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Cost of Capturing CO_2 from Industrial Sources

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Disclaimer

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Acronyms and Abbreviations

AACE	Association for the Advancement of Cost Engineering
ADM	Archer Daniel's Midland
ADNOC	Abu Dhabi National Oil Company
AGR	Acid gas removal
bbl	Barrel
Bcf	Billion cubic feet
BEC	Bare erected cost
BFD	Block flow diagram
BFG	Blast furnace gas
BFS	Blast furnace stove
BOF	Basic oxygen furnace
BOS	Blown oxygen steelmaking
Btu	British thermal unit
B Btu	Billion British thermal units
Btu/hr	British thermal units per hour
Btu/kWh	British thermal units per kilowatt hour
Btu/lb	British thermal units per pound
CCF	Capital charge factor
CCS	Carbon capture and sequestration
CFPP	Coal-fired power plant
CO_2	Carbon dioxide
COE	Cost of electricity
COG	Coke oven gas
CTL	Coal-to-liquids
DDGS	Dried distiller grain solids
DOE	Department of Energy
EAF	Electric arc furnace
EF	Emission factor
EIA	Energy Information Administration
EO	Ethylene oxide
EOR	Enhanced oil recovery
EPA	Environmental Protection Agency
EPC	Engineering/procurement/ construction
FCC	Fluid catalytic cracking
FGD	Flue gas desulfurization
FT	Fischer-Tropsch
ft	Foot, Feet
G	Giga
gal	Gallon
GJ	Gigajoule
GJ/hr	Gigajoules per hour
GTI	Gas Technology Institute
GTL	Gas-to-liquids
h, hr	Hour
H_2	Hydrogen
HHV	Higher heating value
IEA	International Energy Agency
IEA GHG	International Energy Agency's Greenhouse Gas R&D Program
kg/GJ	Kilograms per gigajoule
kg/hr	Kilograms per hour
kgmol	Kilogram mole
kgmol/hr	Kilogram moles per hour
kJ	Kilojoule
kJ/hr	Kilojoules per hour
kJ/kg	Kilojoules per kilogram
kPa	Kilopascal
kW, kWe	Kilowatt electric
kWh	Kilowatt-hour
kWt	Kilowatt thermal
lb	Pound
lb/hr	Pounds per hour
lb/ft^2	Pounds per square foot
lb/MMBtu	Pounds per million British thermal units
lbmol	Pound mole
lbmol/hr	Pound moles per hour
lb/MWh	Pounds per megawatt hour
lb/TBtu	Pounds per trillion British thermal units
LP	low pressure
LNG	Liquefied natural gas
MEA	Monoethanolamine
MDEA	Methyldiethanolamine
M	Million
MMBtu	Million British thermal units
MMBtu/hr	Million British thermal units per hour
MMCFD	Million cubic feet per day
MMkJ	Million kilojoules (also shown as 10^6 kJ)
MMkJ/hr	Million kilojoules (also shown as 10^6 kJ) per hour
MMscf	Million standard cubic feet
MMscfd	Million standard cubic feet per day
MMscfy	Million standard cubic feet per year
MPa	Megapascal
MW,MWe	Megawatt electric
MWh	Megawatt-hour
MWt	Megawatt thermal

NETL	National Energy Technology Laboratory
NGCC	Natural gas combined cycle
NGP	Natural gas processing
O&M	Operation and maintenance
PC	Portland cement
PCC	Post-combustion capture
PPS	Power plant stack
PSA	Pressure swing absorption
psi	Pounds per square inch
psia	Pound per square inch absolute
psid	Pound per square inch differential
psig	Pound per square inch gage
QGESS	Quality Guidelines for Energy System Studies
ROE	Required return on equity
SCR	Selective catalytic reduction
SMR	Steam methane reformer
SR	Steam reformer
TASC	Total as-spent cost
Tcf	Trillion cubic feet
TOC	Total overnight cost(s)
Ton	Short tons
Tonne	Metric tonnes
TPC	Total plant cost
U.S.	United States
USD	U.S. dollars
USDA	U.S. Department of Agriculture
USGS	U.S. Geological Survey
°C	Degrees Celsius
°F	Degrees Fahrenheit

Executive Summary

The objective of this National Energy Technology Laboratory (NETL) study is to evaluate the costs associated with capturing carbon dioxide (CO_2) from industrial processes. Only those processes with effluent streams containing CO_2 concentrations higher than those found in flue gas from coal-fired power plants (CFPP, ~17 mol %) were considered. The following nine processes were chosen due to either the high purity or large quantity of CO_2 available. The processes are categorized based on CO_2 purity.

High Purity Sources (≥ 90 vol %)

Ethanol
Ammonia
Natural Gas Processing
Ethylene Oxide (EO)
Coal-to-Liquids (CTL)
Gas-to-Liquids (GTL)

Low Purity Sources (< 90 vol. %)

Hydrogen (Refinery)
Iron/Steel
Cement

The key features and results for a representative size reference plant are shown in Exhibit ES-1. The cost metric of interest is the breakeven cost of capturing CO_2, in $/tonne. This cost represents the CO_2 selling price that is required for the base plant to recover all of the costs associated with implementing CO_2 separation (where applicable), purification, and compression. The primary market for the CO_2, on this scale, is enhanced oil recovery (EOR) and in that context this breakeven cost represents the minimum plant gate selling price for the CO_2. If the captured CO_2 were to be transported and stored in a saline formation, then costs associated with those activities would need to be added.

A reference plant size was chosen for each process, and factors for the amount of CO_2 generated per amount of product produced were applied based on literature sources for six of the nine processes studied. For each industrial source, the CO_2 separation process (required for low purity sources) and CO_2 compression (all cases) were modeled using Aspen Plus® (Aspen) to generate the material and energy balance data needed for cost estimating. The assumed CO_2 separation process utilizes a methyldiethanolamine (MDEA) solvent. Capital and operating and maintenance (O&M) costs for both greenfield and retrofit facilities were calculated. The costs shown here are for retrofit applications except for CTL and GTL for which there are no existing facilities in the U.S. Greenfield costs are estimated to be no more than 5% less than the retrofit costs. Additional details on reference plant size and cost estimating methodology are in the body of the report.

The results show that point source purity is the primary factor that influences the cost of CO_2 captured. Within the purity groupings, partial pressure and economy of scale account for the observed cost variability. Since the high purity sources are assumed to be of adequate quality to be pipeline-ready without including CO_2 separation or purification, those sources derive a significant cost advantage over the low purity sources.

Exhibit ES-1 Key features and breakeven cost of capturing CO_2 for reference industrial process plants

Industrial Process	Reference Plant Capacity	CO_2 Source Stream	CO_2 to Product Ratio (tonne CO_2/tonne Product)	Source Stream CO_2 Concentration (mol%)	Source Stream CO_2 Partial Pressure (psia)	CO_2 Available for Capture (M tonnes CO_2/year)		Breakeven Cost of Capturing CO_2 (\$/tonne CO_2)
						Reference Plant	All U.S. sources	
High Purity Sources								
Ethanol	*50 M gal/year*	Distillation gas	0.96	100	18.4	0.14	40	30
Ammonia	*907,000 tonnes/year*	Stripping vent	1.9	99	22.8	0.458	6	27
Natural Gas Processing	*500 MMscf/d*	CO_2 vent	N/A[1]	99	23.3	0.649	27	18
Ethylene Oxide	*364,500 tonnes/year*	AGR product stream	0.33	100	43.5	0.122	1	25
Coal-to-Liquids (CTL)	*50,000 bbl/d*	AGR product stream	N/A[2]	100	265	8.74	-	9
Gas-to-Liquids (GTL)	*50,000 bbl/d*	AGR product stream	N/A[2]	100	265	1.86	-	9
Low Purity Sources								
Refinery Hydrogen	*59,000 tonnes/year*	PSA tail gas	10.5	44.5	8.9	0.274	68	118
Iron/Steel	*2.54 M tonnes/year*	Plant Total COG PPS COG/BFG[3]	2.2	N/A 23.2 26.4	N/A 3.4 3.9	3.9 2.75 1.16	49	99 99 101
Cement *SCR/FGD Sensitivity*	*992,500 tonnes/year*	Kiln Off-gas	1.2	22.4	3.3	1.14	80	100 *127*
Coal-fired power plants	*550 MW*	*Flue Gas*	*NA*	*13.5*	*2.0*	*4.13*	*2,545*[4]	*77*[5][6]

[1] *The factor for natural gas processing varies significantly due to varying raw natural gas CO_2 content.*

[2] *Due to multiple products being produced in CTL and GTL plants, a simple ratio was not employed*

[3] *Combined coke oven gas and blast furnace gas*

[4] *Represents total CO_2 available from all plant sizes at any breakeven cost. Approximately 1400 M tonne of CO_2/yr, of the 2,545 M tonne/yr total, is available at or below the \$77/tonne /CO_2 cost.*

[5] *Based on a utility finance structure resulting in a lower capital charge factor (0.11) compared to that used for industrial sources (0.15 and 0.17 for the high purity and the lower purity respectively)*

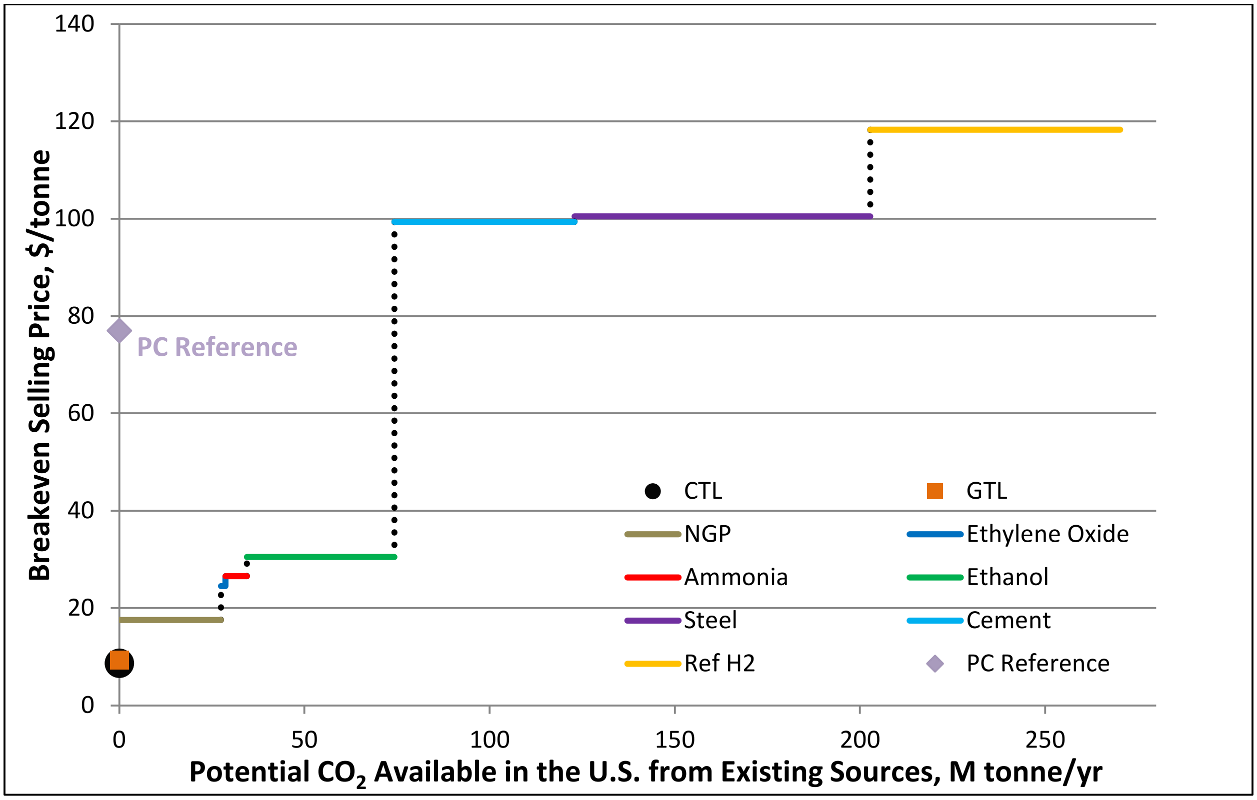
[6] *Assuming capacity factor of 75% and retrofit factor of 1.1*

Two breakeven costs are also given for the cement process – a base cement case and a sensitivity case incorporating SO_2 and NO_x controls. The base cement case assumes that the kiln off-gas is suitable to be sent directly for CO_2 separation; however, a literature search suggested that the kiln off-gas may have higher-than-acceptable levels of SO_2 and NOx, and would require the addition of selective catalytic reduction (SCR) and flue gas desulfurization (FGD). Therefore, a sensitivity to this case was performed. The results show that the addition of SCR and FGD has the potential to add $25/tonne to the breakeven cost.

The capital component of the high purity sources ranges from $3 - $12/tonne of CO_2 captured whereas the capital component of the low purity sources ranges from $33 - $43/tonne CO_2. The difference is due to the need for separation and purification equipment in the low purity cases that is not required in the high purity cases. Additionally cost is added for regeneration steam in the CO_2 separation process. It was assumed that excess plant steam was unavailable for regeneration purposes, and that a standalone natural gas fired boiler was needed for steam production. At a price of $6.13/MMBtu, the natural gas fuel requirement adds $26 - $27/tonne CO_2 to the cost of the low purity sources. The low purity source partial pressure entering the compressor (19 psia) is less than the higher purity sources (23 – 265 psia, except for ethanol at 18 psia), which results in greater compression power and compression costs.

Shown in Exhibit ES-2 is a plot of breakeven selling price versus the cumulative domestic CO_2 supply. The breakeven selling prices are for retrofits, with the exception of CTL and GTL, which are assumed to be greenfield sites. The incremental range on the x-axis for each industry, and not the absolute values, represents the amount of CO_2 available. This plot shows the cost of the source relative to the potentially available supply. It also provides a visual of relative industry standing on the cost/supply curve. For example, while ammonia represents a low cost, it also represents a small supply relative to natural gas processing or ethanol, both of which also project a low cost. Availability from each industrial sector ranges from 1.22 M tonnes/year (ethylene oxide) to 79.8 M tonnes/year (cement). Based on breakeven selling price of existing processes, natural gas processing is the most attractive with a price of $18/tonne, and refinery hydrogen is the least attractive with a price of $118/tonne. CTL and GTL represent attractive options should the technologies be deployed domestically. The point representing a pulverized coal (PC) power plant with carbon capture (1) compares only the reported retrofit breakeven selling price to the industrial cases, and does not imply an amount of CO_2 available from domestic coal fired power units.

Exhibit ES-2 CO_2 Breakeven Selling Price versus CO_2 Supply



Note: The breakeven selling prices used are for a representative plant for each industry and do not account for the size variability and associated economies of scale for the actual sources in each industry.

Source: NETL

Sensitivities were performed on the cost of electricity (COE), plant size in terms of CO_2 emissions per year, and the financing assumptions (represented as the capital charge factor (CCF)). A sensitivity to natural gas price was also performed for the low purity cases. The general results of these sensitivities are as follows:

- As COE increases, the breakeven selling price 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 MDEA CO_2 separation process, the breakeven selling price increase is more dramatic.
- As plant size increases, and thus the CO_2 emissions increase, the breakeven selling price decreases as a result of improved economies of scale. The breakeven cost for the smallest plant is $17 to $110/tonne CO_2 above that of the largest plant in the same industrial sector (this excludes CTL and GTL).
- Financing assumptions are important as different industries may finance capital projects differently. The impact on the more highly capital intensive low purity cases is greater than on the less capital intensive high purity cases. For example, increasing the CCF from 10 percent to 35 percent increases the cost of capture for natural gas processing from $15 to $25/tonne CO_2 while the same CCF increase raises the CO_2 capture cost for refinery hydrogen from $95 to $155/tonne. This change in CCF is equivalent to a change in the weighted average cost of capital (WACC) from around 8% to 20%, assuming a 3 year

capital expenditure period. The reference CCF is 15% for the high purity cases and 17% for the lower purity cases.

- The natural gas price sensitivity impacts only the low purity cases. As natural gas price increases over the range $5-10/MMBtu, the breakeven CO_2 selling price rose as much as $22/tonne (observed in the steel case).

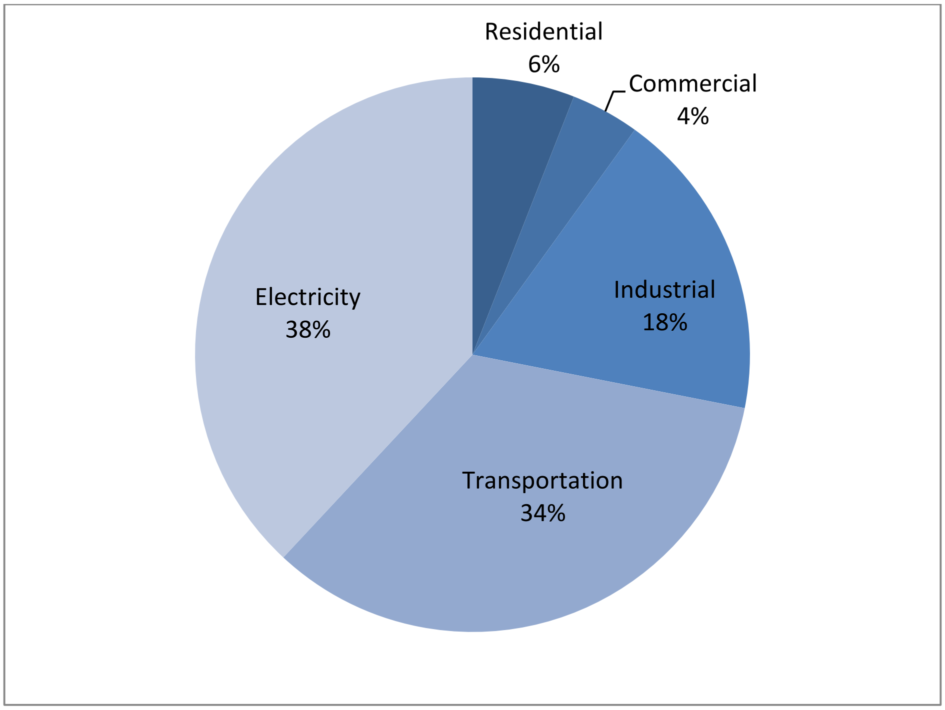
The plant size sensitivity results for each case can be found in the corresponding sections, and details of all other sensitivity results can be found in Section 9.

The study results show a significant CO_2 capture cost disparity between high purity sources that only require compression and low purity sources that require separation and compression. The cost of capture from a conventional utility-scale power plant is intermediate between the high purity and low purity source costs. The four domestically-active high purity industrial sources (ethanol, natural gas processing, ammonia and ethylene oxide) comprise 70 M tonnes/yr of CO_2 availability. Not every domestic plant is a candidate for capture because of scale issues or location relative to EOR fields and existing infrastructure. Demand for CO_2 for EOR beyond what can be satisfied by natural sources or low purity industrial sources will likely be covered by coal-fired power plants before low purity industrial sources would be considered.

1 Introduction

Carbon dioxide has been the focus of domestic greenhouse gas mitigation. The Environmental Protection Agency (EPA) estimates that CO_2 accounted for approximately 82% of all U.S. greenhouse gas emissions attributed to human activities. (2) From these activities, the Energy Information Administration (EIA) estimates that approximately 5.4 billion metric tons of CO_2 were released into the atmosphere in 2014. (3) Exhibit 1-1 illustrates the emission breakdown by sector. The three largest emitters, by sector, are electricity production, transportation, and industry.

Exhibit 1-1 Domestic CO_2 Emissions by Sector



Source: EIA Monthly Energy Review June 25, 2014

In the past, the primary focus of carbon mitigation has been within the electric power generation and transportation sectors through energy efficiency improvements, alternative fuels research, and pre and post-combustion capture research, development and demonstration (RD&D). The purpose of this study is to evaluate CO_2 mitigation opportunities within the industrial sector by applying technologies to separate and compress CO_2 from existing industrial processes.

The major incentive to implement carbon capture into any process is the opportunity to turn the captured CO_2 into a revenue stream. For this reason project developers have begun to look to industrial processes as sources of carbon dioxide (CO_2) to enable CO_2 enhanced oil recovery (EOR) projects. Utilizing industrial sources as a source of CO_2 gives developers the ability to tap highly concentrated CO_2 streams, thereby eliminating or reducing the cost of the separation equipment. This study looks at nine different industrial sources of CO_2 to determine the CO_2 selling price required to "breakeven" with the cost of capture and compression.

2 Assumptions

The following assumptions are used to narrow the field of viable processes. The basis for selection includes processes that: 1) are likely to be at the low end of a cost/supply curve, 2) are representative of a significant amount of supply, and 3) have point sources with CO_2 concentrations above 17 mole percent. Concentrations lower than 17 percent represent flue-gas stream equivalents, which are similar to coal-fired power plants (CFPP). The purpose of this study is to investigate industrial processes that may be more economical than the typical carbon capture systems employed at CFPP, and that are the subject of other National Energy Technology Laboratory (NETL) studies. This study defines an industrial process as any process representing a large amount of CO_2 supply, where power is not the sole product, with gas stream concentrations elevated above that of CFPP flue gas.

3 Literature Search and Results

Exhibit 3-1 shows the major processes identified in the literature search, including information on production rates and quantities of CO_2 produced. The values listed represent the average of all the literature results. These values should not be interpreted as representative of all plants for a specific industry. The literature sources discuss both existing plant and hypothetical plant studies, and use their own individual set of assumptions. Therefore, the values found during the literature search may not be directly relatable to the results of this study. Given the number of sources used to create Exhibit 3-1, in-text citation was not done here; rather, a full citation of the table is given in Appendix C.

