
CO	0.0015
H_2	0.7632
N_2	0.0030
Component	**Liquid Weight Fraction**
CO_2	0.0047
H_2O	0.9952
Parameter	**Value**
Total Stream Vapor Faction	0.658
Temperature	102.2°F
Pressure	400.3 psia

6.1.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the refinery hydrogen process for the purpose of this study:

- The representative refinery hydrogen production unit has a capacity of 87,000 tonnes hydrogen/year
- The raw syngas has a total stream CO_2 concentration of 12.7 mole percent
- The total CO_2 generated at 100 percent CF is 404,700 tonnes CO_2/year
- As a low purity source, separation, compression, and cooling are required. Separation is accomplished using an ADIP-Ultra AGR unit
- The temperature of the CO_2 entering the AGR pre-scrubber is 102.2°F
- The pressure of the stream entering the AGR pre-scrubber is 400.3 psia
- CO_2 capture rates of 90 percent and 99 percent are evaluated
- The CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

6.1.4 CO_2 Capture System

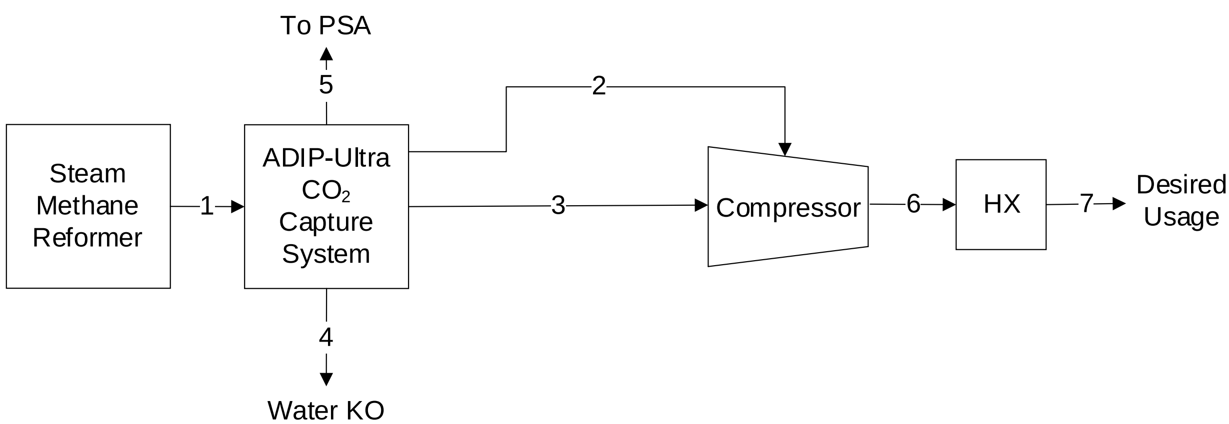
With an assumed concentration of only 12.7 mole percent CO_2 in the raw syngas from SMR, separation is required to meet QGESS EOR pipeline specifications. In addition, water removal, compression, and cooling are necessary to create a CO_2 product stream suitable for EOR end-use. The Shell ADIP-Ultra pre-combustion AGR unit detailed in Section 4.2.2 is modeled to represent CO_2 removal at 90 and 99 percent. AGR auxiliary loads are scaled based on CO_2 flowrate.

The AGR unit requires low pressure steam at 74 psia to regenerate the amine-based solvent. These steam needs are met with the industrial boiler discussed in Section 4.3. In addition, cooling water is required for both the AGR unit and for compression intercooling and after-cooler. The cooling water unit auxiliaries are scaled as described in Section 4.4.

6.1.5 BFD, Stream Table, and Performance Summary

The raw syngas from SMR (stream 1) is fed to ADIP-Ultra capture unit, resulting in four main process streams. Water (stream 4) is removed in the knock-out drum and is routed to waste treatment. In stream 5 of Exhibit 6-2, H_2 and methane (CH_4) (along with other hydrocarbons) are sent to the PSA where the H_2 product is separated for end-use. The remaining process streams are the purified CO_2 streams: one at "mid-pressure" (stream 2) and one at "low-pressure" (stream 30). The CO_2 streams are routed to the centrifugal compressor, like that described in Section 4.1.2, and an aftercooler is used to produce a high purity CO_2 stream at 2,214.7 psia and 86°F for EOR pipeline use.

Exhibit 6-2. CO_2 capture BFD



The stream tables for 99 and 90 percent capture in the refinery hydrogen case are presented in Exhibit 6-3 and Exhibit 6-4, respectively.

Exhibit 6-3. Refinery hydrogen stream table for 99 percent capture

	1	2	3	4	5	6	7
V-L Mole Fraction							
AR	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0179	0.0000	0.0000	0.0000	0.0336	0.0000	0.0000
CO	0.0010	0.0000	0.0000	0.0000	0.0017	0.0000	0.0000
CO_2	0.1268	0.8644	0.9629	0.0020	0.0023	0.9995	0.9995
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.5020	0.0000	0.0000	0.0000	0.9427	0.0000	0.0000
H_2O	0.3438	0.1356	0.0371	0.9980	0.0039	0.0005	0.0005
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0020	0.0000	0.0000	0.0000	0.0034	0.0000	0.0000
C_2H_6	0.0049	0.0000	0.0000	0.0000	0.0091	0.0000	0.0000
C_3H_8	0.0011	0.0000	0.0000	0.0000	0.0021	0.0000	0.0000
C_4H_{10}	0.0006	0.0000	0.0000	0.0000	0.0011	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	8,320	593	546	2,848	4,431	1,040	1,040
V-L Flowrate (kg/hr)	111,368	24,023	23,524	51,457	14,118	45,736	45,736
Temperature (°C)	39	102	40	39	55	121	29
Pressure (MPa, abs)	2.76	0.6	0.2	2.8	2.7	15.3	15.3
Steam Table Enthalpy (kJ/kg)[A]	10,569	9,172	8,865	15,273	1,597	8,760	8,755
Aspen Plus Enthalpy (kJ/kg)[B]	-11,217	-9,144	-9,005	-15,873	-1,501	-8,952	-9,196
Density (kg/m^3)	21.3	8.3	3.3	918.8	3.1	283.1	640.4
V-L Molecular Weight	13.4	40.5	43.0	18.1	3.19	44.0	44.0
V-L Flowrate (lb_{mol}/hr)	18,342	1,308	1,205	6,279	9,768	2,292	2,292
V-L Flowrate (lb/hr)	245,524	52,962	51,861	113,443	31,124	100,830	100,830
Temperature (°F)	102	216	104	102	131	250	85
Pressure (psia)	399.9	90.8	28.3	399.9	394.4	2,215.9	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	4,544	3,943	3,811	6,566	686	3,766	3,764
Aspen Plus Enthalpy (Btu/lb)[B]	-4,822	-3,931	-3,871	-6,824	-645	-3,849	-3,954
Density (lb/ft^3)	1.33	0.518	0.204	57.4	0.196	17.7	44.0

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 6-4. Refinery hydrogen stream table for 90 percent capture

	1	2	3	4	5	6	7
V-L Mole Fraction							
AR	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0179	0.0009	0.0000	0.0000	0.0328	0.0000	0.0000
CO	0.0010	0.0000	0.0000	0.0000	0.0018	0.0000	0.0000
CO_2	0.1268	0.9493	0.9629	0.0020	0.0232	0.9995	0.9995
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.5020	0.0130	0.0000	0.0000	0.9219	0.0000	0.0000
H_2O	0.3438	0.0368	0.0371	0.9980	0.0047	0.0005	0.0005
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0020	0.0000	0.0000	0.0000	0.0036	0.0000	0.0000
C_2H_6	0.0049	0.0000	0.0000	0.0000	0.0090	0.0000	0.0000
C_3H_8	0.0011	0.0000	0.0000	0.0000	0.0020	0.0000	0.0000
C_4H_{10}	0.0006	0.0000	0.0000	0.0000	0.0011	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	8,320	496	492	2,848	4,524	945	945
V-L Flowrate (kg/hr)	111,368	21,076	21,170	51,457	18,375	41,572	41,572
Temperature (°C)	39	102	40	39	55	121	29
Pressure (MPa, abs)	2.76	0.6	0.2	2.8	2.7	15.3	15.3
Steam Table Enthalpy (kJ/kg)[A]	10,569	8,858	8,865	15,273	2,884	8,760	8,755
Aspen Plus Enthalpy (kJ/kg)[B]	-11,217	-8,938	-9,005	-15,873	-3,225	-8,952	-9,196
Density (kg/m^3)	21.3	8.7	3.3	918.8	4.0	283.1	640.4
V-L Molecular Weight	13.4	42.5	43.0	18.1	4.06	44.0	44.0
V-L Flowrate (lb_{mol}/hr)	18,342	1,094	1,084	6,279	9,973	2,083	2,083
V-L Flowrate (lb/hr)	245,524	46,464	46,672	113,443	40,510	91,651	91,651
Temperature (°F)	102	216	104	102	131	250	85
Pressure (psia)	399.9	90.8	28.3	399.9	394.4	2,215.9	2,214.7
Steam Table Enthalpy (Btu/lb)[A]	4,544	3,808	3,811	6,566	1,240	3,766	3,764
Aspen Plus Enthalpy (Btu/lb)[B]	-4,822	-3,843	-3,871	-6,824	-1,387	-3,849	-3,954
Density (lb/ft^3)	1.33	0.542	0.204	57.4	0.250	17.7	40.0

[A]Steam table reference conditions are 32.02°F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance summaries for 90 and 99 percent capture in the refinery hydrogen case are presented in Exhibit 6-5.

Exhibit 6-5. Refinery hydrogen performance summary

Performance Summary		
Item	87,000 tonnes H_2/year with 90 percent CO_2 capture (kWe)	87,000 tonnes H_2/year with 99 percent CO_2 capture (kWe)
CO_2 Capture Auxiliaries	500	500
Steam Boiler Auxiliaries	70	80
CO_2 Compressor	3,160	3,470
Circulating Water Pumps	210	240
Cooling Tower Fans	100	120
Total Auxiliary Load	**4,040**	**4,410**

6.1.6 Capture Integration

The cost and performance implications of adding an NG-fired boiler, as described in Section 4.3, were estimated to meet the steam demands of the Shell ADIP-Ultra CO_2 removal system. However, in real applications at refineries, if steam requirements for the AGR process are met with waste heat from the existing process, an additional boiler for solvent regeneration heating needs may not be necessary. The cooling water system is considered a stand-alone addition; however, there is potential to integrate existing make-up water systems to feed or partially feed the cooling water system, thereby reducing the unit's size, or replacing it completely with a simple HX.

6.1.7 Power Source

The compressor power consumption for the 90 and 99 percent capture cases are 3.16 MW and 3.47 MW, respectively. Power consumption estimates for the cooling water system in each case were scaled as described in Section 4.4. The total power requirements were calculated to be 4.01 MW and 4.42 MW for the 90 and 99 percent capture rates, respectively, which includes all power required by the compression train, cooling water system, and ADIP-Ultra capture unit. Purchased power cost is estimated at a rate of $60/MWh as discussed in Section 3.1.2.4. To satisfy the steam requirements of the AGR system, an industrial boiler was modeled, and fuel consumption costs were estimated at a rate of $4.42/MMBtu as discussed in Section 3.1.2.4.

6.1.8 Economic Analysis Results

The economic results of CO_2 capture application in a refinery hydrogen plant are presented in this section. Owner's costs (Exhibit 6-6), capital costs (Exhibit 6-7 and Exhibit 6-8), and O&M costs are calculated as discussed in Section 3.1. Retrofit costs were determined by applying a retrofit factor to TPC as discussed in Section 3.3. The greenfield TOC for the refinery hydrogen case at 99 percent capture is $159.2 M, while for 90 percent capture, a greenfield TOC of $155.0 M is estimated. The corresponding greenfield COC for the 99 percent and 90 percent capture cases are $57.3/tonne CO_2 and $59.9/tonne CO_2, respectively. The COC is $58.9/tonne CO_2 and $61.7/tonne CO_2 in retrofit applications for 99 percent and 90 percent capture, respectively.

Exhibit 6-6. Owners' costs for refinery hydrogen cases

Description	$/1,000	$/tonnes/yr ($CO_2$)	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs	**99% Capture**		**90% Capture**	
6 Months All Labor	$1,153	$3	$1,139	$3
1-Month Maintenance Materials	$123	$0	$120	$0
1-Month Non-Fuel Consumables	$46	$0	$42	$0
1-Month Waste Disposal	$0	$0	$0	$0
25% of 1-Month Fuel Cost at 100% CF	$89	$0	$78	$0
2% of TPC	$2,613	$7	$2,544	$7
Total	**$4,024**	**$10**	**$3,923**	**$11**
Inventory Capital	**99% Capture**		**90% Capture**	
60-day supply of fuel and consumables at 100% CF	$786	$2	$693	$2
0.5% of TPC (spare parts)	$653	$2	$636	$2
Total	**$1,439**	**$4**	**$1,329**	**$4**
Other Costs	**99% Capture**		**90% Capture**	
Initial Cost for Catalyst and Chemicals	$0	$0	$0	$0
Land	$30	$0	$30	$0
Other Owner's Costs	$19,594	$49	$19,078	$52
Financing Costs	$3,527	$9	$3,434	$9
TOC	**$159,244**	**$397**	**$154,978**	**$426**
TASC Multiplier (Refinery Hydrogen, 33 year)	1.036		1.036	
TASC	**$164,929**	**$412**	**$160,510**	**$441**

Exhibit 6-7. Capital costs for refinery hydrogen greenfield site with 99 percent capture

	Case:	Refinery H_2						Estimate Type:		Conceptual	
	Representative Plant Size:	87,000 tonnes H_2/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$237	$407	$203	$0	$847	$148	$0	$199	$1,195	$3
3.2	Water Makeup & Pretreating	$552	$55	$313	$0	$921	$161	$0	$216	$1,298	$3
3.3	Other Feedwater Subsystems	$82	$27	$25	$0	$134	$23	$0	$32	$189	$0
3.4	Industrial Boiler Package w/Deaerator	$1,090	$0	$317	$0	$1,407	$246	$0	$331	$1,985	$5
3.5	Other Boiler Plant Systems	$19	$7	$18	$0	$44	$8	$0	$10	$62	$0
3.6	NG Pipeline and Start-Up System	$317	$14	$10	$0	$341	$60	$0	$80	$481	$1
3.7	Waste Water Treatment Equipment	$2,898	$0	$1,776	$0	$4,675	$818	$0	$1,099	$6,591	$16
3.9	Miscellaneous Plant Equipment	$68	$9	$34	$0	$111	$19	$0	$26	$157	$0
	Subtotal	**$5,265**	**$518**	**$2,698**	**$0**	**$8,481**	**$1,484**	**$0**	**$1,993**	**$11,958**	**$30**
5	Flue Gas Cleanup										
5.1	ADIP-Ultra CO_2 Removal System	$21,678	$9,377	$19,691	$0	$50,746	$8,881	$8,627	$13,651	$81,904	$204
5.4	CO_2 Compression & Drying	$7,402	$1,110	$2,475	$0	$10,987	$1,923	$0	$2,582	$15,492	$39
5.5	CO_2 Compressor Aftercooler	$81	$13	$35	$0	$129	$23	$0	$30	$182	$0
5.12	Gas Cleanup Foundations	$0	$4	$4	$0	$8	$1	$0	$2	$11	$0
	Subtotal	**$29,162**	**$10,504**	**$22,204**	**$0**	**$61,870**	**$10,827**	**$8,627**	**$16,265**	**$97,589**	**$244**
7	Ductwork & Stack										
7.3	Ductwork	$0	$66	$46	$0	$112	$20	$0	$26	$158	$0
7.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.5	Duct & Stack Foundations	$0	$157	$187	$0	$344	$60	$0	$81	$485	$1
	Subtotal	**$0**	**$223**	**$233**	**$0**	**$456**	**$80**	**$0**	**$107**	**$643**	**$2**
9	Cooling Water System										
9.1	Cooling Towers	$455	$0	$141	$0	$596	$104	$0	$140	$840	$2
9.2	Circulating Water Pumps	$41	$0	$3	$0	$44	$8	$0	$10	$62	$0
9.3	Circulating Water System Aux.	$744	$0	$98	$0	$843	$148	$0	$198	$1,188	$3
9.4	Circulating Water Piping	$0	$344	$312	$0	$656	$115	$0	$154	$925	$2
9.5	Make-up Water System	$100	$0	$128	$0	$228	$40	$0	$54	$322	$1
9.6	Component Cooling Water System	$54	$0	$41	$0	$95	$17	$0	$22	$134	$0
9.7	Circulating Water System Foundations	$0	$41	$68	$0	$109	$19	$0	$26	$154	$0
	Subtotal	**$1,394**	**$385**	**$791**	**$0**	**$2,571**	**$450**	**$0**	**$604**	**$3,624**	**$9**
11	Accessory Electric Plant										
11.2	Station Service Equipment	$1,527	$0	$131	$0	$1,658	$290	$0	$390	$2,338	$6
11.3	Switchgear & Motor Control	$2,371	$0	$411	$0	$2,782	$487	$0	$654	$3,923	$10
11.4	Conduit & Cable Tray	$0	$308	$888	$0	$1,196	$209	$0	$281	$1,687	$4
11.5	Wire & Cable	$0	$816	$1,459	$0	$2,275	$398	$0	$535	$3,208	$8
	Subtotal	**$3,898**	**$1,124**	**$2,890**	**$0**	**$7,912**	**$1,385**	**$0**	**$1,859**	**$11,156**	**$28**

	Case:	Refinery H_2					Estimate Type:			Conceptual	
	Representative Plant Size:	87,000 tonnes H_2/year					Cost Base:			Dec 2018	
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	$/tonnes/yr ($CO_2$)
12						Instrumentation & Control					
12.8	Instrument Wiring & Tubing	$340	$272	$1,089	$0	$1,701	$298	$0	$400	$2,399	$6
12.9	Other I&C Equipment	$418	$0	$969	$0	$1,387	$243	$0	$326	$1,955	$5
	Subtotal	**$759**	**$272**	**$2,057**	**$0**	**$3,088**	**$540**	**$0**	**$726**	**$4,354**	**$11**
13						Improvements to Site					
13.1	Site Preparation	$0	$21	$415	$0	$436	$76	$0	$102	$615	$2
13.2	Site Improvements	$0	$97	$128	$0	$225	$39	$0	$53	$317	$1
13.3	Site Facilities	$111	$0	$116	$0	$227	$40	$0	$53	$320	$1
	Subtotal	**$111**	**$117**	**$660**	**$0**	**$888**	**$155**	**$0**	**$209**	**$1,252**	**$3**
14						Buildings & Structures					
14.5	Circulation Water Pumphouse	$0	$21	$16	$0	$37	$6	$0	$9	$52	$0
	Subtotal	**$0**	**$21**	**$16**	**$0**	**$37**	**$6**	**$0**	**$9**	**$52**	**$0**
	Total	**$40,588**	**$13,166**	**$31,550**	**$0**	**$85,303**	**$14,928**	**$8,627**	**$21,772**	**$130,630**	**$326**

Exhibit 6-8. Capital costs for refinery hydrogen greenfield site with 90 percent capture

	Case:	Refinery H_2					Estimate Type:			Conceptual	
	Representative Plant Size:	87,000 tonnes H_2/year					Cost Base:			Dec 2018	
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	$/tonnes/yr ($CO_2$)
3						Feedwater & Miscellaneous BOP Systems					
3.1	Feedwater System	$217	$372	$186	$0	$775	$136	$0	$182	$1,092	$3
3.2	Water Makeup & Pretreating	$495	$49	$280	$0	$825	$144	$0	$194	$1,163	$3
3.3	Other Feedwater Subsystems	$73	$24	$23	$0	$119	$21	$0	$28	$168	$0
3.4	Industrial Boiler Package w/Deaerator	$971	$0	$282	$0	$1,254	$219	$0	$295	$1,768	$5
3.5	Other Boiler Plant Systems	$17	$6	$16	$0	$39	$7	$0	$9	$55	$0
3.6	NG Pipeline and Start-Up System	$298	$13	$10	$0	$320	$56	$0	$75	$451	$1
3.7	Waste Water Treatment Equipment	$2,807	$0	$1,721	$0	$4,528	$792	$0	$1,064	$6,385	$18
3.9	Miscellaneous Plant Equipment	$66	$9	$33	$0	$108	$19	$0	$25	$152	$0
	Subtotal	**$4,944**	**$473**	**$2,551**	**$0**	**$7,968**	**$1,394**	**$0**	**$1,872**	**$11,234**	**$31**
5						Flue Gas Cleanup					
5.1	ADIP-Ultra CO_2 Removal System	$21,409	$9,260	$19,447	$0	$50,116	$8,770	$8,520	$13,481	$80,888	$222
5.4	CO_2 Compression & Drying	$6,991	$1,049	$2,338	$0	$10,377	$1,816	$0	$2,439	$14,632	$40
5.5	CO_2 Compressor Aftercooler	$75	$12	$32	$0	$120	$21	$0	$28	$169	$0
5.12	Gas Cleanup Foundations	$0	$4	$4	$0	$8	$1	$0	$2	$11	$0
	Subtotal	**$28,476**	**$10,325**	**$21,820**	**$0**	**$60,621**	**$10,609**	**$8,520**	**$15,950**	**$95,700**	**$263**
7						Ductwork & Stack					

	Case:	Refinery H_2						Estimate Type:		Conceptual	
	Representative Plant Size:	87,000 tonnes H_2/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
7.3	Ductwork	$0	$66	$46	$0	$112	$20	$0	$26	$158	$0
7.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.5	Duct & Stack Foundations	$0	$156	$185	$0	$341	$60	$0	$80	$481	$1
	Subtotal	**$0**	**$222**	**$231**	**$0**	**$454**	**$79**	**$0**	**$107**	**$640**	**$2**
9						Cooling Water System					
9.1	Cooling Towers	$405	$0	$125	$0	$530	$93	$0	$124	$747	$2
9.2	Circulating Water Pumps	$36	$0	$3	$0	$38	$7	$0	$9	$54	$0
9.3	Circulating Water System Aux.	$676	$0	$89	$0	$766	$134	$0	$180	$1,079	$3
9.4	Circulating Water Piping	$0	$313	$283	$0	$596	$104	$0	$140	$840	$2
9.5	Make-up Water System	$93	$0	$119	$0	$212	$37	$0	$50	$299	$1
9.6	Component Cooling Water System	$49	$0	$37	$0	$86	$15	$0	$20	$121	$0
9.7	Circulating Water System Foundations	$0	$37	$62	$0	$100	$17	$0	$23	$141	$0
	Subtotal	**$1,258**	**$350**	**$719**	**$0**	**$2,327**	**$407**	**$0**	**$547**	**$3,281**	**$9**
11						Accessory Electric Plant					
11.2	Station Service Equipment	$1,471	$0	$126	$0	$1,597	$280	$0	$375	$2,252	$6
11.3	Switchgear & Motor Control	$2,284	$0	$396	$0	$2,680	$469	$0	$630	$3,779	$10
11.4	Conduit & Cable Tray	$0	$297	$856	$0	$1,152	$202	$0	$271	$1,625	$4
11.5	Wire & Cable	$0	$786	$1,405	$0	$2,191	$384	$0	$515	$3,090	$8
	Subtotal	**$3,755**	**$1,083**	**$2,783**	**$0**	**$7,621**	**$1,334**	**$0**	**$1,791**	**$10,745**	**$30**
12						Instrumentation & Control					
12.8	Instrument Wiring & Tubing	$336	$269	$1,077	$0	$1,682	$294	$0	$395	$2,372	$7
12.9	Other I&C Equipment	$414	$0	$958	$0	$1,371	$240	$0	$322	$1,933	$5
	Subtotal	**$750**	**$269**	**$2,034**	**$0**	**$3,053**	**$534**	**$0**	**$718**	**$4,305**	**$12**
13						Improvements to Site					
13.1	Site Preparation	$0	$20	$408	$0	$428	$75	$0	$101	$604	$2
13.2	Site Improvements	$0	$95	$126	$0	$221	$39	$0	$52	$312	$1
13.3	Site Facilities	$109	$0	$114	$0	$223	$39	$0	$52	$315	$1
	Subtotal	**$109**	**$115**	**$649**	**$0**	**$873**	**$153**	**$0**	**$205**	**$1,231**	**$3**
14						Buildings & Structures					
14.5	Circulation Water Pumphouse	$0	$19	$15	$0	$34	$6	$0	$8	$48	$0
	Subtotal	**$0**	**$19**	**$15**	**$0**	**$34**	**$6**	**$0**	**$8**	**$48**	**$0**
	Total	**$39,291**	**$12,857**	**$30,802**	**$0**	**$82,950**	**$14,516**	**$8,520**	**$21,197**	**$127,184**	**$349**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 6-9 and Exhibit 6-10 for 99 percent and 90 percent capture, respectively, while Exhibit 6-11 shows the COC for greenfield and retrofit sites for the representative refinery hydrogen plants at both capture rates.