Exhibit 3-1 Industrial CO_2 sources from literature search

Process	Average CO_2 Produced per Plant (kTonne/Yr)	Previous NETL Study	Total Estimated U.S. CO_2 Supply Potential (M tonne/year)
Ammonia	710	No	15.3
Coal-to-Liquids	1,445	Yes	*No U.S. Plants*
Ethanol	410	No	69.8
Ethylene Oxide	180	No	8.8
Gas-to-Liquids	*N/A*	Current	*No U.S. Plants*
Hydrogen Plant	600	Yes	100.0
Iron and Steel	7,150	No	105.1

Process	Average CO_2 Produced per Plant (kTonne/Yr)	Previous NETL Study	Total Estimated U.S. CO_2 Supply Potential (M tonne/year)
Natural Gas Processing	1,000	No	*N/A*
Refinery Hydrogen	22,108	No	28.6
Methanol	*N/A*	No	*Feedstock Dependent*
Cement	355	No	79.8

The estimated supply of CO_2 from ammonia plants was calculated by taking the 2006 Environmental Protection Agency (EPA) United States (U.S.) ammonia production value and multiplying by 1.87 tonne CO_2/tonne ammonia. (4) This accounts for all point sources, but does not account for CO_2 captured and used elsewhere in the ammonia plant. The estimated supply of CO_2 from iron and steel was calculated by taking the 2010 U.S. electric arc furnace (EAF) and basic oxygen furnace (BOF) production and multiplying each by their respective tonne CO_2/tonne steel production factors. This resulted in a range of CO_2 supply values for EAF. The median value was added to the BOF total to generate the total steel industry CO_2 supply. The estimated supply of CO_2 from ethanol plants was calculated using Illinois Stage Geological Survey equations for Ethanol plant CO_2 emissions from flue gas fuel burning and distillation off gas. Equations are based on 2012 plant capacity given by the Nebraska Government website that assumes all plants use corn as feedstock. The estimated supply of CO_2 from U.S. natural gas processing plants is difficult to calculate, and no generic CO_2 factor was found in the literature. The estimated supply of CO_2 from hydrogen plants was calculated by taking the median production factor (10.5 tonnes CO_2/tonne H_2 produced) and multiplying by the median H_2 production (9.5 M tons H_2/year). The estimated supply of CO_2 from ethylene oxide (EO) plants was calculated by multiplying 2009 U.S. ethylene oxide capacity by the process stoichiometry of 3:1 moles ethylene oxide:mole CO_2. No value could be estimated for CO_2 available from methanol plants. The estimated supply of CO_2 from refinery hydrogen locations was calculated by taking the 2006 Energy Information Agency (EIA) estimated U.S. refinery hydrogen production capacity and multiplying by the median value (10.5) of the tonnes CO_2/tonne H_2 produced factor. The estimated supply of CO_2 from cement plants was calculated by taking the 2010 United States Geological Survey (USGS) estimated cement production of 66.5 M tonnes of cement and multiplying by a production factor of 1.2 tonnes CO_2/tonne cement.

Methanol is discussed in the literature as a potential source of CO_2. It was not included here as a case study because of the reasons provided below. Seventy-five percent of methanol is produced using natural gas as a feedstock. (5) In this process, methane is reformed to H_2, and the amount of resulting H_2 is in excess of the methanol synthesis stoichiometry. Therefore, natural-gas-based methanol plants may capture CO_2 from dilute flue gas resulting from combustion and use it to boost methanol production. (5) However, when using a heavier feedstock, such as petcoke or coal, CO_2 is produced in excess of optimal methanol synthesis conditions, and addition of CO_2 to the synthesis mixture to boost methanol production proves to not be beneficial, but rather CO_2 must be removed from the syngas. Therefore, CO_2 available to be captured from methanol production would depend on the feedstock, with 75 percent of processes potentially not having abundant or high-purity sources available for capture. The Department of Energy (DOE) has awarded Leucadia funding to develop CO_2 capture from a petcoke-to-methanol facility in Louisiana. This project uses a heavier feedstock and, therefore, will provide suitable CO_2 to be captured. The project is expected to capture 4.5 M tons/year of CO_2 for EOR use at West Hasting's oil field, in Texas. (6)

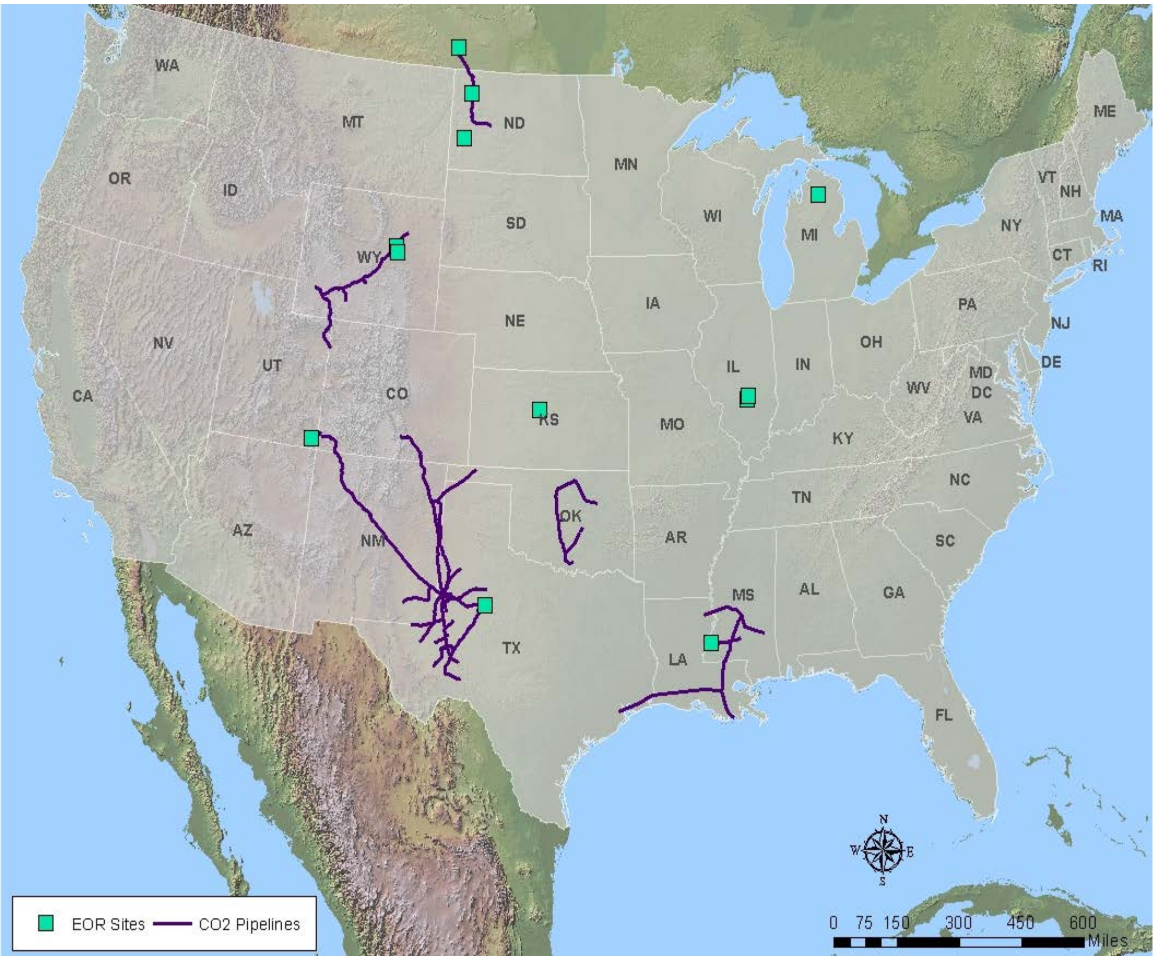
There are currently no plants in the U.S. on which to draw information for gas-to-liquids (GTL) or coal-to-liquids (CTL) facilities. As part of other NETL/DOE studies, both GTL and CTL plants were modeled. These liquid fuel plants require CO_2 capture as an inherent part of the process regardless of the need for CO_2 elsewhere. As such, the incremental cost for using the CO_2 is based solely on compression making them attractive sources.

The results of the literature search yielded nine processes that fit most or all of the criteria set forth in Section 2, and for which suitable information was available. The nine processes covered in this study are ethanol, ammonia, ethylene oxide, natural gas processing, CTL, GTL, refinery hydrogen, steel/iron, and cement.

4 Plant Sites, EOR Fields, and Pipeline Mapping

For this study the final CO_2 product is assumed to be utilized in EOR applications. Depending upon the purity of the CO_2 product, it is possible that other uses for the CO_2 may be available, (7) but alternate possibilities were not considered. Analysis of the base plants for each of the nine processes considered falls outside the scope of this study. It was assumed that all process vent streams with a CO_2 concentration greater than flue gas from a CFPP would constitute the inlet streams for the process models in this study. Leaving the system boundary of this study is a CO_2 stream that has been purified, if necessary, and compressed to pipeline specifications (2,200 psig). The EOR pipeline and EOR fields are considered outside the scope of this study, similar to the base plant. However, while detailed pipeline specifications, such as pressure drop, length, and other characteristics, are not considered here, it is important to highlight potential industrial CO_2 capture locations and their relative location to sites/transport mechanisms that will utilize CO_2 for EOR. There is extensive existing EOR pipeline infrastructure data that was utilized in this study. Exhibit 4-1 below shows the existing CO_2 pipelines and EOR sites.

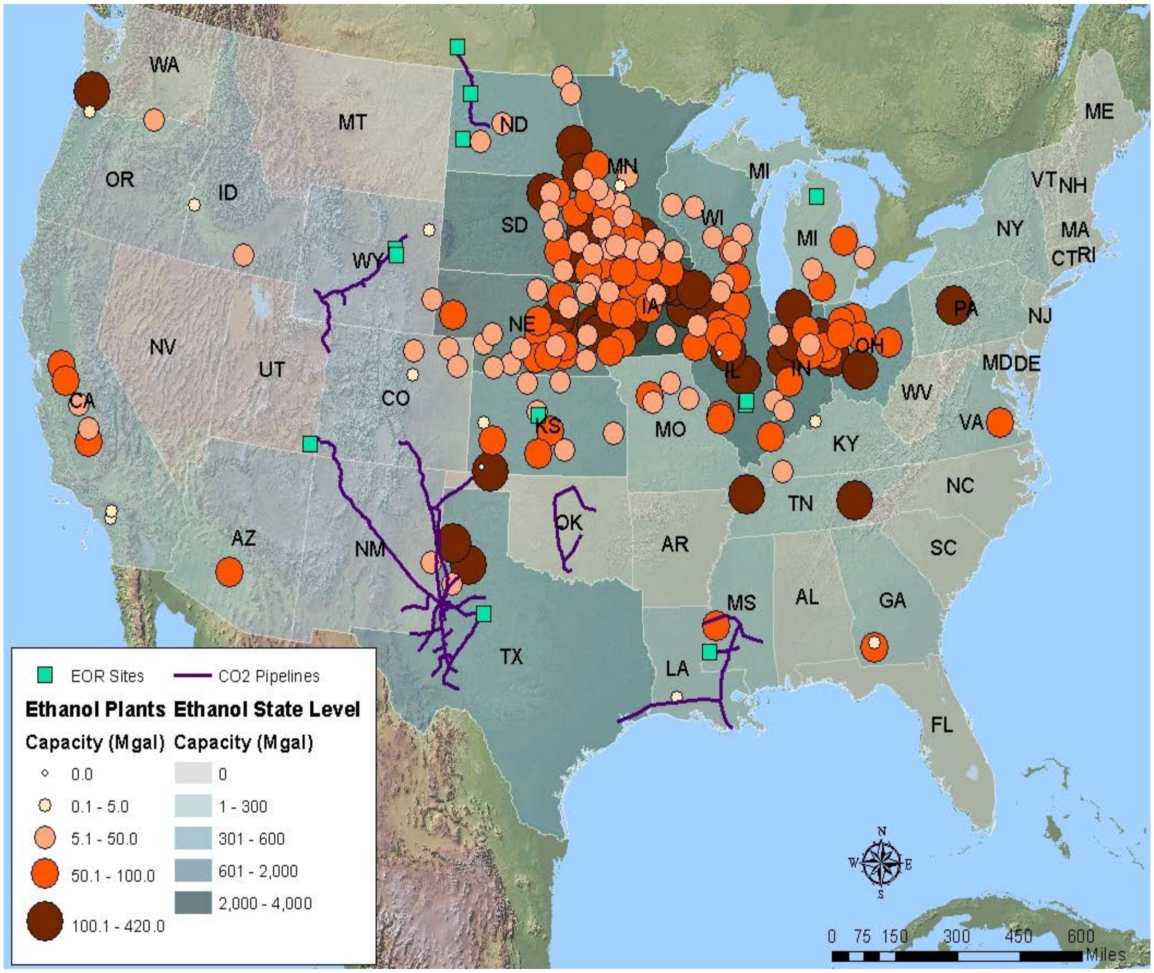
Exhibit 4-1 Existing CO_2 pipeline and active EOR sites



Source: NETL

The following maps, Exhibit 4-2 through Exhibit 4-8, illustrate the proximity of each industrial source to the existing EOR sites and CO_2 pipeline infrastructure.

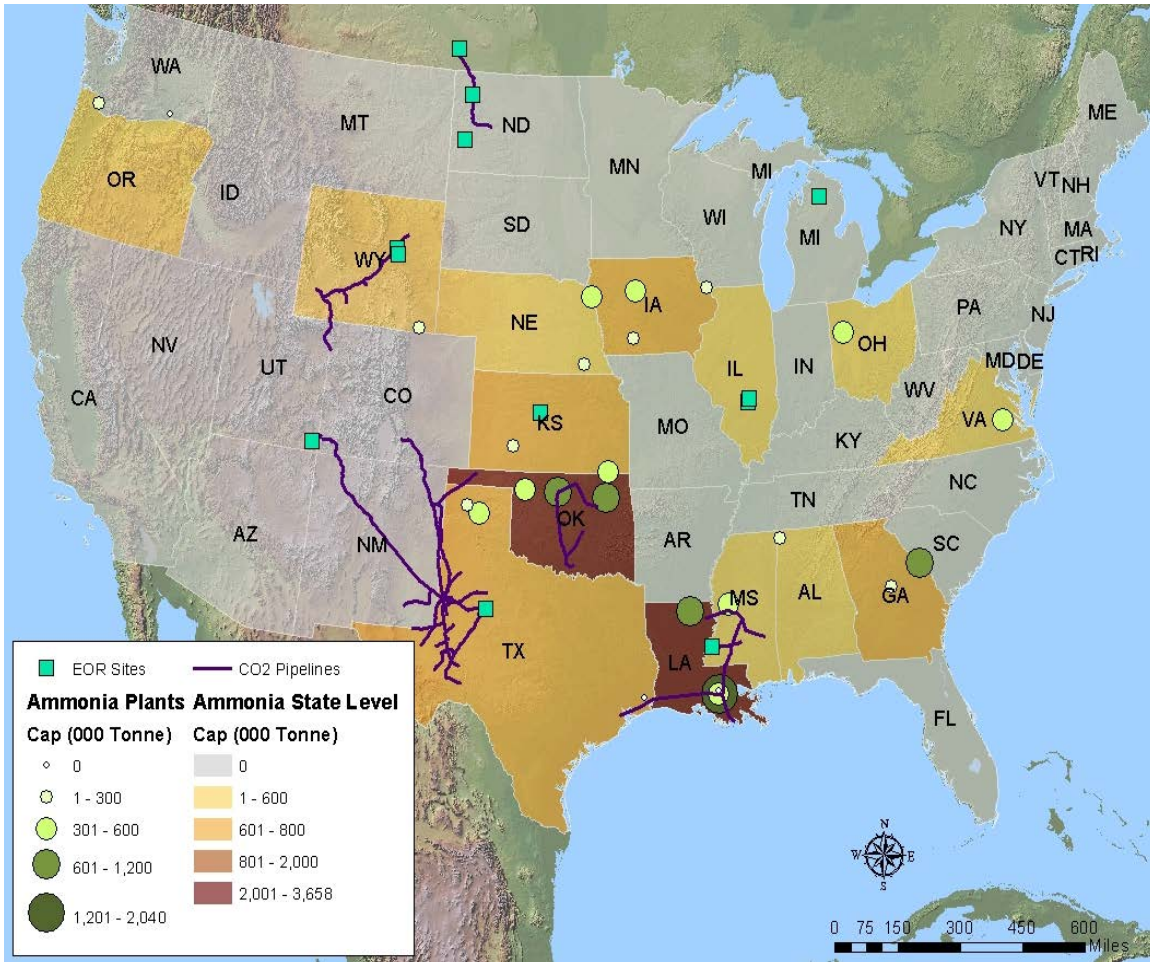
Exhibit 4-2 Ethanol plant locations and existing EOR pipeline



Source: NETL

As shown in Exhibit 4-2, a large percentage of ethanol plant locations do not coincide well with the existing EOR pipeline locations; however, there are several large plants that are conveniently located close to existing infrastructure. While the EOR pipeline falls outside the scope of this work as previously discussed, it is important to point out that for at least the ethanol process, utilization of the majority of captured CO_2 for EOR would be dependent upon the extension of existing CO_2 transport pipelines or the discovery of new injection sites.

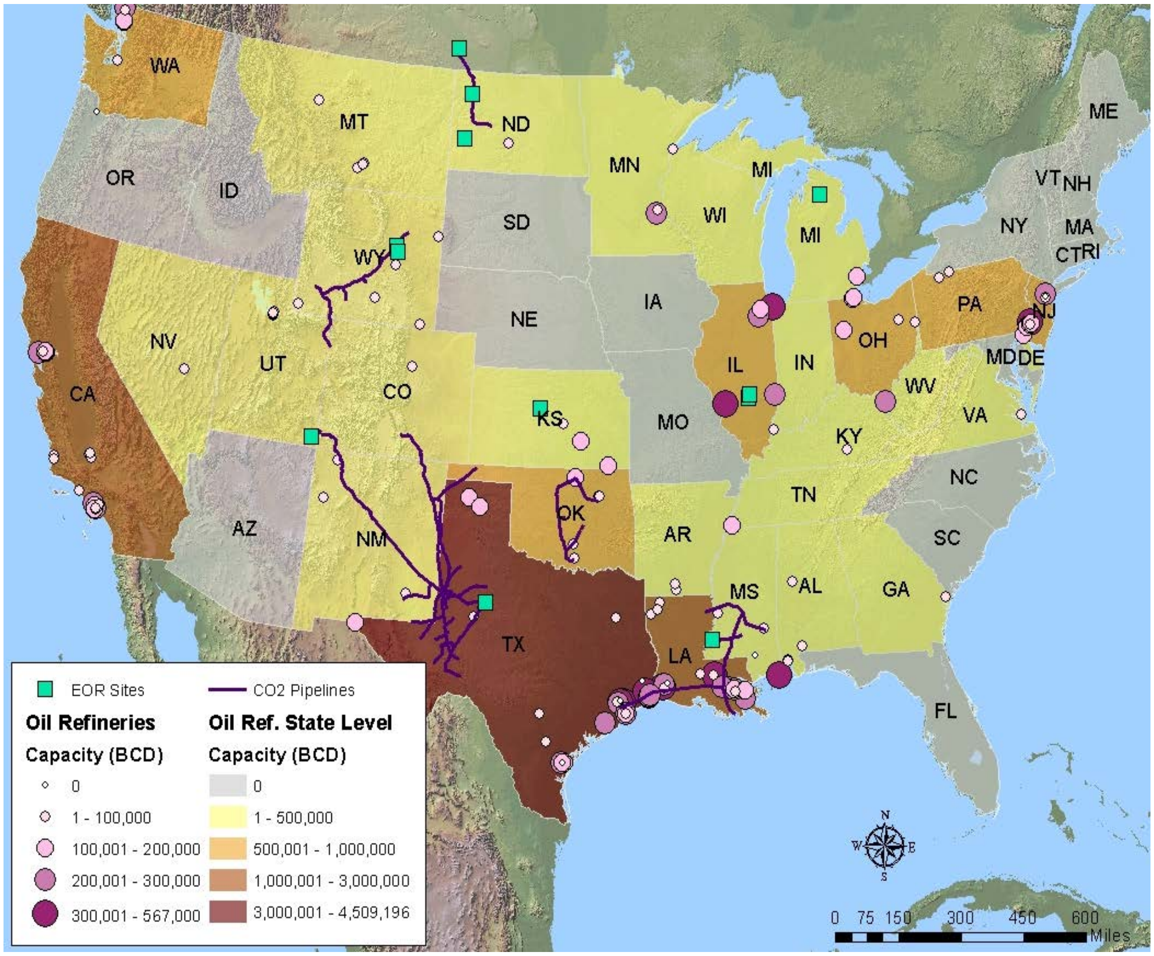
Exhibit 4-3 Ammonia plant locations and existing EOR pipeline



Source: NETL

A large percentage of ammonia plants are located in close proximity to existing EOR pipelines, as shown in Exhibit 4-3.

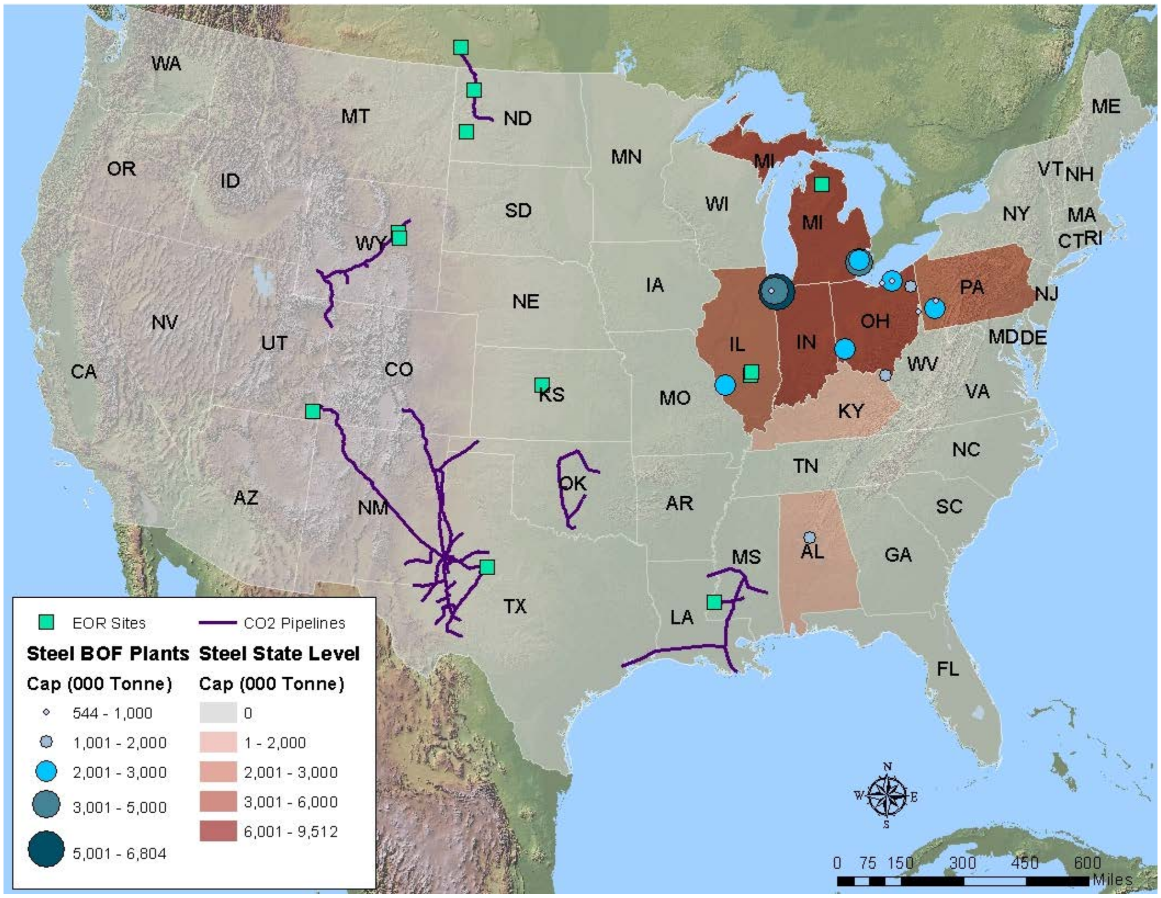
Exhibit 4-4 Refinery hydrogen (U.S. refineries) plant locations and existing EOR pipeline



Source: NETL

Exhibit 4-4 shows the location of U.S. refineries and their proximity to existing EOR pipelines and injection sites. Since this case targets hydrogen produced from refineries, mapping existing refineries is appropriate. There are a large number of existing refineries in close proximity to existing EOR pipelines. However, the map is only intended to show the relative sizes of the refineries, and not the amount of CO_2 available. There is not necessarily a direct relationship between refinery size and CO_2 available.

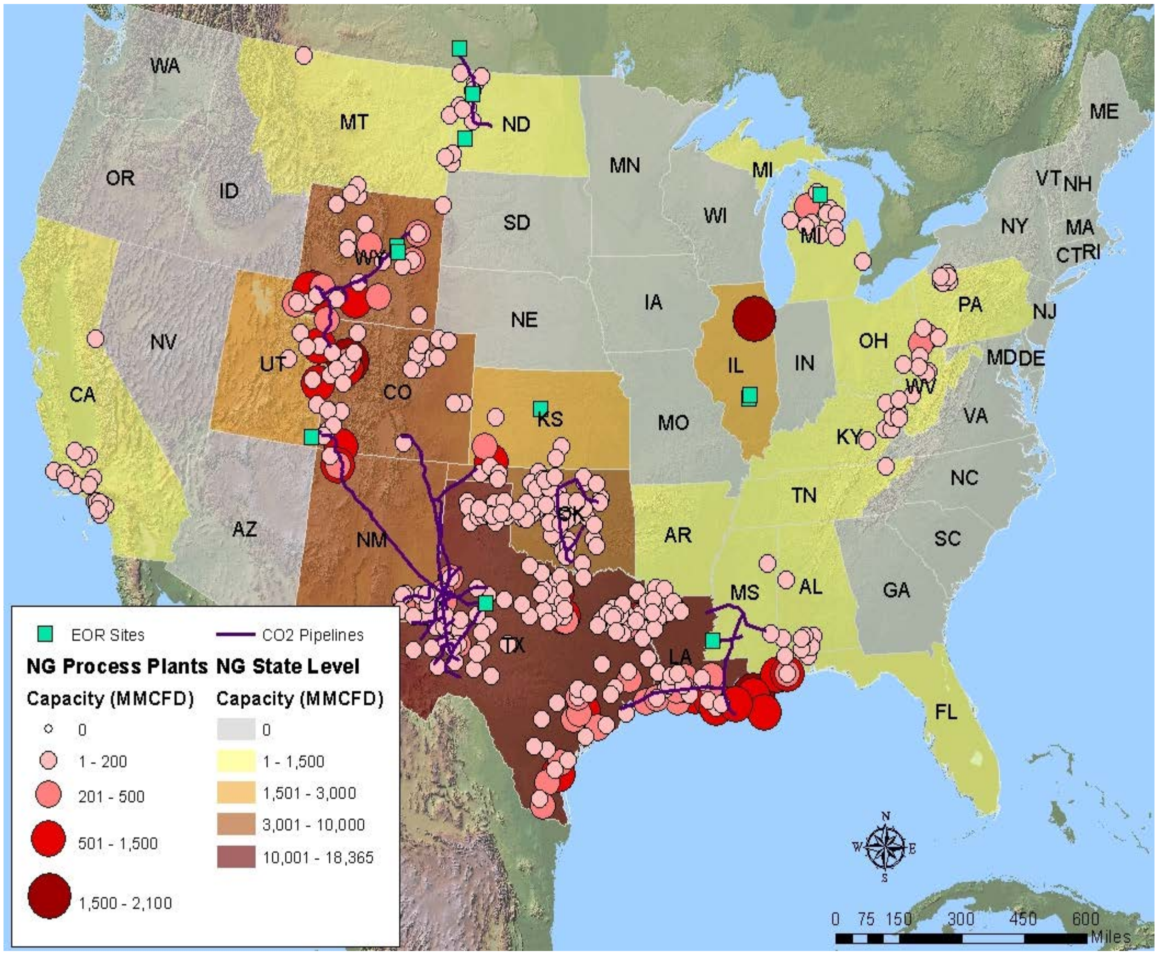
Exhibit 4-5 Steel (BOF) plant locations and existing EOR pipeline



Source: NETL

Exhibit 4-5 shows currently operating steel BOF plants and their relation to existing EOR pipelines and injection sites. Steel does not appear particularly attractive, from the standpoint of ease of implementation, because this case would not be able to utilize any of the existing EOR pipeline infrastructure. However, as demonstrated in Exhibit 4-5, based on this study's assumptions, steel plants represent the largest amount of CO_2 available among industries considered that have currently operating U.S. plants; therefore, construction of the necessary pipelines may be economically viable as long as the market price for CO_2 is high enough to take advantage of the economies of scale.

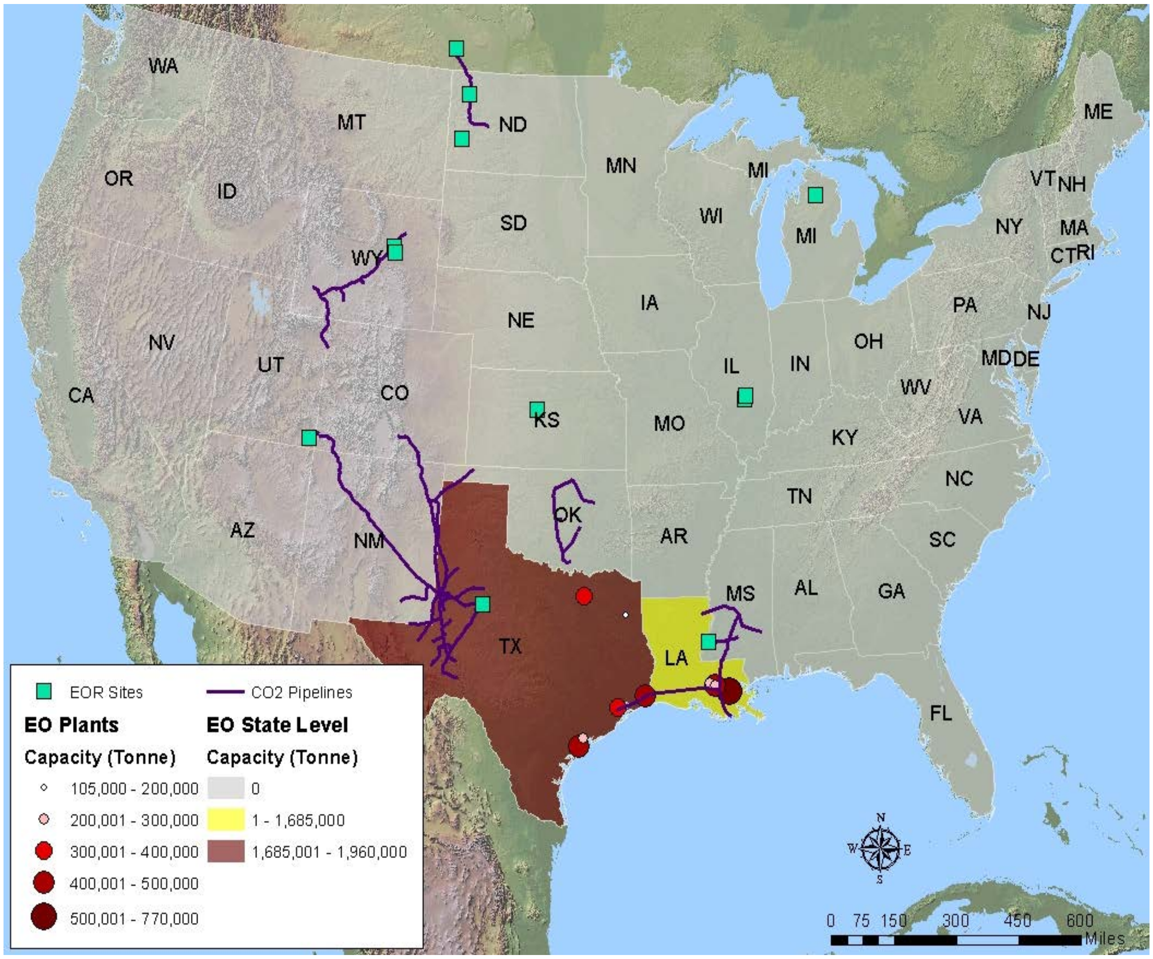
Exhibit 4-6 Natural gas processing plant locations and existing EOR pipeline



Source: NETL

Exhibit 4-6 shows the location of natural gas processing facilities and their relation to existing EOR pipelines and injection sites. Plant capacities are shown on this map; however, given the more than 490 natural gas processing facilities, each treating a different amount of natural gas 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 natural gas with low CO_2 concentration may have less CO_2 available than a smaller facility processing natural gas with a much higher CO_2 concentration. Therefore, the capacity alone can be misleading without the context of the quality of the raw gas, and thus the quantity of CO_2 that the raw gas provides.

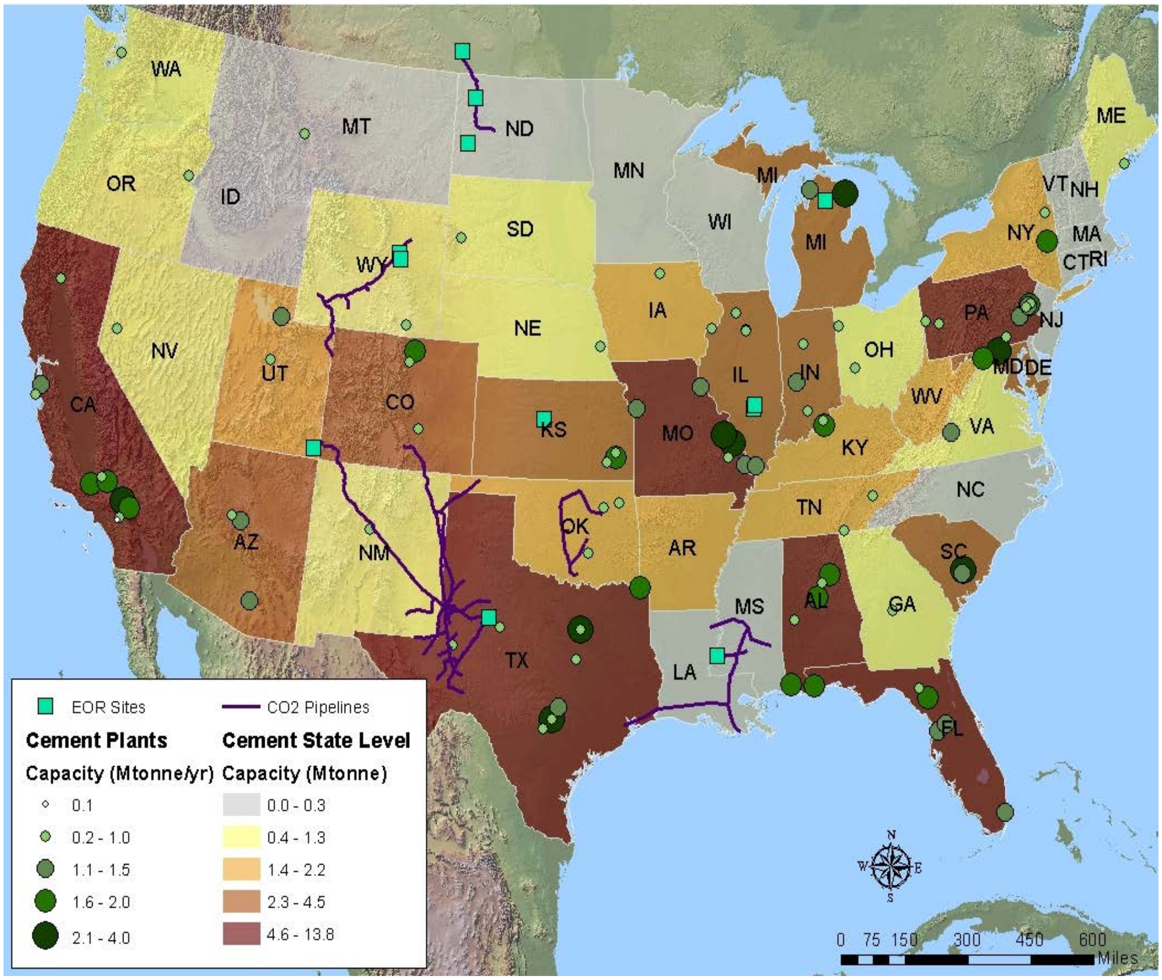
Exhibit 4-7 Ethylene oxide plant locations and existing EOR pipeline



Source: NETL

Exhibit 4-7 shows the location of EO plants and their relation to existing EOR pipelines. U.S. EO production is concentrated in Texas and Louisiana. Of the ten U.S. EO plants, six are located very close to existing EOR pipelines. Therefore, from a location standpoint, EO presents as an attractive option. However, due to the small scale of these EO plants, and therefore the small amount of CO_2 available, while location suggests potential, scale may not.

Exhibit 4-8 Cement plant locations and existing EOR pipeline



Source: NETL

Exhibit 4-8 shows the location of cement plants and their relation to existing EOR pipelines. The cement industry is regional in nature, so there is not one specific region with a high density of cement plants. However, there are several plants located close to existing EOR pipelines, and given the larger scale of cement, and therefore the larger amount of CO_2 emissions available, construction of the necessary pipeline may be economically viable.

There are currently no U.S. CTL or GTL plants in operation, so no map is given.

In conclusion, seven of the nine industries considered in this study have existing U.S. plants in operation, and six have facilities located near existing EOR pipeline infrastructure. The only case that does not fall near an EOR pipeline is steel. However, the size and amount of CO_2 potentially available from one steel facility may justify the construction of an EOR pipeline or development of alternative EOR sites for this case. This economic analysis of amount of CO_2 available versus the cost to construct the required EOR pipeline is not considered in this study.

5 Economic Analysis Overview

The industrial sources considered in this study are grouped in to a ‘High Purity’ and a ‘Low Purity’ group, based on the concentration of CO_2 in the stream to be captured. Each grouping carries with it its own set of financial assumptions. For example, for high purity sources, only compression is required. Therefore, only one year of capital expenditure is assumed, resulting in a specific capital charge factor (CCF). For low purity sources, purification and compression are required. In these cases, a three-year capital expenditure period is assumed, resulting in a slightly higher CCF.

5.1 Capital Charge Factor

The finance structures used are given in Exhibit 5-1 and Exhibit 5-2.

Exhibit 5-1 Financial assumptions for high purity sources

Financial Parameter	Value(s)
Capital Charge Factor	15.2%
Debt/Equity Ratio	50/50
Payback Period	30 years
Interest on Debt	8.0%
Return on Equity	20%
Capital Expenditure Period	1 year
Capital Distribution	1st year – 100%

Exhibit 5-2 Financial assumptions for low purity sources

Financial Parameter	Value(s)
Capital Charge Factor	17.6%
Debt/Equity Ratio	50/50
Payback Period	30 years
Interest on Debt	8.0%
Return on Equity	20%
Capital Expenditure Period	3 years
Capital Distribution	1st year – 10% 2nd year – 60% 3rd year – 30 %

The Association for the Advancement of Cost Engineering (AACE) 16R-90 states that project contingency for a ‘budget-type’ estimate (AACE Class 4 or 5) should be 15 to 30 percent of the sum of bare erected cost (BEC), engineering/procurement/construction (EPC) fees, and process contingency. Therefore, a 20-percent project contingency was added to each account. All process equipment in the estimate is considered to be commercially available, so no process contingencies were added.

The result of the economic analysis is a calculated first year breakeven selling price of CO_2. The first year breakeven selling price for CO_2 is the revenue required by the owner per tonne of CO_2 captured, when escalated at the assumed nominal general inflation rate of 3 percent per year that provides the stipulated rate of return on equity over the entire economic analysis period. Assuming that all annual costs also escalate at the assumed nominal general inflation rate, the breakeven selling price is essentially the sum of the O&M costs and the annualized capital cost charges all normalized to the annual plant CO_2 capture 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 breakeven selling price of CO_2 is expected to be lower due to the economies of scale associated with the cost of the larger equipment.

5.2 Retrofit Factors

As part of its Quality Guidelines for Energy Systems Studies, NETL has a 2013 report, “Estimating Plant Costs Using Retrofit Difficulty Factors” that describes a general cost-estimating procedure for retrofit applications using post-combustion CO_2 capture as an example. (8) However for this report, reference factors of 1.01 and 1.05 were selected using engineering judgement because the factors developed in the QGESS do not translate well for the following:

- The higher purity sources do not require a CO_2 separation system
- CO_2 separation is performed using methyldiethanolamine (MDEA), which is different than the MEA systems that were used to develop the retrofit factors
- The capture rate is 95 percent (as opposed to 90 percent in the referenced study)
- These industrial sources are significantly smaller than the utility scale power plants for which these factors were developed

A retrofit factor of 1.01 was applied to the TOC as a blanket retrofit cost increase for high purity cases requiring only compression, and a retrofit factor of 1.05 was applied to the TOC as a blanket retrofit cost increase for cases requiring compression and purification. A sensitivity to the retrofit factor is performed and discussed in section 9.2.4.

6 Equipment

The different types of equipment used for this study, as well as the methods for scaling are discussed in this section.

The power required for all cases, whether high or low purity, is purchased at a rate of $58.59/MWh. Purchasing power is the strategy used in this study; however, in some cases a plant may benefit from generating and using power produced on-site. This would be determined on a case-by-case basis.

To satisfy any steam requirements a package boiler is used. Natural gas to fuel the boiler is purchased at a cost of $6.13/MMBtu.

6.1 Compression

Two different types of compressors are used, integrally geared centrifugal, and reciprocating. The type of compressor selected for each case is based on the mass flow of CO_2 to the first compression stage as well as the suction conditions at stage one. At conditions of 80°F suction temperature and 16.7 psia suction pressure, the minimum required flow for a centrifugal compressor to reach 2,200 psig is approximately 615,900 tonnes CO_2/yr (678,900 tons CO_2/yr). (9) This sets the lower limit for where centrifugal compression may be used, below which, reciprocating compression is used.

A quote for a five-stage reciprocating compressor was obtained. The suction pressure was quoted as 17.4 psia, and the suction temperature was quoted as 80°F, with an inlet flow to stage one of 36,000 lb/hr. The discharge pressure was quoted as 2,214.7 psia with a total power requirement of 1.815 MW. The BEC for this compressor was quoted as $3.9875 M USD, and costs for each case (where reciprocating compression was used) were scaled on mass flow to compression stage one. The reciprocating compressor was used in the cases shown in Exhibit 6-1 with the following specifications.

Exhibit 6-1 Reciprocating compressor cases specifications

Case	Number of Compression Stages	Inlet Flow to Compression Stage 1 (lb/hr)	Suction Pressure (psia)	Suction Temperature (°F)
Ethanol	5	35,991	17.4	81
Ammonia	5	115,953	23.5	69
Refinery Hydrogen	5	68,923	17.4	81
Ethylene Oxide	4	30,578	43.3	96

Quotes for integrally geared centrifugal compressors were obtained. Two separate quotes, both discharging at 2,214.7 psia, were used and scaled. Quote one has a suction pressure of 23.5 psia, suction temperature of 69°F, and an inlet mass flow of 242,000 lb/hr. The power requirement was quoted as 10.1 MW and the BEC quoted as $13.04 M USD. Quote two has a suction pressure of 19.0 psia, a suction temperature of 80°F, and an inlet mass flow of 695,000 lb/hr. The power requirement was quoted as 30 MW and the BEC quoted as $21.894 M USD.

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. The compression trains use centrifugal compressors similar to the quoted compressors. This approach is advantageous particularly for CTL, where CO_2 is available at multiple pressures, and requires a special compression train that can accommodate multiple suction pressures. Exhibit 6-2 shows the cases using integrally geared centrifugal compression and their case specifications.

Exhibit 6-2 Integrally geared centrifugal compressor cases specifications

Case	Compressor Quote Used	Number of Compression Stages	Inlet Flow to Compression Stage 1 (lb/hr)	Suction Pressure (psia)	Suction Temperature (°F)
Natural Gas Processing	1	8	164,059	23.5	69
Steel/Iron COG/BFS	1	8	291,302	19.0	80

Steel/Iron COG PPS	2	8	695,418	19.0	80
Cement	1	8	287,873	19.0	80
Coal-to-Liquids	NETL Study	N/A[1]	2,200,4232	160/265/300	N/A
Gas-to-Liquids	NETL Study	N/A[1]	467,794	265	100

[1] – Both CTL and GTL are assumed to use eight total compression stages, but this is not explicitly stated in the respective reports.