Exhibit 6-9. Initial and annual O&M costs for refinery hydrogen greenfield site with 99 percent capture

Case:	Refinery Hydrogen				Cost Base:	Dec 2018
Representative Plant Size:	87,000 tonnes H_2/year				Capacity Factor (%):	85
Operating & Maintenance Labor						
Operating Labor				Operating Labor Requirements per Shift		
Operating Labor Rate (base):		38.50	$/hour	Skilled Operator:		0.0
Operating Labor Burden:		30.00	% of base	Operator:		2.3
Labor O-H Charge Rate:		25.00	% of labor	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**2.3**
Fixed Operating Costs						
					Annual Cost	
					($)	($/tonnes/yr CO_2)
Annual Operating Labor:					$1,008,407	$2.52
Maintenance Labor:					$836,029	$2.09
Administrative & Support Labor:					$461,109	$1.15
Property Taxes and Insurance:					$2,612,591	$6.52
Total:					**$4,918,137**	**$12.28**
Variable Operating Costs						
					($)	($/tonnes/yr CO_2)
Maintenance Material:					$1,254,044	$3.68
Consumables						
	Initial Fill	Per Day	Per Unit	Initial Fill		
Water (/1000 gallons):	0	176	$1.90	$0	$103,543	$0.30
Makeup and Waste Water Treatment Chemicals (ton):	0	1.5	$550.00	$0	$261,093	$0.77
CO_2 Capture System Chemicals[A]:	Proprietary				$29,128	$0.09
Triethylene Glycol (gal):	w/equip.	37	$6.80	$0	$78,191	$0.23
Subtotal:				**$0**	$471,954	$1.39
Waste Disposal						
Triethylene Glycol (gal):		37	$0.35	$0	$4,025	$0.01
Subtotal:				**$0**	**$4,025**	**$0.01**
Variable Operating Costs Total:				**$0**	**$1,730,022**	**$5.08**
Fuel Cost						
Natural Gas (MMBtu)	0	2,653	$4.42	$0	$3,638,461	$10.68
Total:				**$0**	**$3,638,461**	**$10.68**

[A]CO_2 capture system chemicals includes ADIP-Ultra Solvent

Exhibit 6-10. Initial and annual O&M costs for refinery hydrogen greenfield site with 90 percent capture

Case:	**Refinery Hydrogen**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**87,000 tonnes H_2/year**				**Capacity Factor (%):**	**85**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		2.3
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**2.3**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$1,008,407	$2.77
Maintenance Labor:					$813,977	$2.24
Administrative & Support Labor:					$455,596	$1.25
Property Taxes and Insurance:					$2,543,679	$6.98
Total:					**$4,821,660**	**$13.24**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$1,220,966	$3.94
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	151	$1.90	$0	$89,100	$0.29
Makeup and Waste Water Treatment Chemicals (ton):	0	1.4	$550.00	$0	$244,702	$0.79
CO_2 Capture System Chemicals[A]:	Proprietary				$23,994	$0.08
Triethylene Glycol (gal):	w/equip.	34	$6.80	$0	$71,551	$0.23
Subtotal:				**$0**	**$429,347**	**$1.39**
Waste Disposal						
Triethylene Glycol (gal):		34	$0.35	$0	$3,683	$0.01
Subtotal:				**$0**	**$3,683**	**$0.01**
Variable Operating Costs Total:				**$0**	**$1,653,996**	**$5.34**
Fuel Cost						
Natural Gas (MMBtu)	0	2,330	$4.42	$0	$3,194,817	$10.32
Total:				**$0**	**$3,194,817**	**$10.32**

[A]CO_2 capture system chemicals includes ADIP-Ultra Solvent

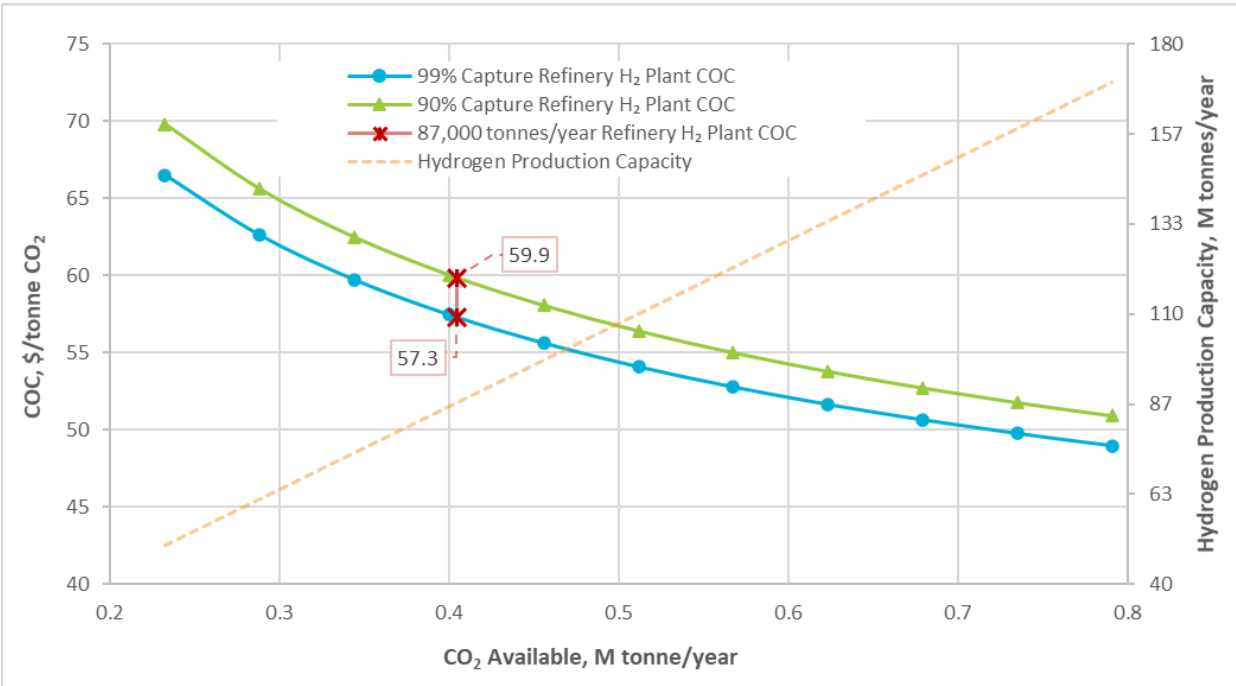
Exhibit 6-11. COC for 87,000 tonnes H_2/year refinery hydrogen cases

	99% Capture COC, $/tonne CO_2		**90% Capture COC, $/tonne CO_2**	
Component	**Greenfield**	**Retrofit**	**Greenfield**	**Retrofit**
Capital	21.3	22.2	22.8	23.8
Fixed	14.4	15.0	15.6	16.2
Variable	5.1	5.3	5.3	5.5
Purchased Power and Fuel	16.5	16.5	16.2	16.2
Total COC	**57.3**	**58.9**	**59.9**	**61.7**

6.1.9 Plant Capacity Sensitivity Analysis

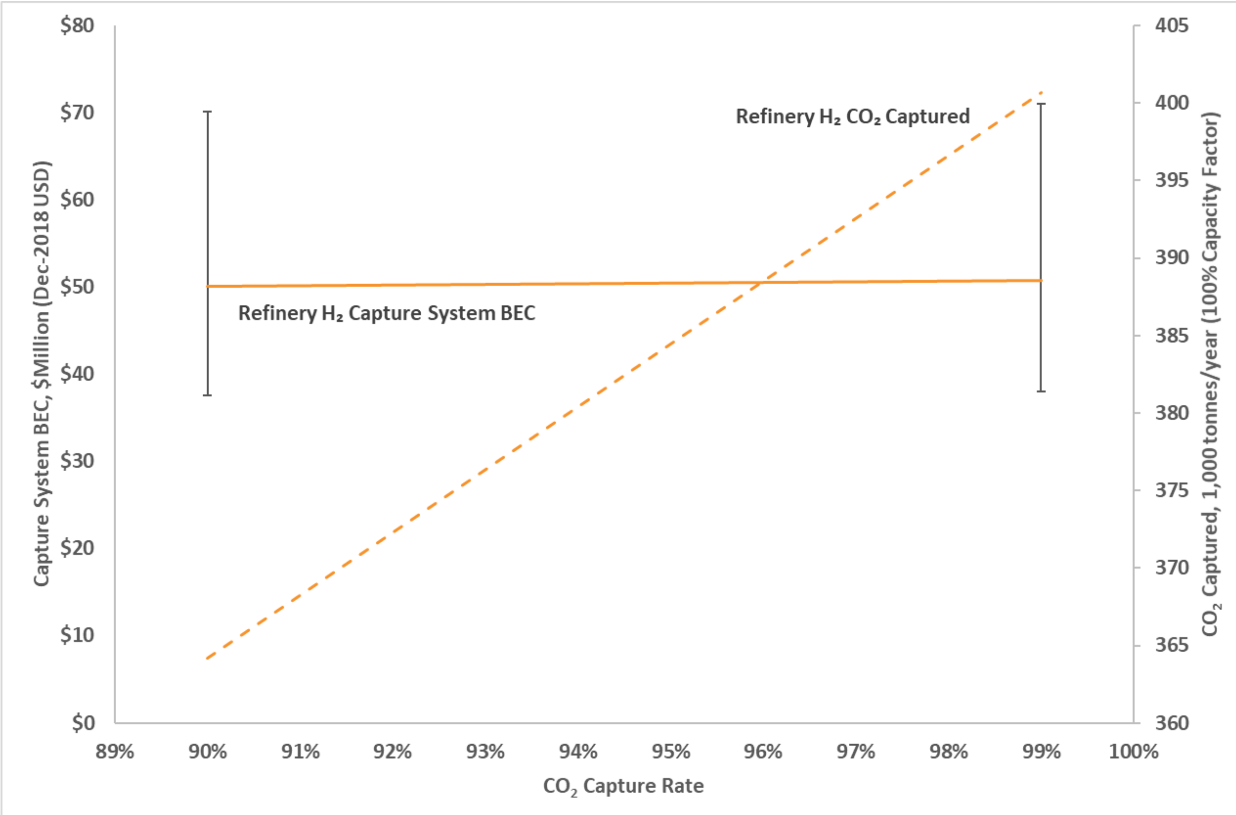
An analysis of the sensitivity of greenfield COC to refinery hydrogen plant capacity is shown in Exhibit 6-12. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 6-12. Refinery hydrogen plant capacity sensitivity



As the cost of capturing CO_2 is a normalized cost (i.e., \$/tonne CO_2), higher capture rates appear to cost less than lower capture rates. Comparing the true capital and O&M costs (i.e., not as normalized costs) shows that capital and O&M expenditures increase at higher capture rates. The cost of the capture system and associated consumables increases at a lesser rate than that of the amount of CO_2 captured (i.e., a 10 percent increase from 90 to 99 percent capture). The margin of error associated with the financial assumptions and cost scaling methodology employed in this study indicate that with increasing capture rate in the low purity cases, the COC is effectively the same. The reported minor increase in capital cost with increased capture rate (up to 99 percent for sources with CO_2 purity greater than 12 percent) has been validated by independent modeling performed by the carbon capture simulation initiative (CCSI) team at NETL and has been reported independently in literature. [4] Exhibit 6-13 shows the error in the calculated capture system BEC associated with the vendor's quoted uncertainty rate (-25/+40 percent) alongside the amount of CO_2 captured in the refinery H_2 case from 90 to 99 percent capture rate.

Exhibit 6-13. Refinery H_2 capture system BEC and amount of CO_2 captured versus capture rate



6.1.10 Refinery Hydrogen Conclusion

The low purity CO_2 stream produced in a refinery hydrogen plant results in a higher COC when compared to the high purity cases evaluated in this report, but the quantity of CO_2 to be captured from refinery H_2 production processes makes them attractive industrial processes for CCS as it would represent a large GHG reduction at a relatively low cost. A CO_2 capture and compression system for an 87,000 tonnes/year hydrogen plant was modeled to estimate the COC of capturing CO_2 from the SMR raw syngas. The results showed the COC of CO_2 to be $57.3/tonne CO_2 and $59.9/tonne CO_2 for a greenfield site with 99 and 90 percent capture, respectively. For a retrofit application, the COC is $58.9/tonne CO_2 and $61.7/tonne CO_2 for 99 and 90 percent capture, respectively. The small disparities between greenfield and retrofit cases are the result of unknown difficulties required for a retrofit installation versus a greenfield application, assuming adequate plot plan space for the retrofit case exists.

The plant size sensitivity showed that as plant size decreased from 170,000 to 50,000 tonnes/year, the COC increased by $17.5/tonne CO_2 and $18.9/tonne CO_2, for 99 and 90 percent capture, respectively. As the plant size decreases, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC. Though Shell indicates that for capture rates lower than 99 percent, post-combustion capture would be the optimal design, the pre-combustion capture system performance and cost was applied for the 90 percent capture case in this study for comparative purposes. As demonstrated by the resulting COCs and the sensitivity analysis, the normalized cost of 99 percent CO_2 capture is less than that of 90 percent capture.

6.2 Cement

Concrete is formed with a mixture of sand, gravel, water, and cement. Cement, when activated with water, is the binder that holds the concrete mixture together. In 2020, the U.S. cement industry produced approximately 89.3 M tonnes of Portland cement (PC) and masonry cement, with sales at approximately $12.7 billion (B). [43] In the same year, the U.S. apparent consumption of cement was 102 M tonnes of cement, meaning that imported cement filled the production gap. The United States Geological Survey (USGS) asserts in their 2021 *Minerals Commodity Summary* that U.S. cement production growth has been continuously constrained in recent years "by closed or idle plants, underutilized capacity at others, production disruptions from plant upgrades, and relatively inexpensive imports." Production trends for cement, as reported by the USGS, are shown in Exhibit 6-14. [43]

Exhibit 6-14. USGS cement production trends

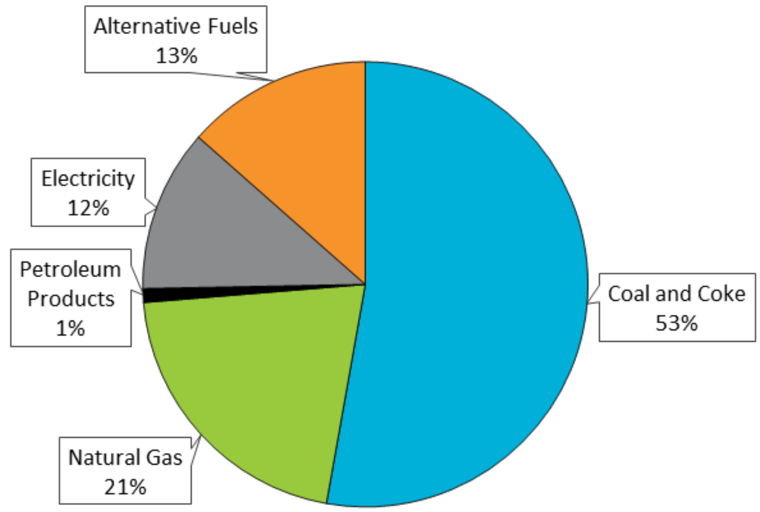
Year	2016	2017	2018	2019	2020[A]
PC Production, M tonnes	84.7	86.4	86.4	88.0[A]	89.0
Apparent PC Consumption, M tonnes	95.2	97.2	98.5	103.0[A]	102.0
U.S. Market Satisfied by U.S. Production, %	89.0	88.9	87.7	85.4	87.3
PC Price, $/tonne[B]	111	117	121	123[A]	124

[A] Estimated
[B] Average mill value

There are two processes for producing PC: wet kiln and dry kiln. The number of the more energy-intensive wet process kilns in the United States has declined by 96 percent from 234, in 1974, to 10, in 2019, while the number of dry process kilns was reduced from 198 to 110 over the same period. [44] Since 2008, approximately 85 percent of U.S. cement is produced using the dry-kiln process. [45]

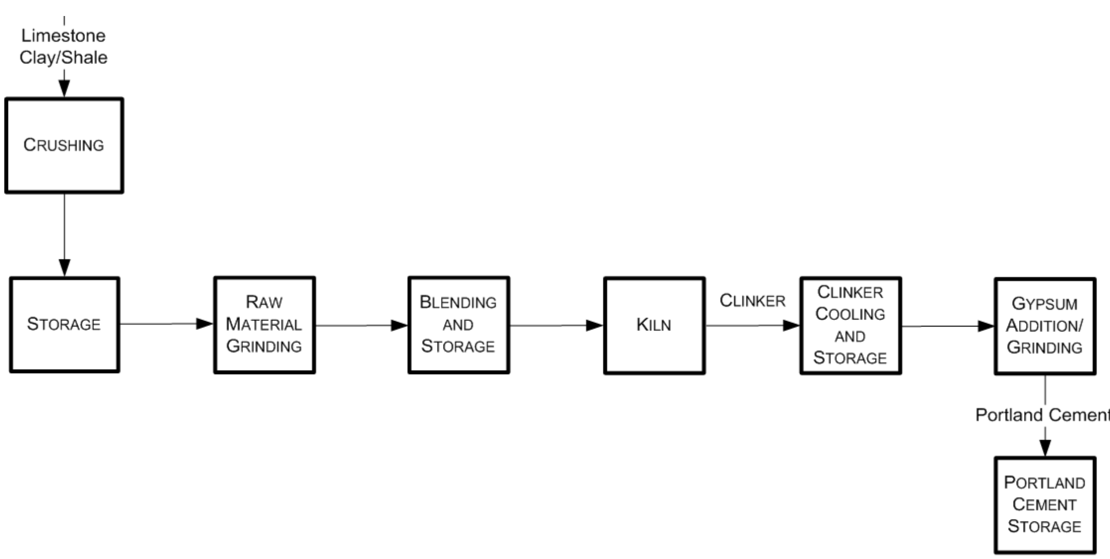
Both the dry- and wet-kiln processes utilize a multitude of different fuels to provide the heat necessary for drying, calcination, and sintering. Shown in Exhibit 6-15 is a breakdown of the fuel type consumed for 2019 as reported by the Portland Cement Association. [44] The values are given as a percentage of Btu consumed.

Exhibit 6-15. 2019 U.S. PC fuel consumption



Fuel burning to provide kiln heat is one of two CO_2 emissions sources, with the second resulting from the calcinations of calcium carbonate to form calcium oxide/calcium silicate species during the manufacturing process itself. PC is manufactured by crushing limestone and clay/shale raw materials to a powder, and then feeding in dry or slurry form to a kiln. Inside the kiln, the raw materials are heated to 2,600–3,000°F (1,430–1,650°C) and a chemical reaction takes place, fusing the raw materials into PC clinker, thus, generating CO_2. The clinker exits the kiln, is cooled, and is ground with gypsum to form PC. [46] Exhibit 6-16 shows the traditional PC production process, as adapted from Hassan (2005). [47]

Exhibit 6-16. PC production process



6.2.1 Size Range

In 2020, there were 96 U.S. cement plants, including both wet and dry processing kilns, in operation, with a total production capacity of 89.3 M tonnes/year. [43] The representative plant for this study is assumed to produce 1.3 M tonnes/year of PC and masonry cement. Of the 96 cement plants in 2020, 69 plants fall within the range of 0.5–1.5 M tonnes cement/year, and 31 plants fall within the range of 0.75–1.25 M tonnes cement/year, which adequately brackets the assumed plant size for this study.

Cement production creates on average 0.922 tonnes CO_2 per tonne cement [48]; however, this emissions factor may be broken down to two separate factors: an emissions factor for fuel burning and an emissions factor for calcium carbonate calcinations. The average fuel-burning emissions factor is 0.48 tonnes CO_2 per tonne cement, and the average calcination emissions factor is 0.44 tonne CO_2 per tonne cement. [48] For the reference plant capacity in this report, at 100 percent CF, these emissions factors give 631,737 tonnes CO_2/year from calcinations of raw materials, and 579,092 tonnes CO_2/year from fuel burning, totaling 1,210,829 tonnes CO_2/year from one point source. It is assumed that there is no air in-leakage in the kiln off-gas.

6.2.2 CO_2 Point Sources

A techno-economic analysis of CO_2 capture from a cement plant used the St. Mary's cement plant located in Ontario, Canada, as a reference plant. Specifics given for that plant as of 2004 are shown below, in Exhibit 6-17. [47]

Exhibit 6-17. St. Mary's cement plant characteristics

St. Mary's Cement Plant Characteristics	
Kiln Off-Gas Temperature (°F)	320
Kiln Off-Gas Pressure (psia)	14.7
Composition (mole %)	
H_2O	7.2
CO_2	22.4
N_2	68.1
O_2	2.3

For this study, the main point source of CO_2 available for capture is the kiln off-gas, and the concentrations given for the St. Mary's cement plant are assumed as representative. It is assumed that the kiln off-gas requires only CO_2 removal and compression and no other clean-up; however, it is possible that other treatment of the off-gas would be necessary prior to AGR.

A study done by the IEAGHG in 2009 estimated the cost per tonne of CO_2 avoided and the cost per tonne of cement product when adding CO_2 capture to a reference cement plant. [49] Their analysis points out that for post-combustion CO_2 capture to be implemented, there are several issues that must be addressed, as operational problems may arise from: the SO_2 concentration in the off-gas stream, which is dependent on the sulfide concentration in the raw meal; NO_2 concentration in the off-gas stream, which may cause solvent degradation; and dust present in the off-gas, which will reduce the efficiency of the post-combustion capture process. These issues are not considered in this study's base case; rather, the kiln off-gas is assumed suitable for post-combustion amine capture. However, a sensitivity case is evaluated to account for these issues with the addition of a selective catalytic reduction (SCR) unit to treat NO_x and flue gas desulfurization (FGD) to remove oxides of sulfur (SOx).

6.2.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the cement process for the purpose of this study:

- The representative cement plant has a production capacity of 1.3 M tonnes cement/year
- The CO_2 generated is 1,210,829 tonnes CO_2/year at 100 percent CF
- The CO_2 stream available for capture is 22.4 mole percent CO_2
- As a low purity source, separation, compression, and cooling are required. Separation is accomplished using a Cansolv AGR unit
- The temperature of the CO_2 available is 320°F
- The pressure of the CO_2 available is 14.7 psia
- CO_2 capture rates of 90 percent and 99 percent are evaluated
- The CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

6.2.4 CO_2 Capture System

The kiln off-gas stream CO_2 concentration is relatively low requires purification before compression to meet EOR pipeline standards. The purification system used is Shell's Cansolv post-combustion capture system discussed in Section 4.2.1. Steam for solvent regeneration is provided by the industrial boiler discussed in Section 4.3. One integrally geared centrifugal compression train as discussed in Section 4.1.2 is employed and costs for the compressor are scaled based on product CO_2 flow.

6.2.5 BFD, Stream Table, and Performance Summary

As shown in Exhibit 6-18, the kiln off-gas is sent to the Cansolv separation unit. Water and solids recovered from the AGR process are sent to waste treatment. The CO_2 stream is then compressed with interstage cooling and then after-cooled before reaching the EOR pipeline. Exhibit 6-18 shows the BFD for this process, and Exhibit 6-19 and Exhibit 6-20 show the stream table for this process with 99 percent and 90 percent capture, respectively.

Exhibit 6-18. Cement CO_2 capture BFD

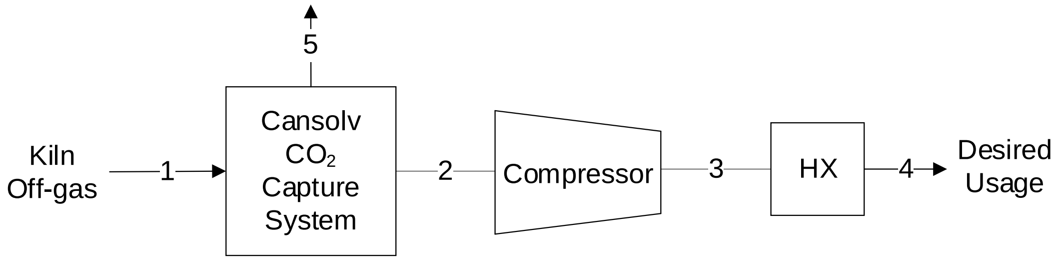


Exhibit 6-19. Cement stream table for 99 percent capture

	1	2	3	4	5
V-L Mole Fraction					
AR	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2240	0.9885	0.9995	0.9995	0.0032
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0720	0.0115	0.0005	0.0005	0.0205
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6810	0.0000	0.0000	0.0000	0.9444
O_2	0.0230	0.0000	0.0000	0.0000	0.0319
Total	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	14,012	3,142	3,107	3,107	10,104
V-L Flowrate (kg/hr)	433,946	137,356	136,707	136,707	282,775
Temperature (°C)	160	31	80	30	38
Pressure (MPa, abs)	0.10	0.2	15.3	15.3	0.1
Steam Table Enthalpy (kJ/kg)[A]	3,442	8,791	8,758	8,755	274
Aspen Plus Enthalpy (kJ/kg)[B]	-3,269	-8,959	-9,042	-9,195	-209.2
Density (kg/m^3)	0.9	3.5	432.5	630.1	1.1
V-L Molecular Weight	31.0	43.7	44.0	44.0	28.0
V-L Flowrate (lb_{mol}/hr)	30,891	6,928	6,850	6,850	22,276
V-L Flowrate (lb/hr)	956,688	302,818	301,387	301,387	623,412
Temperature (°F)	320	88	177	86	100
Pressure (psia)	14.7	28.9	2,216.9	2,214.7	14.8
Steam Table Enthalpy (Btu/lb)[A]	1,480	3,780	3,765	3,764	118
Aspen Plus Enthalpy (Btu/lb)[B]	-1,406	-3,852	-3,887	-3,953	-89.9
Density (lb/ft^3)	0.054	0.217	27.0	39.3	0.069

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 6-20. Cement stream table for 90 percent capture

	1	2	3	4	5
V-L Mole Fraction					
AR	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2240	0.9887	0.9995	0.9995	0.0302
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0720	0.0113	0.0005	0.0005	0.0207
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6810	0.0000	0.0000	0.0000	0.9181
O_2	0.0230	0.0000	0.0000	0.0000	0.0310
Total	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	14,012	2,857	2,826	2,826	10,393
V-L Flowrate (kg/hr)	433,946	124,914	124,334	124,334	295,281
Temperature (°C)	160	31	80	30	38
Pressure (MPa, abs)	0.10	0.2	15.3	15.3	0.1
Steam Table Enthalpy (kJ/kg)[A]	3,442	8,791	8,758	8,755	631
Aspen Plus Enthalpy (kJ/kg)[B]	-3,269	-8,959	-9,042	-9,195	-580.6
Density (kg/m^3)	0.9	3.5	432.5	630.1	1.1
V-L Molecular Weight	31.0	43.7	44.0	44.0	28.4
V-L Flowrate (lb_{mol}/hr)	30,891	6,300	6,230	6,230	22,912
V-L Flowrate (lb/hr)	956,688	275,388	274,110	274,110	650,984
Temperature (°F)	320	88	177	86	100
Pressure (psia)	14.7	28.9	2,216.9	2,214.7	14.8
Steam Table Enthalpy (Btu/lb)[A]	1,480	3,779	3,765	3,764	271
Aspen Plus Enthalpy (Btu/lb)[B]	-1,406	-3,852	-3,887	-3,953	-249.6
Density (lb/ft^3)	0.054	0.217	27.0	39.3	0.070

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance summary for both 90 and 99 percent capture cases is provided in Exhibit 6-21.

Exhibit 6-21. Performance summary

Performance Summary		
Item	**1.3 M tonnes cement/year with 90 percent CO_2 capture (kWe)**	**1.3 M tonnes cement/year with 99 percent CO_2 capture (kWe)**
CO_2 Capture Auxiliaries	3,100	3,500
Steam Boiler Auxiliaries	330	370
CO_2 Compressor	9,570	10,460
Circulating Water Pumps	980	1,040
Cooling Tower Fans	500	540
Total Auxiliary Load	**14,480**	**15,910**

6.2.6 Capture Integration

The cooling water system in this study is a stand-alone unit; however, there is potential to integrate make-up water to feed or partially feed the cooling water system, thereby reducing the unit's size. This would be evaluated on case-by-case basis depending on the size of the plant, its layout, and size of the plant's current cooling system. This evaluation is outside of the scope of this report.