[2] – Flow reported is total. The individual flows at the three suction pressures given sum to the total flow.

As mentioned, all compressors discharge at a pressure of 2,214.7 psia (2,200 psig). This is the target pipeline specification assumed in this study, which is given in the QGESS. (10) 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. (11)

6.2 CO_2 Separation and Purification

For cases requiring separation and purification prior to compression, a MDEA acid gas removal (AGR) unit was used. The MDEA performance and cost information was obtained from a previous NETL study which investigated hydrogen production with CO_2 capture. (12) The MDEA unit cost and performance are scaled based on CO_2 product mass flow. Cases where an MDEA AGR is used include

- Refinery Hydrogen
- Steel/Iron
- Cement

The CO_2 capture efficiency of the MDEA unit is constant at 95 percent for all cases, as it was stated in the referenced report.

6.3 Boiler Steam Production

The MDEA AGR unit requires low pressure steam at 71 psia for solvent regeneration. Since no assumptions, such as available steam, are made about the base plants, cases requiring CO_2 separation and purification also require a boiler for steam production.

A quote for a package boiler was obtained from CleaverBrooks. The boiler produces superheated steam at 600 psig, which is significantly higher than required for this application. However, for each case requiring an MDEA unit, the total heat required from 71 psia steam for solvent regeneration was calculated, and that amount of heat delivered was used to calculate the amount of steam the CleaverBrooks package boiler must deliver. The boiler auxiliaries for pumps and compressors were scaled from the amount of heat delivered, and consumables such as makeup water and natural gas were also scaled on heat delivered. The quoted BEC for the CleaverBrooks boiler was \$4.25 M USD. The costs were also scaled on heat delivered.

6.4 Cooling Water Unit

As previously stated, no assumptions were made regarding the base plant for each process, and so no assumptions about the existing plant's cooling water system are made. Therefore, any cooling required by the compression train, and in some cases the MDEA AGR unit, must be supplied by a stand-alone cooling water unit.

A performance quote for a cooling water unit was not available. Power consumption and cost estimates for the cooling water unit were scaled from the Bituminous Baseline Report Case 12 cooling tower based on cooling water flow required.

Each case considered in this report requires a cooling water unit.

6.5 Heat Exchangers

Cooling of the product CO_2 is required for all cases following compression to meet the pipeline temperature specification of 120°F, and only for some cases preceding compression. Depending on the type of compressor used, post-cooling of the compressed product CO_2 may be included in the compressor quote. 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.

The pre-cooler heat exchanger and post-cooler heat exchanger for this study are based on heat exchanger performance and cost information obtained from a previous NETL report, "Advancing Oxycombustion Technology for Bituminous Coal Power plants: An R&D Guide." (13)

The heat exchanger from that study operates at 14.8 psia, and heats a stream from 135°F to 150°F. The reference heat duty is 52 MMBtu/hr, and the BEC is $4.5 M USD.

No adjustments for changes in operating conditions or materials of construction were made for this report. The BEC is scaled on heat duty.

7 High Purity Sources

The sources discussed in this section are considered high purity sources, meaning the available CO_2 does not require purification to meet EOR pipeline specifications. In some cases, however, dehydration of the CO_2 stream may be required, but this is not considered purification.

7.1 Ethanol

Ethanol production generates as a byproduct a high-purity CO_2 stream > 85 percent by volume. (14) Though not a large-scale CO_2 producer, the cost of CO_2 captured 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 (ADM) project in Decatur, IL. The purpose of the project is to demonstrate how the next generation of technologies capture and sequester or reuse industrial CO_2 emissions. (15) The CO_2 to be captured is a byproduct of the ethanol production process. The project will capture approximately 1 M tons of CO_2/year using dehydration and compression, and will sequester the captured CO_2 in the Mt. Simon Sandstone Formation saline reservoir. (15)

7.1.1 Size Range

There are 209 ethanol refineries in the U.S. demonstrating a wide range of production, with 80 percent of these refineries using the dry-mill process. (16) It was found that 82 of the plants (approximately 39 percent) fall between 40-60 M gal/year. (17) However, multiple plants are expected to expand in the near future to produce ethanol in the range of 100 to 420 M gal/year, including four plants that produce in excess of 215 M gal/year. The EIA's Annual Energy Outlook for 2012 projects ethanol consumption in the transportation sector to increase by 0.8 M gal/year. (18)

CO_2 produced from a 50 M gal/year plant versus a 215+ M gal/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 6.1.

Since a large portion of existing ethanol plants, 82 plants, have smaller production capacities between 40-60 M gal/year, the plant size chosen was 50 M gal/year, and utilized reciprocating compression. It was also assumed that the plant uses the dry-mill process with corn as the feedstock of choice.

7.1.2 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 point sources, only the fermentation off gas stream is considered high purity, and was, therefore, the focus. The fuel burning stream is consistent with flue gas compositions, and is, therefore, outside the scope of this study. A study done by the Illinois State Geological Survey (19) 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 below to calculate the amount of CO_2 emissions from the fermentation point source:

$$CO_{2\ \text{Fermentation}}\ (\text{ton/year}) = (\text{ethanol production [gal/year]} \times \text{EF } [\text{lbCO}_2/\text{gal}])/\ 2000\ [\text{lb/ton}]$$

where EF stands for the emission factor dependent on the feedstock. The generic plant assumed in this study utilizes corn as the feedstock, giving an EF = 6.31 lb CO_2/gal ethanol. The EF was formulated in the reference cited (19) through communication with representatives from existing ethanol plants in the Illinois Basin. Using this relationship, the ethanol plant will generate approximately 143,045 tonnes CO_2/year from fermentation for a 50 M gal/year ethanol plant.

In a report published by the Global Carbon Capture and Sequestration (CCS) Institute in 2010, they state 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." (20) Therefore, the fermentation stream will be assumed to be 100 percent CO_2 and may be sent directly for cooling and compression. Other sources (15) 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, therefore, further purification before processing would be unnecessary.

The fermentation process occurs at a temperature of 140-180 °C. (21)

7.1.3 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:

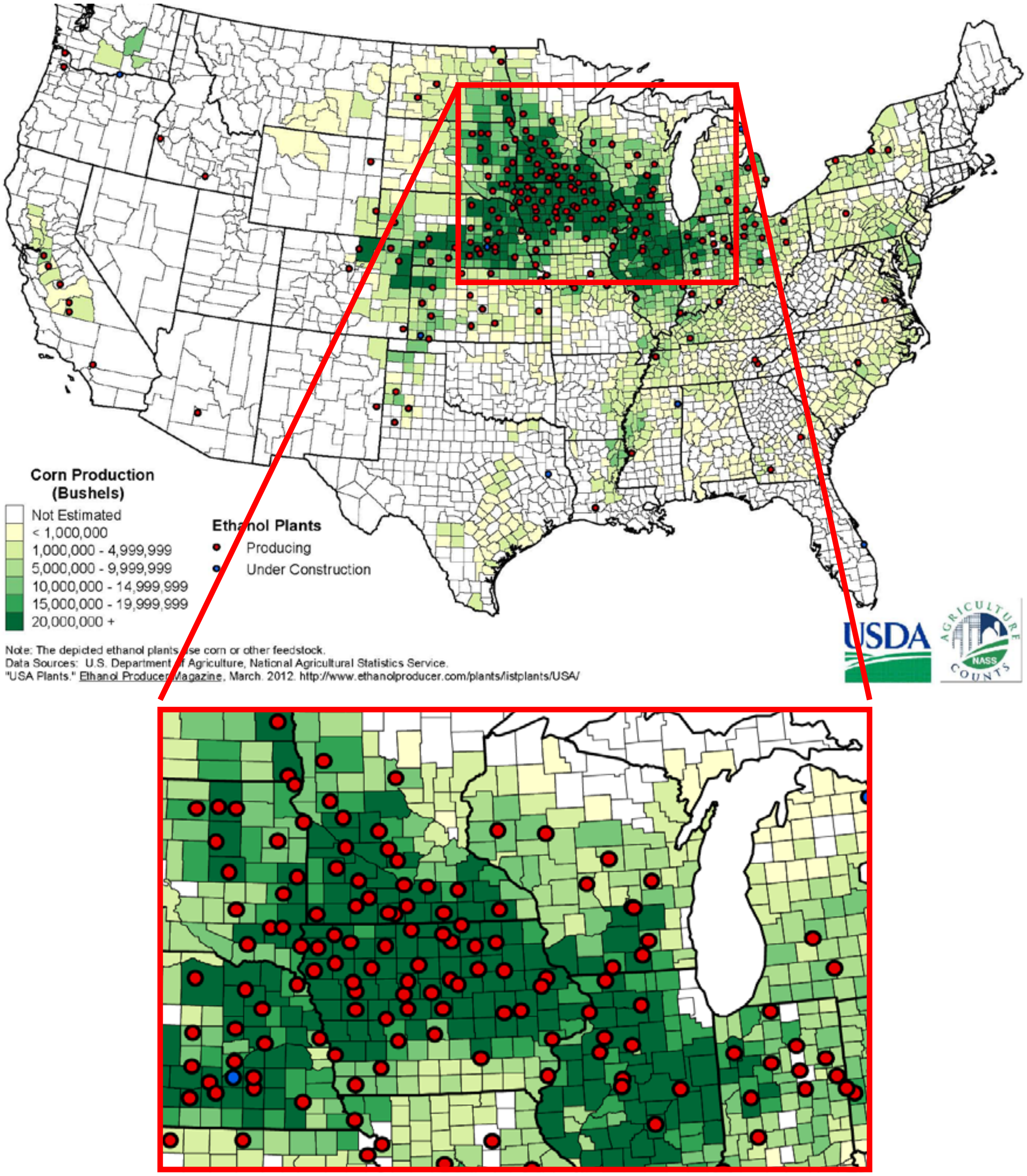
- The design plant is represented by an ethanol production rate of 50 M gal/year
- The plant uses the dry-mill process with corn as the feedstock
- The fermentation off-gas, assumed to be 100 percent CO_2, is the only high-purity point source chosen
- The CO_2 amount, as calculated above, is 143,045 tonnes CO_2/year; CO_2 temperature is 160 °C
- The CO_2 pressure is 17.4 psia
- The CO_2 quality is based on the EOR "pipeline" standard as mentioned in the NETL Quality Guidelines for Energy System Studies (QGESS) for CO_2 Impurity Design Parameters. (10) For the purpose of this study, 100 percent purity is assumed

7.1.4 CO_2 Capture System

Exhibit 7-1 (22) 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 contained in the area of high corn production, namely the Midwest states. The highest density of ethanol plants occurs in Illinois, Iowa, Minnesota, and Nebraska.

The trend for the ethanol industry is smaller plants, which in turn, will produce smaller CO_2 streams, and will require compression equipment capable of handling smaller flows. This requirement is satisfied by using reciprocating compression discussed in Section 6.1; however, an alternative to smaller equipment could be to aggregate the emissions from multiple nearby plants for a single EOR project. The possibility of combining multiple small CO_2 streams to take advantage of economies of scale and larger equipment is discussed below.

Exhibit 7-1 U.S. ethanol plant locations



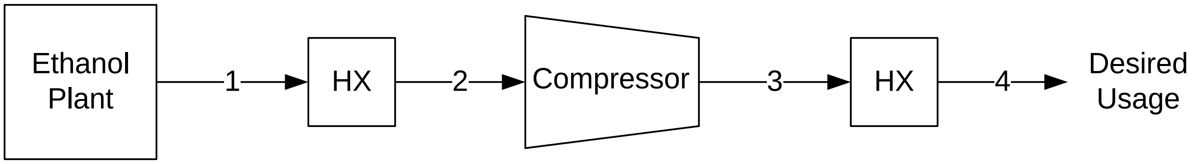
Source: USDA

The potential for combining CO_2 streams from plants with production flow rates of less than 215 M gal/year to take advantage of economies of scale seems feasible; however, this option requires multiple plants located in very close proximity to one another. This is because the pressure drop incurred would need to be overcome either by utilizing plants in close proximity or by unreasonably large diameter pipe. Combining these technical obstacles with negotiating multiple parties' pipeline usage, maintenance, and purchase agreements makes the option almost impossible, and it was, therefore, not considered further.

7.1.5 Block Flow Diagram, 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 7-2, the fermentation vent is cooled through a heat exchanger, compressed (with inter-stage cooling), and then after-cooled to meet EOR pipeline specification.

Exhibit 7-2 Ethanol CO_2 capture block flow diagram



Source: NETL

Exhibit 7-3 Ethanol stream table

	1	2	3	4
V-L Mole Fraction				
AR	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000
CO_2	1.0000	1.0000	1.0000	1.0000
COS	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000
H_2O	0.0000	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	371	371	371	371
V-L Flowrate (kg/hr)	16,325	16,325	16,325	16,325
Solids Flowrate (kg/hr)	0	0	0	0
Temperature (°C)	160	27	144	49
Pressure (MPa, abs)	0.13	0.1	15.4	15.3
Enthalpy (kJ/kg)^A	143.56	21.4	45.4	-171.6
Density (kg/m^3)	1.6	2.1	247.6	673.5
V-L Molecular Weight	44.010	44.010	44.010	44.010
V-L Flowrate (lb_{mol}/hr)	818	818	818	818
V-L Flowrate (lb/hr)	35,991	35,991	35,991	35,991
Solids Flowrate (lb/hr)	0	0	0	0
Temperature (°F)	320	81	292	120
Pressure (psia)	18.4	17.4	2,237.8	2,215.6
Enthalpy (Btu/lb)^A	61.7	9.2	19.5	-73.8
Density (lb/ft^3)	0.097	0.133	15.459	42.044

The performance results are based on compressor quotes discussed in Section 6.1. The performance summary is provided in Exhibit 7-4.

Exhibit 7-4 Performance summary

Performance Summary	
Item	**50 M Gal Ethanol/year (kW$_e$)**
CO_2 Compressor	1,880

7.1.6 Capture Integration

The fermentation process occurs at a temperature of 140-180 °C. 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. This would have to be evaluated on a case-by-case basis.

For the purposes of this study, it will be 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 (DDGS). This product is produced by drying the solids that remain after fermentation. Heat for DDGS drying is generally provided by natural gas.

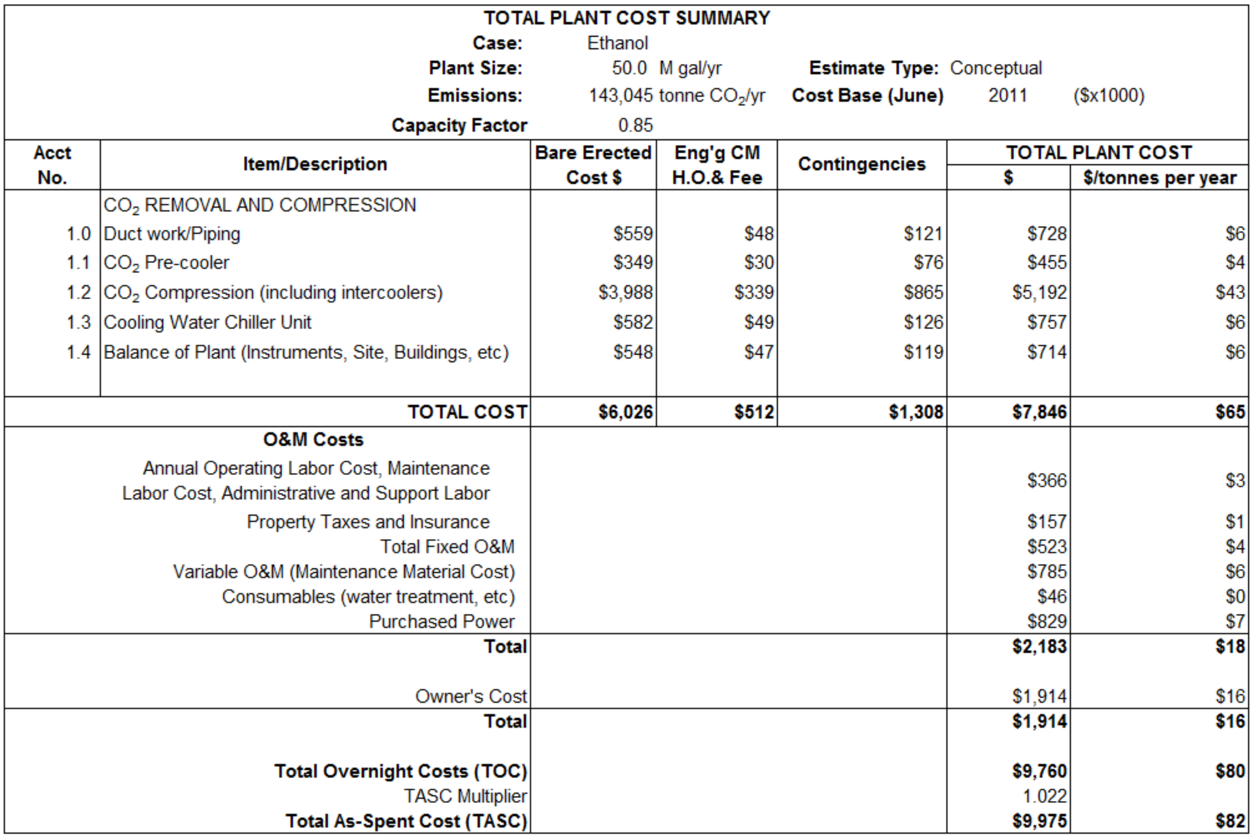
7.1.7 Power Source

Given the relatively small amount of CO_2 to be compressed, the power consumption of the compressor is 1.88 MW. Power consumption estimates for the stand-alone cooler were approximated by scaling from Bituminous Baseline (BB) Case 12 as discussed in Section 6.4. The total power requirement was calculated to be 1.9 MWh/hr, which includes all power required by the compression train and the cooling system. Power will be purchased at a rate of $58.59/MWh as discussed in Section 6.

7.1.8 Economic Analysis Results

The operation and maintenance (O&M), TOC for Greenfield and retrofit, and breakeven costs are given in this section for analysis and discussion. Owner's costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner's costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

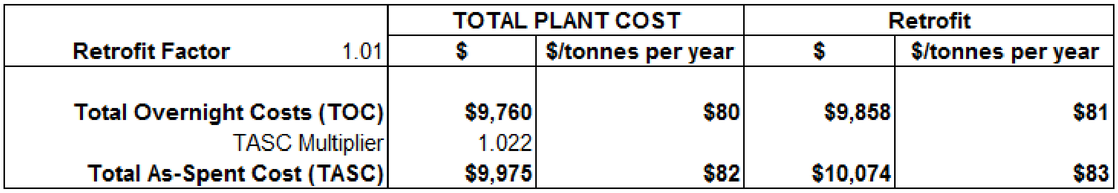
Exhibit 7-5 Capital and O&M costs for ethanol Greenfield site



TOTAL PLANT COST SUMMARY
Case: Ethanol
Plant Size: 50.0 M gal/yr **Estimate Type:** Conceptual
Emissions: 143,045 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST $	TOTAL PLANT COST $/tonnes per year
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$559	$48	$121	$728	$6
1.1	CO_2 Pre-cooler	$349	$30	$76	$455	$4
1.2	CO_2 Compression (including intercoolers)	$3,988	$339	$865	$5,192	$43
1.3	Cooling Water Chiller Unit	$582	$49	$126	$757	$6
1.4	Balance of Plant (Instruments, Site, Buildings, etc)	$548	$47	$119	$714	$6
	TOTAL COST	**$6,026**	**$512**	**$1,308**	**$7,846**	**$65**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$366	$3
	Property Taxes and Insurance				$157	$1
	Total Fixed O&M				$523	$4
	Variable O&M (Maintenance Material Cost)				$785	$6
	Consumables (water treatment, etc)				$46	$0
	Purchased Power				$829	$7
	Total				**$2,183**	**$18**
	Owner's Cost				$1,914	$16
	Total				**$1,914**	**$16**
	Total Overnight Costs (TOC)				**$9,760**	**$80**
	TASC Multiplier				1.022	
	Total As-Spent Cost (TASC)				**$9,975**	**$82**

Exhibit 7-6 Total overnight costs for retrofit



		TOTAL PLANT COST		Retrofit	
Retrofit Factor	1.01	**$**	**$/tonnes per year**	**$**	**$/tonnes per year**
Total Overnight Costs (TOC)		**$9,760**	**$80**	**$9,858**	**$81**
TASC Multiplier		1.022			
Total As-Spent Cost (TASC)		**$9,975**	**$82**	**$10,074**	**$83**

The first year breakeven costs for a Greenfield site and retrofit site were calculated and are shown in Exhibit 7-7 and Exhibit 7-8 below. The financial assumptions and retrofit factor used represent the assumptions detailed in Section 5.

Exhibit 7-7 Breakeven cost for 50 M gal/year Greenfield

BREAKEVEN COST		
Case:	Ethanol	
Plant Size:	50.0	M gal/yr
Emissions:	121,588	tonnes CO_2/yr
Capacity Factor:	0.85	
Cost Base (June):	2011	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	12.20	\$/tonne CO_2
Fixed O&M	4.30	\$/tonne CO_2
Variable O&M	6.46	\$/tonne CO_2
Consumables	0.38	\$/tonne CO_2
Purchased Power	6.82	\$/tonne CO_2
First year Breakeven Cost	30.15	\$/tonne CO_2

Exhibit 7-8 Breakeven cost for 50 M gal/year retrofit

RETROFIT BREAKEVEN COST		
Case:	Ethanol	
Plant Size:	50.0	M gal/yr
Emissions:	121,588	tonnes CO_2/yr
Capacity Factor:	0.85	
Cost Base (June):	2011	
Retrofit Factor:	1.01	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	12.32	\$/tonne CO_2
Fixed O&M	4.34	\$/tonne CO_2
Variable O&M	6.52	\$/tonne CO_2
Consumables	0.38	\$/tonne CO_2
Purchased Power	6.89	\$/tonne CO_2
First year Breakeven Cost	30.46	\$/tonne CO_2

7.1.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

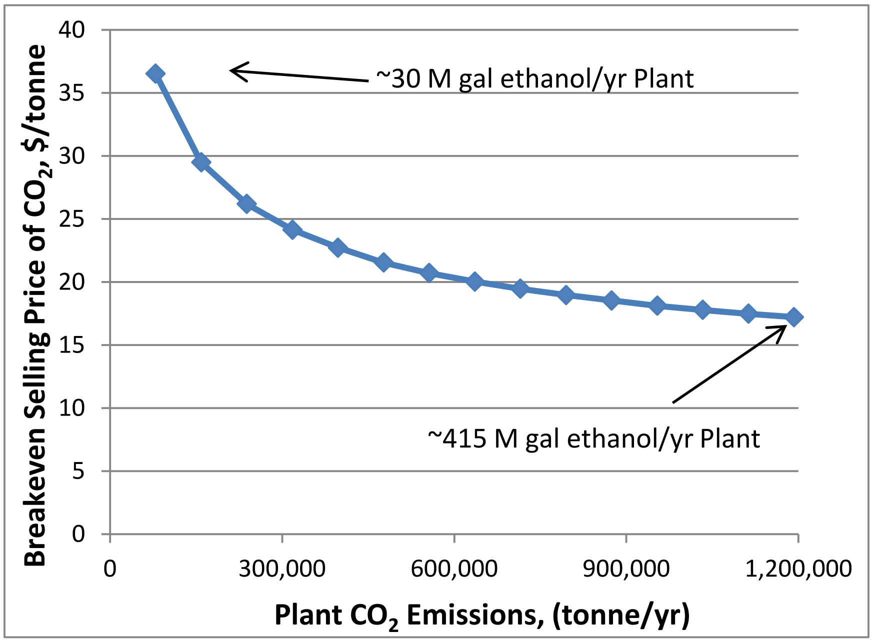
- Plant Size
- Cost of Electricity (COE)
- CO_2 Purity
- CCF

The plant size sensitivity is process specific and difficult to compare across cases, and is, therefore, the only sensitivity discussed in this section. The other three sensitivities, COE, CO_2 purity, and CCF, are discussed and compared across the other eight cases in Section 9.2.

7.1.9.1 Plant Size

Plant size was varied over the range of ethanol production between 415 M gal/yr to 30 M gal/yr. The x-axis refers to plant emissions from fermentation only. Emissions from combustion are not included. Using the equation given in section 7.1.2, a 415 M gal/yr plant would produce approximately 1.19 M tonnes CO_2/yr, and a 30 M gal/yr plant would produce approximately 80,000 tonne CO_2/yr. As shown in Exhibit 7-9, as the plant size decreases over the given range, the breakeven cost increases by \$19.30/tonne CO_2.

Exhibit 7-9 Plant size sensitivity



Source: NETL

7.1.10 Ethanol Conclusion

The high-purity CO_2 streams produced from ethanol plants makes them an attractive industrial process since they require no costly separation equipment. A compression system for a 50 M gal/yr ethanol plant was modeled, and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be \$30.15 /tonne CO_2 for a Greenfield site, and \$30.46/ tonne CO_2 for a retrofit site. The small disparity is the result of small differences in equipment required for a Greenfield versus retrofit site, assuming adequate plot plan space for the retrofit case. Sensitivities were performed on the following variables: COE, CCF, CO_2 purity, and plant size. The sensitivity tests for COE, CCF, and CO_2 purity are detailed in Section 9.2, but showed that changes to COE and CCF have essentially the same impact. Over the \$20-\$140/MWh range, COE increased by \$13.96/MWh, and over a 10 percent to 35 percent range, CCF increased by \$20.06/MWh. It should be noted that for the CO_2 purity sensitivity, greater than 10 percent change in CO_2 purity could result in a larger change in the breakeven price. This was assumed to represent a different case with a different set of base plant assumptions, and possibly a different process altogether, and was, therefore, not considered.

The plant size sensitivity showed interesting results. As the plant size decreased from 415 M gal/yr to 30 M gal/yr, the breakeven selling price increased by $19.30/tonne CO_2. As the plant size is decreased, less CO_2 is produced, and economies of scale are lost, resulting in a higher breakeven selling price.

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 research is required.

Further ethanol examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

7.2 Ammonia

It is estimated that the U.S. production of ammonia in 2006 was over 8.2 M tonnes. (23) Ammonia can be thought of as a high-purity industrial source of CO_2. In all but one plant in the U.S., the ammonia process reforms natural gas to form H_2, CO, and CO_2; however international ammonia production may be more largely based on coal as a feedstock over natural gas. 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, and, 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 in many cases is captured and reused to produce urea. Urea is synthesized by reacting ammonia with CO_2. The amount of CO_2 captured and reused will vary from plant to plant. With CO_2 removal crucial to the ammonia process, coupled with the need for CO_2 to convert ammonia into urea, ammonia processing is a viable option.

7.2.1 Size Range

As of 2006, there were 24 ammonia plants in the U.S. Of these, 17 fell in the range of 0.09 – 0.635 M tonnes/year (0.1 - 0.7 M tons/year) production capacity, and five had a capacity of 690,000 tonnes/year or greater. (23) The largest U.S. ammonia plant has a capacity of 2.04 M tonnes/year. (23) The top 20 percent of ammonia plants by capacity are best represented with a capacity of 907,000 tonnes/year (1 M tons/year). It is reasoned that future ammonia plant construction will have a capacity based on the top 20 percent of existing plants. Therefore, for the purposes of this study, ammonia will be represented with a production capacity of 907,000 tonnes/year. Using the capacity utilization factor of 72 percent, as given by U.S. ammonia producers in 2006, (23) the representative plant production is 653,000 tonnes ammonia/year.

7.2.2 CO_2 Point Sources

The main point source of CO_2 emissions in an ammonia plant comes from the flue gas from the primary reformer and the CO_2 stripper vent which separates CO_2 from the ammonia syngas. Of these two, only the CO_2 stripper vent is considered a high purity source. The primary reformer flue gas has a CO_2 partial pressure of 0.5 MPa in a total stream pressure of 2.8 MPa. (24) (25) Since this partial pressure is consistent with CFPP flue gas partial pressures, the CO_2 stripper vent will be the focus of this study. An article published by KBR Technology (4) concerning CO_2 capture in the ammonia industry stated that for an average ammonia plant producing 660,000 tonne/year ammonia, 34 percent of the total CO_2 would be vented from the primary reformer flue gas and 66 percent would be vented from the CO_2 stripper vent. The ratio of CO_2 to ammonia produced is 1.87 tonnes CO_2/tonne ammonia. (4) Applying these values to this study, the reference ammonia plant produces 805,932 tonnes CO_2 vented from the CO_2 stripper. It will be assumed that the stripper vent CO_2 concentration is 99 percent by volume. (26)

Of the 24 U.S. ammonia plants, the capacity and production assumed in this study closely represents two plants; the Koch Nitrogen Co. plant located in Enid, OK and the Terra Industries Inc. plant located in Verdigris, OK. The capacities for these two plants were found to be 930,000 and 953,000 tonnes ammonia/year with a production of 669,600 and 686,160 tonnes ammonia/year, respectively. The reported urea capacity of these two plants was given as 346,527 tonnes/year and 495,614 tonnes/year, respectively. The chemical process of converting ammonia to urea through the intermediate ammonium carbamate gives a molar ratio of $2NH_3{:}1CO_2{:}1$Urea, as shown in the chemical reaction below.

$$2\ NH_3 + CO_2 \leftarrow\rightarrow H_2N\text{-}COONH_4 \leftarrow\rightarrow (NH_2)_2CO + H_2O$$

Using this stoichiometry of 1 mole CO_2 is used to create 1 mole of urea, a ratio of 0.733 tonnes CO_2/tonne urea is calculated. For the two existing plants previously referenced, using their known ammonia production, the assumed CO_2 production factor of 1.87 tonnes CO_2/tonne ammonia, the assumption of 66 percent CO_2 is related to process emissions, and the calculated 0.733 tonnes CO_2/tonne urea, the percentage of total CO_2 captured and recycled within the plant to produce urea can be approximated by matching the known urea capacity. The result of this calculation is that for the Koch Nitrogen Co. plant, approximately 20.3 percent of the total CO_2 produced is used in urea synthesis, and for the Terra Industries Inc. plant, approximately 28.3 percent of the total CO_2 produced is used in urea synthesis.

The calculated percentages of total plant CO_2 used in urea production are approximate numbers, and are subject to several different variables that affect these percentages. For example, the 0.733 tonnes CO_2/tonne urea is a stoichiometric ratio, and does not account for process losses, undesired products such as ammonium carbamate, or other factors within the urea synthesis loop such as recycle, that may increase the percentage of total plant CO_2 allocated for urea synthesis. Another factor that will affect the percentage of total plant CO_2 used will be the distribution of process versus combustion CO_2, which may deviate from the assumption of 66 percent used in this study. Yet another variable effecting this calculation is the urea plant capacity utilization factor for each plant, which is assumed to be 100 percent in this calculation. It is known that the reference Terra Industries Inc. plant does sell food grade liquid CO_2. The alternate use of CO_2 as a sellable food grade product is another consideration when approximating the amount of CO_2 used, as this will further reduce the amount of CO_2 available to be sold for EOR. However, the economics of a food grade product versus an EOR product would play a role in determining in which market the CO_2 would be sold.

Based on the two reference plants previously discussed, it is assumed that the ammonia plant in this study currently captures and recycles 28 percent of total plant CO_2 produced, for use in urea production, and this CO_2 is captured from the CO_2 stripper vent. This equates to approximately 43 percent of the CO_2 stripper vent stream that is already captured and re-used. This will leave approximately 57 percent of the total CO_2 stripper vent outlet stream still remaining and available for capture. The compression system for this case will be sized to the remaining 57 percent of the stream, or 458,399 tonnes CO_2/year.

7.2.3 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:

- An average plant will be represented by a capacity of 907,000 tonnes ammonia/year, with an actual production rate of 653,000 tonnes ammonia/year
- The ammonia process feedstock will be natural gas
- The gas from the stripper vent is assumed 99 volume percent CO_2 as given in the reference, and confirmed independently from vendor quotes utilizing a stripping column, and will be the one high-purity point source. The balance of the stream (1 volume percent) will be assumed as water
- The total high purity CO_2 amount produced by the plant is 805,932 tonnes CO_2/year; the amount available for capture after urea production is 458,399 tonnes CO_2/year
- 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 CO_2 quality will be based on the EOR “pipeline” standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters. (10) For the purpose of this study, 99 percent purity is assumed

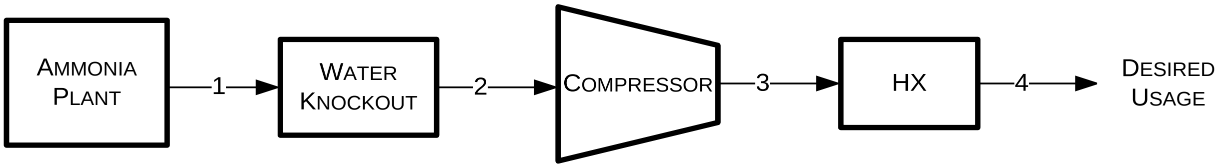
7.2.4 CO_2 Capture System

Only cooling and compression is required for this ammonia case. Reciprocating compression discussed previously in Section 6.1 will be employed, and scaled. Based on mass flow rate, this represents a large scale up of 3.24 times the quoted flow rate.

7.2.5 Block Flow Diagram, Stream Table, and Performance Summary

There is no cooling of the inlet stream for this ammonia case 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. After compression, the CO_2 product stream is cooled and sent directly for EOR or other usage. Exhibit 7-10 gives the block flow diagram (BFD) for this process.

Exhibit 7-10 Ammonia CO_2 capture block flow diagram



Source: NETL

Exhibit 7-11 Ammonia stream table

	1	2	3	4
V-L Mole Fraction				
AR	0.0000	0.0000	0.0000	0.0000
CH_4	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	0.0000	0.0000

CO_2	0.9709	0.9877	1.0000	1.0000
COS	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000
H_2O	0.0291	0.0123	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000
N_2	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	1,225	1,204	1,189	1,189
V-L Flowrate (kg/hr)	52,971	52,595	52,329	52,329
Solids Flowrate (kg/hr)	0	0	0	0
Temperature (°C)	21	21	144	49
Pressure (MPa, abs)	0.16	0.2	15.4	15.3
Enthalpy (kJ/kg)^A	27.89	28.2	45.4	-171.7
Density (kg/m^3)	2.9	2.9	247.9	673.8
V-L Molecular Weight	43.253	43.690	44.010	44.010
V-L Flowrate (lb_{mol}/hr)	2,700	2,654	2,621	2,621
V-L Flowrate (lb/hr)	116,780	115,953	115,365	115,365
Solids Flowrate (lb/hr)	0	0	0	0
Temperature (°F)	69	69	292	120
Pressure (psia)	23.5	23.5	2,239.5	2,217.3
Enthalpy (Btu/lb)^A	12.0	12.1	19.5	-73.8
Density (lb/ft^3)	0.184	0.183	15.474	42.064

The performance results are based on the reciprocating compressor quote and are provided in Exhibit 7-12.

Exhibit 7-12 Performance summary

Performance Summary	
Item	**907,000 tonnes ammonia/year (kW_e)**
CO_2 Compressor	5,610

7.2.6 Capture Integration

Integration of a capture system to the stripper vent of an ammonia plant should be relatively simple. Several plants in the ammonia industry capture and utilize CO_2 that they produce for urea production, as stated previously. Therefore, the required utilities, and at some plants, the necessary equipment will already be present. This provides complications with how to approach the implementation of CO_2 capture to this type of facility. For a Greenfield site that also produces urea, a compression system would be sized to 100 percent of the stripping vent flow; however, for a retrofit case the compression system could be sized to only the remaining uncompressed portion of the stripping vent, or sized to 100 percent as in the Greenfield case. For the purposes of this analysis, it is assumed that a portion of this stream is already compressed using existing equipment, and, therefore, the additional compression system implemented here is sized for the remaining portion of the stripping vent stream.

The potential integration of both the CO_2 capture for urea production and CO_2 capture for EOR into one compression train has both advantages and disadvantages. For the current case where the compression system is sized to only 57 percent of the stripping vent stream, reciprocating compression is required to be used due to the low mass flow. If these point sources were combined and compressed in one train, centrifugal compression may be used, and economies of scale may be gained due to the larger flow. However, this might also dictate the pressure at which the ammonia synthesis loop and urea synthesis loop within the base plant must operate, assuming no additional booster compression for these respective loops is available. Therefore, this problem becomes an optimization of gains in economies of scale resulting from one compression train versus gains or losses in ammonia/urea synthesis loop efficiency. Due to the complexity of this scenario, the base plant specific nature of this sensitivity, and the fact that the base plant is outside the scope of this study, this case it is not considered. However, it is important to point out as it represents a major area of integration for this case.

If the existing ammonia plant produces urea, as is assumed here, there will be an existing cooling water system in place to satisfy the condenser cooling duty for the CO_2 removal system, as well as other cooling loads in the urea synthesis loop. However, for the purposes of this study, it is not assumed that any existing system has the remaining capacity to handle the cooling required for the additional compression system. Therefore, it is assumed that a stand-alone cooling system will be required. The inclusion of an additional cooling water system would have to be evaluated on a case-by-case basis.

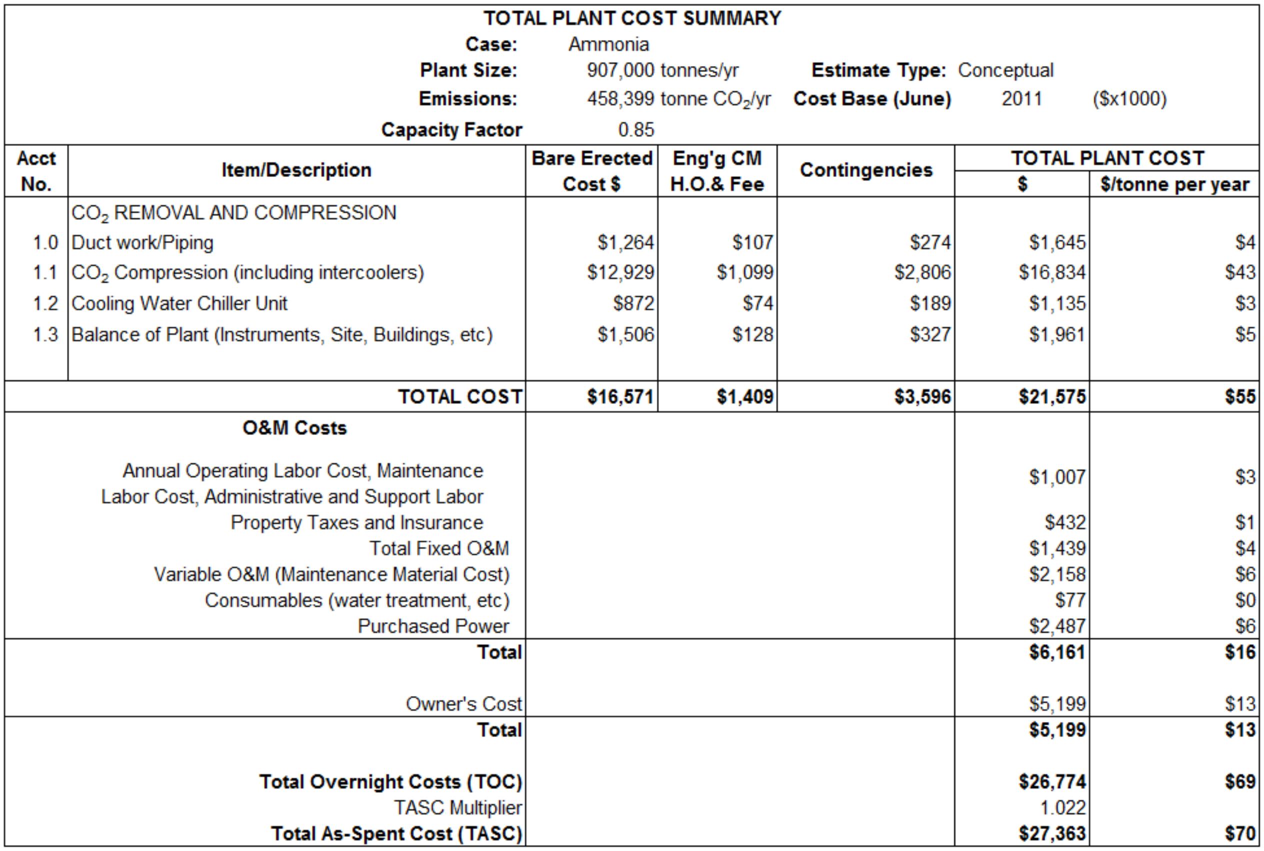
7.2.7 Power Source

Given the relatively small amount of CO_2, the compression power consumption is 5.61 MW. Power consumption estimates for the cooling system were scaled as described in Section 6.4. The total power requirement was calculated to be 5.7 MWh/hr, which includes all power required by the compression train and the cooling system. Power will be purchased at a rate of $58.59/MWh as discussed in Section 6.

7.2.8 Economic Analysis Results

The O&M, TOC for Greenfield and retrofit, and breakeven costs are given in this section for analysis and discussion. Owner’s costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner’s costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

Exhibit 7-13 Capital and O&M costs for ammonia Greenfield site

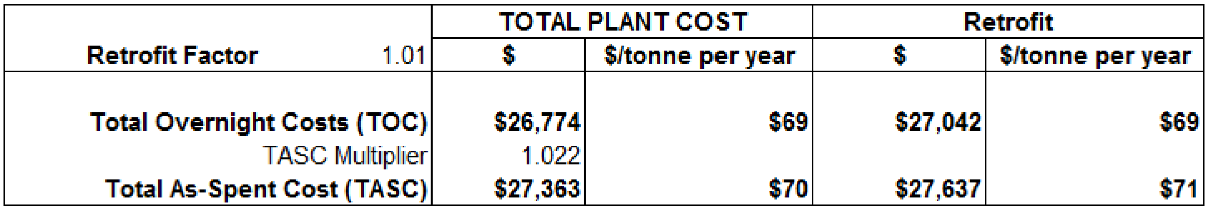


TOTAL PLANT COST SUMMARY

Case: Ammonia
Plant Size: 907,000 tonnes/yr **Estimate Type:** Conceptual
Emissions: 458,399 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST	
					$	$/tonne per year
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$1,264	$107	$274	$1,645	$4
1.1	CO_2 Compression (including intercoolers)	$12,929	$1,099	$2,806	$16,834	$43
1.2	Cooling Water Chiller Unit	$872	$74	$189	$1,135	$3
1.3	Balance of Plant (Instruments, Site, Buildings, etc)	$1,506	$128	$327	$1,961	$5
	TOTAL COST	**$16,571**	**$1,409**	**$3,596**	**$21,575**	**$55**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$1,007	$3
	Property Taxes and Insurance				$432	$1
	Total Fixed O&M				$1,439	$4
	Variable O&M (Maintenance Material Cost)				$2,158	$6
	Consumables (water treatment, etc)				$77	$0
	Purchased Power				$2,487	$6
	Total				**$6,161**	**$16**
	Owner's Cost				$5,199	$13
	Total				**$5,199**	**$13**
	Total Overnight Costs (TOC)				**$26,774**	**$69**
	TASC Multiplier				1.022	
	Total As-Spent Cost (TASC)				**$27,363**	**$70**

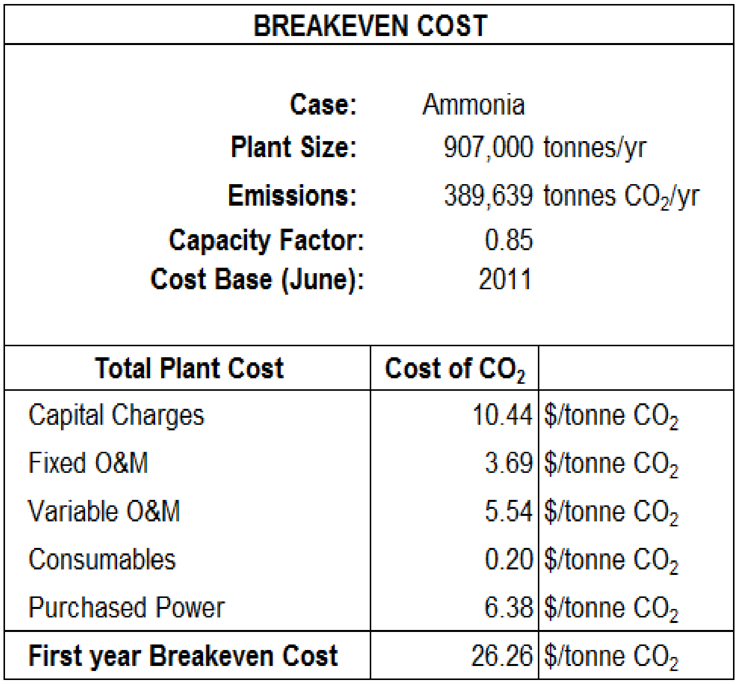
Exhibit 7-14 Total overnight costs for retrofit



		TOTAL PLANT COST		Retrofit	
Retrofit Factor	1.01	**$**	**$/tonne per year**	**$**	**$/tonne per year**
Total Overnight Costs (TOC)		**$26,774**	**$69**	**$27,042**	**$69**
TASC Multiplier		1.022			
Total As-Spent Cost (TASC)		**$27,363**	**$70**	**$27,637**	**$71**

The first-year breakeven cost for a Greenfield site and a retrofit site was calculated and is shown in Exhibit 7-15 and Exhibit 7-16 below. The financial assumptions and retrofit factor used represent the assumptions detailed in Section 5.