6.2.7 Power Source

The compressor power consumption for the 90 and 99 percent capture cases are 9.57 MW and 10.46 MW, respectively. Power consumption estimates for the cooling water system in each case were scaled as described in Section 4.4. The total power requirements were calculated to be 14.48 MW and 15.91 MW for the 90 and 99 percent capture rates, respectively, which includes all power required by the compression train, cooling system, and Cansolv capture unit. Purchased power cost is estimated at a rate of $60/MWh as discussed in Section 3.1.2.4. To satisfy the steam requirements of the AGR system, an industrial boiler was modeled, and fuel consumption costs were estimated at a rate of $4.42/MMBtu as discussed in Section 3.1.2.4.

6.2.8 Economic Analysis Results

The economic results of CO_2 capture application in a cement plant are presented in this section. Owner's costs (Exhibit 6-22), capital costs (Exhibit 6-23 and Exhibit 6-24), and O&M costs are calculated as discussed in Section 3.1. Retrofit costs were determined by applying a retrofit factor to TPC as discussed in Section 3.3. The greenfield TOC for the cement case at 99 percent capture is $414.0 M, while for 90 percent capture, a greenfield TOC of $386.0 M is estimated. The corresponding greenfield COC for the 99 percent and 90 percent capture cases are $60.8/tonne CO_2 and $62.7 /tonne CO_2, respectively. The COC is $62.4/tonne CO_2 and $64.3/tonne CO_2 in retrofit applications for 99 percent and 90 percent capture, respectively.

Exhibit 6-22. Owners' costs for cement cases

Description	$/1,000	$/tonnes/yr ($CO_2$)	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs	**99% Capture**		**90% Capture**	
6 Months All Labor	$1,986	$2	$1,922	$2
1-Month Maintenance Materials	$319	$0	$304	$0
1-Month Non-Fuel Consumables	$257	$0	$240	$0
1-Month Waste Disposal	$11	$0	$11	$0
25% of 1-Month Fuel Cost at 100% CF	$391	$0	$355	$0
2% of TPC	$6,779	$6	$6,457	$6
Total	**$9,742**	**$8**	**$9,289**	**$9**
Inventory Capital	**99% Capture**		**90% Capture**	
60-day supply of fuel and consumables at 100% CF	$3,550	$3	$3,239	$3
0.5% of TPC (spare parts)	$1,695	$1	$1,614	$1
Total	**$5,245**	**$4**	**$4,853**	**$4**
Other Costs	**99% Capture**		**90% Capture**	
Initial Cost for Catalyst and Chemicals	$0	$0	$0	$0
Land	$30	$0	$30	$0
Other Owner's Costs	$50,842	$42	$48,431	$44
Financing Costs	$9,152	$8	$8,718	$8
TOC	**$413,960**	**$346**	**$394,192**	**$362**
TASC Multiplier (Cement, 33 year)	1.054		1.054	
TASC	**$436,252**	**$364**	**$415,418**	**$381**

Exhibit 6-23. Capital costs for cement greenfield site with 99 percent capture

	Case:	Cement					Estimate Type:			Conceptual	
	Representative Plant Size:	1.3 M tonnes cement/year					Cost Base:			Dec 2018	
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	$/tonnes/yr ($CO_2$)
3					Feedwater & Miscellaneous BOP Systems						
3.1	Feedwater System	$658	$1,127	$564	$0	$2,349	$411	$0	$552	$3,311	$3
3.2	Water Makeup & Pretreating	$1,633	$163	$925	$0	$2,722	$476	$0	$640	$3,837	$3
3.3	Other Feedwater Subsystems	$305	$100	$95	$0	$500	$87	$0	$117	$704	$1
3.4	Industrial Boiler Package w/Deaerator	$4,061	$0	$1,181	$0	$5,242	$917	$0	$1,232	$7,391	$6
3.5	Other Boiler Plant Systems	$73	$27	$67	$0	$167	$29	$0	$39	$235	$0
3.6	NG Pipeline and Start-Up System	$654	$28	$21	$0	$703	$123	$0	$165	$992	$1
3.7	Waste Water Treatment Equipment	$3,003	$0	$1,840	$0	$4,843	$848	$0	$1,138	$6,829	$6
3.9	Miscellaneous Plant Equipment	$98	$13	$50	$0	$161	$28	$0	$38	$227	$0
	Subtotal	**$10,485**	**$1,458**	**$4,743**	**$0**	**$16,686**	**$2,920**	**$0**	**$3,921**	**$23,527**	**$20**
5					Flue Gas Cleanup						
5.1	Cansolv CO_2 Removal System	$58,671	$25,377	$53,292	$0	$137,340	$24,034	$23,348	$36,944	$221,667	$185
5.4	CO_2 Compression & Drying	$17,147	$2,572	$5,733	$0	$25,452	$4,454	$0	$5,981	$35,887	$30
5.5	CO_2 Compressor Aftercooler	$137	$22	$59	$0	$218	$38	$0	$51	$307	$0
5.12	Gas Cleanup Foundations	$0	$65	$57	$0	$122	$21	$0	$29	$172	$0
	Subtotal	**$75,955**	**$28,036**	**$59,141**	**$0**	**$163,132**	**$28,548**	**$23,348**	**$43,006**	**$258,033**	**$215**
7					Ductwork & Stack						
7.3	Ductwork	$0	$1,608	$1,117	$0	$2,725	$477	$0	$640	$3,842	$3
7.4	Stack	$7,699	$0	$4,474	$0	$12,174	$2,130	$0	$2,861	$17,165	$14
7.5	Duct & Stack Foundations	$0	$172	$204	$0	$376	$66	$0	$88	$530	$0
	Subtotal	**$7,699**	**$1,779**	**$5,795**	**$0**	**$15,274**	**$2,673**	**$0**	**$3,589**	**$21,537**	**$18**
9					Cooling Water System						
9.1	Cooling Towers	$1,426	$0	$441	$0	$1,867	$327	$0	$439	$2,632	$2
9.2	Circulating Water Pumps	$147	$0	$10	$0	$157	$27	$0	$37	$221	$0
9.3	Circulating Water System Aux.	$1,895	$0	$251	$0	$2,146	$376	$0	$504	$3,025	$3
9.4	Circulating Water Piping	$0	$876	$794	$0	$1,670	$292	$0	$392	$2,355	$2
9.5	Make-up Water System	$207	$0	$265	$0	$472	$83	$0	$111	$666	$1
9.6	Component Cooling Water System	$137	$0	$105	$0	$241	$42	$0	$57	$340	$0
9.7	Circulating Water System Foundations	$0	$97	$161	$0	$258	$45	$0	$61	$363	$0
	Subtotal	**$3,811**	**$973**	**$2,027**	**$0**	**$6,811**	**$1,192**	**$0**	**$1,600**	**$9,603**	**$8**
11					Accessory Electric Plant						
11.2	Station Service Equipment	$2,650	$0	$227	$0	$2,878	$504	$0	$676	$4,058	$3
11.3	Switchgear & Motor Control	$4,114	$0	$714	$0	$4,828	$845	$0	$1,135	$6,808	$6
11.4	Conduit & Cable Tray	$0	$535	$1,541	$0	$2,076	$363	$0	$488	$2,927	$2

Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/tonnes/yr ($CO_2$)
	Case:	Cement						Estimate Type:		Conceptual	
	Representative Plant Size:	1.3 M tonnes cement/year						Cost Base:		Dec 2018	
11.5	Wire & Cable	$0	$1,416	$2,532	$0	$3,948	$691	$0	$928	$5,567	$5
	Subtotal	**$6,765**	**$1,951**	**$5,014**	**$0**	**$13,730**	**$2,403**	**$0**	**$3,227**	**$19,360**	**$16**
12	Instrumentation & Control										
12.8	Instrument Wiring & Tubing	$402	$322	$1,286	$0	$2,010	$352	$0	$472	$2,834	$2
12.9	Other I&C Equipment	$494	$0	$1,144	$0	$1,638	$287	$0	$385	$2,310	$2
	Subtotal	**$896**	**$322**	**$2,431**	**$0**	**$3,648**	**$638**	**$0**	**$857**	**$5,144**	**$4**
13	Improvements to Site										
13.1	Site Preparation	$0	$27	$537	$0	$563	$99	$0	$132	$794	$1
13.2	Site Improvements	$0	$125	$166	$0	$291	$51	$0	$68	$410	$0
13.3	Site Facilities	$143	$0	$150	$0	$293	$51	$0	$69	$414	$0
	Subtotal	**$143**	**$152**	**$853**	**$0**	**$1,148**	**$201**	**$0**	**$270**	**$1,618**	**$1**
14	Buildings & Structures										
14.5	Circulation Water Pumphouse	$0	$50	$40	$0	$90	$16	$0	$21	$127	$0
	Subtotal	**$0**	**$50**	**$40**	**$0**	**$90**	**$16**	**$0**	**$21**	**$127**	**$0**
	Total	**$105,754**	**$34,722**	**$80,043**	**$0**	**$220,519**	**$38,591**	**$23,348**	**$56,491**	**$338,949**	**$283**

Exhibit 6-24. Capital costs for cement greenfield site with 90 percent capture

Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/tonnes/yr ($CO_2$)
	Case:	Cement						Estimate Type:		Conceptual	
	Representative Plant Size:	1.3 M tonnes cement/year						Cost Base:		Dec 2018	
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$616	$1,056	$528	$0	$2,199	$385	$0	$517	$3,101	$3
3.2	Water Makeup & Pretreating	$1,543	$154	$874	$0	$2,571	$450	$0	$604	$3,626	$3
3.3	Other Feedwater Subsystems	$280	$92	$87	$0	$459	$80	$0	$108	$647	$1
3.4	Industrial Boiler Package w/Deaerator	$3,731	$0	$1,085	$0	$4,816	$843	$0	$1,132	$6,790	$6
3.5	Other Boiler Plant Systems	$67	$25	$61	$0	$153	$27	$0	$36	$216	$0
3.6	NG Pipeline and Start-Up System	$624	$27	$20	$0	$671	$117	$0	$158	$946	$1
3.7	Waste Water Treatment Equipment	$2,872	$0	$1,760	$0	$4,632	$811	$0	$1,088	$6,531	$6
3.9	Miscellaneous Plant Equipment	$96	$13	$49	$0	$157	$27	$0	$37	$221	$0
	Subtotal	**$9,829**	**$1,366**	**$4,464**	**$0**	**$15,659**	**$2,740**	**$0**	**$3,680**	**$22,079**	**$20**
5	Flue Gas Cleanup										
5.1	Cansolv CO_2 Removal System	$55,656	$24,073	$50,554	$0	$130,284	$22,800	$22,148	$35,046	$210,278	$193
5.4	CO_2 Compression & Drying	$16,242	$2,436	$5,430	$0	$24,108	$4,219	$0	$5,665	$33,993	$31
5.5	CO_2 Compressor Aftercooler	$127	$20	$54	$0	$201	$35	$0	$47	$284	$0
5.12	Gas Cleanup Foundations	$0	$65	$57	$0	$122	$21	$0	$29	$172	$0

	Case:	Cement					Estimate Type:			Conceptual	
	Representative Plant Size:	1.3 M tonnes cement/year					Cost Base:			Dec 2018	
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	$/tonnes/yr ($CO_2$)
	Subtotal	**$72,025**	**$26,595**	**$56,096**	**$0**	**$154,716**	**$27,075**	**$22,148**	**$40,788**	**$244,727**	**$225**
	7					**Ductwork & Stack**					
7.3	Ductwork	$0	$1,608	$1,117	$0	$2,725	$477	$0	$640	$3,842	$4
7.4	Stack	$7,712	$0	$4,482	$0	$12,194	$2,134	$0	$2,866	$17,194	$16
7.5	Duct & Stack Foundations	$0	$171	$203	$0	$374	$65	$0	$88	$527	$0
	Subtotal	**$7,712**	**$1,778**	**$5,802**	**$0**	**$15,293**	**$2,676**	**$0**	**$3,594**	**$21,563**	**$20**
	9					**Cooling Water System**					
9.1	Cooling Towers	$1,343	$0	$415	$0	$1,759	$308	$0	$413	$2,480	$2
9.2	Circulating Water Pumps	$137	$0	$10	$0	$147	$26	$0	$35	$207	$0
9.3	Circulating Water System Aux.	$1,805	$0	$239	$0	$2,043	$358	$0	$480	$2,881	$3
9.4	Circulating Water Piping	$0	$834	$756	$0	$1,590	$278	$0	$374	$2,242	$2
9.5	Make-up Water System	$199	$0	$255	$0	$454	$80	$0	$107	$641	$1
9.6	Component Cooling Water System	$130	$0	$100	$0	$230	$40	$0	$54	$324	$0
9.7	Circulating Water System Foundations	$0	$93	$154	$0	$246	$43	$0	$58	$347	$0
	Subtotal	**$3,614**	**$927**	**$1,929**	**$0**	**$6,470**	**$1,132**	**$0**	**$1,520**	**$9,122**	**$8**
	11					**Accessory Electric Plant**					
11.2	Station Service Equipment	$2,545	$0	$218	$0	$2,763	$484	$0	$649	$3,896	$4
11.3	Switchgear & Motor Control	$3,951	$0	$685	$0	$4,636	$811	$0	$1,090	$6,537	$6
11.4	Conduit & Cable Tray	$0	$514	$1,480	$0	$1,994	$349	$0	$469	$2,811	$3
11.5	Wire & Cable	$0	$1,360	$2,431	$0	$3,791	$663	$0	$891	$5,346	$5
	Subtotal	**$6,496**	**$1,874**	**$4,815**	**$0**	**$13,185**	**$2,307**	**$0**	**$3,098**	**$18,590**	**$17**
	12					**Instrumentation & Control**					
12.8	Instrument Wiring & Tubing	$397	$318	$1,271	$0	$1,985	$347	$0	$467	$2,799	$3
12.9	Other I&C Equipment	$488	$0	$1,130	$0	$1,618	$283	$0	$380	$2,282	$2
	Subtotal	**$885**	**$318**	**$2,401**	**$0**	**$3,604**	**$631**	**$0**	**$847**	**$5,081**	**$5**
	13					**Improvements to Site**					
13.1	Site Preparation	$0	$26	$527	$0	$553	$97	$0	$130	$780	$1
13.2	Site Improvements	$0	$123	$163	$0	$285	$50	$0	$67	$402	$0
13.3	Site Facilities	$140	$0	$147	$0	$288	$50	$0	$68	$406	$0
	Subtotal	**$140**	**$149**	**$837**	**$0**	**$1,126**	**$197**	**$0**	**$265**	**$1,588**	**$1**
	14					**Buildings & Structures**					
14.5	Circulation Water Pumphouse	$0	$48	$38	$0	$86	$15	$0	$20	$122	$0
	Subtotal	**$0**	**$48**	**$38**	**$0**	**$86**	**$15**	**$0**	**$20**	**$122**	**$0**
	Total	**$100,701**	**$33,054**	**$76,381**	**$0**	**$210,137**	**$36,774**	**$22,148**	**$53,812**	**$322,871**	**$296**

The initial and annual O&M costs for a greenfield site were calculated and are shown in Exhibit 6-25 and Exhibit 6-26 for 99 percent and 90 percent capture, respectively, while Exhibit 6-27 shows the COC for greenfield and retrofit sites for the representative cement plants at both capture rates.

Exhibit 6-25. Initial and annual O&M costs for cement greenfield site with 99 percent capture

Case:	**Cement**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**1.3 M tonnes cement/year**				**Capacity Factor (%):**	**85**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		2.3
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**2.3**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$1,008,407	$0.84
Maintenance Labor:					$2,169,273	$1.81
Administrative & Support Labor:					$794,420	$0.66
Property Taxes and Insurance:					$6,778,980	$5.66
Total:					**$10,751,081**	**$8.98**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$3,253,910	$3.20
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	775	$1.90	$0	$457,112	$0.45
Makeup and Waste Water Treatment Chemicals (ton):	0	2.6	$550.00	$0	$440,049	$0.43
CO_2 Capture System Chemicals[A]:			Proprietary		$1,215,644	$1.19
Triethylene Glycol (gal):	w/equip.	240	$6.80	$0	$507,172	$0.50
Subtotal:				**$0**	**$2,619,977**	**$2.57**
Waste Disposal						
Triethylene Glycol (gal):		240	$0.35	$0	$26,104	$0.03
Thermal Reclaimer Unit Waste (ton)		0.69	$38.00	$0	$8,077	$0.01
Prescrubber Blowdown Waste (ton)		6.7	$38.00	$0	$78,627	$0.08
Subtotal:				**$0**	**$112,809**	**$0.11**
Variable Operating Costs Total:				**$0**	**$5,986,696**	**$5.88**
Fuel Cost						
Natural Gas (MMBtu)	0	11,625	$4.42	$0	$15,941,580	$15.66
Total:				**$0**	**$15,941,580**	**$15.66**

[A] CO_2 capture system chemicals includes NaOH and Cansolv solvent

Exhibit 6-26. Initial and annual O&M costs for cement greenfield site with 90 percent capture

Case:	Cement				Cost Base:	Dec 2018
Representative Plant Size:	1.3 M tonnes cement/year				Capacity Factor (%):	85
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		2.3
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**2.3**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$1,008,407	$0.93
Maintenance Labor:					$2,066,376	$1.90
Administrative & Support Labor:					$768,696	$0.71
Property Taxes and Insurance:					$6,457,426	$5.93
Total:					**$10,300,905**	**$9.46**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$3,099,564	$3.35
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	717	$1.90	$0	$422,931	$0.46
Makeup and Waste Water Treatment Chemicals (ton):	0	2.4	$550.00	$0	$410,207	$0.44
CO_2 Capture System Chemicals[A]:			Proprietary		$1,152,997	$1.25
Triethylene Glycol (gal):	w/equip.	219	$6.80	$0	$461,270	$0.50
Subtotal:				**$0**	**$2,447,404**	**$2.64**
Waste Disposal						
Triethylene Glycol (gal):		219	$0.35	$0	$23,742	$0.03
Thermal Reclaimer Unit Waste (ton)		0.65	$38.00	$0	$7,713	$0.01
Prescrubber Blowdown Waste (ton)		6.7	$38.00	$0	$78,627	$0.08
Subtotal:				**$0**	**$110,082**	**$0.12**
Variable Operating Costs Total:				**$0**	**$5,657,051**	**$6.11**
Fuel Cost						
Natural Gas (MMBtu)	0	10,569	$4.42	$0	$14,493,467	$15.66
Total:				**$0**	**$14,493,467**	**$15.66**

[A]CO_2 capture system chemicals includes NaOH and Cansolv solvent

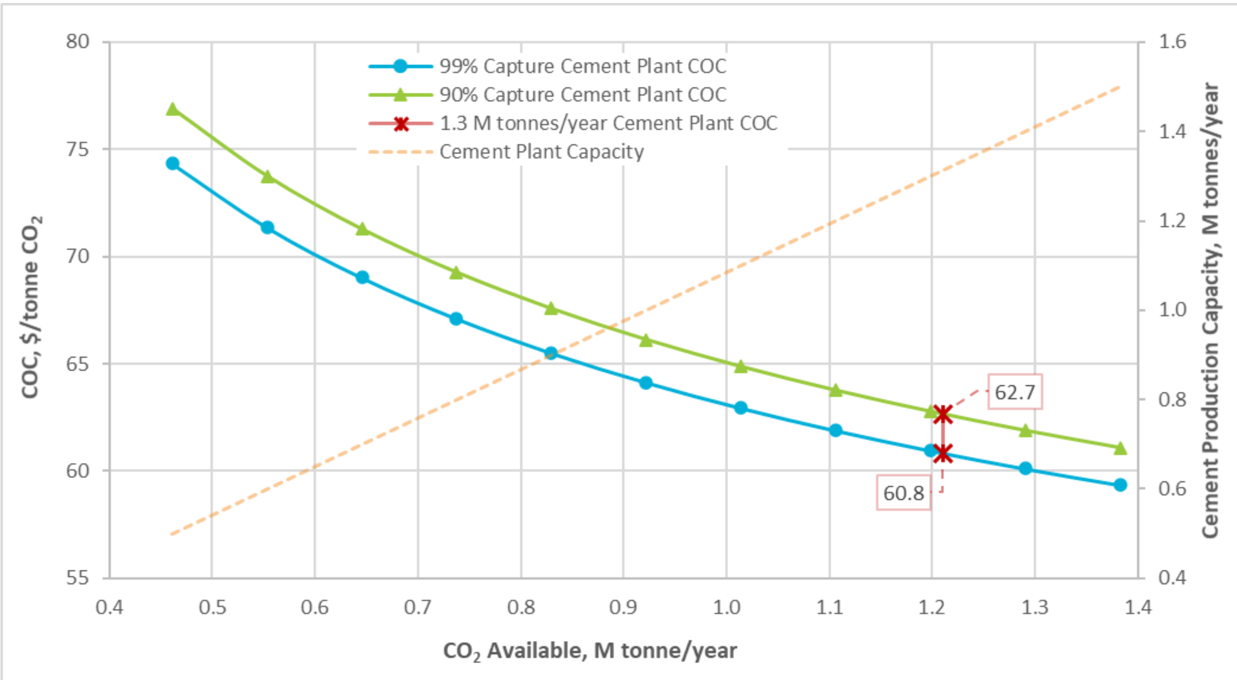
Exhibit 6-27. COC for 1.3 M tonnes/year cement cases

	99% Capture COC, $/tonne CO_2		**90% Capture COC, $/tonne CO_2**	
Component	**Greenfield**	**Retrofit**	**Greenfield**	**Retrofit**
Capital	21.8	22.6	22.8	23.7
Fixed	10.6	11.0	11.1	11.6
Variable	5.9	6.0	6.1	6.3
Purchased Power and Fuel	22.6	22.6	22.6	22.6
Total COC	**60.8**	**62.4**	**62.7**	**64.3**

6.2.9 Plant Capacity Sensitivity Analysis

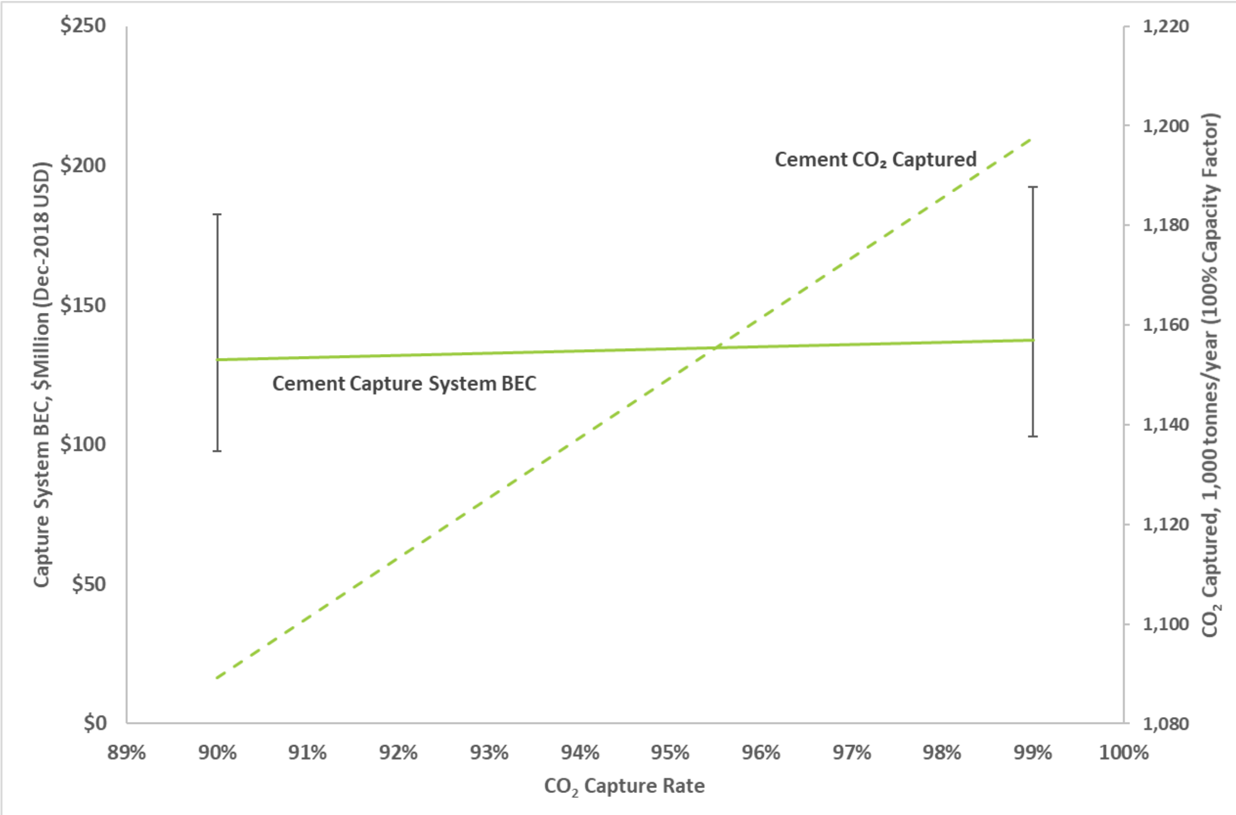
An analysis of the sensitivity of greenfield COC to cement plant capacity is shown in Exhibit 6-28. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 6-28. Cement plant capacity sensitivity



As the cost of capturing CO_2 is a normalized cost (i.e., \$/tonne CO_2), higher capture rates appear to cost less than lower capture rates. Comparing the true capital and O&M costs (i.e., not as normalized costs) shows that capital and O&M expenditures increase at higher capture rates. The cost of the capture system and associated consumables increases at a lesser rate than that of the amount of CO_2 captured (i.e., a 10 percent increase from 90 to 99 percent capture). The margin of error associated with the financial assumptions and cost scaling methodology employed in this study indicate that with increasing capture rate in the low purity cases, the COC is effectively the same. The reported minor increase in capital cost with increased capture rate (up to 99 percent for sources with CO_2 purity greater than 12 percent) has been validated by independent modeling performed by the CCSI team at NETL and has been reported independently in literature. [4] Exhibit 6-29 shows the error in the calculated capture system BEC associated with the vendor's quoted uncertainty rate (-25/+40 percent) alongside the amount of CO_2 captured in the cement case from 90 to 99 percent capture rate.

Exhibit 6-29. Cement capture system BEC and amount of CO_2 captured versus capture rate



6.2.10 FGD + SCR Sensitivity Case

As stated previously, a cement plant's kiln off-gas may require additional treatment prior to purification to maximize the efficiency of the amine-based CO_2 removal system and prevent solvent degradation. Definitive concentrations for cement kiln off-gas SOx and oxides of nitrogen (NOx) are not available, as SOx is highly dependent upon the sulfide concentration of the raw meal used, and NOx content varies widely. Therefore, to account for the addition of SCR and FGD units in terms of capital cost, as well as power and chemical requirements/costs, these values were scaled from BBR4 Case B12B [5] based on quantity of gas treated. The FGD employed in the reference case is a wet FGD; however, if SOx concentrations were low enough, a lower cost option, such as a dry FGD could also be used, which would reduce cost compared to the wet FGD estimated in this report.