Exhibit 7-15 Breakeven cost for 907,000 tonnes ammonia/year Greenfield

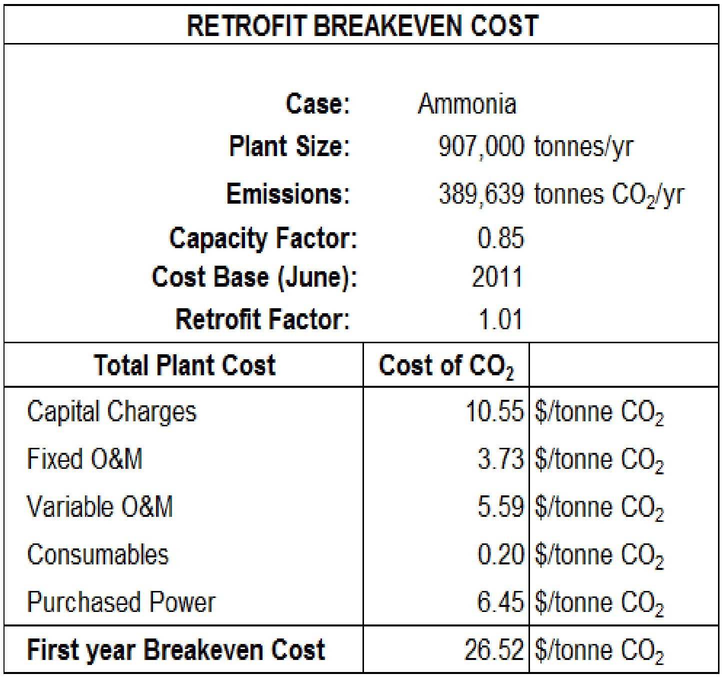


BREAKEVEN COST

Case: Ammonia
Plant Size: 907,000 tonnes/yr
Emissions: 389,639 tonnes CO_2/yr
Capacity Factor: 0.85
Cost Base (June): 2011

Total Plant Cost	Cost of CO_2	
Capital Charges	10.44	\$/tonne CO_2
Fixed O&M	3.69	\$/tonne CO_2
Variable O&M	5.54	\$/tonne CO_2
Consumables	0.20	\$/tonne CO_2
Purchased Power	6.38	\$/tonne CO_2
First year Breakeven Cost	26.26	\$/tonne CO_2

Exhibit 7-16 Breakeven cost for 907,000 tonnes ammonia/year retrofit



RETROFIT BREAKEVEN COST

Case: Ammonia
Plant Size: 907,000 tonnes/yr
Emissions: 389,639 tonnes CO_2/yr
Capacity Factor: 0.85
Cost Base (June): 2011
Retrofit Factor: 1.01

Total Plant Cost	Cost of CO_2	
Capital Charges	10.55	\$/tonne CO_2
Fixed O&M	3.73	\$/tonne CO_2
Variable O&M	5.59	\$/tonne CO_2
Consumables	0.20	\$/tonne CO_2
Purchased Power	6.45	\$/tonne CO_2
First year Breakeven Cost	26.52	\$/tonne CO_2

7.2.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

- Plant Size

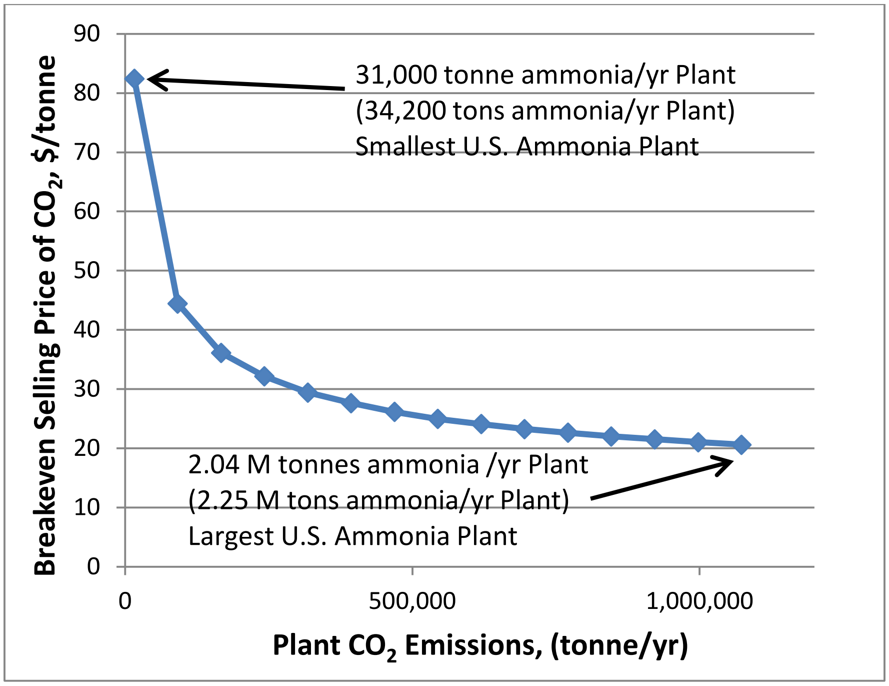
- COE
- CO_2 Purity
- CCF

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore the only sensitivity discussed in this section. The other three sensitivities, COE, CO_2 purity, and CCF, are discussed and compared across the other eight cases in Section 9.2.

7.2.9.1 Plant Size

Plant size was varied over the capacity range of 31,000 tonnes ammonia/year to 2.04 M tonnes ammonia/year. These values use the same set of Ammonia plant assumptions set forth in section 7.2.2, and represent the smallest and largest U.S. Ammonia plant capacities as given by the U.S. EPA in 2009, respectively. As shown in Exhibit 7-17, as the plant size decreases over the given range, the breakeven cost increases by $61.81/tonne CO_2. This variation over the range is much more significant than in the ethanol case; however, the variable range is larger. The large jump in breakeven selling price is attributed to the large decrease in plant capacity.

Exhibit 7-17 Plant size sensitivity



Source: NETL

7.2.10 Ammonia Conclusion

The high-purity CO_2 stream produced from ammonia plants makes them an attractive industrial process for CO_2 capture and use since the plant itself acts as the separation medium.

The results showed the breakeven price of CO_2 to be $26.26/tonne CO_2 for a Greenfield site, and $26.52/ tonne CO_2 for a retrofit site. The small disparity is the result of small differences in equipment required for a Greenfield versus retrofit site, assuming adequate plot-plan space for the retrofit case. Sensitivities were performed on the following variables: COE, CCF, and plant size. The sensitivity tests for plant size showed the largest change in breakeven selling price, where when varying over the entire 2006 U.S. ammonia fleet plant capacity, a change in price of \$61.81/tonne CO_2 was observed. However, the low end of U.S. plant capacity may likely be too small to justify this type of application.

The sensitivity test results for COE and CCF, detailed in Section 9.2, demonstrated a change in breakeven selling price. Variation of COE over a \$120/MWh range resulted in a \$13.07/tonne CO_2 increase in the breakeven price. Variation of CCF over a 25 percent range, 10-35 percent, resulted in a \$17.18/tonne CO_2 increase in the breakeven price.

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 such as in the beverage industry. For example, in addition to ammonia and urea, the Terra Industries Inc. plant previously mentioned in section 7.2.2 also produces food grade liquid CO_2 as a sellable product. The amount and respective selling price was unspecified; however this would reduce or eliminate the amount of high purity CO_2 potentially available for EOR. This scenario was not considered in this study as it would need to be evaluated on a case by case basis; however it is important to note that minor sellable products like food grade CO_2 will reduce the amount of CO_2 available from ammonia, and other high purity CO_2 producing industries.

In parallel with potential CO_2 sold to the food industry, there is also the potential for a combined urea and ammonia plant to expand urea production to utilize all available CO_2 instead of selling for EOR purposes. However, if the current urea capacity utilization rate is at 100 percent, this would require a large capital investment to develop additional urea capacity, and this urea capital investment would most likely be higher than the capital investment required to provide a sellable CO_2 EOR product. This specific scenario would require further research to develop a suitable characterization of the potential choices existing ammonia plants would need to evaluate.

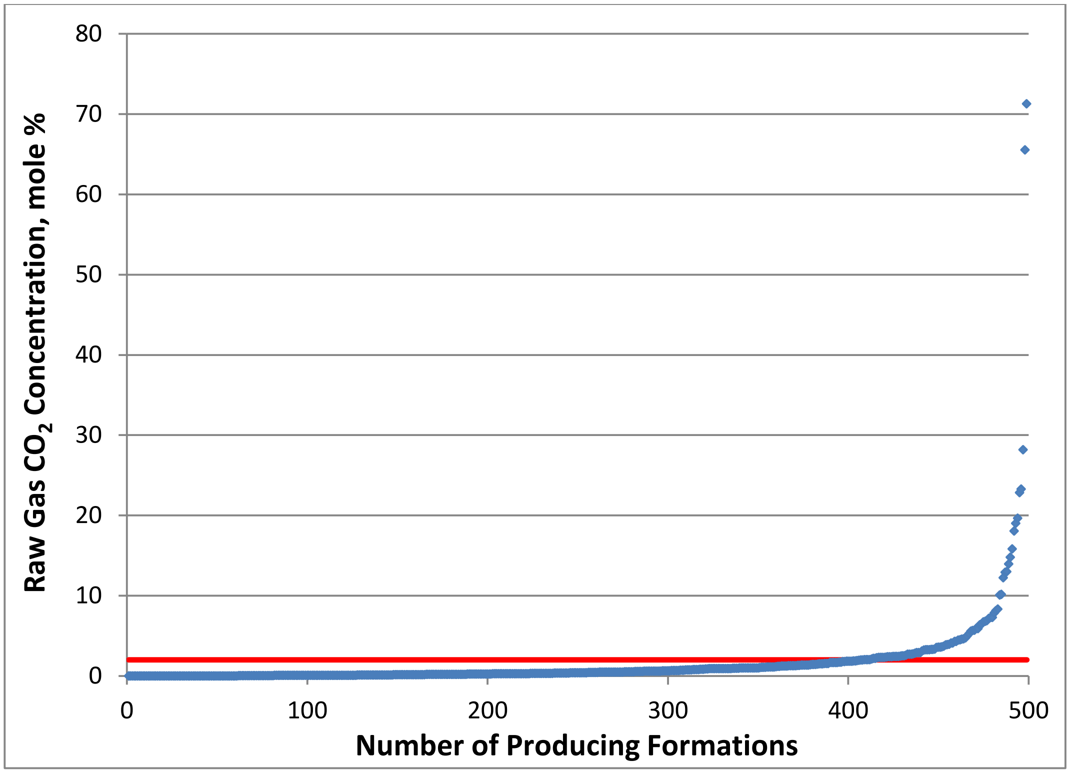
Further ammonia examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

7.3 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 CO_2 separation and removal are inherently necessary to the processing of natural gas, given a higher raw natural gas CO_2 content than finished product specifications would allow, natural gas processing presents as an attractive, potentially low cost source of industrial CO_2. One example of where natural gas processing has been utilized as an industrial source of CO_2 capture for EOR use is the LaBarge Field-Shute Creek processing facility in Wyoming. (27) At this facility, raw natural gas with a high inlet sulfur concentration of five percent is processed to produce a sellable CO_2 product for EOR applications. This facility also uses acid gas removal and acid gas injection to remove unwanted sulfur species. Along with CO_2, other sellable products include processed natural gas, electricity, and helium. (27) This specific case represents what may be considered an outlier data point, as the raw natural gas has a high inlet concentration of sulfur species, and the facility also produces electricity for export, which isn’t common in natural gas processing facilities.

While the actual processing facility serves as the separation medium, producing a high purity CO_2 stream, not all raw natural gas requires processing to meet pipeline specifications. In some instances, high-quality raw gas can be sent directly to a product pipeline or borderline high-quality raw gas may be blended with high-quality gas to form a product suitable to be sent directly for sale. The Gas Technology Institute (GTI) published and updates a natural gas database which gives the natural gas compositions from producing formations for the lower-48 states. They report a total of 499 producing formations with associated gas compositions. Shown in Exhibit 7-18 is a plot of each producing formation with its associated CO_2 composition. The red line represents 2 mole percent CO_2 concentrations.

Exhibit 7-18 GTI's reported gas composition



Source: NETL

Based only on allowable CO_2 content and neglecting H_2S or N_2 content limitations, and assuming that less than 2 percent CO_2 is a standard pipeline specification, only 90 of the 499 reported producing formations would require CO_2 removal processing. However, the previously mentioned blending technique may lower the number of producing formations requiring processing even further.

EIA reported in 2004 that 24.2 Trillion cubic feet (Tcf) of raw natural gas was produced. (28) Of this total, 0.41 percent was flared, 15.29 percent was re-injected to reservoirs to maintain pressure, and 6.2 percent was removed as a wet fraction, totaling 21.9 percent of raw gas extracted that was removed. (28) The remaining balance was dry gas. The GTI 1998 database update reports a total Lower-48 formation production of 23.4 Tcf. Applying the previous values to the GTI database wellhead production for 1998 as reasonable approximations, gives a dry gas production of 18.3 Tcf. EIA reports U.S. dry gas production from gas wells in 1998 to be 19 Tcf, which shows good agreement between EIA data and the GTI database. (29) Additionally, no shale gas processing is considered in this study.

Of the higher content CO_2 producing formations (5 percent CO_2 and up), direct correlation between a producing formation and a processing site is not possible. For example, EIA lists eight gas processing facilities in the state of Michigan. The GTI database lists 5 producing formations in the Michigan Basin, but does not state which facilities these formations feed. However, of the 5 producing formations, 4 have CO_2 concentrations at or below 1.2 percent CO_2, with the fifth having a CO_2 concentration of 10.2 percent. It could be reasoned that this fifth producing formation is the only one requiring processing, when considering only CO_2 content, and therefore this formation feeds and satisfies the majority of processing plants in the Michigan Basin.

For the purposes of this study, the 10 percent CO_2 content of the producing formation found in Michigan will be used as the reference raw gas concentration.

7.3.1 Size Range

According to the U.S. EIA, in 2009 there were 493 natural gas processing plants in operation in the U.S. with a total operating capacity of 77 Billion cubic feet (Bcf)/day. The average 2009 utilization rate was given at 66 percent. (30) Of the total, 53 percent fall within the range of 0-50 mmcf/day, and 28 percent fall within the range of 50-200 mmcf/day. (30) The average 2009 plant capacity was given as 139 mmcf/day. (30) The general trend appears to be small processing facilities, with 81 percent of plants falling into the 0-200 mmcf/d range. Factors determining the amount of CO_2 available for capture include plant size, as well as CO_2 concentration in the raw natural gas; therefore, a specific combination of these two factors is necessary to warrant capture.

Of the eight Michigan processing facilities previously mentioned, two fall in the 500-800 mmcf/day range, with the remaining six in the 0-50 mmcf/day range. For the purposes of this study, it will be assumed that the reference plant has a capacity of 500 mmcf/day. This is based on the Michigan processing facilities, as well as the fact that a 500 mmcf/day facility is large enough to be able to justify the addition of capture, with a large enough CO_2 supply to adequately drive down the CO_2 breakeven selling price.

The composition of the raw gas processed will be represented by the Michigan Basin producing formation with 10.2 percent CO_2. The full raw gas characteristics are given in Exhibit 7-19, and represent average concentrations which are not expected to sum to 100 percent.

Exhibit 7-19 Michigan basin producing formation raw gas characteristics

Michigan Basin Raw Gas Characteristics	
Component	**Average Mole %**
CH_4	82.35
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	10.17
N_2	2.23
He	0.00
Other	0.00

Given this plant capacity and applying the 2009 utilization rate of 66 percent, coupled with the raw natural gas CO_2 composition, this plant would have approximately 649,198 tonnes CO_2/year available for capture.

It should be noted that the assumptions for this study's reference plant are not limited to only 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 GTI database.

7.3.2 CO_2 Point Sources

Natural gas processing (or gas sweetening) takes raw natural gas, which can contain 2 to 70 percent CO_2 by volume as previously stated, and removes CO_2 and other impurities to meet the required pipeline or liquefaction specifications. The single point source is the separated CO_2 stream that is generally vented to the atmosphere. The variation in raw natural gas CO_2 content will affect the amount of CO_2 available for capture; however, the concentration of the CO_2 stream to be captured will be very high at 96 to 99 percent. The plant specifications for the natural gas product composition, pipeline or liquefied natural gas (LNG), will also affect, but to a lesser degree, the amount of CO_2 to be captured.

7.3.3 Design Input and Assumptions

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

- An average plant will be represented by a capacity of 500 mmcf/d of raw gas processed, with an actual production rate of 330 mmcf/d
- The raw gas CO_2 content is 10.2 mole percent

- The CO_2 amount, as calculated above after applying the 2009 average capacity utilization factor of 66 percent is 649,198 tonnes CO_2/year, neglecting processing losses
- The CO_2 stream temperature is 69°F
- The CO_2 stream pressure is 23.52 psia
- The CO_2 stream is 99 volume % CO_2, balanced with H_2O
- The CO_2 quality will be based on the EOR "pipeline" standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters. (10) For the purpose of this study, 99 volume percent purity is assumed

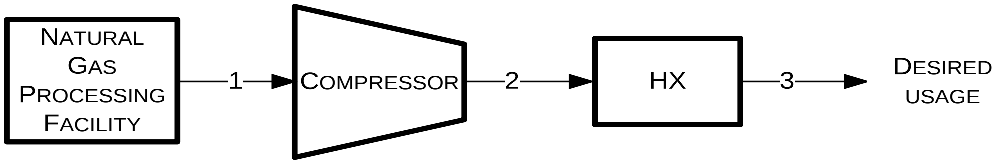
7.3.4 CO_2 Capture System

Only cooling and compression is required for this natural gas processing case. Given the amount of CO_2 available for capture, a centrifugal compressor, discussed in Section 6.1, is used to attain 2,200 psig EOR pipeline pressure as specified in QGESS. (10)

7.3.5 Block Flow Diagram, Stream Table, and Performance Summary

Since the stripping column releases 99 volume percent CO_2 balanced with water, only compression 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 7-20 gives the BFD for this process.

Exhibit 7-20 CO_2 capture block flow diagram



Source: NETL

Exhibit 7-21 Natural gas processing 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	1.0000	1.0000
COS	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000
H_2O	0.0100	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)	1,701	1,684	1,684
V-L Flowrate (kg/hr)	74,416	74,109	74,109
Solids Flowrate (kg/hr)	0	0	0
Temperature (°C)	21	103	49
Pressure (MPa, abs)	0.16	15.3	15.3
Enthalpy (kJ/kg)^A	25.83	-23.2	-171.6
Density (kg/m^3)	2.9	324.9	673.4
V-L Molecular Weight	43.750	44.010	44.010
V-L Flowrate (lb_{mol}/hr)	3,750	3,712	3,712
V-L Flowrate (lb/hr)	164,059	163,383	163,383
Solids Flowrate (lb/hr)	0	0	0
Temperature (°F)	69	218	120
Pressure (psia)	23.5	2,216.0	2,215.3
Enthalpy (Btu/lb)^A	11.1	-10.0	-73.8
Density (lb/ft^3)	0.183	20.283	42.040

The performance results are based on the centrifugal compressor discussed in Section 6.1. The performance summary is provided in Exhibit 7-22.

Exhibit 7-22 Performance summary

Performance Summary	
Item	**500 mmscf/d (kW_e)**
CO_2 Compressor	7,100

7.3.6 Capture Integration

In this instance, the base plant is also considered the capture system. Therefore, there is little opportunity for 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 inter-cooling for the compression process. However, the necessity for a standalone cooling water system would need to be evaluated on a case-by-case basis. There could be a potential to integrate make-up water to feed or partially feed the cooling system thereby reducing the unit size or replacing it completely with a simple heat exchanger.