The economic results of this sensitivity case are presented in Exhibit 6-30 and Exhibit 6-31 for the 99 and 90 percent capture cases, respectively. The addition of SCR and FGD increases the TPC over the base case greenfield cost by approximately $124.5 M. Most of this additional capital is attributed to the FGD absorber vessels and accessories, which account for $110.7 M of the TPC increase.

Fixed O&M and maintenance materials costs also increase, as some are calculated based on TPC. Consumables costs also increase by $2.3 M, due to the requirement of limestone for the FGD, as well as 19 weight percent ammonia for SCR injection. The initial SCR catalyst is assumed to be included with equipment purchase, but catalyst makeup cost is calculated on a 3-year replacement cycle. The auxiliary requirements for the FGD and SCR are scaled linearly from the BBR4 Case B12B, adding 672 kW to the auxiliary load requirements for capture integration in the representative cement plant. O&M costs for each cement sensitivity case are shown in Exhibit 6-33 and Exhibit 6-34 for 99 and 90 percent capture cases, respectively, while owner's cost summaries for both cases are shown in Exhibit 6-32.

Exhibit 6-30. Capital costs for cement greenfield site with FGD and SCR and 99 percent CO_2 capture

Case:		Cement with FGD and SCR					Estimate Type:			Conceptual	
Representative Plant Size:		1.3 M tonnes/year					Cost Base:			Dec 2018	
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	$/tonnes/yr ($CO_2$)
3		Feedwater & Miscellaneous BOP Systems									
3.1	Feedwater System	$658	$1,127	$564	$0	$2,349	$411	$0	$552	$3,311	$3
3.2	Water Makeup & Pretreating	$1,633	$163	$925	$0	$2,722	$476	$0	$640	$3,837	$3
3.3	Other Feedwater Subsystems	$305	$100	$95	$0	$500	$87	$0	$117	$704	$1
3.4	Industrial Boiler Package w/Deaerator	$4,061	$0	$1,181	$0	$5,242	$917	$0	$1,232	$7,391	$6
3.5	Other Boiler Plant Systems	$73	$27	$67	$0	$167	$29	$0	$39	$235	$0
3.6	NG Pipeline and Start-Up System	$654	$28	$21	$0	$703	$123	$0	$165	$992	$1
3.7	Waste Water Treatment Equipment	$3,003	$0	$1,840	$0	$4,843	$848	$0	$1,138	$6,829	$6
3.9	Miscellaneous Plant Equipment	$98	$13	$50	$0	$161	$28	$0	$38	$227	$0
	Subtotal	**$10,485**	**$1,458**	**$4,743**	**$0**	**$16,686**	**$2,920**	**$0**	**$3,921**	**$23,527**	**$20**
4		Cement Kiln Accessories									
4.1	Selective Catalytic Reduction System	$5,660	$0	$3,225	$0	$8,885	$1,555	$0	$2,088	$12,528	$10
	Subtotal	**$5,660**	**$0**	**$3,225**	**$0**	**$8,885**	**$1,555**	**$0**	**$2,088**	**$12,528**	**$10**
5		Flue Gas Cleanup									
5.1	Cansolv CO_2 Removal System	$58,671	$25,377	$53,292	$0	$137,340	$24,034	$23,348	$36,944	$221,667	$185
5.2	FGD Absorber Vessels & Accessories	$64,703	$0	$13,834	$0	$78,537	$13,744	$0	$18,456	$110,737	$92
5.3	Other FGD	$290	$0	$327	$0	$617	$108	$0	$145	$870	$1
5.4	CO_2 Compression & Drying	$17,147	$2,572	$5,733	$0	$25,452	$4,454	$0	$5,981	$35,887	$30
5.5	CO_2 Compressor Aftercooler	$137	$22	$59	$0	$218	$38	$0	$51	$307	$0
5.12	Gas Cleanup Foundations	$0	$65	$57	$0	$122	$21	$0	$29	$172	$0
	Subtotal	**$140,948**	**$28,036**	**$73,302**	**$0**	**$242,286**	**$42,400**	**$23,348**	**$61,607**	**$369,640**	**$309**
7		Ductwork & Stack									
7.3	Ductwork	$0	$1,608	$1,117	$0	$2,725	$477	$0	$640	$3,842	$3
7.4	Stack	$7,699	$0	$4,474	$0	$12,174	$2,130	$0	$2,861	$17,165	$14
7.5	Duct & Stack Foundations	$0	$172	$204	$0	$376	$66	$0	$88	$530	$0
	Subtotal	**$7,699**	**$1,779**	**$5,795**	**$0**	**$15,274**	**$2,673**	**$0**	**$3,589**	**$21,537**	**$18**
9		Cooling Water System									
9.1	Cooling Towers	$1,426	$0	$441	$0	$1,867	$327	$0	$439	$2,632	$2
9.2	Circulating Water Pumps	$147	$0	$10	$0	$157	$27	$0	$37	$221	$0
9.3	Circulating Water System Aux.	$1,895	$0	$251	$0	$2,146	$376	$0	$504	$3,025	$3
9.4	Circulating Water Piping	$0	$876	$794	$0	$1,670	$292	$0	$392	$2,355	$2
9.5	Make-up Water System	$207	$0	$265	$0	$472	$83	$0	$111	$666	$1

Case:	Cement with FGD and SCR						Estimate Type:		Conceptual	
Representative Plant Size:	1.3 M tonnes/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
9.6 Component Cooling Water System	$137	$0	$105	$0	$241	$42	$0	$57	$340	$0
9.7 Circulating Water System Foundations	$0	$97	$161	$0	$258	$45	$0	$61	$363	$0
Subtotal	**$3,811**	**$973**	**$2,027**	**$0**	**$6,811**	**$1,192**	**$0**	**$1,600**	**$9,603**	**$8**
11					**Accessory Electric Plant**					
11.2 Station Service Equipment	$2,698	$0	$231	$0	$2,929	$513	$0	$688	$4,130	$3
11.3 Switchgear & Motor Control	$4,188	$0	$727	$0	$4,915	$860	$0	$1,155	$6,930	$6
11.4 Conduit & Cable Tray	$0	$544	$1,569	$0	$2,113	$370	$0	$497	$2,980	$2
11.5 Wire & Cable	$0	$1,442	$2,577	$0	$4,019	$703	$0	$945	$5,667	$5
Subtotal	**$6,886**	**$1,986**	**$5,104**	**$0**	**$13,977**	**$2,446**	**$0**	**$3,285**	**$19,707**	**$16**
12					**Instrumentation & Control**					
12.8 Instrument Wiring & Tubing	$404	$323	$1,293	$0	$2,021	$354	$0	$475	$2,849	$2
12.9 Other I&C Equipment	$497	$0	$1,150	$0	$1,647	$288	$0	$387	$2,323	$2
Subtotal	**$901**	**$323**	**$2,444**	**$0**	**$3,668**	**$642**	**$0**	**$862**	**$5,172**	**$4**
13					**Improvements to Site**					
13.1 Site Preparation	$0	$27	$541	$0	$568	$99	$0	$133	$801	$1
13.2 Site Improvements	$0	$126	$167	$0	$293	$51	$0	$69	$414	$0
13.3 Site Facilities	$144	$0	$151	$0	$296	$52	$0	$70	$417	$0
Subtotal	**$144**	**$153**	**$860**	**$0**	**$1,157**	**$203**	**$0**	**$272**	**$1,632**	**$1**
14					**Buildings & Structures**					
14.5 Circulation Water Pumphouse	$0	$50	$40	$0	$90	$16	$0	$21	$127	$0
Subtotal	**$0**	**$50**	**$40**	**$0**	**$90**	**$16**	**$0**	**$21**	**$127**	**$0**
Total	**$176,534**	**$34,760**	**$97,540**	**$0**	**$308,834**	**$54,046**	**$23,348**	**$77,245**	**$463,473**	**$387**

Exhibit 6-31. Capital costs for cement greenfield site with FGD and SCR and 90 percent CO_2 capture

Case:	Cement with FGD and SCR						Estimate Type:		Conceptual	
Representative Plant Size:	1.3 M tonnes/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
3					**Feedwater & Miscellaneous BOP Systems**					
3.1 Feedwater System	$616	$1,056	$528	$0	$2,199	$385	$0	$517	$3,101	$3
3.2 Water Makeup & Pretreating	$1,543	$154	$874	$0	$2,571	$450	$0	$604	$3,626	$3
3.3 Other Feedwater Subsystems	$280	$92	$87	$0	$459	$80	$0	$108	$647	$1
3.4 Industrial Boiler Package w/Deaerator	$3,731	$0	$1,085	$0	$4,816	$843	$0	$1,132	$6,790	$6
3.5 Other Boiler Plant Systems	$67	$25	$61	$0	$153	$27	$0	$36	$216	$0
3.6 NG Pipeline and Start-Up System	$624	$27	$20	$0	$671	$117	$0	$158	$946	$1

	Case:	Cement with FGD and SCR							Estimate Type:		Conceptual
	Representative Plant Size:	1.3 M tonnes/year							Cost Base:		Dec 2018
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	$/tonnes/yr ($CO_2$)
3.7	Waste Water Treatment Equipment	$2,872	$0	$1,760	$0	$4,632	$811	$0	$1,088	$6,531	$6
3.9	Miscellaneous Plant Equipment	$96	$13	$49	$0	$157	$27	$0	$37	$221	$0
	Subtotal	**$9,829**	**$1,366**	**$4,464**	**$0**	**$15,659**	**$2,740**	**$0**	**$3,680**	**$22,079**	**$20**
	4					Cement Kiln Accessories					
4.1	Selective Catalytic Reduction System	$5,660	$0	$3,225	$0	$8,885	$1,555	$0	$2,088	$12,528	$12
	Subtotal	**$5,660**	**$0**	**$3,225**	**$0**	**$8,885**	**$1,555**	**$0**	**$2,088**	**$12,528**	**$12**
	5					Flue Gas Cleanup					
5.1	Cansolv CO_2 Removal System	$55,656	$24,073	$50,554	$0	$130,284	$22,800	$22,148	$35,046	$210,278	$193
5.2	FGD Absorber Vessels & Accessories	$64,703	$0	$13,834	$0	$78,537	$13,744	$0	$18,456	$110,737	$102
5.3	Other FGD	$290	$0	$327	$0	$617	$108	$0	$145	$870	$1
5.4	CO_2 Compression & Drying	$16,242	$2,436	$5,430	$0	$24,108	$4,219	$0	$5,665	$33,993	$31
5.5	CO_2 Compressor Aftercooler	$127	$20	$54	$0	$201	$35	$0	$47	$284	$0
5.12	Gas Cleanup Foundations	$0	$65	$57	$0	$122	$21	$0	$29	$172	$0
	Subtotal	**$137,018**	**$26,595**	**$70,257**	**$0**	**$233,869**	**$40,927**	**$22,148**	**$59,389**	**$356,334**	**$327**
	7					Ductwork & Stack					
7.3	Ductwork	$0	$1,608	$1,117	$0	$2,725	$477	$0	$640	$3,842	$4
7.4	Stack	$7,712	$0	$4,482	$0	$12,194	$2,134	$0	$2,866	$17,194	$16
7.5	Duct & Stack Foundations	$0	$171	$203	$0	$374	$65	$0	$88	$527	$0
	Subtotal	**$7,712**	**$1,778**	**$5,802**	**$0**	**$15,293**	**$2,676**	**$0**	**$3,594**	**$21,563**	**$20**
	9					Cooling Water System					
9.1	Cooling Towers	$1,343	$0	$415	$0	$1,759	$308	$0	$413	$2,480	$2
9.2	Circulating Water Pumps	$137	$0	$10	$0	$147	$26	$0	$35	$207	$0
9.3	Circulating Water System Aux.	$1,805	$0	$239	$0	$2,043	$358	$0	$480	$2,881	$3
9.4	Circulating Water Piping	$0	$834	$756	$0	$1,590	$278	$0	$374	$2,242	$2
9.5	Make-up Water System	$199	$0	$255	$0	$454	$80	$0	$107	$641	$1
9.6	Component Cooling Water System	$130	$0	$100	$0	$230	$40	$0	$54	$324	$0
9.7	Circulating Water System Foundations	$0	$93	$154	$0	$246	$43	$0	$58	$347	$0
	Subtotal	**$3,614**	**$927**	**$1,929**	**$0**	**$6,470**	**$1,132**	**$0**	**$1,520**	**$9,122**	**$8**
	11					Accessory Electric Plant					
11.2	Station Service Equipment	$2,595	$0	$223	$0	$2,818	$493	$0	$662	$3,973	$4
11.3	Switchgear & Motor Control	$4,029	$0	$699	$0	$4,728	$827	$0	$1,111	$6,666	$6
11.4	Conduit & Cable Tray	$0	$524	$1,509	$0	$2,033	$356	$0	$478	$2,866	$3
11.5	Wire & Cable	$0	$1,387	$2,479	$0	$3,866	$677	$0	$909	$5,451	$5
	Subtotal	**$6,624**	**$1,911**	**$4,910**	**$0**	**$13,444**	**$2,353**	**$0**	**$3,159**	**$18,957**	**$17**
	12					Instrumentation & Control					

	Case:	Cement with FGD and SCR						Estimate Type:		Conceptual	
	Representative Plant Size:	1.3 M tonnes/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
12.8	Instrument Wiring & Tubing	$399	$320	$1,278	$0	$1,997	$350	$0	$469	$2,816	$3
12.9	Other I&C Equipment	$491	$0	$1,137	$0	$1,628	$285	$0	$383	$2,295	$2
	Subtotal	**$890**	**$320**	**$2,415**	**$0**	**$3,625**	**$634**	**$0**	**$852**	**$5,111**	**$5**
	13					**Improvements to Site**					
13.1	Site Preparation	$0	$26	$532	$0	$558	$98	$0	$131	$787	$1
13.2	Site Improvements	$0	$124	$164	$0	$288	$50	$0	$68	$406	$0
13.3	Site Facilities	$142	$0	$149	$0	$291	$51	$0	$68	$410	$0
	Subtotal	**$142**	**$150**	**$845**	**$0**	**$1,136**	**$199**	**$0**	**$267**	**$1,602**	**$1**
	14					**Buildings & Structures**					
14.5	Circulation Water Pumphouse	$0	$48	$38	$0	$86	$15	$0	$20	$122	$0
	Subtotal	**$0**	**$48**	**$38**	**$0**	**$86**	**$15**	**$0**	**$20**	**$122**	**$0**
	Total	**$171,489**	**$33,095**	**$93,884**	**$0**	**$298,468**	**$52,232**	**$22,148**	**$74,570**	**$447,417**	**$411**

Exhibit 6-32. Owners' costs for cement cases with FGD and SCR

Description	$/1,000	$/tonnes/yr ($CO_2$)	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs	**99% Capture**		**90% Capture**	
6 Months All Labor	$2,484	$2	$2,420	$2
1-Month Maintenance Materials	$436	$0	$421	$0
1-Month Non-Fuel Consumables	$366	$0	$349	$0
1-Month Waste Disposal	$11	$0	$11	$0
25% of 1-Month Fuel Cost at 100% CF	$391	$0	$355	$0
2% of TPC	$9,269	$8	$8,948	$8
Total	**$12,958**	**$11**	**$12,505**	**$11**
Inventory Capital	**99% Capture**		**90% Capture**	
60-day supply of fuel and consumables at 100% CF	$3,768	$3	$3,457	$3
0.5% of TPC (spare parts)	$2,317	$2	$2,237	$2
Total	**$6,086**	**$5**	**$5,694**	**$5**
Other Costs	**99% Capture**		**90% Capture**	
Initial Cost for Catalyst and Chemicals	$0	$0	$0	$0
Land	$30	$0	$30	$0
Other Owner's Costs	$69,521	$58	$67,113	$62
Financing Costs	$12,514	$10	$12,080	$11
TOC	**$564,581**	**$471**	**$544,839**	**$500**
TASC Multiplier (Cement, 33 year)	1.054		1.054	
TASC	**$594,983**	**$497**	**$574,178**	**$527**

Exhibit 6-33. Initial and annual O&M costs for cement greenfield site with FGD and SCR at 99 percent capture

Case:	**Cement with FGD and SCR**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**1.3 M tonnes/year**				**Capacity Factor (%):**	**85**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		2.3
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**2.3**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$1,008,407	$0.84
Maintenance Labor:					$2,966,225	$2.48
Administrative & Support Labor:					$993,658	$0.83
Property Taxes and Insurance:					$9,269,455	$7.74
Total:					**$14,237,746**	**$11.89**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$4,449,338	$4.37
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	775	$1.90	$0	$457,112	$0.45
Makeup and Waste Water Treatment Chemicals (ton):	0	2.6	$550.00	$0	$440,049	$0.43
SCR Catalyst (ft^3):	w/equip.	0.0	$150.00	$0	$104,464	$0.10
CO_2 Capture System Chemicals[A]:	Proprietary				$1,215,644	$1.19
Triethylene Glycol (gal):	w/equip.	240	$6.80	$0	$507,172	$0.50
Limestone (ton):	0	0	$22.00	$0	$0	$0.00
Ammonia (19 wt%, ton):	0.00	10.8	$300.00	$0	$1,008,681	$0.99
Subtotal:				**$0**	**$3,733,121**	**$3.67**
Waste Disposal						
Triethylene Glycol (gal):		240	$0.35	$0	$26,104	$0.03
Thermal Reclaimer Unit Waste (ton)		0.69	$38.00	$0	$8,077	$0.01
SCR Catalyst (ft^3):		0	$2.50	$0	$1,741	$0.00
Prescrubber Blowdown Waste (ton)		6.7	$38.00	$0	$78,627	$0.08
Subtotal:				**$0**	**$114,550**	**$0.11**
Variable Operating Costs Total:				**$0**	**$8,297,009**	**$8.15**
Fuel Cost						
Natural Gas (MMBtu)	0	11,625	$4.42	$0	$15,941,580	$15.66
Total:				**$0**	**$15,941,580**	**$15.66**

[A]CO_2 capture system chemicals includes NaOH and Cansolv solvent

Exhibit 6-34. Initial and annual O&M costs for cement greenfield site with FGD and SCR at 90 percent capture

Case:	**Cement with FGD and SCR**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**1.3 M tonnes/year**				**Capacity Factor (%):**	**85**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		2.3
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**2.3**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$1,008,407	$0.93
Maintenance Labor:					$2,863,470	$2.63
Administrative & Support Labor:					$967,969	$0.89
Property Taxes and Insurance:					$8,948,343	$8.22
Total:					**$13,788,190**	**$12.66**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$4,295,205	$4.64
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	717	$1.90	$0	$422,931	$0.46
Makeup and Waste Water Treatment Chemicals (ton):	0	2.4	$550.00	$0	$410,207	$0.44
SCR Catalyst (ft³):	w/equip.	0.0	$150.00	$0	$104,464	$0.11
CO_2 Capture System Chemicals[A]:	Proprietary				$1,152,997	$1.25
Triethylene Glycol (gal):	w/equip.	219	$6.80	$0	$461,270	$0.50
Limestone (ton):	0	0	$22.00	$0	$0	$0.00
Ammonia (19 wt%, ton):	0.00	10.8	$300.00	$0	$1,008,681	$1.09
Subtotal:				**$0**	**$3,560,548**	**$3.85**
Waste Disposal						
Triethylene Glycol (gal):		219	$0.35	$0	$23,742	$0.03
Thermal Reclaimer Unit Waste (ton)		0.65	$38.00	$0	$7,713	$0.01
SCR Catalyst (ft³):		0	$2.50	$0	$1,741	$0.00
Prescrubber Blowdown Waste (ton)		6.7	$38.00	$0	$78,627	$0.08
Subtotal:				**$0**	**$111,823**	**$0.12**
Variable Operating Costs Total:				**$0**	**$7,967,577**	**$8.61**
Fuel Cost						
Natural Gas (MMBtu)	0	10,569	$4.42	$0	$14,493,467	$15.66
Total:				**$0**	**$14,493,467**	**$15.66**

[A] CO_2 capture system chemicals includes NaOH and Cansolv solvent

The COC for the greenfield FGD + SCR sensitivity cases at 99 and 90 percent capture are presented in Exhibit 6-35 alongside corresponding values for the base case cement plants.

Exhibit 6-35. COC for 1.3 M tonnes/year cement greenfield cases (base cases and FGD + SCR cases)

	COC at 99 percent capture, \$/tonne CO_2		COC at 90 percent capture, \$/tonne CO_2	
Component	**Base Case**	**FGD + SCR Case**	**Base Case**	**FGD + SCR Case**
Capital	21.8	29.7	22.8	31.5
Fixed	10.6	14.0	11.1	14.9
Variable	5.9	8.2	6.1	8.6
Purchased Power and Fuel	22.6	22.9	22.6	23.0
Total COC	**60.8**	**74.8**	**62.7**	**78.0**

The result of this sensitivity is that the total COC increases by \$14.0/tonne CO_2 and \$15.3/tonne CO_2 for 99 and 90 percent capture, respectively, with the addition of FGD and SCR systems for flue gas treating prior to AGR. At \$78.0/tonne CO_2, this cement sensitivity case with 90 percent capture is the highest COC of any of the processes considered in this report. This COC sensitivity is an approximation, as actual plant SOx/NOx concentrations were not available, and it is not clear whether this sensitivity would be common occurrence in U.S. cement plants, or a special isolated case due to raw materials used in a specific plant or region.

6.2.11 Cement Conclusion

The low purity CO_2 stream produced in a cement plant results in a higher COC when compared to the high purity cases evaluated in this report, but the quantity of CO_2 to be captured from such a process makes them attractive industrial processes for CCS as it would represent a significant GHG reduction. A CO_2 capture and compression system for a 1.3 M tonnes/year cement plant was modeled to estimate the COC of capturing CO_2 from the kiln off-gas. The results showed the COC of CO_2 to be \$60.8/tonne CO_2 and \$62.7/tonne CO_2 for a greenfield site with 99 and 90 percent capture, respectively. For a retrofit application, the COC is \$62.4/tonne CO_2 and \$64.3/tonne CO_2 for 99 and 90 percent capture, respectively. The small disparities between greenfield and retrofit cases are the result of unknown difficulties required for a retrofit installation versus a greenfield application as discussed in Section 3.3, assuming adequate plot plan space for the retrofit case exists.

The plant size sensitivity showed that as plant size decreased from 1.5 M tonnes/year to 0.5 M tonnes/year of cement production, the COC increased by \$15.0/tonne CO_2 and \$15.8/tonne CO_2, for 99 and 90 percent capture, respectively. As the plant size is decreased, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC. As demonstrated by the resulting COCs and the sensitivity analysis, the normalized cost of 99 percent CO_2 capture is less than that of 90 percent capture.

For plants with SOx and/or NOx contaminants above that which is acceptable at the inlet of the AGR, an FGD and/or SCR system would be required to purify the stream before entering the CO_2 capture unit. An approximation of the additional cost of adding these systems showed an increase in greenfield COC by 23–25 percent. This approximation does not account for actual SOx/NOx concentrations in the kiln off-gas, and could be substantially higher or lower, depending on off-gas conditions and specific requirements of the AGR system deployed.

6.3 Iron/Steel

According to the Environmental Protection Agency, in 2019 the industrial sector emitted 1.51 B tonnes of CO_2, representing 23 percent of U.S. GHG emissions. [50] The Iron and Steel industry accounted for 4.8 percent or about 72 M tonnes of CO_2 emissions in 2019. [6] Due to the large amounts of emissions available for capture from the iron and steel industry, these facilities present a great opportunity for the consideration of industrial decarbonization.

6.3.1 Size Range

According to the World Steel Association, there were 132 steel plants in the United States, accounting for approximately 86.6 M tonnes of steel production in 2018. Of these 86.6 M tonnes of steel produced, 32 percent was produced using an electric arc furnace (EAF) and the balance was produced using the more traditional BOF. [51] The main difference between the EAF and BOF processes involves the raw materials used as inputs as well as the furnace design. The resulting steel product from an EAF process contains approximately 100 percent recycled steel, whereas the BOF product contains 25 percent recycled steel on average. [51] The utilization of scrap steel results in lower CO_2 emissions for an EAF process (0.6–0.9 tonne CO_2 per tonne steel) versus the BOF process (2.2 tonne CO_2 per tonne steel). [52] The combination of generally smaller EAF plants and lower concentration of EAF plant CO_2 emissions projects to a higher COC from an EAF process. Therefore, this study focuses on CO_2 capture from BOF process steel plants. The total production capacity, as given by the World Steel Association for BOF plants in the United States in 2018, was 58.9 M tonnes. [51]

6.3.2 CO_2 Point Sources

A study by Wiley, et al., ("Wiley Study") published in 2010, assessed the opportunities for CO_2 capture in Australian iron and steel mills. [52] The Wiley Study utilized stream data from an Australian BOF steel mill, and within the base plant, the largest source of CO_2 comes from the top gas of the blast furnace as is typical in an integrated steel mill; however, this stream is not directly vented. Instead, the blast furnace gas is cleaned and used in the plant as low-grade fuel, and instead of having a high-content CO_2 point source from the blast furnace gas, the CO_2 is distributed throughout the plant as smaller CO_2 point sources. The resulting CO_2 point sources available to be captured include the power plant stack (PPS), coke oven gas (COG), blast furnace stove (BFS), sinter stack, blown oxygen steelmaking stack, hot strip mill stack, plate mill stack, and lime kiln, based on the configuration detailed in the Wiley Study. [52] The three highest CO_2 concentrations of these point sources are the COG at 27 volume percent, the BFS at 21 volume percent, and the PPS at 23 volume percent.

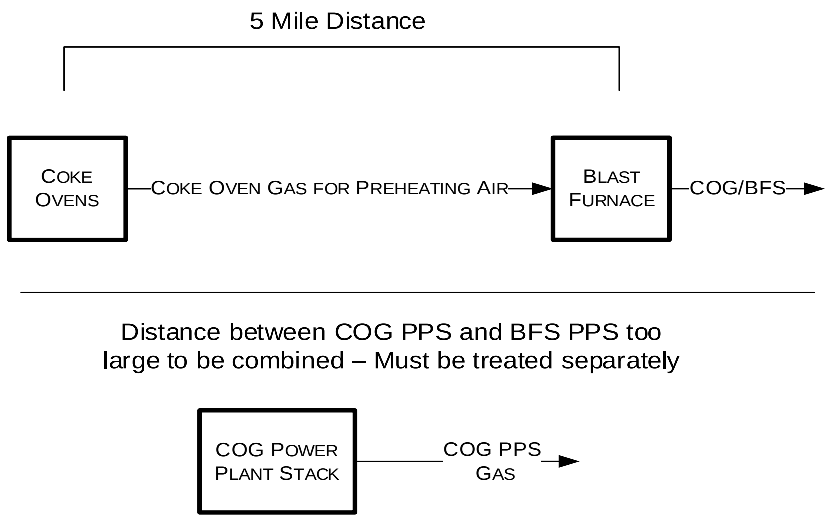
Of the eight CO_2 point sources listed by the Wiley Study, five have CO_2 concentrations that would have capture costs comparable to those in a typical coal-fired power plant flue gas stream and are not included in this analysis. Only the three higher CO_2 concentration streams, the PPS, COG, and BFS are evaluated, as shown in Exhibit 6-36.

Exhibit 6-36. BOF iron and steel plant characteristics [52]

Description	PPS	COG	BFS
CO_2 Emitted/Tonne Steel produced	0.74	0.35	0.39
Pressure (psia)	14.7	14.7	14.7
Temperature (°F)	572	212	572
Composition (vol %)			
N_2	67.00	67.00	68.00
H_2O	8.00	5.00	10.00
CO_2	23.00	27.00	21.00
O_2	1.00	1.00	1.00

Personal communication with a former U.S. Steel Braddock, PA, facility employee indicated that while the coke ovens are approximately five miles from the blast furnace, the COG is circulated back to the blast furnace to preheat the incoming air. Therefore, these two streams are located relatively close to one another and may be combined. Exhibit 6-37 is a simplified BFD of the plot plan description of the Braddock steel mill.

Exhibit 6-37. Braddock steel mill plot plan



6.3.3 Design Input and Assumptions

The following is a list of design inputs and assumptions made specific to the iron/steel process for the purpose of this study:

- The representative BOF integrated steel mill has a production capacity of 2.54 M tonnes/year
- The CO_2 generated is 3,738,928 tonnes CO_2/year at 100 percent CF
- There are three high purity point sources: COG, BFS, and COG PPS. The COG and BFS will be combined into one stream due to plot plan and totals 1,864,388 tonnes CO_2/year (at 100 percent CF); COG PPS will utilize its own separation and compression facility and generates 1,874,540 tonnes CO_2/year at 100 percent CF
- Since there are two separate capture systems, 4.6 operators are considered (i.e., 2.3 operators per capture system)
- As a low purity source, separation, compression, and cooling are required. Separation is accomplished using a Cansolv AGR unit
- CO_2 capture rates of 90 percent and 99 percent are evaluated
- The CO_2 quality is based on the EOR pipeline standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters [1]

6.3.4 CO_2 Capture System

The COG/BFS and COG PPS stream CO_2 concentrations require purification before compression to meet EOR pipeline standards. The purification system used is Shell's Cansolv post-combustion capture system discussed in Section 4.2.1. Steam for solvent regeneration is provided by the industrial boiler discussed in Section 4.3. A separate capture unit, boiler, and ancillary equipment is modeled for each COG/BFS and COG PPS stream. One integrally geared centrifugal compression train as discussed in Section 4.1.2 is employed for the COG/BFS stream and a second is used to compress the COG PPS stream. Costs for the compressors are scaled based on product CO_2 flow.

6.3.5 BFD, Stream Table, and Performance Summary

For the COG/BFS case, the COG stream and BFS stream are mixed and sent to the CO_2 capture system. Water and solids recovered from the AGR process are sent to waste treatment. The CO_2 stream is then compressed with interstage cooling and after-cooled before reaching the EOR pipeline. Exhibit 6-38 shows the BFD for this process, and Exhibit 6-39 and Exhibit 6-40 show the stream table for this process with 99 percent and 90 percent capture, respectively.

Exhibit 6-38. CO_2 capture BFD for COG/BFS

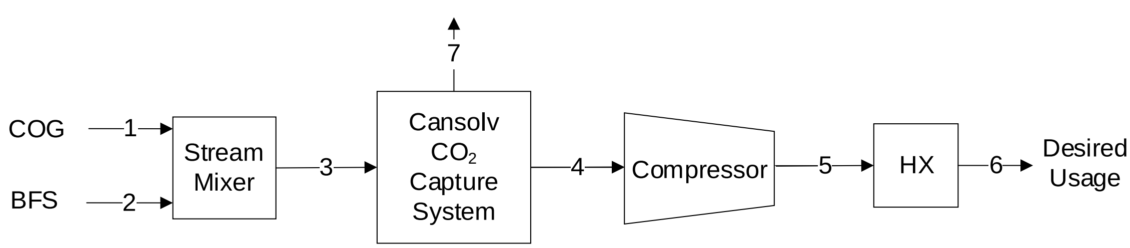


Exhibit 6-39. Iron/steel COG/BFS stream table with 99 percent capture

	1	2	3	4	5	6	7
V-L Mole Fraction							
AR	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2700	0.2100	0.2346	0.9879	0.9995	0.9995	0.0034
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0500	0.1000	0.0795	0.0121	0.0005	0.0005	0.0237
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6700	0.6800	0.6759	0.0000	0.0000	0.0000	0.9588
O_2	0.0100	0.0100	0.0100	0.0000	0.0000	0.0000	0.0141
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	8,443	12,173	20,616	4,845	4,788	4,788	14,533
V-L Flowrate (kg/hr)	269,106	370,224	639,331	211,692	210,637	210,637	405,309
Temperature (°C)	100	300	219	31	80	30	38
Pressure (MPa, abs)	0.10	0.1	0.1	0.2	15.3	15.3	0.1
Steam Table Enthalpy (kJ/kg)[A]	3,700	3,593	3,638	8,793	8,758	8,755	309.0
Aspen Plus Enthalpy (kJ/kg)[B]	-3,638	-3,217	-3,394	-8,961	-9,042	-9,195	-240.1
Density (kg/m^3)	1.0	0.6	0.8	3.5	432.5	630.1	1.1
V-L Molecular Weight	31.9	30.4	31.0	43.7	44.0	44.0	27.9
V-L Flowrate (lb_{mol}/hr)	18,613	26,837	45,450	10,681	10,555	10,555	32,041
V-L Flowrate (lb/hr)	593,278	816,205	1,409,483	466,701	464,375	464,375	893,553
Temperature (°F)	212	572	426	88	177	86	100
Pressure (psia)	14.7	14.7	14.7	28.9	2,216.9	2,214.7	14.8
Steam Table Enthalpy (Btu/lb)[A]	1,591	1,545	1,564	3,780	3,765	3,764	132.8
Aspen Plus Enthalpy (Btu/lb)[B]	-1,564	-1,383	-1,459	-3,852	-3,887	-3,953	-103.2
Density (lb/ft^3)	0.065	0.040	0.048	0.217	27.0	39.3	0.069

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 6-40. Iron/steel COG/BFS stream table with 90 percent capture

	1	2	3	4	5	6	7
V-L Mole Fraction							
AR	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2700	0.2100	0.2346	0.9881	0.9995	0.9995	0.0322
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0500	0.1000	0.0795	0.0119	0.0005	0.0005	0.0237
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6700	0.6800	0.6759	0.0000	0.0000	0.0000	0.9303
O_2	0.0100	0.0100	0.0100	0.0000	0.0000	0.0000	0.0137
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	8,443	12,173	20,616	4,405	4,354	4,354	14,978
V-L Flowrate (kg/hr)	269,106	370,224	639,331	192,516	191,573	191,573	424,582
Temperature (°C)	100	300	219	31	80	30	38
Pressure (MPa, abs)	0.10	0.1	0.1	0.2	15.3	15.3	0.1
Steam Table Enthalpy (kJ/kg)[A]	3,700	3,593	3,638	8,793	8,758	8,755	691.0
Aspen Plus Enthalpy (kJ/kg)[B]	-3,638	-3,217	-3,394	-8,960	-9,042	-9,195	-636.8
Density (kg/m^3)	1.0	0.6	0.8	3.5	432.5	630.1	1.1
V-L Molecular Weight	31.9	30.4	31.0	43.7	44.0	44.0	28.3
V-L Flowrate (lb_{mol}/hr)	18,613	26,837	45,450	9,712	9,599	9,599	33,021
V-L Flowrate (lb/hr)	593,278	816,205	1,409,483	424,424	422,347	422,347	936,044
Temperature (°F)	212	572	426	88	177	86	100
Pressure (psia)	14.7	14.7	14.7	28.9	2,216.9	2,214.7	14.8
Steam Table Enthalpy (Btu/lb)[A]	1,591	1,545	1,564	3,780	3,765	3,764	297.1
Aspen Plus Enthalpy (Btu/lb)[B]	-1,564	-1,383	-1,459	-3,852	-3,887	-3,953	-273.8
Density (lb/ft^3)	0.065	0.040	0.048	0.217	27.0	39.3	0.070

[A]Steam table reference conditions are 32.02°F & 0.089 psia
[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

In the same manner, the COG PPS stream is sent to the Cansolv CO_2 capture system. Water and solids recovered from the AGR process are sent to waste treatment. The CO_2 stream is then compressed with interstage cooling and after-cooled before reaching the EOR pipeline. Exhibit

6-41 shows the BFD for this process, and Exhibit 6-42 and Exhibit 6-43 show the stream table for this process with 99 percent and 90 percent capture, respectively.

Exhibit 6-41. CO_2 capture BFD for COG PPS

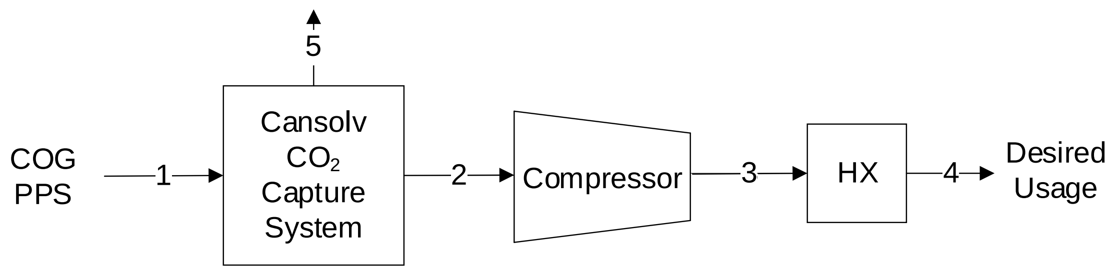


Exhibit 6-42. Iron/steel COG PPS stream table with 99 percent capture

	1	2	3	4	5
V-L Mole Fraction					
AR	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2323	0.9875	0.9995	0.9995	0.0034
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0808	0.0125	0.0005	0.0005	0.0242
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6768	0.0000	0.0000	0.0000	0.9581
O_2	0.0101	0.0000	0.0000	0.0000	0.0142
Total	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	20,931	4,873	4,814	4,814	14,785
V-L Flowrate (kg/hr)	648,081	212,873	211,784	211,784	412,236
Temperature (°C)	300	31	80	30	38
Pressure (MPa, abs)	0.1	0.2	15.3	15.3	0.1
Steam Table Enthalpy (kJ/kg)[A]	3,630	8,794	8,758	8,755	314.2
Aspen Plus Enthalpy (kJ/kg)[B]	-3,292	-8,961	-9,042	-9,195	-244.5
Density (kg/m^3)	0.7	3.5	432.5	630.1	1.1
V-L Molecular Weight	31.0	43.7	44.0	44.0	27.9
V-L Flowrate (lb_{mol}/hr)	46,145	10,743	10,612	10,612	32,595
V-L Flowrate (lb/hr)	1,428,775	469,304	466,905	466,905	908,825

	1	2	3	4	5
V-L Mole Fraction					
Temperature (°F)	572	88	177	86	100
Pressure (psia)	14.7	28.9	2,216.9	2,214.7	14.8
Steam Table Enthalpy (Btu/lb)[A]	1,561	3,781	3,765	3,764	135.1
Aspen Plus Enthalpy (Btu/lb)[B]	-1,415	-3,853	-3,887	-3,953	-105.1
Density (lb/ft^3)	0.041	0.217	27.0	39.3	0.069

[A]Steam table reference conditions are 32.02°F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

Exhibit 6-43. Iron/steel COG PPS stream table with 90 percent capture

	1	2	3	4	5
V-L Mole Fraction					
AR	0.0000	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.2323	0.9878	0.9995	0.9995	0.0319
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0808	0.0122	0.0005	0.0005	0.0243
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6768	0.0000	0.0000	0.0000	0.9300
O_2	0.0101	0.0000	0.0000	0.0000	0.0138
Total	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg$_{mol}$/hr)	20,931	4,431	4,378	4,378	15,232
V-L Flowrate (kg/hr)	648,081	193,589	192,617	192,617	431,610
Temperature (°C)	300	31	80	30	38
Pressure (MPa, abs)	0.1	0.2	15.3	15.3	0.1
Steam Table Enthalpy (kJ/kg)[A]	3,630	8,793	8,758	8,755	691.6
Aspen Plus Enthalpy (kJ/kg)[B]	-3,292	-8,961	-9,042	-9,195	-636.6
Density (kg/m^3)	0.7	3.5	432.5	630.1	1.1
V-L Molecular Weight	31.0	43.7	44.0	44.0	28.3
V-L Flowrate (lb$_{mol}$/hr)	46,145	9,768	9,652	9,652	33,581
V-L Flowrate (lb/hr)	1,428,775	426,791	424,647	424,647	951,538

	1	2	3	4	5
V-L Mole Fraction					
Temperature (°F)	572	88	177	86	100
Pressure (psia)	14.7	28.9	2,216.9	2,214.7	14.8
Steam Table Enthalpy (Btu/lb)[A]	1,561	3,781	3,765	3,764	297.3
Aspen Plus Enthalpy (Btu/lb)[B]	-1,415	-3,853	-3,887	-3,953	-273.7
Density (lb/ft^3)	0.041	0.217	27.0	39.3	0.070

[A]Steam table reference conditions are 32.02°F & 0.089 psia

[B]Aspen thermodynamic reference state is the component's constituent elements in an ideal gas state at 25°C and 1 atm

The performance summary for both 90 and 99 percent capture cases in the COG/BFS section of the steel mill is provided in Exhibit 6-44, while that of the COG PPS section is shown in Exhibit 6-45.

Exhibit 6-44. Performance summary for iron/steel COG/BFS section

Performance Summary		
Item	**2.54 M tonnes steel/year with 90 percent CO_2 capture (kWe)**	**2.54 M tonnes steel/year with 99 percent CO_2 capture (kWe)**
CO_2 Capture Auxiliaries	4,800	5,400
Steam Boiler Auxiliaries	510	560
CO_2 Compressor	14,660	16,120
Circulating Water Pumps	1,480	1,610
Cooling Tower Fans	770	830
Total Auxiliary Load	**22,220**	**24,520**

Exhibit 6-45. Performance summary for iron/steel COG PPS section

Performance Summary		
Item	**2.54 M tonnes steel/year with 90 percent CO_2 capture (kWe)**	**2.54 M tonnes steel/year with 99 percent CO_2 capture (kWe)**
CO_2 Capture Auxiliaries	4,900	5,400
Steam Boiler Auxiliaries	520	570
CO_2 Compressor	14,750	16,210
Circulating Water Pumps	1,490	1,620
Cooling Tower Fans	770	830
Total Auxiliary Load	**22,430**	**24,630**

6.3.6 Capture Integration

The BOF process integrated steel mill makes use of the BFS and COG as low-grade fuel for electricity generation. Due to this set-up, integrating equipment with additional auxiliary needs, such as power, steam, or cooling loads for the capture system, into the existing plant systems may be capacity limited.

The cooling water system considered in this study is a stand-alone unit; however, there is potential to integrate make-up water to feed or partially feed the cooling system thereby reducing the unit's size or replacing it completely with a simple HX. This would be evaluated on case-by-case basis depending on the size of the plant, its layout, and size of the plant's current cooling system, and such an evaluation is outside of the scope of this study.

6.3.7 Power Source

The compressor power consumption for the 90 and 99 percent capture cases in the COG/BFS section of the plant are 14.66 MW and 16.12 MW, respectively. The compressor power consumption for the 90 and 99 percent capture cases in the COG PPS section of the plant are 14.75 MW and 16.21 MW, respectively. Power consumption estimates for the cooling water system in each case were scaled as described in Section 4.4. The total power requirements were calculated to be 22.22 MW and 24.52 MW for the 90 and 99 percent capture rates in the COG/BFS section, respectively, while the total power requirements were calculated to be 22.43 MW and 24.63 MW for the 90 and 99 percent capture rates in the COG PPS section, respectively. These estimates include all power required by the compression train, cooling water system, and Cansolv capture unit. Purchased power cost is estimated at a rate of $60/MWh as discussed in Section 3.1.2.4. To satisfy the steam requirements of the AGR system, an industrial boiler was modeled, and fuel consumption costs were estimated at a rate of $4.42/MMBtu as discussed in Section 3.1.2.4.

6.3.8 Economic Analysis Results

The economic results of CO_2 capture application in an iron/steel mill are presented in this section. Owner's costs, capital costs, and O&M costs are calculated as discussed in Section 3.1. Retrofit costs were determined by applying a retrofit factor to TPC as discussed in Section 3.3. The retrofit TOC for the iron/steel case at 99 percent capture is $1,151 M, while for 90 percent capture, a retrofit TOC of $1,055 M is estimated. The corresponding retrofit COC for the 99 percent and 90 percent capture cases are $65.4/tonne CO_2 and $65.9/tonne CO_2, respectively. Greenfield cost estimates for the iron/steel case are not estimated, as BOF steel mills are no longer being constructed; thus, any capture application in a BOF mill as evaluated in this study would be implemented as a retrofit. Capital and O&M costs for each section (COG/BFS and COG PPS) are presented separately (Exhibit 6-47 through Exhibit 6-50), while owners costs and COCs are presented in whole for 99 and 90 percent capture cases in Exhibit 6-46 and Exhibit 6-53, respectively.

It should be noted that line-item capital costs were not estimated for retrofit cases, as the retrofit costs were estimated by applying a retrofit factor to the TPC of a greenfield plant as described in Section 3.3. As such, the account specific capital costs reported in this section are for a hypothetical greenfield plant but could be estimated for each account by applying a retrofit factor TPC as described in Section 3.3. As some O&M and owner's costs are estimated based on TPC, the retrofit TPC value was used to estimate the owner's costs and O&M costs presented in Exhibit 6-46 through Exhibit 6-52; thus, those values are indicative of a retrofit installation.

Exhibit 6-46. Owners' costs for iron/steel retrofit cases

Description	$/1,000	$/tonnes/yr ($CO_2$)	$/1,000	$/tonnes/yr ($CO_2$)
Pre-Production Costs	**99% Capture**		**90% Capture**	
6 Months All Labor	$5,095	$3	$4,776	$3
1-Month Maintenance Materials	$902	$0	$827	$0
1-Month Non-Fuel Consumables	$802	$0	$750	$0
1-Month Waste Disposal	$33	$0	$32	$0
25% of 1-Month Fuel Cost at 100% CF	$0	$0	$0	$0
2% of TPC	$19,171	$10	$17,151	$10
Total	**$26,003**	**$14**	**$23,536**	**$14**
Inventory Capital				
60-day supply of fuel and consumables at 100% CF	$1,327	$1	$1,243	$1
0.5% of TPC (spare parts)	$4,793	$3	$4,394	$3
Total	**$6,120**	**$3**	**$5,637**	**$3**
Other Costs				
Initial Cost for Catalyst and Chemicals	$0	$0	$0	$0
Land	$0	$0	$0	$0
Other Owner's Costs	$143,780	$78	$131,820	$78
Financing Costs	$25,880	$14	$23,728	$14
TOC	**$1,160,313**	**$627**	**$1,063,524**	**$632**
TASC Multiplier (Iron/Steel, 33 year)	1.091		1.091	
TASC	**$1,266,188**	**$684**	**$1,160,567**	**$690**

Exhibit 6-47. Capital costs for iron/steel COG/BFS section retrofit with 99 percent capture

	Case:	Iron/Steel COG/BFS Section						Estimate Type:		Conceptual	
	Representative Plant Size:	2.54 M tonnes steel/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$886	$1,519	$760	$0	$3,165	$554	$0	$744	$4,463	$2
3.2	Water Makeup & Pretreating	$2,239	$224	$1,269	$0	$3,732	$653	$0	$877	$5,263	$3
3.3	Other Feedwater Subsystems	$448	$147	$139	$0	$734	$128	$0	$173	$1,035	$1
3.4	Industrial Boiler Package w/Deaerator	$5,968	$0	$1,735	$0	$7,703	$1,348	$0	$1,810	$10,861	$6
3.5	Other Boiler Plant Systems	$108	$39	$99	$0	$246	$43	$0	$58	$347	$0
3.6	NG Pipeline and Start-Up System	$808	$35	$26	$0	$869	$152	$0	$204	$1,226	$1
3.7	Waste Water Treatment Equipment	$4,113	$0	$2,521	$0	$6,633	$1,161	$0	$1,559	$9,353	$5
3.9	Miscellaneous Plant Equipment	$109	$14	$56	$0	$179	$31	$0	$42	$253	$0
	Subtotal	**$14,680**	**$1,979**	**$6,604**	**$0**	**$23,263**	**$4,071**	**$0**	**$5,467**	**$32,801**	**$18**
5	Flue Gas Cleanup										
5.1	Cansolv CO_2 Removal System	$81,899	$35,424	$74,391	$0	$191,714	$33,550	$32,591	$51,571	$309,426	$168
5.4	CO_2 Compression & Drying	$22,324	$3,349	$7,464	$0	$33,136	$5,799	$0	$7,787	$46,722	$25
5.5	CO_2 Compressor Aftercooler	$196	$31	$84	$0	$312	$55	$0	$73	$440	$0
5.12	Gas Cleanup Foundations	$0	$89	$78	$0	$166	$29	$0	$39	$234	$0
	Subtotal	**$104,419**	**$38,893**	**$82,017**	**$0**	**$225,328**	**$39,432**	**$32,591**	**$59,470**	**$356,822**	**$193**
7	Ductwork & Stack										
7.3	Ductwork	$0	$2,303	$1,600	$0	$3,903	$683	$0	$917	$5,503	$3
7.4	Stack	$7,869	$0	$4,573	$0	$12,442	$2,177	$0	$2,924	$17,543	$10
7.5	Duct & Stack Foundations	$0	$176	$209	$0	$386	$68	$0	$91	$544	$0
	Subtotal	**$7,869**	**$2,479**	**$6,382**	**$0**	**$16,731**	**$2,928**	**$0**	**$3,932**	**$23,590**	**$13**
9	Cooling Water System										
9.1	Cooling Towers	$1,990	$0	$615	$0	$2,605	$456	$0	$612	$3,673	$2
9.2	Circulating Water Pumps	$213	$0	$15	$0	$228	$40	$0	$54	$321	$0
9.3	Circulating Water System Aux.	$2,489	$0	$329	$0	$2,818	$493	$0	$662	$3,974	$2
9.4	Circulating Water Piping	$0	$1,151	$1,042	$0	$2,193	$384	$0	$515	$3,092	$2
9.5	Make-up Water System	$255	$0	$328	$0	$583	$102	$0	$137	$823	$0
9.6	Component Cooling Water System	$179	$0	$138	$0	$317	$55	$0	$74	$447	$0
9.7	Circulating Water System Foundations	$0	$124	$207	$0	$331	$58	$0	$78	$467	$0
	Subtotal	**$5,126**	**$1,275**	**$2,674**	**$0**	**$9,076**	**$1,588**	**$0**	**$2,133**	**$12,797**	**$7**
11	Accessory Electric Plant										
11.2	Station Service Equipment	$3,192	$0	$274	$0	$3,466	$606	$0	$814	$4,887	$3

Case:	Iron/Steel COG/BFS Section						Estimate Type:		Conceptual	
Representative Plant Size:	2.54 M tonnes steel/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
11.3	Switchgear & Motor Control	$4,955	$0	$860	$0	$5,815	$1,018	$0	$1,366	$8,199	$4
11.4	Conduit & Cable Tray	$0	$644	$1,856	$0	$2,500	$438	$0	$588	$3,526	$2
11.5	Wire & Cable	$0	$1,706	$3,049	$0	$4,755	$832	$0	$1,117	$6,704	$4
	Subtotal	**$8,147**	**$2,350**	**$6,039**	**$0**	**$16,535**	**$2,894**	**$0**	**$3,886**	**$23,315**	**$13**
12	Instrumentation & Control										
12.8	Instrument Wiring & Tubing	$425	$340	$1,361	$0	$2,126	$372	$0	$500	$2,998	$2
12.9	Other I&C Equipment	$523	$0	$1,210	$0	$1,733	$303	$0	$407	$2,444	$1
	Subtotal	**$948**	**$340**	**$2,571**	**$0**	**$3,859**	**$675**	**$0**	**$907**	**$5,441**	**$3**
13	Improvements to Site										
13.1	Site Preparation	$0	$29	$585	$0	$614	$107	$0	$144	$866	$0
13.2	Site Improvements	$0	$136	$181	$0	$317	$56	$0	$75	$447	$0
13.3	Site Facilities	$156	$0	$164	$0	$320	$56	$0	$75	$451	$0
	Subtotal	**$156**	**$165**	**$930**	**$0**	**$1,251**	**$219**	**$0**	**$294**	**$1,764**	**$1**
14	Buildings & Structures										
14.5	Circulation Water Pumphouse	$0	$65	$52	$0	$117	$20	$0	$28	$165	$0
	Subtotal	**$0**	**$65**	**$52**	**$0**	**$117**	**$20**	**$0**	**$28**	**$165**	**$0**
	Total	**$141,345**	**$47,546**	**$107,269**	**$0**	**$296,160**	**$51,828**	**$32,591**	**$76,116**	**$456,696**	**$248**
					Retrofit Values	**$310,968**	**$54,419**	**$34,221**	**$79,922**	**$479,530**	**$260**

Exhibit 6-48. Capital costs for iron/steel COG PPS retrofit with 99 percent capture

Case:	Iron/Steel COG PPS Section						Estimate Type:		Conceptual	
Representative Plant Size:	2.54 M tonnes steel/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$890	$1,525	$763	$0	$3,177	$556	$0	$747	$4,480	$2
3.2	Water Makeup & Pretreating	$2,249	$225	$1,274	$0	$3,748	$656	$0	$881	$5,284	$3
3.3	Other Feedwater Subsystems	$450	$148	$140	$0	$738	$129	$0	$173	$1,040	$1
3.4	Industrial Boiler Package w/Deaerator	$5,998	$0	$1,744	$0	$7,741	$1,355	$0	$1,819	$10,915	$6
3.5	Other Boiler Plant Systems	$109	$40	$99	$0	$248	$43	$0	$58	$349	$0
3.6	NG Pipeline and Start-Up System	$811	$35	$26	$0	$872	$153	$0	$205	$1,229	$1
3.7	Waste Water Treatment Equipment	$4,144	$0	$2,540	$0	$6,683	$1,170	$0	$1,571	$9,424	$5
3.9	Miscellaneous Plant Equipment	$109	$14	$56	$0	$179	$31	$0	$42	$253	$0
	Subtotal	**$14,759**	**$1,986**	**$6,641**	**$0**	**$23,387**	**$4,093**	**$0**	**$5,496**	**$32,975**	**$18**
5	Flue Gas Cleanup										
5.1	Cansolv CO_2 Removal System	$81,456	$35,233	$73,989	$0	$190,678	$33,369	$32,415	$51,292	$307,755	$166
5.4	CO_2 Compression & Drying	$22,399	$3,360	$7,489	$0	$33,249	$5,819	$0	$7,813	$46,881	$25