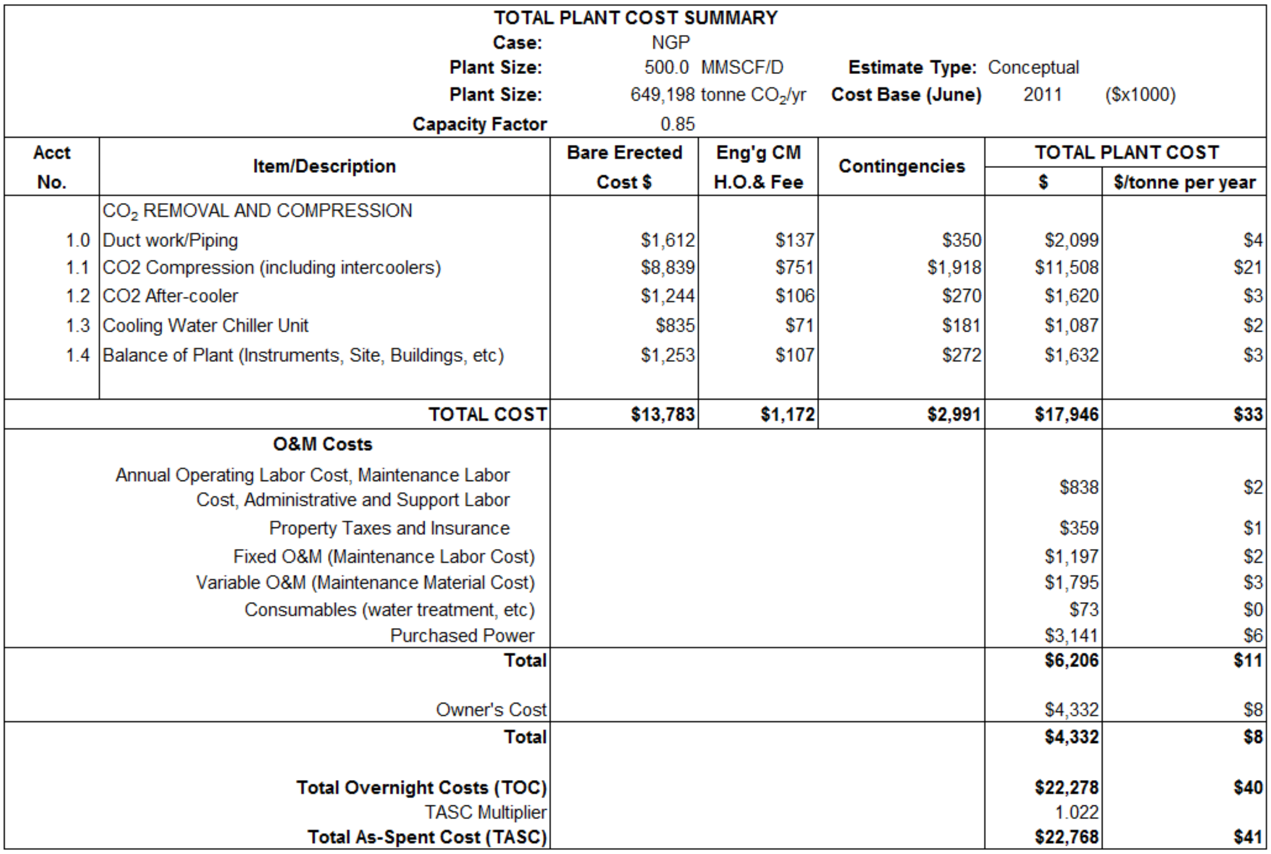
7.3.7 Power Source

The compressor power consumption for this case is 7.1 MW. Power consumption estimates for the cooling water system were scaled as described in Section 6.4. The total power requirement was calculated to be 7.2 MWh/hr, which includes all power required by the compression train and the cooling water system. Power will be purchased at a rate of $58.59/MWh as discussed in Section 6. For practical applications for this type of facility with natural gas readily available, the power required to operate the cooling system as well as the compression system could easily be generated on site, but this scenario should be looked at on a case-by-case basis. Depending on the size and location of the facility, there could be multiple reasons to produce the required power on-site.

7.3.8 Economic Analysis Results

The O&M, TOC for Greenfield and retrofit, and breakeven costs are given in this section for analysis and discussion. Owner’s costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner’s costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

Exhibit 7-23 Capital and O&M costs for natural gas processing Greenfield site



TOTAL PLANT COST SUMMARY

Case: NGP

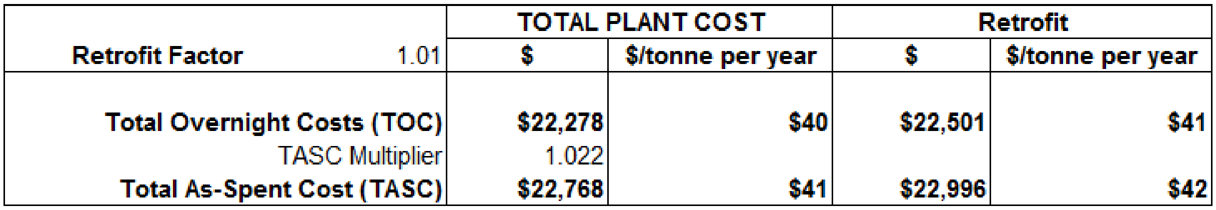
Plant Size: 500.0 MMSCF/D   Estimate Type: Conceptual

Plant Size: 649,198 tonne CO_2/yr   Cost Base (June) 2011 ($x1000)

Capacity Factor 0.85

Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST $	TOTAL PLANT COST $/tonne per year
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$1,612	$137	$350	$2,099	$4
1.1	CO2 Compression (including intercoolers)	$8,839	$751	$1,918	$11,508	$21
1.2	CO2 After-cooler	$1,244	$106	$270	$1,620	$3
1.3	Cooling Water Chiller Unit	$835	$71	$181	$1,087	$2
1.4	Balance of Plant (Instruments, Site, Buildings, etc)	$1,253	$107	$272	$1,632	$3
	TOTAL COST	**$13,783**	**$1,172**	**$2,991**	**$17,946**	**$33**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$838	$2
	Property Taxes and Insurance				$359	$1
	Fixed O&M (Maintenance Labor Cost)				$1,197	$2
	Variable O&M (Maintenance Material Cost)				$1,795	$3
	Consumables (water treatment, etc)				$73	$0
	Purchased Power				$3,141	$6
	Total				**$6,206**	**$11**
	Owner's Cost				$4,332	$8
	Total				**$4,332**	**$8**
	Total Overnight Costs (TOC)				**$22,278**	**$40**
	TASC Multiplier				1.022	
	Total As-Spent Cost (TASC)				**$22,768**	**$41**

Exhibit 7-24 Total overnight costs for retrofit



		TOTAL PLANT COST		Retrofit	
Retrofit Factor	1.01	**$**	**$/tonne per year**	**$**	**$/tonne per year**
Total Overnight Costs (TOC)		**$22,278**	**$40**	**$22,501**	**$41**
TASC Multiplier		1.022			
Total As-Spent Cost (TASC)		**$22,768**	**$41**	**$22,996**	**$42**

The first-year breakeven cost for a Greenfield site and a retrofit site was calculated and is shown in Exhibit 7-25 and Exhibit 7-26 below. The financial assumptions and retrofit factor used represent the assumptions detailed in Economic Analysis Overview.

Exhibit 7-25 Breakeven cost for 500 MMscfd Greenfield

BREAKEVEN COST		
Case:	NGP	
Plant Size:	500.0 MMSCFD	
Plant Size:	551,818 tonne CO_2/yr	
Capacity Factor:	0.85	
Cost Base (June):	2011	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	6.14	$/tonne CO_2
Fixed O&M	2.17	$/tonne CO_2
Variable O&M	3.25	$/tonne CO_2
Consumables	0.13	$/tonne CO_2
Purchased Power	5.69	$/tonne CO_2
First year Breakeven Cost	17.38	$/tonne CO_2

Exhibit 7-26 Breakeven cost for 500 MMscfd retrofit

RETROFIT BREAKEVEN COST		
Case:	NGP	
Plant Size:	500.0 MMSCFD	
Plant Size:	551,818 tonne CO_2/yr	
Capacity Factor:	0.85	
Cost Base (June):	2011	
Retrofit Factor:	1.01	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	6.20	$/tonne CO_2
Fixed O&M	2.19	$/tonne CO_2
Variable O&M	3.29	$/tonne CO_2
Consumables	0.13	$/tonne CO_2
Purchased Power	5.75	$/tonne CO_2
First year Breakeven Cost	17.56	$/tonne CO_2

7.3.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

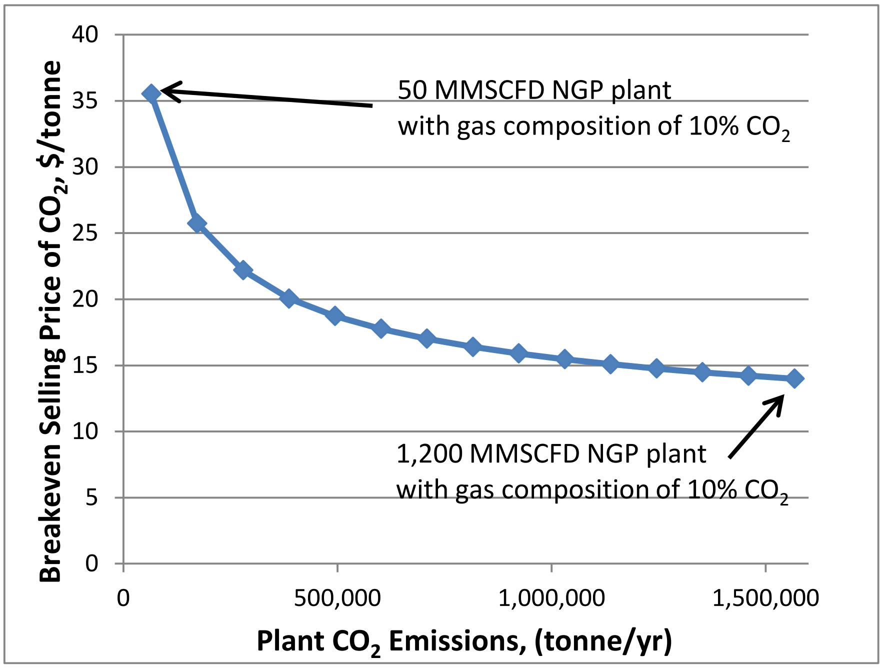
- Plant Size
- COE
- CO_2 Purity
- CCF

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore the only sensitivity discussed in this section. The other three sensitivities, COE, CO_2 purity, and CCF, are discussed and compared across the other eight cases in section 9.2.

7.3.9.1 Plant Size

Plant size was varied over the range of 50 MMscfd to 1,200 MMscfd. The x-axis refers to plant CO_2 emissions from the stripping column only, and does not include emission from combustion, onsite power generation, or other. The plant size range uses the same set of natural gas processing plant assumptions set forth in section 7.3.3, specifically the 10 percent inlet raw natural gas CO_2 concentration assumption. As shown in Exhibit 7-27, as the plant size decreases over the given range, the breakeven cost increases by \$21.54/tonne CO_2.

Exhibit 7-27 Plant size sensitivity



Source: NETL

The general trend of the plant size sensitivity results shown here for fixed concentration, variable plant size, may also be interpreted as an analogous sensitivity for fixed plant size, variable CO_2 concentration. For a fixed plant size, as CO_2 concentration decreases, the amount of CO_2 available also decreases. This roughly relates to the 50 MMscfd sensitivity result. The opposite, increasing CO_2 concentration, roughly compares to the 1,200 MMscfd case. The conclusion to be drawn is that the amount of CO_2 available to be captured and sold is the important parameter, and any combination of changing plant size and CO_2 concentration will have an effect on the breakeven selling price.

7.3.10 Natural Gas Processing Conclusion

The high-purity CO_2 stream produced from natural gas processing plants makes them an attractive industrial process since CO_2 separation is inherent to normal operations.

A compression system for a 500 MMscfd natural gas processing plant was modeled and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be \$17.38/tonne CO_2 for a Greenfield site, and \$17.56/tonne CO_2 for a retrofit site. The small disparity is the result of small differences in equipment required for a Greenfield versus retrofit site, assuming adequate plot plan space for the retrofit case. Sensitivities were performed on the following variables: COE, CCF, and plant size. The sensitivity tests for plant size showed the largest change in breakeven selling price, where a range of 50 – 1,200 MMscfd resulted in a rise in the breakeven selling price of \$21.54/tonne CO_2.

The sensitivity tests for COE and CCF, discussed in detail in Section 9.2, demonstrated a change in breakeven selling price. Variation of COE over a \$20-140/MWh range resulted in an \$11.66/tonne CO_2 increase in the breakeven price. Variation of CCF over a 25 percent range, 10-35 percent, resulted in a \$10.09/tonne CO_2 increase in the breakeven price.

Further natural gas processing examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

7.4 Ethylene Oxide

EO is a colorless flammable gas that is mainly used as a raw material for production of several industrial chemical intermediates. (31) 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. (31)

Global EO production in 2009 was approximated to be 19 M tonnes (31). North American production capacity in 2007 was reported as 4.8 M tonnes, with U.S. production capacity accounting for 3.6 M tonnes. (32)

EO is produced by direct oxidation of ethylene in the presence of a silver catalyst. The reaction conditions range from 200-300°C and 10-30 bar. (31) The reaction is exothermic, and therefore steam may be generated by excess heat, and the product stream can use physical sorbents to remove CO_2. The reaction stoichiometry suggests that CO_2 is produced during the oxidation reaction in a ratio of 6:2 EO:CO_2; however, there is limited data available on the rate of CO_2 generation during reaction. Other literature suggests the reactor product stream CO_2 concentration to be 8 percent CO_2, (24) but these values do not represent the stream where CO_2 may be captured from. The large potential variation in concentration of CO_2 in the reactor product stream, however, will skew the calculation of CO_2 available for capture and use on a U.S. or global scale.

7.4.1 Size Range

Current EO U.S. plant sizes range from 105,000 tonnes to 770,000 tonnes. Exhibit 7-28 shows the ten U.S. EO production facilities and their associated capacity as of 2007.

Exhibit 7-28 2007 U.S. EO production facility capacities

Company	Location	Capacity (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 tons ethylene oxide, which is representative of the majority of EO plants, and is, therefore, the basis for this study. With a 6:2 ratio of EO:CO_2, this plant size will produce 121,500 tonnes CO_2/year. The International Energy Agency's Greenhouse Gas R&D Program (IEA GHG) database gives an average emission for the 52 worldwide EO production sites of 150,000 tonnes CO_2/year (31), which is in-line with the assumptions of this study.

7.4.2 CO_2 Point Sources

EO is considered a high purity source. The process has a single CO_2 point source: the CO_2 removal system. The removal system may be one of several types—physical sorbents such as Rectisol or Selexol, chemical sorbents such as MEA, or cryogenic separation. 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.

Some reports state that the CO_2 concentration can range from 30-100 percent CO_2 (33); however, it is unclear what streams this CO_2 concentration range considers. Most references give a range of 95-100 percent CO_2 concentration for the stream to be captured. (34) For the purpose of this study, the concentration is assumed as 100 percent CO_2.

7.4.3 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:

- An average plant will be represented by a capacity of 364,500 tonnes of EO/year
- The CO_2 amount, as calculated above, is 121,500 tonnes CO_2/year at 100 percent utilization and neglecting processing losses
- The CO_2 stream is 100 volume % 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 CO_2 quality will be based on the EOR "pipeline" standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters. (10) For the purpose of this study, 100 volume percent purity is assumed

7.4.4 CO_2 Capture System

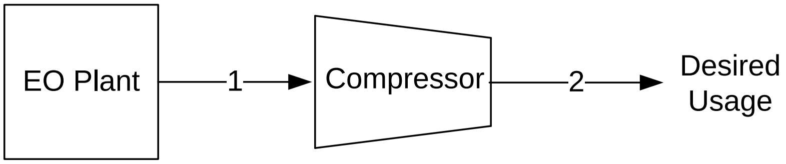
For this EO case, CO_2 separation is an inherent part of base plant operations, and, therefore, only compression and cooling is required. Given the low CO_2 flow rate, reciprocating compression will be employed and scaled for this case. Based on mass flow rate, this represents a scale down of 15 percent versus the quoted flow rate as given previously in Section 6.1.

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 matches the quoted 43.3 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 15 percent reduction in cost.

7.4.5 Block Flow Diagram, 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 7-29, 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. The compressed, cooled CO_2 is then sent to the EOR pipeline or elsewhere for end usage.

Exhibit 7-29 Ethylene oxide CO_2 capture block flow diagram



Source: NETL

Exhibit 7-30 Ethylene oxide 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
COS	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
Solids Flowrate (kg/hr)	0	0
Temperature (°C)	36	49
Pressure (MPa, abs)	0.30	15.3
Enthalpy (kJ/kg)^A	27.10	-171.6
Density (kg/m^3)	5.2	673.5
V-L Molecular Weight	44.010	44.010
V-L Flowrate (lb_{mol}/hr)	695	695
V-L Flowrate (lb/hr)	30,578	30,578
Solids Flowrate (lb/hr)	0	0
Temperature (°F)	96	120
Pressure (psia)	43.3	2,215.4
Enthalpy (Btu/lb)^A	11.7	-73.8
Density (lb/ft^3)	0.324	42.042

The performance summary is provided in Exhibit 7-31.

Exhibit 7-31 Performance summary

Performance Summary	
Item	**364,500 tonnes/yr (kW_e)**
CO_2 Compressor	1,250

7.4.6 Capture Integration

The reactor effluent is received by the AGR absorber at a temperature of 410°F (35) and will require cooling, indicating an existing cooling water system. 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. This would have to be evaluated on a case-by-case basis. It is interesting to note that if a power plant using a steam cycle is present, an efficient heat exchanger could capture this energy to heat condensate make-up.

For the purposes of this study, it will be assumed that an additional, stand-alone cooling unit will perform the necessary cooling for capture and compression. 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.

7.4.7 Power Source

Given the relatively small amount of CO_2, the compressor power consumption is 1.25 MW. Power consumption estimates for the cooling water system were scaled as described in Section 6.4. The total power requirement was approximated to be 1.3 MWh/hr, which includes all power required by the compression train and the cooling system. Power will be purchased at a rate of $58.59/MWh as discussed in Section 6. Given that the EO reaction is exothermic, and this additional heat is 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.

7.4.8 Economic Analysis Results

The O&M, TOC for Greenfield and retrofit and breakeven costs are given in this section for analysis and discussion. Owner's costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner's costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

Exhibit 7-32 Capital and O&M costs for EO Greenfield site

TOTAL PLANT COST SUMMARY

Case: EO
Plant Size: 364,500 tonnes/yr **Estimate Type:** Conceptual
Emissions: 121,501 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

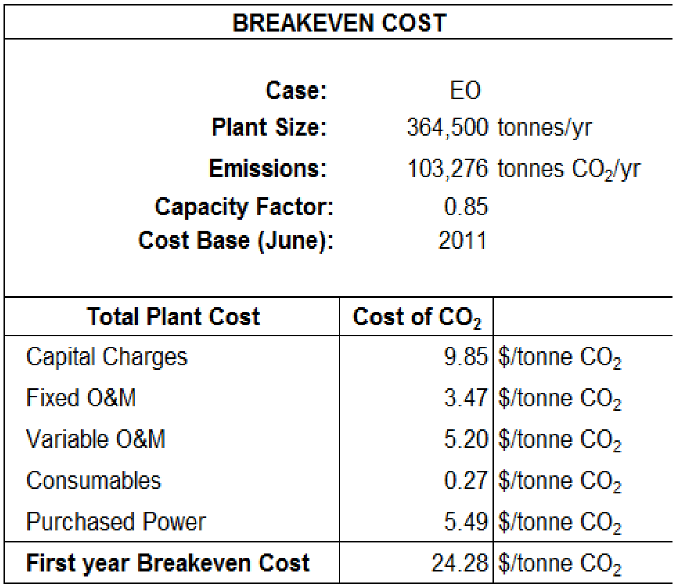
Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST $	TOTAL PLANT COST $/tonnes per year
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$499	$42	$108	$649	$6
1.1	CO_2 Compression (including intercoolers)	$2,869	$244	$623	$3,736	$36
1.2	Cooling Water Chiller Unit	$385	$33	$84	$502	$5
1.3	Balance of Plant (Instruments, Site, Buildings, etc)	$375	$32	$81	$488	$5
	TOTAL COST	**$4,128**	**$351**	**$896**	**$5,375**	**$52**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$251	$2
	Property Taxes and Insurance				$107	$1
	Total Fixed O&M				$358	$3
	Variable O&M (Maintenance Material Cost)				$537	$5
	Consumables (water treatment, etc)				$28	$0
	Purchased Power				$567	$5
	Total				**$1,490**	**$14**
	Owner's Cost				$1,320	$13
	Total				**$1,320**	**$13**
	Total Overnight Costs (TOC)				**$6,695**	**$65**
	TASC Multiplier				1.022	
	Total As-Spent Cost (TASC)				**$6,842**	**$66**

Exhibit 7-33 Total overnight costs for retrofit

		TOTAL PLANT COST		Retrofit	
Retrofit Factor	1.01	**$**	**$/tonnes per year**	**$**	**$/tonnes per year**
Total Overnight Costs (TOC)		**$6,695**	**$65**	**$6,762**	**$65**
TASC Multiplier		1.022			
Total As-Spent Cost (TASC)		**$6,842**	**$66**	**$6,910**	**$67**

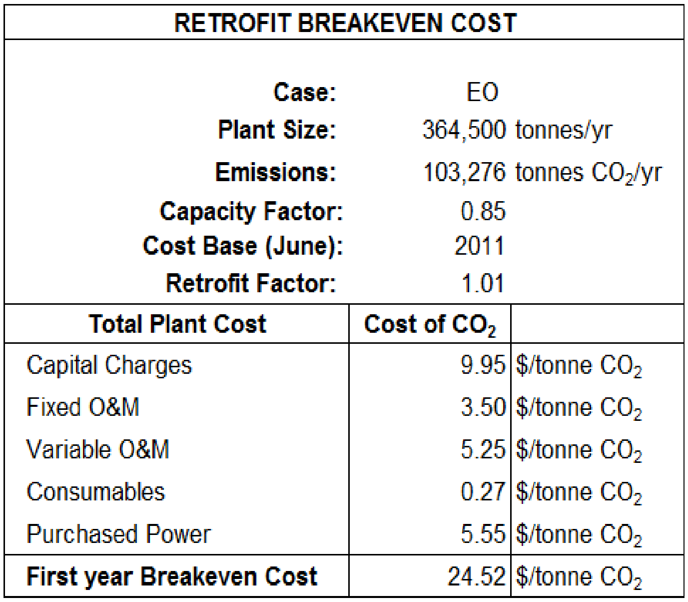
The first-year breakeven cost for a Greenfield site and a retrofit site was calculated and is shown in Exhibit 7-34 and Exhibit 7-35 below. The financial assumptions and retrofit factor used represent the assumptions detailed in Section 5.