Case:	Iron/Steel COG PPS Section						Estimate Type:		Conceptual		
Representative Plant Size:	2.54 M tonnes steel/year						Cost Base:		Dec 2018		
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	$/tonnes/yr ($CO_2$)
5.5	CO_2 Compressor Aftercooler	$197	$31	$85	$0	$313	$55	$0	$74	$442	$0
5.12	Gas Cleanup Foundations	$0	$98	$86	$0	$184	$32	$0	$43	$259	$0
	Subtotal	**$104,053**	**$38,722**	**$81,649**	**$0**	**$224,424**	**$39,274**	**$32,415**	**$59,223**	**$355,337**	**$192**
7						**Ductwork & Stack**					
7.3	Ductwork	$0	$2,591	$1,800	$0	$4,391	$768	$0	$1,032	$6,191	$3
7.4	Stack	$7,877	$0	$4,577	$0	$12,455	$2,180	$0	$2,927	$17,561	$9
7.5	Duct & Stack Foundations	$0	$176	$210	$0	$386	$68	$0	$91	$544	$0
	Subtotal	**$7,877**	**$2,767**	**$6,587**	**$0**	**$17,232**	**$3,016**	**$0**	**$4,049**	**$24,297**	**$13**
9						**Cooling Water System**					
9.1	Cooling Towers	$1,998	$0	$618	$0	$2,616	$458	$0	$615	$3,689	$2
9.2	Circulating Water Pumps	$214	$0	$15	$0	$229	$40	$0	$54	$323	$0
9.3	Circulating Water System Aux.	$2,498	$0	$330	$0	$2,828	$495	$0	$665	$3,988	$2
9.4	Circulating Water Piping	$0	$1,155	$1,046	$0	$2,201	$385	$0	$517	$3,103	$2
9.5	Make-up Water System	$256	$0	$329	$0	$585	$102	$0	$137	$825	$0
9.6	Component Cooling Water System	$180	$0	$138	$0	$318	$56	$0	$75	$448	$0
9.7	Circulating Water System Foundations	$0	$125	$207	$0	$332	$58	$0	$78	$468	$0
	Subtotal	**$5,146**	**$1,280**	**$2,684**	**$0**	**$9,110**	**$1,594**	**$0**	**$2,141**	**$12,845**	**$7**
11						**Accessory Electric Plant**					
11.2	Station Service Equipment	$3,198	$0	$274	$0	$3,472	$608	$0	$816	$4,896	$3
11.3	Switchgear & Motor Control	$4,965	$0	$861	$0	$5,826	$1,020	$0	$1,369	$8,215	$4
11.4	Conduit & Cable Tray	$0	$645	$1,860	$0	$2,505	$438	$0	$589	$3,532	$2
11.5	Wire & Cable	$0	$1,709	$3,055	$0	$4,764	$834	$0	$1,120	$6,718	$4
	Subtotal	**$8,163**	**$2,355**	**$6,051**	**$0**	**$16,568**	**$2,899**	**$0**	**$3,893**	**$23,361**	**$13**
12						**Instrumentation & Control**					
12.8	Instrument Wiring & Tubing	$425	$340	$1,362	$0	$2,127	$372	$0	$500	$3,000	$2
12.9	Other I&C Equipment	$523	$0	$1,211	$0	$1,734	$303	$0	$408	$2,445	$1
	Subtotal	**$948**	**$340**	**$2,573**	**$0**	**$3,861**	**$676**	**$0**	**$907**	**$5,445**	**$3**
13						**Improvements to Site**					
13.1	Site Preparation	$0	$29	$586	$0	$615	$108	$0	$144	$867	$0
13.2	Site Improvements	$0	$137	$181	$0	$317	$56	$0	$75	$448	$0
13.3	Site Facilities	$156	$0	$164	$0	$320	$56	$0	$75	$451	$0
	Subtotal	**$156**	**$166**	**$931**	**$0**	**$1,252**	**$219**	**$0**	**$294**	**$1,766**	**$1**
14						**Buildings & Structures**					
14.5	Circulation Water Pumphouse	$0	$66	$52	$0	$117	$21	$0	$28	$166	$0
	Subtotal	**$0**	**$66**	**$52**	**$0**	**$117**	**$21**	**$0**	**$28**	**$166**	**$0**
	Total	**$141,103**	**$47,682**	**$107,167**	**$0**	**$295,952**	**$51,792**	**$32,415**	**$76,032**	**$456,190**	**$246**
					Retrofit Values	**$310,749**	**$54,381**	**$34,036**	**$79,833**	**$479,000**	**$258**

Exhibit 6-49. Capital costs for iron/steel COG/BFS section retrofit with 90 percent capture

Case:		Iron/Steel COG/BFS Section					Estimate Type:			Conceptual	
Representative Plant Size:		2.54 M tonnes steel/year					Cost Base:			Dec 2018	
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	$/tonnes/yr ($CO_2$)
3	Feedwater & Miscellaneous BOP Systems										
3.1	Feedwater System	$830	$1,423	$711	$0	$2,964	$519	$0	$697	$4,179	$2
3.2	Water Makeup & Pretreating	$2,116	$212	$1,199	$0	$3,527	$617	$0	$829	$4,972	$3
3.3	Other Feedwater Subsystems	$411	$135	$128	$0	$674	$118	$0	$158	$951	$1
3.4	Industrial Boiler Package w/Deaerator	$5,483	$0	$1,594	$0	$7,077	$1,238	$0	$1,663	$9,979	$6
3.5	Other Boiler Plant Systems	$100	$36	$90	$0	$226	$40	$0	$53	$319	$0
3.6	NG Pipeline and Start-Up System	$772	$33	$25	$0	$830	$145	$0	$195	$1,170	$1
3.7	Waste Water Treatment Equipment	$3,935	$0	$2,412	$0	$6,346	$1,111	$0	$1,491	$8,948	$5
3.9	Miscellaneous Plant Equipment	$107	$14	$54	$0	$175	$31	$0	$41	$247	$0
	Subtotal	**$13,753**	**$1,853**	**$6,214**	**$0**	**$21,819**	**$3,818**	**$0**	**$5,127**	**$30,765**	**$18**
5	Flue Gas Cleanup										
5.1	Cansolv CO_2 Removal System	$71,707	$31,016	$65,134	$0	$167,857	$29,375	$28,536	$45,154	$270,921	$161
5.4	CO_2 Compression & Drying	$21,067	$3,160	$7,044	$0	$31,272	$5,473	$0	$7,349	$44,093	$26
5.5	CO_2 Compressor Aftercooler	$182	$29	$78	$0	$288	$50	$0	$68	$406	$0
5.12	Gas Cleanup Foundations	$0	$89	$78	$0	$166	$29	$0	$39	$234	$0
	Subtotal	**$92,956**	**$34,294**	**$72,333**	**$0**	**$199,583**	**$34,927**	**$28,536**	**$52,609**	**$315,655**	**$188**
7	Ductwork & Stack										
7.3	Ductwork	$0	$2,303	$1,600	$0	$3,903	$683	$0	$917	$5,503	$3
7.4	Stack	$7,883	$0	$4,581	$0	$12,464	$2,181	$0	$2,929	$17,575	$10
7.5	Duct & Stack Foundations	$0	$175	$208	$0	$384	$67	$0	$90	$541	$0
	Subtotal	**$7,883**	**$2,478**	**$6,389**	**$0**	**$16,751**	**$2,931**	**$0**	**$3,936**	**$23,619**	**$14**
9	Cooling Water System										
9.1	Cooling Towers	$1,874	$0	$580	$0	$2,454	$429	$0	$577	$3,460	$2
9.2	Circulating Water Pumps	$199	$0	$14	$0	$213	$37	$0	$50	$301	$0
9.3	Circulating Water System Aux.	$2,370	$0	$314	$0	$2,684	$470	$0	$631	$3,784	$2
9.4	Circulating Water Piping	$0	$1,096	$993	$0	$2,089	$365	$0	$491	$2,945	$2
9.5	Make-up Water System	$246	$0	$316	$0	$562	$98	$0	$132	$792	$0
9.6	Component Cooling Water System	$171	$0	$131	$0	$302	$53	$0	$71	$426	$0
9.7	Circulating Water System Foundations	$0	$119	$198	$0	$317	$55	$0	$74	$446	$0
	Subtotal	**$4,860**	**$1,215**	**$2,544**	**$0**	**$8,619**	**$1,508**	**$0**	**$2,026**	**$12,153**	**$7**
11	Accessory Electric Plant										
11.2	Station Service Equipment	$3,060	$0	$262	$0	$3,322	$581	$0	$781	$4,684	$3

Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/tonnes/yr ($CO_2$)
	Case:	Iron/Steel COG/BFS Section					Estimate Type:				Conceptual
	Representative Plant Size:	2.54 M tonnes steel/year					Cost Base:				Dec 2018
11.3	Switchgear & Motor Control	$4,750	$0	$824	$0	$5,574	$975	$0	$1,310	$7,859	$5
11.4	Conduit & Cable Tray	$0	$617	$1,779	$0	$2,397	$419	$0	$563	$3,380	$2
11.5	Wire & Cable	$0	$1,635	$2,923	$0	$4,558	$798	$0	$1,071	$6,427	$4
	Subtotal	**$7,810**	**$2,253**	**$5,789**	**$0**	**$15,851**	**$2,774**	**$0**	**$3,725**	**$22,350**	**$13**
	12					**Instrumentation & Control**					
12.8	Instrument Wiring & Tubing	$420	$336	$1,343	$0	$2,099	$367	$0	$493	$2,960	$2
12.9	Other I&C Equipment	$516	$0	$1,195	$0	$1,711	$299	$0	$402	$2,413	$1
	Subtotal	**$936**	**$336**	**$2,538**	**$0**	**$3,810**	**$667**	**$0**	**$895**	**$5,372**	**$3**
	13					**Improvements to Site**					
13.1	Site Preparation	$0	$28	$574	$0	$602	$105	$0	$142	$849	$1
13.2	Site Improvements	$0	$134	$177	$0	$311	$54	$0	$73	$438	$0
13.3	Site Facilities	$153	$0	$161	$0	$314	$55	$0	$74	$442	$0
	Subtotal	**$153**	**$162**	**$912**	**$0**	**$1,227**	**$215**	**$0**	**$288**	**$1,730**	**$1**
	14					**Buildings & Structures**					
14.5	Circulation Water Pumphouse	$0	$62	$49	$0	$112	$20	$0	$26	$158	$0
	Subtotal	**$0**	**$62**	**$49**	**$0**	**$112**	**$20**	**$0**	**$26**	**$158**	**$0**
	Total	**$128,351**	**$42,652**	**$96,769**	**$0**	**$267,772**	**$46,860**	**$28,536**	**$68,634**	**$411,802**	**$245**
					Retrofit Values	**$281,161**	**$49,203**	**$29,963**	**$72,065**	**$432,392**	**$258**

Exhibit 6-50. Capital costs for iron/steel COG PPS retrofit with 90 percent capture

Item No.	Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Bare Erected Cost	Eng'g CM H.O. & Fee	Contingencies Process	Contingencies Project	Total Plant Cost $/1,000	Total Plant Cost $/tonnes/yr ($CO_2$)
	Case:	Iron/Steel COG PPS Section					Estimate Type:				Conceptual
	Representative Plant Size:	2.54 M tonnes steel/year					Cost Base:				Dec 2018
	3					**Feedwater & Miscellaneous BOP Systems**					
3.1	Feedwater System	$833	$1,428	$714	$0	$2,975	$521	$0	$699	$4,195	$2
3.2	Water Makeup & Pretreating	$2,125	$212	$1,204	$0	$3,541	$620	$0	$832	$4,993	$3
3.3	Other Feedwater Subsystems	$413	$136	$129	$0	$678	$119	$0	$159	$956	$1
3.4	Industrial Boiler Package w/Deaerator	$5,510	$0	$1,602	$0	$7,112	$1,245	$0	$1,671	$10,028	$6
3.5	Other Boiler Plant Systems	$100	$36	$91	$0	$227	$40	$0	$53	$321	$0
3.6	NG Pipeline and Start-Up System	$774	$33	$25	$0	$832	$146	$0	$196	$1,173	$1
3.7	Waste Water Treatment Equipment	$3,965	$0	$2,430	$0	$6,396	$1,119	$0	$1,503	$9,018	$5
3.9	Miscellaneous Plant Equipment	$107	$14	$54	$0	$175	$31	$0	$41	$247	$0
	Subtotal	**$13,827**	**$1,860**	**$6,249**	**$0**	**$21,937**	**$3,839**	**$0**	**$5,155**	**$30,931**	**$18**
	5					**Flue Gas Cleanup**					
5.1	Cansolv CO_2 Removal System	$74,921	$32,406	$68,053	$0	$175,379	$30,691	$29,814	$47,177	$283,062	$168
5.4	CO_2 Compression & Drying	$21,146	$3,172	$7,070	$0	$31,389	$5,493	$0	$7,376	$44,258	$26
5.5	CO_2 Compressor Aftercooler	$182	$29	$78	$0	$289	$51	$0	$68	$408	$0

	Case:	Iron/Steel COG PPS Section						Estimate Type:		Conceptual	
	Representative Plant Size:	2.54 M tonnes steel/year						Cost Base:		Dec 2018	
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	$/tonnes/yr ($CO_2$)
5.12	Gas Cleanup Foundations	$0	$98	$86	$0	$184	$32	$0	$43	$259	$0
	Subtotal	**$96,249**	**$35,705**	**$75,287**	**$0**	**$207,241**	**$36,267**	**$29,814**	**$54,665**	**$327,987**	**$194**
	7	**Ductwork & Stack**									
7.3	Ductwork	$0	$2,591	$1,800	$0	$4,391	$768	$0	$1,032	$6,191	$4
7.4	Stack	$7,891	$0	$4,586	$0	$12,477	$2,183	$0	$2,932	$17,593	$10
7.5	Duct & Stack Foundations	$0	$175	$208	$0	$384	$67	$0	$90	$541	$0
	Subtotal	**$7,891**	**$2,766**	**$6,594**	**$0**	**$17,252**	**$3,019**	**$0**	**$4,054**	**$24,325**	**$14**
	9	**Cooling Water System**									
9.1	Cooling Towers	$1,882	$0	$582	$0	$2,465	$431	$0	$579	$3,475	$2
9.2	Circulating Water Pumps	$200	$0	$14	$0	$214	$37	$0	$50	$302	$0
9.3	Circulating Water System Aux.	$2,379	$0	$315	$0	$2,693	$471	$0	$633	$3,797	$2
9.4	Circulating Water Piping	$0	$1,100	$996	$0	$2,096	$367	$0	$493	$2,955	$2
9.5	Make-up Water System	$247	$0	$317	$0	$563	$99	$0	$132	$794	$0
9.6	Component Cooling Water System	$171	$0	$132	$0	$303	$53	$0	$71	$427	$0
9.7	Circulating Water System Foundations	$0	$119	$198	$0	$318	$56	$0	$75	$448	$0
	Subtotal	**$4,879**	**$1,219**	**$2,553**	**$0**	**$8,652**	**$1,514**	**$0**	**$2,033**	**$12,199**	**$7**
	11	**Accessory Electric Plant**									
11.2	Station Service Equipment	$3,072	$0	$264	$0	$3,336	$584	$0	$784	$4,703	$3
11.3	Switchgear & Motor Control	$4,769	$0	$827	$0	$5,597	$979	$0	$1,315	$7,891	$5
11.4	Conduit & Cable Tray	$0	$620	$1,787	$0	$2,407	$421	$0	$566	$3,393	$2
11.5	Wire & Cable	$0	$1,642	$2,935	$0	$4,577	$801	$0	$1,076	$6,453	$4
	Subtotal	**$7,841**	**$2,262**	**$5,813**	**$0**	**$15,916**	**$2,785**	**$0**	**$3,740**	**$22,441**	**$13**
	12	**Instrumentation & Control**									
12.8	Instrument Wiring & Tubing	$420	$336	$1,345	$0	$2,102	$368	$0	$494	$2,963	$2
12.9	Other I&C Equipment	$517	$0	$1,196	$0	$1,713	$300	$0	$403	$2,416	$1
	Subtotal	**$937**	**$336**	**$2,542**	**$0**	**$3,815**	**$668**	**$0**	**$897**	**$5,379**	**$3**
	13	**Improvements to Site**									
13.1	Site Preparation	$0	$28	$575	$0	$603	$106	$0	$142	$851	$1
13.2	Site Improvements	$0	$134	$178	$0	$312	$55	$0	$73	$439	$0
13.3	Site Facilities	$153	$0	$161	$0	$314	$55	$0	$74	$443	$0
	Subtotal	**$153**	**$162**	**$913**	**$0**	**$1,229**	**$215**	**$0**	**$289**	**$1,733**	**$1**
	14	**Buildings & Structures**									
14.5	Circulation Water Pumphouse	$0	$63	$50	$0	$112	$20	$0	$26	$158	$0
	Subtotal	**$0**	**$63**	**$50**	**$0**	**$112**	**$20**	**$0**	**$26**	**$158**	**$0**
	Total	**$131,779**	**$44,373**	**$100,001**	**$0**	**$276,154**	**$48,327**	**$29,814**	**$70,859**	**$425,154**	**$252**
					Retrofit Values	**$289,961**	**$50,743**	**$31,305**	**$74,402**	**$446,411**	**$265**

The initial and annual O&M costs for an iron/steel retrofit site were calculated and are shown in Exhibit 6-51 and Exhibit 6-52 for 99 percent and 90 percent capture, respectively, while Exhibit 6-53 shows the retrofit COC of the representative iron/steel plants at both capture rates.

Exhibit 6-51. Initial and annual O&M costs for iron/steel site with 99 percent capture

Case:	**Iron/Steel**				**Cost Base:**	**Dec 2018**
Representative Plant Size:	**2.54 M tonnes steel/year**				**Capacity Factor (%):**	**85**
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		4.6
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**4.6**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$2,016,815	$1.09
Maintenance Labor:					$6,134,594	$3.32
Administrative & Support Labor:					$2,037,852	$1.10
Property Taxes and Insurance:					$19,170,607	$10.36
Total:					**$29,359,868**	**$15.87**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$9,201,891	$5.85
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	2,397	$1.90	$0	$1,413,167	$0.90
Makeup and Waste Water Treatment Chemicals (ton):	0	8.0	$550.00	$0	$1,369,239	$0.87
CO_2 Capture System Chemicals[A]:	Proprietary				$3,832,974	$2.44
Triethylene Glycol (gal):	w/equip.	743	$6.80	$0	$1,567,150	$1.00
Subtotal:				**$0**	**$8,182,530**	**$5.20**
Waste Disposal						
Triethylene Glycol (gal):		743	$0.35	$0	$80,662	$0.05
Thermal Reclaimer Unit Waste (ton)		2.12	$38.00	$0	$24,958	$0.02
Prescrubber Blowdown Waste (ton)		19.8	$38.00	$0	$233,136	$0.15
Subtotal:				**$0**	**$338,756**	**$0.22**
Variable Operating Costs Total:				**$0**	**$17,723,177**	**$11.27**
Fuel Cost						
Natural Gas (MMBtu)	0	35,931	$4.42	$0	$49,275,013	$31.33
Total:				**$0**	**$49,275,013**	**$31.33**

[A]CO_2 capture system chemicals includes NaOH and Cansolv solvent

Exhibit 6-52. Initial and annual O&M costs for an iron/steel retrofit site with 90 percent capture

Case:	Iron/Steel				Cost Base:	Dec 2018
Representative Plant Size:	2.5 M tonnes steel/year				Capacity Factor (%):	85
Operating & Maintenance Labor						
Operating Labor				**Operating Labor Requirements per Shift**		
Operating Labor Rate (base):		38.50	**$/hour**	Skilled Operator:		0.0
Operating Labor Burden:		30.00	**% of base**	Operator:		4.6
Labor O-H Charge Rate:		25.00	**% of labor**	Foreman:		0.0
				Lab Techs, etc.:		0.0
				Total:		**4.6**
Fixed Operating Costs						
					Annual Cost	
					($)	**($/tonnes/yr CO_2)**
Annual Operating Labor:					$2,016,815	$1.20
Maintenance Labor:					$5,624,341	$3.34
Administrative & Support Labor:					$1,910,289	$1.14
Property Taxes and Insurance:					$17,576,066	$10.44
Total:					**$27,127,511**	**$16.12**
Variable Operating Costs						
					($)	**($/tonnes/yr CO_2)**
Maintenance Material:					$8,436,512	$5.90
Consumables						
	Initial Fill	**Per Day**	**Per Unit**	**Initial Fill**		
Water (/1000 gallons):	0	2,218	$1.90	$0	$1,307,481	$0.91
Makeup and Waste Water Treatment Chemicals (ton):	0	7.5	$550.00	$0	$1,276,955	$0.89
CO_2 Capture System Chemicals[A]:	Proprietary				$3,635,446	$2.54
Triethylene Glycol (gal):	w/equip.	676	$6.80	$0	$1,425,314	$1.00
Subtotal:				**$0**	**$7,645,197**	**$5.35**
Waste Disposal						
Triethylene Glycol (gal):		676	$0.35	$0	$73,362	$0.05
Thermal Reclaimer Unit Waste (ton)		1.98	$38.00	$0	$22,754	$0.02
Prescrubber Blowdown Waste (ton)		19.8	$38.00	$0	$233,136	$0.16
Subtotal:				**$0**	**$329,251**	**$0.23**
Variable Operating Costs Total:				**$0**	**$16,410,960**	**$11.47**
Fuel Cost						
Natural Gas (MMBtu)	0	32,667	$4.42	$0	$44,798,673	$31.32
Total:				**$0**	**$44,798,673**	**$31.32**

[A]CO_2 capture system chemicals includes NaOH and Cansolv solvent

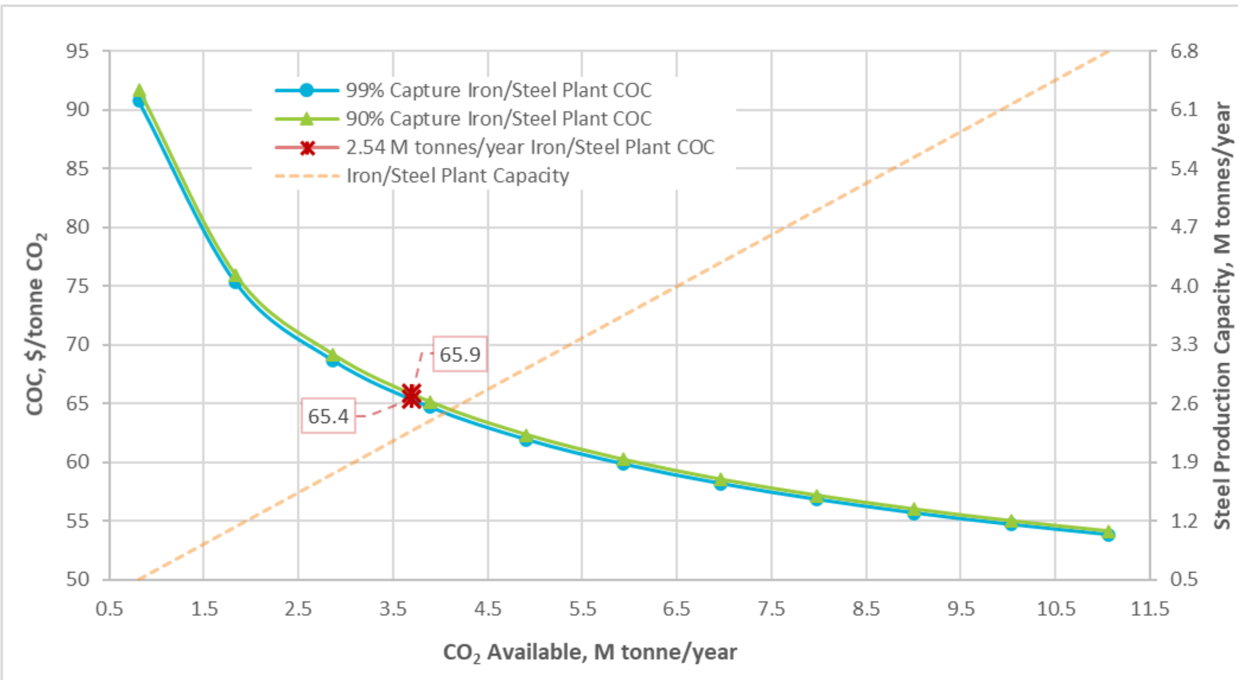
Exhibit 6-53. COC for 2.54 M tonnes/year iron/steel retrofit cases

Component	99% capture COC, $/tonne CO_2	90% capture COC, $/tonne CO_2
Capital	27.8	28.0
Fixed	9.3	9.5
Variable	5.6	5.7
Purchased Power and Fuel	22.6	22.6
Total COC	**65.4**	**65.9**

6.3.9 Plant Capacity Sensitivity Analysis

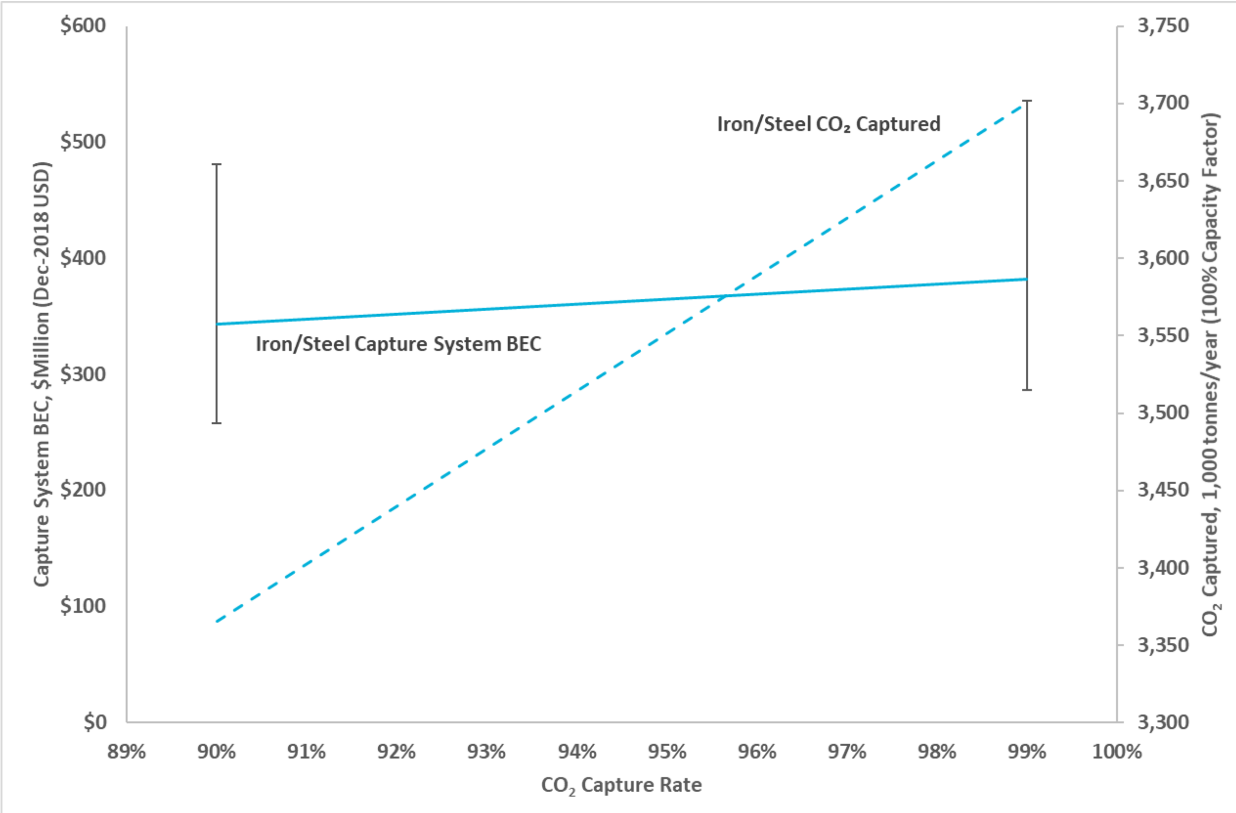
An analysis of the sensitivity of retrofit COC to iron/steel plant capacity is shown in Exhibit 6-54. As the plant capacity increases, more CO_2 is available for capture, thus realizing economies of scale. This generic scaling exercise assumes that equipment is available at continuous capacities; however, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of this sensitivity analysis.

Exhibit 6-54. Iron/steel plant capacity sensitivity



As the cost of capturing CO_2 is a normalized cost (i.e., \$/tonne CO_2), higher capture rates appear to cost less than lower capture rates. Comparing the true capital and O&M costs (i.e., not as normalized costs) shows that capital and O&M expenditures increase at higher capture rates. The cost of the capture system and associated consumables increases at a lesser rate than that of the amount of CO_2 captured (i.e., a 10 percent increase from 90 to 99 percent capture). The margin of error associated with the financial assumptions and cost scaling methodology employed in this study indicate that with increasing capture rate in the low purity cases, the COC is effectively the same. The reported minor increase in capital cost with increased capture rate (up to 99 percent for sources with CO_2 purity greater than 12 percent) has been validated by independent modeling performed by the CCSI team at NETL and has been reported independently in literature. [4] Exhibit 6-55 shows the error in the calculated capture system BEC associated with the vendor's quoted uncertainty rate (-25/+40 percent) alongside the amount of CO_2 captured in the cement case from 90 to 99 percent capture rate.

Exhibit 6-55. Iron/steel capture system BEC and amount of CO_2 captured versus capture rate



6.3.10 Iron/Steel Conclusion

The low purity CO_2 streams produced in an iron/steel mill results in a higher COC when compared to the high purity cases evaluated in this report, but the quantity of CO_2 to be captured from such a process makes them attractive industrial processes for CCS as it would represent a significant GHG reduction. Two CO_2 capture and compression systems for a 2.54 M tonnes/year integrated steel mill were modeled to estimate the COC of capturing CO_2 from the COG and BFS combined flue gas stream and from the COG PPS exhaust. The results showed the COC of CO_2 to be $65.4/tonne CO_2 and $65.9/tonne CO_2 for a retrofit site with 99 and 90 percent capture, respectively. No greenfield COC is calculated, as BOF steel mills are no longer being constructed; thus, any application of CO_2 capture in such a facility would be a retrofit application.

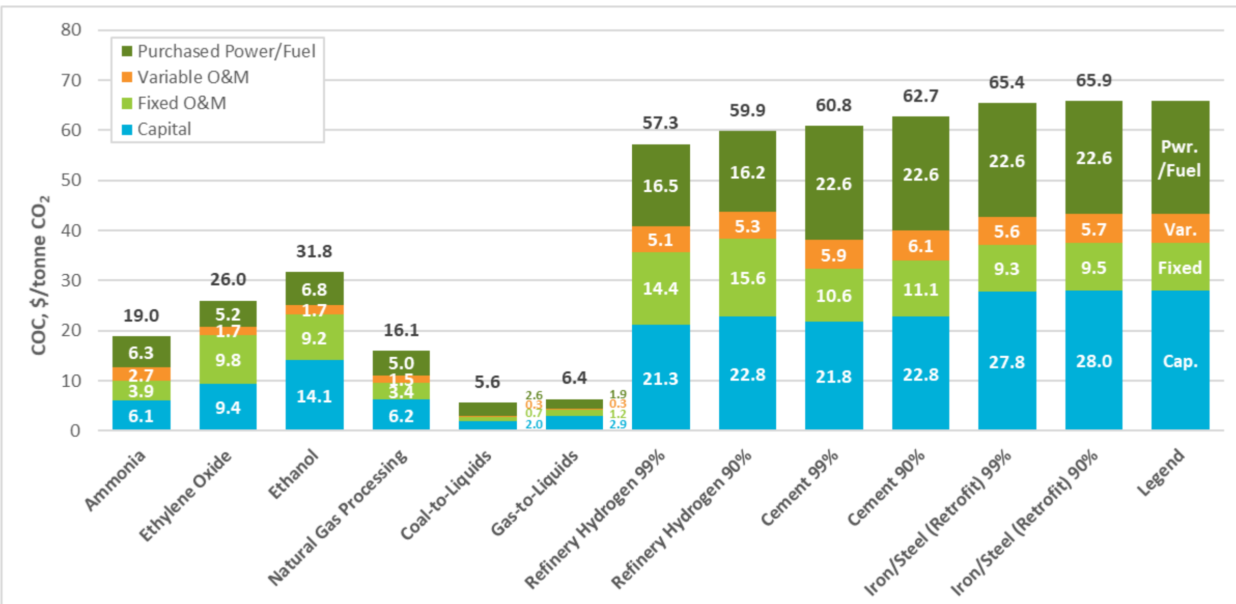
The plant size sensitivity showed that as plant size decreased from 6.8 M tonnes/year to 0.5 M tonnes/year of iron/steel production, the COC increased by $36.9/tonne CO_2 and $37.6/tonne CO_2, for 99 and 90 percent capture, respectively. As the plant size is decreased, less CO_2 is produced, and economies of scale are lost, resulting in a higher COC. As demonstrated by the resulting COCs and the sensitivity analysis, the normalized cost of 99 percent CO_2 capture is less than that of 90 percent capture.

7 Economic Analysis

7.1 Economic Results

Exhibit 7-1 shows the COC results of each industry considered in this study. When comparing high purity to low purity industrial sources, the former show lower COCs, as they require less equipment (i.e., no capture unit or boiler) and consumables (i.e., no solvents or NG fuel and less purchased power) than the low purity industrial sources. The low purity sources higher COC is notable not only in the additional capital costs, but in the O&M and purchased power and fuel costs as well. These cases require an industrial boiler, which is fueled by purchased NG, and the CO_2 capture systems add consumables and additional electrical auxiliary loads that increase purchased power costs over that of high purity sources.

Exhibit 7-1. COC summary



Evaluating the capital portion of the COC for each source shows the effects of capital intensity. The financial assumptions assumed in this report are industry specific. For instance, ethanol financial factors suggest that ethanol facilities would incur higher capital intensity compared to the cement, steel, and refining industries due to the return on equity and financing scenarios prevalent within the ethanol production market. Another interesting observation regarding capital intensity is the relationship between the EO and ethanol results. Although ethanol presents a higher amount available CO_2 for capture, its capital and power costs are higher than EO. This is counter-intuitive to the notion of economies of scale but illustrates the role that capture stream conditions (i.e., temperature, pressure, composition, and flow rate) plays on capture costs. In the ethanol case, the pure CO_2 stream must first be cooled, due to the high temperature from the fermentation process, and then has a higher compression ratio (compared to the EO case) to reach the required pipeline pressure of 2,200 psig. The additional stage of compression and the additional HX impact the auxiliary load as well as the capital expenditure.

Lastly, the CO_2 available for capture is both process and market dependent. The process emissions detailed for each case throughout the report are average constants; however, as each individual market dictates production capacities, the total CO_2 available from a plant could, with increasing market demand (e.g., plant expansions, increased CF, etc.), drive down the COC for that representative case. This trend could be estimated from the results of the plant size sensitivities for each case, but it should be noted that these estimates, and the sensitivities to plant size for each case, are dependent upon the assumption that equipment is available at any and every capacity or rating. However, equipment is often manufactured in discrete sizes, which would possibly affect the advantages of economies of scale and skew the results of the estimates provided herein. A general observation made under the assumptions of this report is demonstrated in the normalized COC elements and the total normalized COCs calculated: more CO_2 available results in lower normalized costs and realizes economies of scale.

7.1.1 Cost and Performance Summaries

The cost and performance results presented in this study are summarized in Exhibit 7-2 and Exhibit 7-3 for the high purity and low purity cases, respectively. Of all cases examined in this study, the lowest COC of $5.6/tonne CO_2 is achieved in a representative CTL facility. There are no CTL facilities currently in operation in the United States, but the low COC in such a facility implies that any new builds would include carbon capture in its greenfield design.

Of the existing industrial plant types available in the United States, the lowest COC of $16.1/tonne CO_2 is indicated at a representative NGP facility. The amount of CO_2 available for capture in an NGP facility is dependent upon the raw gas CO_2 content at the inlet of the plant. Capture costs for such a facility account for costs of CO_2 compression and cooling, based on the design assumptions regarding the base NGP plant. Although COC would increase with decreasing CO_2 availability, it is expected that integrating CO_2 capture for EOR would be feasible in most NGP facilities since the AGR unit is often inherent to the facility design.

Of the low purity cases, which require CO_2 purification (i.e., AGR units) along with compression and cooling, the pre-combustion capture in the refinery hydrogen 99 percent capture case represents the lowest COC at $57.3/tonne CO_2. In pre-combustion capture units, variable costs such as consumables, waste disposal, purchased power, and fuel are lower on a normalized basis when compared to post-combustion capture applications. It should be noted that the pre-combustion capture system described in Section 4.2.2 would not be installed for design capture rates lower than approximately 99 percent. As such, the values reported for the 90 percent capture rate in the refinery hydrogen case are meant for comparison purposes only and likely represent a deviation from the optimal design operation.

Exhibit 7-2. Cost and performance summary comparison – high purity cases

	Industrial Source Facilities					
	Ammonia	EO	Ethanol	NGP	CTL	GTL
PERFORMANCE						
Capacity Factor	85%	85%	85%	85%	85%	85%
Representative Plant Size	394,000 tonnes EO/year	364,500 tonnes EO/year	50 M gallons ethanol/year	330 MMSCFD natural gas	50,000 barrels F-T liquids/day	50,000 barrels F-T liquids/day
CO_2 Captured (at 85% CF), tonnes/year[A]	413,163	103,275	121,588	551,815	7,431,825	1,579,952
CO_2 Captured (at 85% CF), tonnes/hour	47	12	14	63	848	180
CO_2 Compressor Load, kW	5,770	1,180	1,810	6,010	43,480	6,700
Cooling Water Flowrate, gpm	2,994	673	1,098	3,479	25,172	3,823
Cooling Tower Duty, MMBtu/hour	30	7	11	35	252	38
COST						
TPC, $/1,000	37,347	16,636	20,187	46,690	162,840	49,170
BEC	26,487	11,799	14,317	33,114	115,490	34,872
Home Office Expenses	4,635	2,065	2,505	5,795	20,211	6,103
Project Contingency	6,225	2,773	3,364	7,782	27,140	8,195
Process Contingency	0	0	0	0	0	0
TOC, $M	46	20	25	57	197	60
TOC, $/1,000	45,587	20,385	24,672	56,764	196,924	59,661
Owner's Costs	8,240	3,749	4,485	10,074	34,084	10,491
TASC, $/1,000	47,162	20,892	25,840	58,977	207,583	62,890
Capital Costs, $/tonne CO_2	6.1	9.4	14.1	6.2	2.0	2.9
Fixed Costs, $/tonne CO_2	3.9	9.8	9.2	3.4	0.7	1.2
Variable Costs, $/tonne CO_2	2.7	1.7	1.7	1.5	0.3	0.3
Purchased Power and/or Fuel, $/tonne CO_2	6.3	5.2	6.8	5.0	2.6	1.9
COC (ex. T&S), $/tonne CO_2	**19.0**	**26.0**	**31.8**	**16.1**	**5.6**	**6.4**

[A]Due to simplification of BFDs and stream tables throughout the body of the report where minor process streams are omitted, actual CO_2 captured as calculated in summary tables may be slightly less than that calculated at the capture rates applied in each case. This is due primarily to trace amounts of CO_2 entrained in water vapor generated during dehydration. Such differences, where they appear, are not expected to have any meaningful impact on the key results of this study, as they account for less than 1 percent of the CO_2 generated by the emitter.

Exhibit 7-3. Cost and performance summary comparison – low purity cases

	Industrial Source Facilities					
	Refinery H_2 99%	Refinery H_2 90%	Cement 99%	Cement 90%	Iron/Steel (Retrofit) 99%	Iron/Steel (Retrofit) 90%
PERFORMANCE						
Capacity Factor	85%	85%	85%	85%	85%	85%
Representative Plant Size	87,000 tonnes H_2/year	87,000 tonnes H_2/year	1.29 M tonnes cement/year	1.29 M tonnes cement/year	2.54 M tonnes steel/year	2.54 M tonnes steel/year
CO_2 Captured (at 85% CF), tonnes/year[A]	340,550	309,548	1,017,920	925,793	3,145,352	2,860,681
CO_2 Captured (at 85% CF), tonnes/hour	39	35	116	106	359	327
CO_2 Compressor Load, kW	3,470	3,160	10,460	9,570	32,330	29,410
Cooling Water Flowrate, gpm	11,367	9,757	50,096	46,356	154,873	143,309
Cooling Tower Duty, MMBtu/hour	11	10	20	18	61	56
COST						
TPC, $/1,000	130,630	127,184	338,949	322,871	958,530	878,803
BEC	85,303	82,950	220,519	210,137	621,718	571,122
Home Office Expenses	14,928	14,516	38,591	36,774	108,801	99,946
Project Contingency	21,772	21,197	56,491	53,812	159,755	146,467
Process Contingency	8,627	8,520	23,348	22,148	68,257	61,268
TOC, $M	159	155	414	394	1,160	1,064
TOC, $/1,000	159,244	154,978	413,960	394,192	1,160,313	1,063,524
Owner's Costs	28,614	27,794	75,011	71,320	201,783	184,720
TASC, $/1,000	164,929	160,510	436,252	415,418	1,266,188	1,160,567
Capital Costs, $/tonne CO_2	21.3	22.8	21.8	22.8	27.8	28.0
Fixed Costs, $/tonne CO_2	14.4	15.6	10.6	11.1	9.3	9.5
Variable Costs, $/tonne CO_2	5.1	5.3	5.9	6.1	5.6	5.7
Purchased Power and/or Fuel, $/tonne CO_2	16.5	16.2	22.6	22.6	22.6	22.6
COC (ex. T&S), $/tonne CO_2	**57.3**	**59.9**	**60.8**	**62.7**	**65.4**	**65.9**

[A]Due to simplification of BFDs and stream tables throughout the body of the report where minor process streams are omitted, actual CO_2 captured as calculated in summary tables may be slightly less than that calculated at the capture rates applied in each case. This is due primarily to trace amounts of CO_2 entrained in water vapor generated during dehydration. Such differences, where they appear, are not expected to have any meaningful impact on the key results of this study, as they account for less than 1 percent of the CO_2 generated by the emitter.

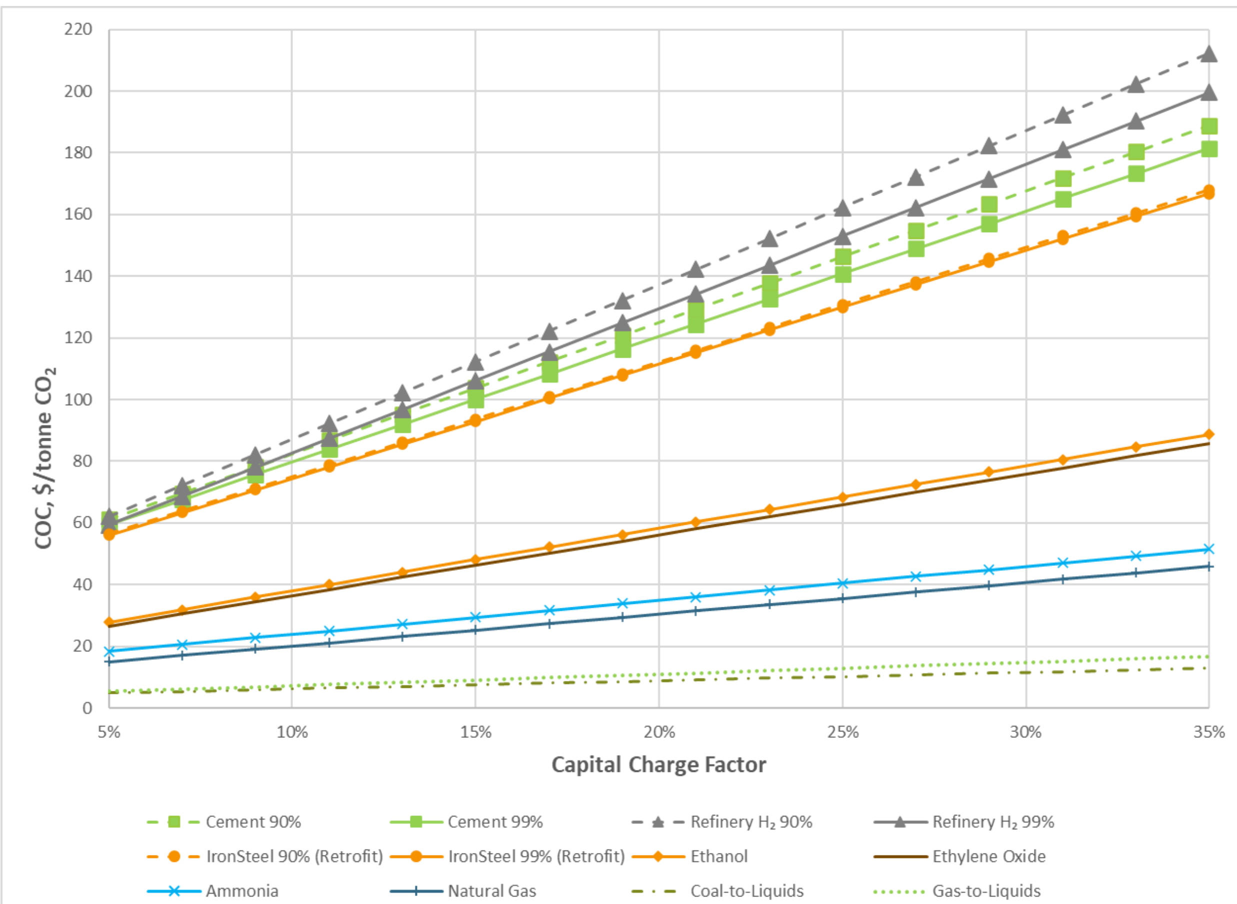
7.2 Sensitivity Analyses

In addition to the sensitivity analyses regarding plant capacities presented throughout Section 5 and Section 6 for each case, evaluations of the COC effects of varying assumptions made in this report are presented in this section.

7.2.1 Capital Charge Factor

The CCFs used to estimate the capital portion of the COC for each case were determined by the NETL Energy Markets Analysis Team and are market dependent. The financial assumptions are detailed in Section 3.2, but those factors could vary depending on economic conditions, among other aspects. For instance, changing payback period assumptions (i.e., 20-year payback period instead of 30-year), debt-to-equity ratios, rates of return and taxes could each affect the capital charge factor. Ultimately, the result of the financial assumptions would be applied as the capital charge factor. As such, the COC for each case was evaluated across a range of CCFs of 5–35 percent (Exhibit 7-4).

Exhibit 7-4. COC vs. CCF



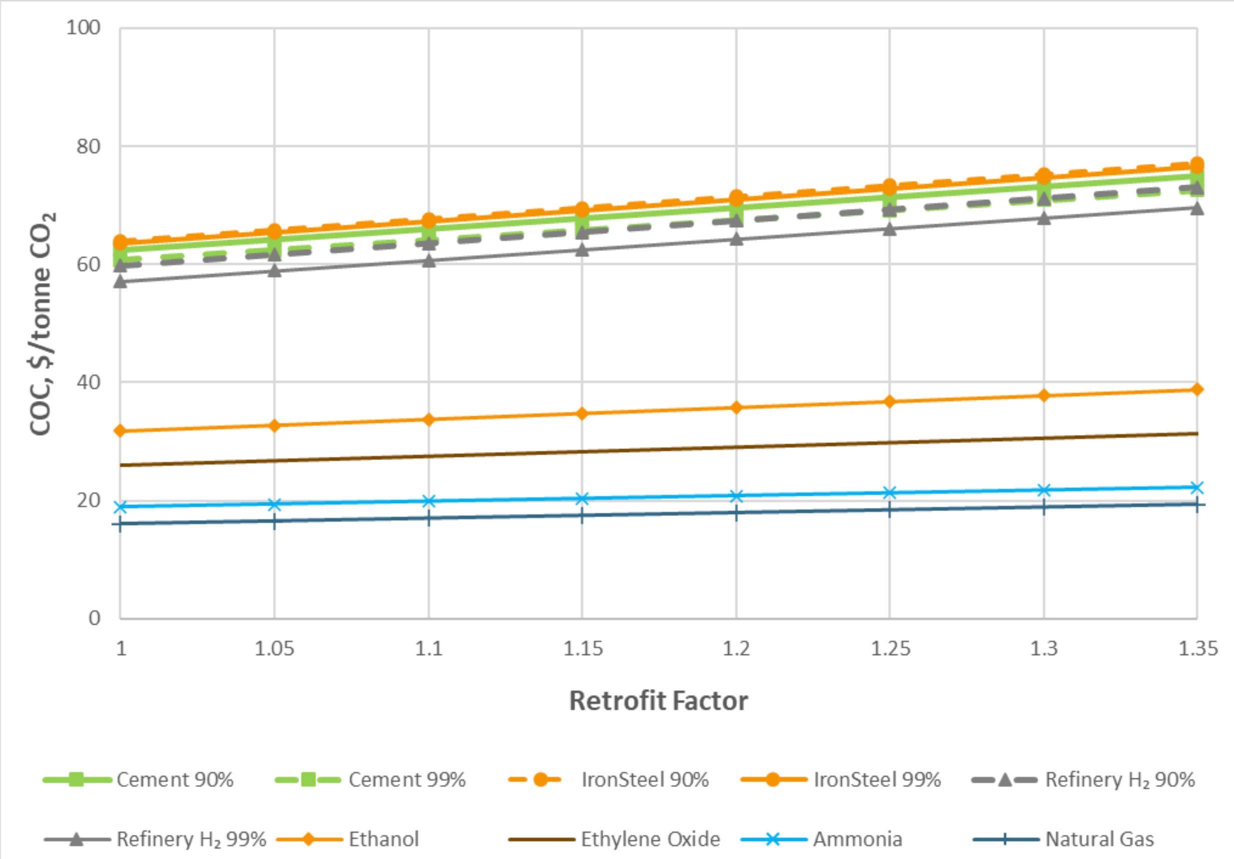
The results show that changing financial assumptions can have a very large effect on the COC. In the high purity cases, the largest change when varying the CCF over a range of 5–35 percent is observed in the ethanol case, where an increase of \$60.9/tonne CO_2 is noted. In the low purity cases, the effect is larger, as the low purity cases require more capital investment due to the need for AGR equipment. The largest COC increase in the low purity cases when varying the CCF occurs in the refinery hydrogen cases, where a \$140.3/tonne CO_2 change in the COC is observed for the 99 percent capture case and a \$150.2/tonne CO_2 increase is noted in the 90 percent capture case.

The CCFs used for the high purity and low purity cases, details of which have been given previously in Section 3.2, are representative of a project-specific CCF in each individual industrial sector. In addition to the industrial sectors' market influences on CCF, the maturity of a technology, specifically a capture technology like the AGR units employed in this study, may also affect the CCF. As capture systems are becoming more prevalent, and the project learning curve has improved, the low end of the CCF sensitivity curve demonstrated in this analysis may be a more reasonable representation.

7.2.2 Retrofit Factor

The retrofit factors used to estimate retrofit COC for each case, excluding CTL and GTL, were applied as a multiplier to TPC. The basis for this methodology is detailed in Section 3.3, but such an overall retrofit factor could vary depending on installation specifics, technology considerations, existing site constraints, and other determinants. As such, the COC for each case was evaluated across a retrofit factor range of 1.0–1.35, where the values corresponding to a 1.0 retrofit factor are indicative of a greenfield COC in each case (Exhibit 7-5).

Exhibit 7-5. COC vs. retrofit factor

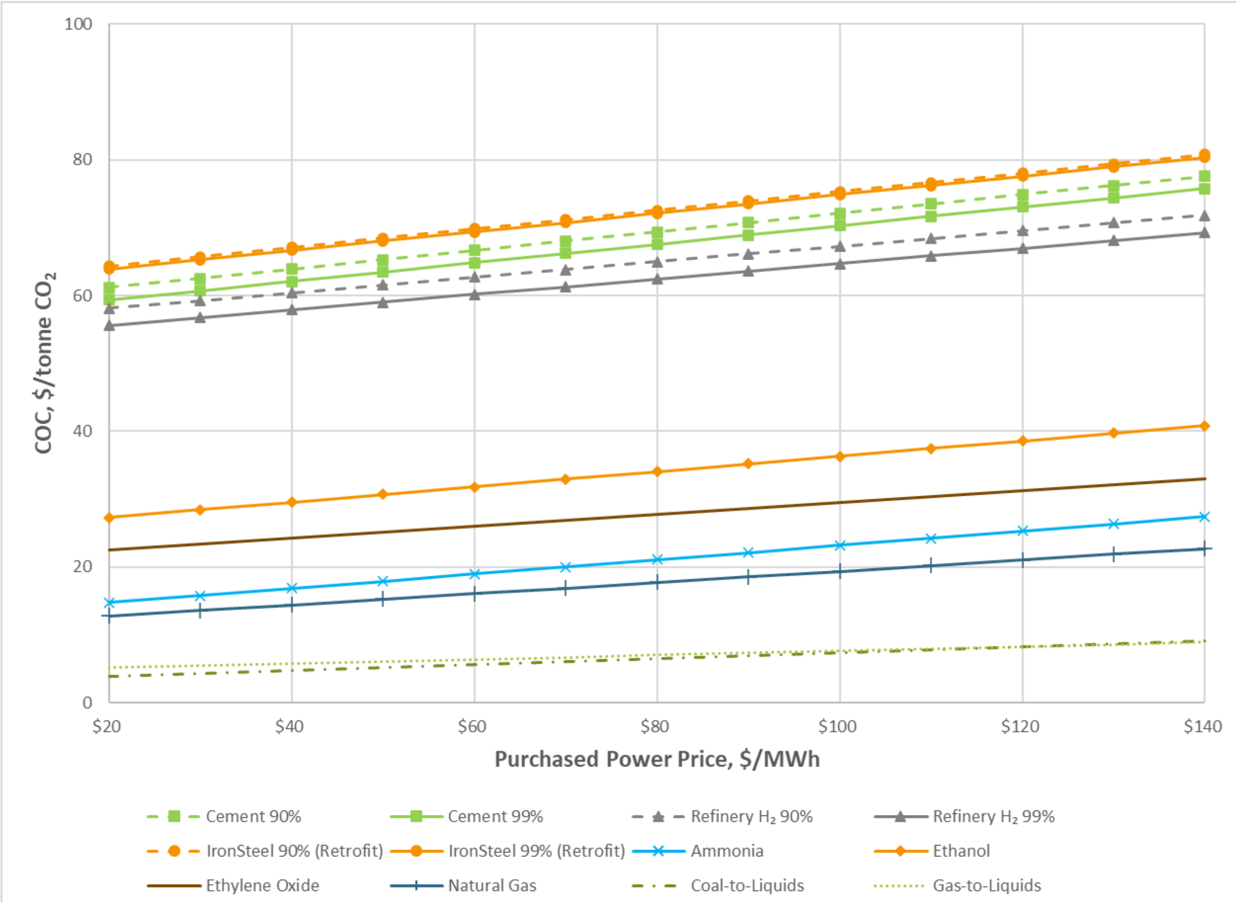


Because the retrofit factors in this study are applied as a multiplier to TPC, the effect of varying those factors across a range of values is an increasing COC with increasing retrofit factor for all cases. An interesting observation from this sensitivity analysis is the differing slopes of the lines between the low purity and high purity cases, meaning that the retrofit factors applied do not have equal magnitude of effect on all cases. For instance, the change in COC for the high purity cases ranged \$3.3–7.1/tonne CO_2 with increasing retrofit factor, whereas that of the low purity cases ranged \$11.9–13.3/tonne CO_2. This is due to the higher capital costs required for purifying the CO_2 prior to compression creating a larger TPC, which is the figure that is affected by the addition of the retrofit difficulty factor.

7.2.3 Purchased Power Price

The purchased power cost for each case is directly dependent upon the purchased power price assumed. For each case, a \$60/MWh price was used to estimate the purchased power costs, but price can vary widely depending upon market scenario, location, economic conditions, fuel pricing, and more. As such, the total COC for each case was estimated across a range of \$20–140/MWh purchased power price. Purchased power price increase has the most dramatic effect in the cement and iron/steel cases, where an increase of \$16.4/tonne CO_2 is observed across the sensitivity range (Exhibit 7-6).

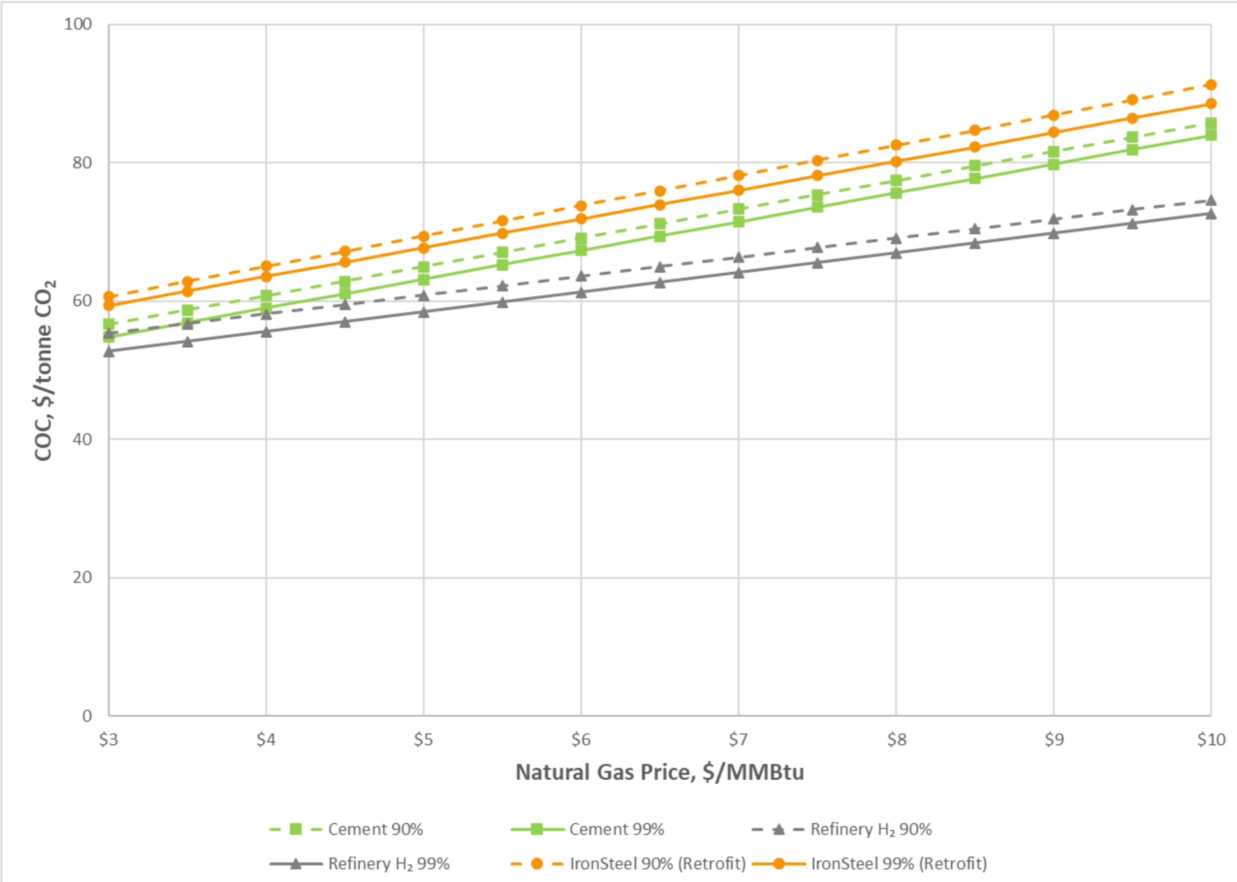
Exhibit 7-6. COC vs. purchased power price



7.2.4 Natural Gas Price

The fuel cost required for the industrial boiler in each low purity case is directly dependent upon the NG price assumed. For each case, $4.42/MMBtu was used for the NG price but can vary widely depending upon market scenario, location, economic conditions, fuel availability, oil prices, and more. As such, the total COC for each case was estimated across a fuel price range of $3–10/MMBtu. NG price increase has the most dramatic effect in the iron/steel 90 percent capture case, where an increase of $30.6/tonne CO_2 is observed across the sensitivity range (Exhibit 7-7).

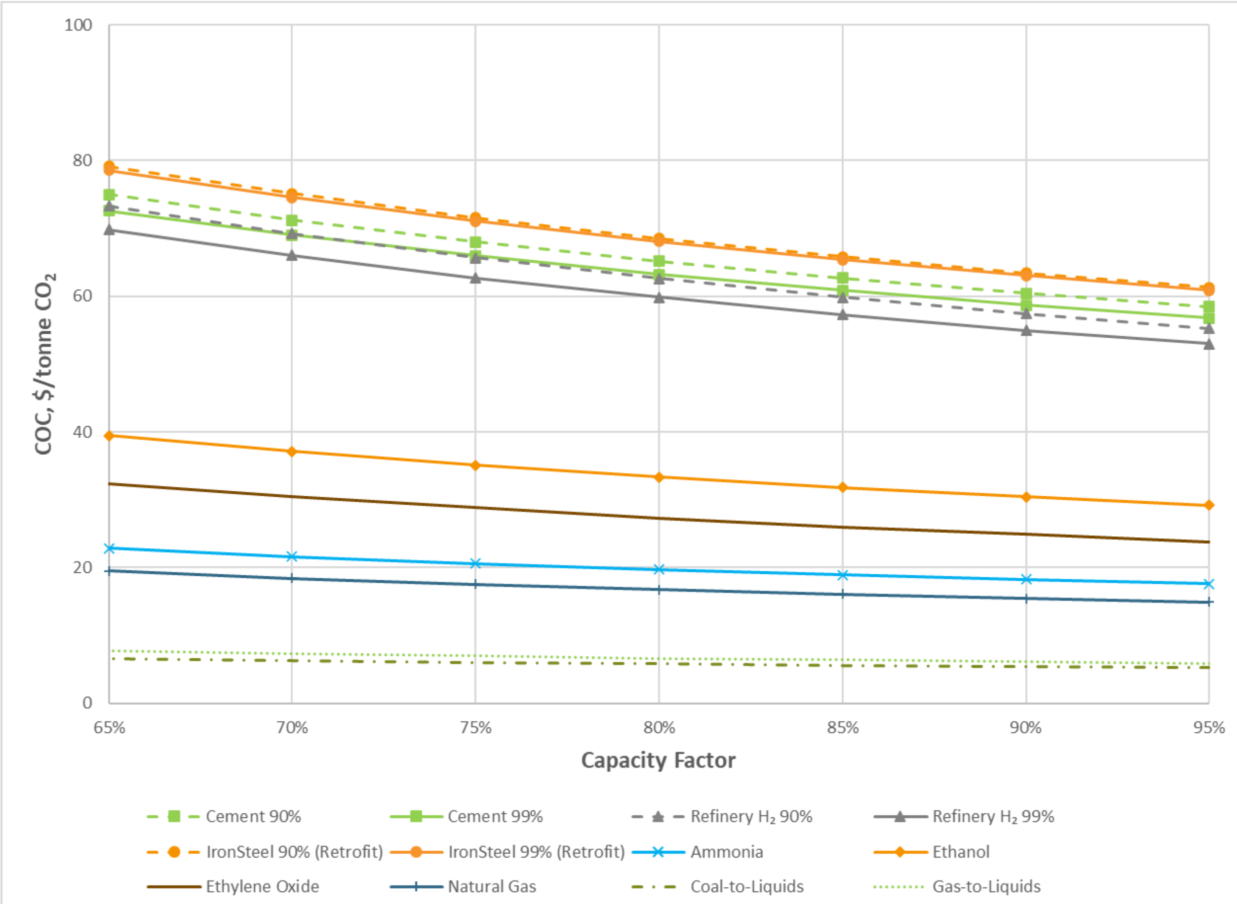
Exhibit 7-7. COC vs. NG price



7.2.5 Capacity Factor

Average capacity factors at industrial plants are variable, due to market fluctuations, differences in production cycles, operational upsets and planned shutdown requirements, regulatory constraints, and more. An 85 percent CF was assumed for the cases in this study, but it is important to consider how CFs affect the COCs calculated in this analysis. As CF varies from 65 to 95 percent, the COC for each case decreases, most notably in the Refinery H_2 90 percent capture case where a $18.0/tonne CO_2 decrease is observed across the sensitivity range.

Exhibit 7-8. COC vs. CF



8 CONCLUSION

Nine different industrial sources were examined in this study: ammonia, EO, ethanol, NGP, CTL, GTL, refinery hydrogen, cement, and iron/steel. Plant sizes were chosen based on different factors, including representative plant sizes expected to be built or already built in the industry (ammonia, refinery hydrogen), plant sizes representative of most of the production for the industry (ethanol, steel/iron, EO, cement), or plant sizes that would justify the addition of capture equipment (NGP). Plant sizes for CTL and GTL were determined based on those presented in previous NETL studies. Both greenfield and retrofit application costs were determined. The retrofit costs were derived by application of a retrofit factor to calculated total greenfield plant cost.

The results of this study show that CTL gives the lowest greenfield COC for the CO_2 product, a value of $5.6/tonne. This result is driven by the highly pure CO_2 sources produced from the CTL plant, as well as the largest amount of CO_2 available for capture across the cases considered. This combination of high availability coupled with high purity results in the lowest COC. The costliest option for capturing CO_2 in the group of industrial plants evaluated is iron/steel production, with a retrofit cost of $65.4/tonne CO_2 and $65.9/tonne CO_2 at 99 and 90 percent capture rates, respectively. The low purity CO_2 emission streams from iron and steel mills require purification equipment to attain EOR pipeline standards.

The greenfield COCs for the remaining cases fall in between the maximum and minimum cases as follows: GTL at $6.4/tonne, NGP at $16.1/tonne, ammonia at $19.0/tonne, EO at 26.0/tonne, ethanol at $31.8/tonne, refinery hydrogen with 99 percent capture at $57.3/tonne and with 90 percent capture at $59.9/tonne, and finally, cement at $60.8/tonne and $62.7/tonne for 99 and 90 percent capture, respectively. The assumed CO_2 concentrations for GTL, NGP, EO, ammonia, and ethanol were relatively high purity, either equivalent to or nearly the same purity as the lowest-COC CTL case. The reason for the increasing COC given similar purity is related to the amount of CO_2 available for capture, or economies of scale.

Economies of scale have a notable impact when comparing 99 and 90 percent capture rates in the low purity cases. On a normalized (i.e., $/tonne CO_2) basis, COC appears lower for higher capture rates in the refinery hydrogen, cement, and iron/steel analyses. This is also indicated in the plant size sensitivity analyses for each low purity case. As discussed in Section 6.1.8, Section 6.2.8, and Section 6.3.8, capital and O&M costs rise with increasing capture rates, but as there is more CO_2 captured, those costs result in a lower normalized costs at higher capture rates as presented. It is important to note that given the margin of error associate with the AACE Class 4 estimates applied in this study, and the margin of error assigned to the quotation from the capture system vendor (-25/+40 percent), the change in normalized cost from 90 to 99 percent is insignificant.

Sensitivity analyses of retrofit factor and purchased power price show minimal change in the COC for all cases. The most noticeable sensitivity effect is observed with plant size (economy of scale). For all cases, as the plant size is increased and, therefore, the amount of CO_2 available for capture increased, the COC decreased. The largest effect is observed with the iron/steel plant size sensitivity, where the COC increased by $36.9/tonne CO_2 and $37.6/tonne CO_2, for 99 and

90 percent capture cases, respectively, was observed when plant size was varied over the range of 0.5–6.8 M tonnes of steel per year. The base case production was 2.54 M tonnes of steel per year. All sensitivity analyses were evaluated in isolation, and it is possible that if individual design assumption changes were considered in combination, impacts on the COCs would potentially differ from the additive values of each change in design assumption.

CO_2 purity, as expected, plays a large role in the normalized COC; however, the amount of CO_2 and, therefore, the varying economies of scale from one industrial process to another, also has a dramatic effect on the cost of capturing CO_2. This analysis evaluated potential decarbonization opportunities in representative industrial plant applications, and the results show that capturing CO_2 can be cost-effective in the industrial sector, especially when a facility has two specific emissions stream characteristics: 1) high CO_2 purity so that further purification is not required, and (2) large amounts of CO_2 available.

9 FUTURE WORK

Future work in this area should look to plants with the characteristics of relatively high CO_2 purity and large CO_2 supply to expand upon the findings in the report. Potential recommendations include plants where CO_2 removal is inherent to the base plant process. A perfect example of this is ammonia and urea production, where not only is CO_2 removal crucial for maximizing ammonia synthesis loop efficiency and, therefore, production, but also reuse of the CO_2 for producing urea justifies this removal and recycle. The following items are potential future work that could expand on the analysis presented in this study.

9.1 IN-DEPTH PROCESS ANALYSIS

There are several opportunities where the results herein could be used as a starting point for a more in-depth analysis of the industries covered in this study. For example, the ammonia case does not account for in calculations how the base ammonia plant might allocate CO_2 for reuse in the urea or other derivative production processes. In addition, lesser products such as food-grade liquid CO_2, presumably captured from the high purity stripping vent point source, may also affect the amount of CO_2 available for capture from any one plant. The potential for food-grade liquid CO_2 also appears in the literature as an option for ethanol plants. These types of lesser-known factors could be investigated to better frame the amount of CO_2 available from different industries.

In addition to alternate CO_2 uses in the base plants, heat integration opportunities may exist, especially in greenfield cases or in plants where combined heat and power systems are in place or considered in the plant design. In retrofit cases, heat integration opportunities might increase retrofit difficulty factors, affecting capital expenditures, but lessening O&M costs. The heat requirements of the capture systems employed in the low purity cases analyzed in this study elicit the need for a standalone boiler, as discussed in Section 4.3. The flue gas from this NG-fired boiler contains additional CO_2 emissions over that of the base process, which were not captured based on the assumptions made in this analysis. Future work might consider an additional capture process or a mixing of this flue gas stream with the base plant emissions source to reduce those greenhouse gas emissions necessary for steam generation. Such scenarios may be evaluated with a more in-depth process analysis.

9.2 MULTIPLE PROCESS SCENARIO

Many chemical plants have two or more of the processes discussed in this analysis at the same industrial facility location. This could decrease the cost for CO_2 capture and make some processes more feasible when combined with others. Combining processes could be viewed from the perspective of mixing flue gas streams to take advantage of the economy of scale of building a single, larger capture unit, versus multiple smaller units, or from the perspective of combining CO_2 product streams in a larger trunk line to limit transport costs. Transport costs were not considered in this study.

9.3 Additional Processes

Methanol and a variety of other commodity chemical manufacturing facilities could be potential processes for consideration, assuming appropriate feedstock to justify capture. Additionally, as mentioned in Section 6.1, the fluid catalytic cracking unit at refineries is another viable point source for CO_2 capture. This may be investigated separately, or it could be included as a multiple process scenario, where the fluid catalytic cracking unit and the refinery hydrogen unit are combined to take advantages of economies of scale.

Another means of hydrogen production that could be considered for decarbonation is hydrogen from coal gasification. NETL recently evaluated the cost of capturing CO_2 in hydrogen production via gasification applications as part of the report "Comparison of Commercial, State-of-the-Art, Fossil-Based Hydrogen Production Technologies." [40] Lastly, only the BOF steel plant configuration was considered in this study, but EAF plants make up 32 percent of steel production in current industry and are expected to be the only greenfield steel plants to be constructed. An analysis of EAF steel production for decarbonization would likely be impactful.

9.4 Techno-economic Analysis of CO_2 Distribution to EOR Fields

As stated previously in Section 4.1.2, pressures as low as 1,200 psig may be acceptable for EOR field usage. Reducing the pressure to which CO_2 needs to be compressed would reduce the COC. A reduction in pressure would result in a lower compressor capital cost, as well as reduced power consumption resulting in a lower cost associated with purchasing power from the grid. The economics of CO_2 transport with the existing pipeline infrastructure was not part of this analysis but does contribute to the true COC.

9.5 Life Extension Costs for Existing Facilities

The implicit assumption for the cases presented in this report is that the plants that have been retrofitted (i.e., cement, steel, etc.) have sufficient remaining life, such that the base plant remaining life will match the expected life of the retrofitted equipment (i.e., capture system, compression), assumed to be 30 years. This study does not consider, or include any costs to represent, life extension projects that a plant (i.e., a cement plant) may consider if adding capture and compression. Future work could include an analysis to identify the average age of the various industry's plants, characterize the standard expected life for these plants by industry, and characterize the cost of typical life extension projects that would be considered as part of a capture retrofit. This would allow for a more complete cost for a retrofit project, when considering factors outside of just the capture and/or compression equipment.

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APPENDIX: CARBON BALANCES

Note: All convergence tolerance values in the tables within this appendix are calculated by difference.

The carbon balance[i] for the ethanol case is shown in Exhibit A-1.

Exhibit A-1. Ethanol case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		**kg/hr (lb/hr)**
Fermentation Stream	4,457 (9,825)	CO_2 Captured Stream	4,457 (9,825)
Total	**4,457 (9,825)**	**Total**	**4,457 (9,825)**

The carbon balance for the ammonia case is shown in Exhibit A-2.

Exhibit A-2. Ammonia case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		**kg/hr (lb/hr)**
Stripping Vent	15,149 (33,398)	CO_2 Captured Stream	15,140 (33,379)
		TEG Vent	9 (19)
Total	**15,149 (33,398)**	**Total**	**15,149 (33,398)**

The carbon balance for the natural gas processing (NGP) case is shown in Exhibit A-3.

Exhibit A-3. NGP case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		**kg/hr (lb/hr)**
Stripping Vent	20,266 (44,590)	CO_2 Captured Stream	20,221 (44,581)
		TEG Vent	4 (9)
Total	**26,266 (44,590)**	**Total**	**26,226 (44,590)**

The carbon balance for the ethylene oxide (EO) case is shown in Exhibit A-4.

[i] Carbon balances may show carbon content of minor process streams, including the CO_2 entrained in the water vapor vent from the TEG dehydration system and CO_2 entrained in process water knockouts, that are not represented in the block flow diagrams throughout the report body. These process streams were omitted from the report body for simplicity and brevity. Cases where this simplification applies include ammonia, NGP, refinery H_2, iron/steel, and cement.

Exhibit A-4. EO case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Rectisol Stream	3,785 (8,345)	CO_2 Captured Stream	3,785 (8,345)
Total	**3,785 (8,345)**	**Total**	**3,785 (8,345)**

The carbon balance for the coal-to-liquids (CTL) case is shown in Exhibit A-5.

Exhibit A-5. CTL case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Gasification AGR Unit	110,862 (244,411)	CO_2 Captured Stream	272,397 (600,525)
FT AGR Unit	161,536 (356,114)		
Total	**272,397 (600,525)**	**Total**	**272,397 (600,525)**

The carbon balance for the gas-to-liquids (GTL) case is shown in Exhibit A-6.

Exhibit A-6. GTL case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Stripping Vent	57,905 (127,665)	CO_2 Captured Stream	57,905 (127,665)
Total	**57,905 (127,665)**	**Total**	**57,905 (127,665)**

The carbon balance for the refinery hydrogen case with 99 percent capture is shown in Exhibit A-7.

Exhibit A-7. Refinery hydrogen case with 99 percent capture carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
SMR Off-Gas Stream	16,102 (35,499)	CO_2 Captured Stream	12,480 (27,513)
Amine Recycle	405 (893)	TEG Vent	1 (2)
		Gas to PSA	3,543 (7,812)
		Recycle	378 (832)
		Process Knockout Entrainment	106 (233)
Total	**16,507 (36,392)**	**Total**	**16,507 (36,392)**

The carbon balance for the refinery hydrogen case with 90 percent capture is shown in Exhibit A-8.

Exhibit A-8. Refinery hydrogen case with 90 percent capture carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
SMR Off-Gas Stream	16,102 (35,499)	CO_2 Captured Stream	11,343 (25,008)
Amine Recycle	368 (811)	TEG Vent	6 (14)
		Gas to PSA	4,675 (10,307)
		Recycle	378 (832)
		Process Knockout Entrainment	67 (149)
Total	**16,470 (36,310)**	**Total**	**16,470 (36,310)**

The carbon balance for the iron/steel case coke oven gas (COG)/blast furnace stove (BFS) stream with 99 percent capture is shown in Exhibit A-9.

Exhibit A-9. Iron/steel case COG/BFS stream with 99 percent capture carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
COG Stream	27,380 (60,363)	CO_2 Captured Stream	57,475 (126,710)
BFS Stream	30,704 (67,690)	TEG Vent	10 (23)
		Clean Flue Gas	599 (1,320)
Total	**58,084 (128,053)**	**Total**	**58,084 (128,053)**

The carbon balance for the iron/steel case COG/BFS stream with 90 percent capture is shown in Exhibit A-10.

Exhibit A-10. Iron/steel case COG/BFS stream with 90 percent capture carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
COG Stream	27,380 (60,363)	CO_2 Captured Stream	52,273 (115,242)
BFS Stream	30,704 (67,690)	TEG Vent	9 (21)
		Clean Flue Gas	5,802 (12,790)
Total	**58,084 (128,053)**	**Total**	**58,084 (128,053)**

The carbon balance for the steel case COG power plant stack (PPS) stream with 99 percent capture is shown in Exhibit A-11.

Exhibit A-11. Steel case COG PPS stream with 99 percent capture carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
COG PPS Stream	58,400 (128,751)	CO_2 Captured Stream	57,788 (127,400)
		TEG Vent	10 (23)
		Clean Flue Gas	602 (1,328)
Total	**58,400 (128,751)**	**Total**	**58,400 (128,751)**

The carbon balance for the steel case COG PPS stream with 90 percent capture is shown in Exhibit A-12.

Exhibit A-12. Steel case COG PPS stream with 90 percent capture carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
COG PPS Stream	58,400 (128,751)	CO_2 Captured Stream	52,558 (115,870)
		TEG Vent	9 (21)
		Clean Flue Gas	5,833 (12,860)
Total	**58,400 (128,751)**	**Total**	**58,400 (128,751)**

The carbon balance for the cement 99 percent capture case is shown in Exhibit A-13.

Exhibit A-13. Cement 99 percent capture case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Kiln Off-Gas Stream	37,697 (83,108)	CO_2 Captured Stream	37,302 (82,237)
		TEG Vent	7 (15)
		Clean Flue Gas	389 (857)
Total	**37,697 (83,108)**	**Total**	**37,697 (83,108)**

The carbon balance for the cement 90 percent capture case is shown in Exhibit A-14.

Exhibit A-14. Cement 90 percent capture case carbon balance

Carbon In		Carbon Out	
	kg/hr (lb/hr)		kg/hr (lb/hr)
Kiln Off-Gas Stream	37,697 (83,108)	CO_2 Captured Stream	33,926 (74,794)
		TEG Vent	6 (13)
		Clean Flue Gas	3,765 (8,301)
Total	**37,697 (83,108)**	**Total**	**37,697 (83,108)**

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