Exhibit 7-34 Breakeven cost for 364,500 tonnes/yr Greenfield



BREAKEVEN COST		
Case:	EO	
Plant Size:	364,500 tonnes/yr	
Emissions:	103,276 tonnes CO_2/yr	
Capacity Factor:	0.85	
Cost Base (June):	2011	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	9.85	\$/tonne CO_2
Fixed O&M	3.47	\$/tonne CO_2
Variable O&M	5.20	\$/tonne CO_2
Consumables	0.27	\$/tonne CO_2
Purchased Power	5.49	\$/tonne CO_2
First year Breakeven Cost	24.28	\$/tonne CO_2

Exhibit 7-35 Breakeven cost for 364,500 tonnes/yr retrofit



RETROFIT BREAKEVEN COST		
Case:	EO	
Plant Size:	364,500 tonnes/yr	
Emissions:	103,276 tonnes CO_2/yr	
Capacity Factor:	0.85	
Cost Base (June):	2011	
Retrofit Factor:	1.01	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	9.95	\$/tonne CO_2
Fixed O&M	3.50	\$/tonne CO_2
Variable O&M	5.25	\$/tonne CO_2
Consumables	0.27	\$/tonne CO_2
Purchased Power	5.55	\$/tonne CO_2
First year Breakeven Cost	24.52	\$/tonne CO_2

7.4.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

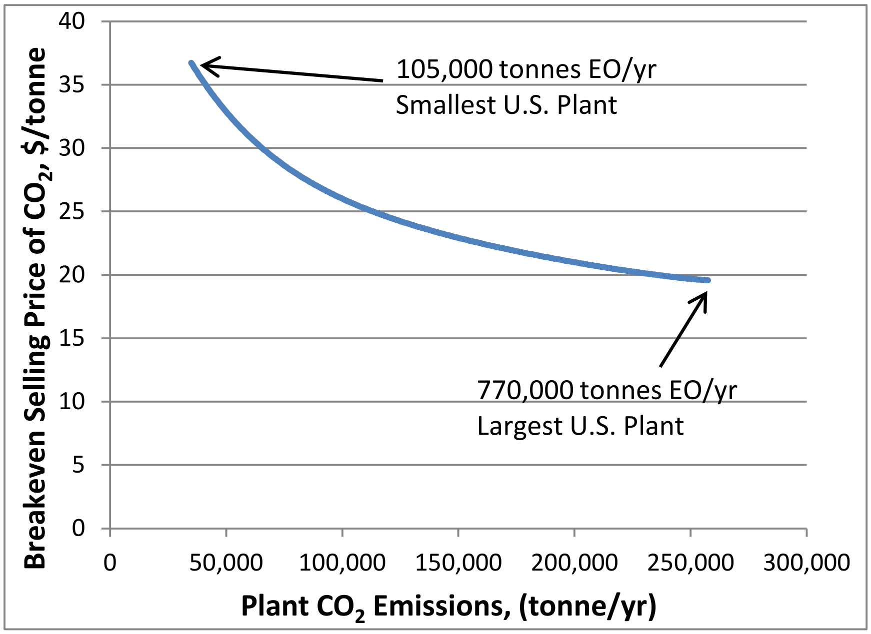
- Plant Size
- COE
- CO_2 Purity
- CCF

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore the only sensitivity discussed in this section. The other three sensitivities, COE, CO_2 purity, and CCF, are discussed and compared across the other eight cases in Section 9.2.

7.4.9.1 Plant Size

Plant size was varied over the range of 105,000 to 770,000 tonnes EO/yr, which represents the smallest and largest U.S. EO plants, respectively. These values use the same set of EO plant assumptions set forth in Section 7.4.3. As shown in Exhibit 7-36, as the plant size decreases over the given range, the breakeven cost increases by \$17.16/tonne CO_2.

Exhibit 7-36 Plant size sensitivity



Source: NETL

7.4.10 Ethylene Oxide Conclusion

The high-purity CO_2 stream produced from EO plants makes them an attractive industrial process since the plant itself performs the separation of CO_2 under normal operating conditions.

A compression system for a 364,500 tonne/year EO plant was modeled, and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be \$24.28/tonne CO_2 for a Greenfield site, and \$24.52/tonne CO_2 for a retrofit site. The small disparity is the result of small differences in equipment required for a Greenfield versus retrofit site, assuming adequate plot plan space for the retrofit case. Sensitivities were performed on the following variables: COE, CCF, and plant size. The sensitivity tests for plant size showed the largest change in breakeven selling price, where a range of 105,000 – 770,000 tonnes EO/yr, representing U.S. plant capacity range, resulted in a rise in the breakeven selling price of \$17.16/tonne CO_2.

The sensitivity tests for COE and CCF also demonstrated a change in breakeven selling price. Variation of COE over a \$20-140/MWh range resulted in an \$11.25/tonne CO_2 increase in the breakeven price. Variation of CCF over a 25 percent range, 10-35 percent, resulted in a \$16.20/tonne CO_2 increase in the breakeven price.

Further EO examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

7.5 Coal-to-Liquids (CTL)

Economic and national security concerns related to liquid fuels have revived national interest in alternative liquid fuel sources. Coal-to-Fischer-Tropsch fuels production has emerged as a major technology option for many states and the Department of Energy.

The 2007 NETL report 'Baseline Technical and Economic Assessment of a Commercial Scale Fischer-Tropsch Liquids Facility' (36) examined the technical and economic feasibility of a commercial 50,000 barrel per day (bbl/d) CTL facility. The facility employs gasification and Fischer-Tropsch (FT) technology to produce commercial-grade diesel and naptha liquids from medium-sulfur bituminous coal.

In a recent NETL study to be released this year, updating the 2007 CTL study, the base configuration was modified to emphasize diesel fuel production over naptha. Additional updates include compression of all available CO_2 in a single train. These studies will be used as the basis for this report.

7.5.1 Size Range

The previous NETL studies focused on a 50,000 bbl/d CTL production facility, and, therefore, this will be the plant size assumed here. The previous studies also considered power production where the gas turbine and steam turbine produced power in excess of what base plant operations would require, and this excess power was exported to the grid. The 2007 study had available 124 MWe for export, with the study update projected to have significantly less available, on the order of 25 MWe.

These reported excess power quantities to be exported are net, and do include auxiliary loads for CO_2 compressors. For the purposes of this study, all power requirements will continue to be 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.

With the given size range, the CTL facility will produce 8.74 M tonnes/year (9.64 M tons/year) of CO_2 that will be available for capture.

7.5.2 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 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.

7.5.3 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:

- An average plant will be represented by a capacity of 50,000 bbl/d.
- The CO_2 amount, as stated above, is 8.74 M tonnes CO_2/year at 100 percent utilization and neglecting processing losses.
- 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 psia, 265 psia, and 300 psia.
- The CO_2 quality will be based on the EOR "pipeline" standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters. (10) For the purpose of this study, 100 volume percent purity is assumed

7.5.4 CO_2 Capture System

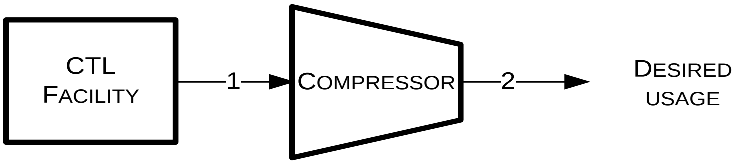
The previous NETL CTL studies (36) consider cases with CO_2 compression for EOR export, and therefore the base plant acts as the separation medium. The specific AGR units used in the previous study (36) discharge CO_2 at multiple pressures, and therefore the compression trains used are configured specifically to handle these compression requirements. Of the vendor quotes discussed in Section 6.1, 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 previous NETL CTL studies (36) will be used here. This will require that the amount of cooling water necessary for inter-stage cooling must be approximated.

It should be noted that in the previous NETL CTL studies, after the CO_2 streams are combined, a portion is removed and sent back to the gasifier, the purpose for which is not discussed. 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 NETL CTL studies.

7.5.5 Block Flow Diagram, 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. While there are three separate CO_2 streams at different pressures, and two of those streams are introduced to the compression train at different compression stages, the BFD shown represents that all three streams enter the compression train at one point for simplicity. The compression train used discharges the product CO_2 at 2,200 psig and 121°F, and therefore no after-cooling is required.

Exhibit 7-37 CO_2 capture block flow diagram



Source: NETL

Exhibit 7-38 CTL 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
COS	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 (lbmol/hr)	49,998	49,998
V-L Flowrate (lb/hr)	2,200,423	2,200,423
Solids Flowrate (lb/hr)	0	0
Temperature (°F)	N/A	121
Pressure (psia)	N/A	2,214.70
Enthalpy (Btu/lb)	N/A	-169.59
Density (lb/ft^3)	N/A	41.7

The performance results are taken from the previous NETL CTL study (36) cases that considered CO_2 capture. The performance summary is provided in Exhibit 7-39.

Exhibit 7-39 Performance summary

Performance Summary	
Item	**50,000 bbl/d (kW_e)**
CO_2 Compressor	43,480

7.5.6 Capture Integration

For the purposes of this study, it will be assumed that an additional, stand-alone cooling 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 compression. However, in order to make this case comparable to the other cases considered in this study, the cost for cooling must be included in the Greenfield breakeven price. Therefore, a stand-alone cooling system is included.

7.5.7 Power Source

The power consumption for this case is 43.48 MW. Power consumption estimates for the cooling water system were scaled as described in Section 6.4. The total power requirement was approximated to be 45.1 MWh/hr, which includes all power required by the compression train and the cooling water system. Power will be purchased at a rate of $58.59/MWh as discussed in

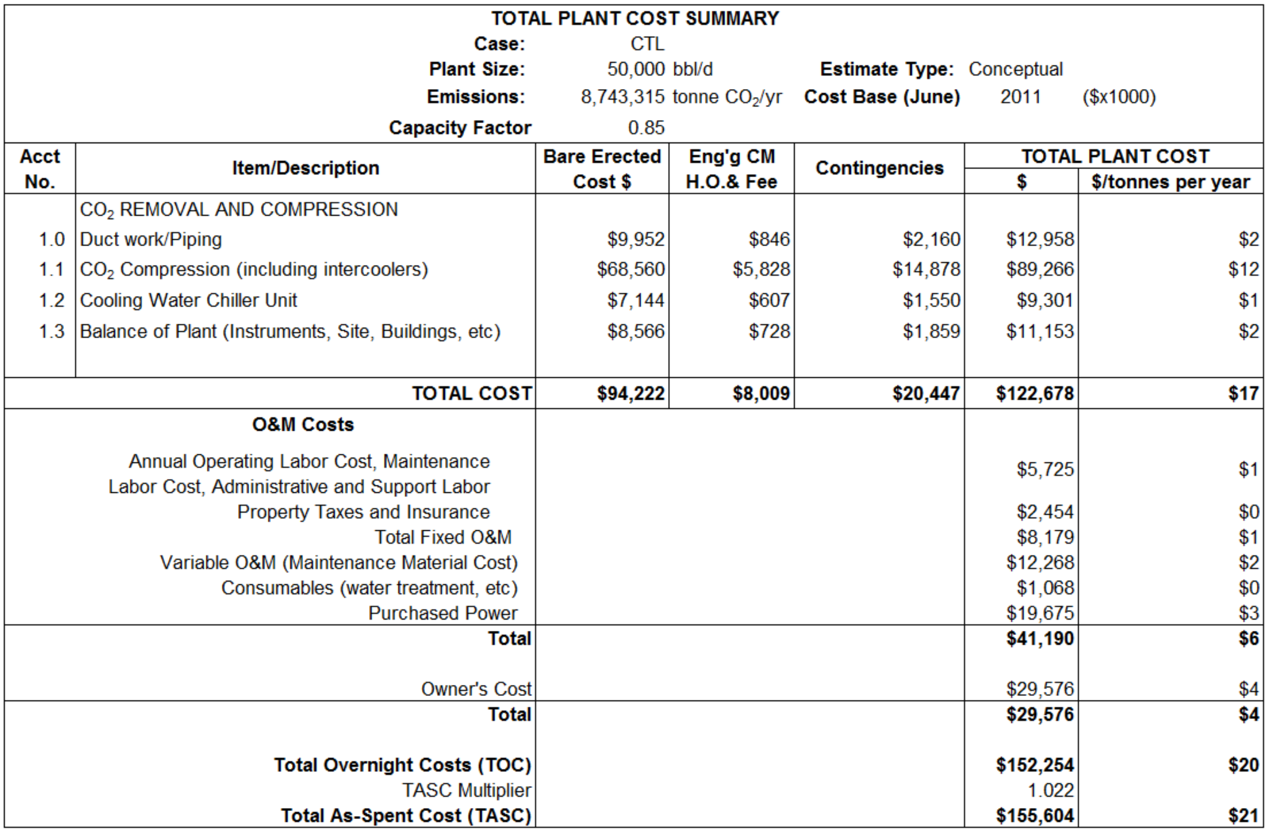
Section 6. For practical applications for this type of facility with power produced on-site, and excess power sent to the grid, the power requirements may be met with power generated on-site. The base plant power export values given in Section 7.5.1 already account for compression power requirements. Therefore, using the 2007 NETL CTL study as an example, the 124 MWe to be exported to the grid would only need to satisfy the cooling water unit power requirement.

While the specific NETL CTL case has excess power that would be able to satisfy this study's power requirements, this scenario should be looked at on a case-by-case basis. Depending on the size and location of the facility, there could be multiple reasons to produce the required power on-site.

7.5.8 Economic Analysis Results

The O&M, TOC for Greenfield, and breakeven costs are given in this section for analysis and discussion. Owner's costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner's costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

Exhibit 7-40 Capital and O&M costs for CTL Greenfield site



TOTAL PLANT COST SUMMARY

Case: CTL
Plant Size: 50,000 bbl/d **Estimate Type:** Conceptual
Emissions: 8,743,315 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST $	TOTAL PLANT COST $/tonnes per year
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$9,952	$846	$2,160	$12,958	$2
1.1	CO_2 Compression (including intercoolers)	$68,560	$5,828	$14,878	$89,266	$12
1.2	Cooling Water Chiller Unit	$7,144	$607	$1,550	$9,301	$1
1.3	Balance of Plant (Instruments, Site, Buildings, etc)	$8,566	$728	$1,859	$11,153	$2
	TOTAL COST	**$94,222**	**$8,009**	**$20,447**	**$122,678**	**$17**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$5,725	$1
	Property Taxes and Insurance				$2,454	$0
	Total Fixed O&M				$8,179	$1
	Variable O&M (Maintenance Material Cost)				$12,268	$2
	Consumables (water treatment, etc)				$1,068	$0
	Purchased Power				$19,675	$3
	Total				**$41,190**	**$6**
	Owner's Cost				$29,576	$4
	Total				**$29,576**	**$4**
	Total Overnight Costs (TOC)				**$152,254**	**$20**
	TASC Multiplier				1.022	
	Total As-Spent Cost (TASC)				**$155,604**	**$21**

The first year breakeven cost for a Greenfield site was calculated and is shown in Exhibit 7-41. The financial assumptions represent the assumptions detailed in Section 5.

Exhibit 7-41 Breakeven cost for 50,000 bbl/d CTL Greenfield

BREAKEVEN COST		
Case:	CTL	
Plant Size:	50,000	bbl/d
Emissions:	7,431,818	tonnes CO_2/yr
Capacity Factor:	0.85	
Cost Base (June):	2011	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	3.11	$\$$/tonne CO_2
Fixed O&M	1.10	$\$$/tonne CO_2
Variable O&M	1.65	$\$$/tonne CO_2
Consumables	0.14	$\$$/tonne CO_2
Purchased Power	2.65	$\$$/tonne CO_2
First year Breakeven Cost	8.66	$\$$/tonne CO_2

7.5.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

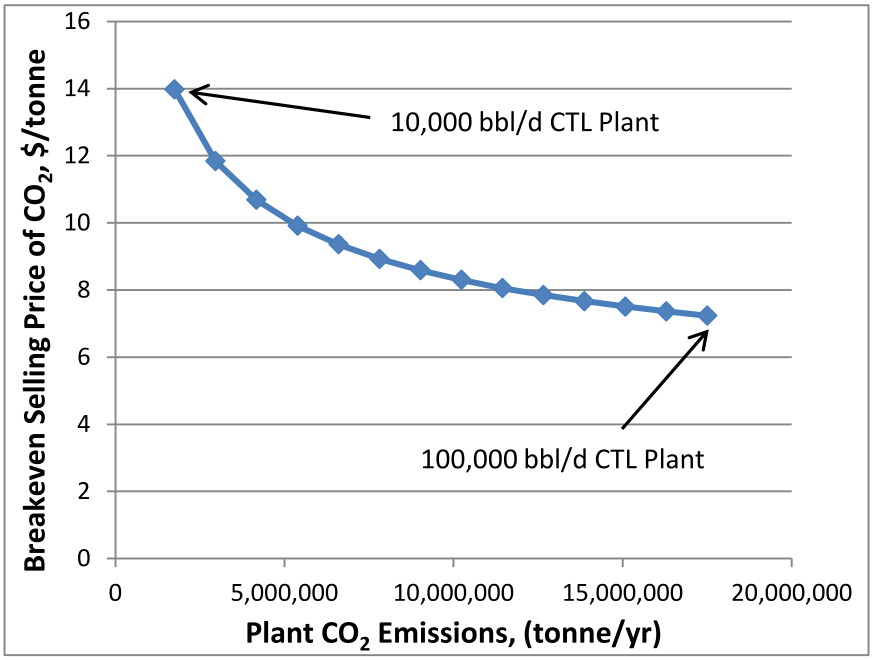
- Plant Size
- COE
- CO_2 Purity
- CCF

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore the only sensitivity discussed in this section. The other three sensitivities, COE, CO_2 purity, and CCF, are discussed and compared across the other eight cases in Section 9.2.

7.5.9.1 Plant Size

Plant size was varied over the range of 10,000 bbl/d to 100,000 bbl/d. This range is an approximation, as the plant CO_2 emissions are varied, and then the plant production in bbl/d is back-calculated using a CO_2/bbl emission factor. The emission factor is obtained from the CTL reference study base case by dividing the CO_2 stream flow by the liquid production, yielding a 0.479 tonnes CO_2/bbl factor. As shown in Exhibit 7-42, as the plant size decreases over the given range, the breakeven cost increases by $\$$6.74/tonne CO_2.

Exhibit 7-42 Plant size sensitivity



Source: NETL

7.5.10 CTL Conclusion

The high-purity CO_2 stream produced from CTL plants makes them an attractive industrial process since the plant performs the CO_2 separation as a part of normal operations.

A compression system for a 50,000 bbl/d CTL plant was modeled and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be $8.66/tonne CO_2 for a Greenfield site. A retrofit case was not considered, as it is assumed that any new CTL plant builds would include CO_2 capture and compression, thus not presenting an opportunity for retrofit. Sensitivities were performed on the following variables: COE, CCF, and plant size. The sensitivity tests for plant size showed the largest change in breakeven selling price, where a range of 10,000 – 100,000 bbl/d resulted in a rise in the breakeven selling price of $6.74/tonne CO_2.

The sensitivity tests for COE and CCF also demonstrated a change in breakeven selling price. Variation of COE over a $20-140/MWh range resulted in a $3.68/tonne CO_2 increase in the breakeven price. Variation of CCF over a 25 percent range, 10-35 percent, resulted in a $5.12/tonne CO_2 increase in the breakeven price.

Further CTL examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

7.6 Gas-to-Liquid (GTL)

Domestic, Fischer-Tropsch GTL technology provides options for using the United States' increasing supplies of domestic natural gas. As with CTL, GTL can create a significant economic value while increasing the country's energy security.

The 2013 NETL report ‘Analysis of Natural Gas-to Liquid Transportation Fuels via Fischer-Tropsch’ (37) evaluated the cost and performance of a 50,000 bbl/d 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 41 MWe to the grid. This study also considers CO_2 capture and compression with associated performance and cost. This current GTL study will be used as the basis for this report. (37)

7.6.1 Size Range

The NETL GTL study plant size is a 50,000 bbl/d GTL production facility, and therefore this will be the plant size assumed here. It 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 has a net of 41 MWe available for export. (37) Therefore, as will be addressed later in section 7.6.7, while this study assumes that all power requirements will be met with power purchased from the grid, in some cases the base plant will have excess power available to meet compression and cooling power requirements.

With the given size range, the GTL facility will produce 1.86 M tonnes/year (2.05 M tons/year) of CO_2 that will be available for capture. (37)

7.6.2 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 section AGR generates CO_2 at 265 psia and 100°F, with a purity of 100 percent CO_2. (37)

7.6.3 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:

- An average plant will be represented by a capacity of 50,000 bbl/d.
- The CO_2 amount, as stated above, is 1.86 M tonnes CO_2/year at 100 percent utilization and neglecting processing losses.
- The CO_2 stream is100 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 will be based on the EOR “pipeline” standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters. (10) For the purpose of this study, 100 volume percent purity is assumed

7.6.4 CO_2 Capture System

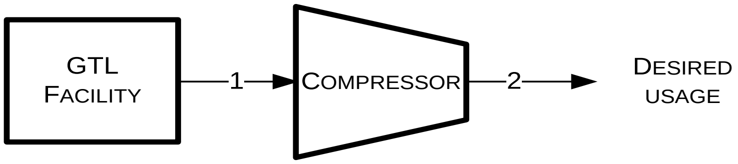
The NETL GTL study considers CO_2 removal and compression for EOR export, and therefore the base plant separates CO_2 due to the production of the liquid product. The specific AGR unit used discharges CO_2 at 265 psia, and therefore the compression train used is configured specifically to handle this higher inlet suction pressure. Of the vendors quotes discussed in

Section 6.1, 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 NETL GTL study will be replicated here. This will require that the amount of cooling water necessary for inter-stage cooling must be approximated, similar to the CTL case in this study.

7.6.5 Block Flow Diagram, 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 no after-cooling is required.

Exhibit 7-43 CO_2 capture block flow diagram



Source: NETL

Exhibit 7-44 GTL 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
COS	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 (kgmol/hr)	4,821	4,821
V-L Flowrate (kg/hr)	212,188	212,188
Solids Flowrate (kg/hr)	0	0
Temperature (°C)	38	47
Pressure (MPa, abs)	1.83	15.27
Density (kg/m^3)	34.2	686.3
V-L Molecular Weight	44.010	44.010
V-L Flowrate (lbmol/hr)	10,629	10,629
V-L Flowrate (lb/hr)	467,794	467,794

Solids Flowrate (lb/hr)	0	0
Temperature (°F)	100	117
Pressure (psia)	265	2,214.7
Density (lb/ft^3)	2.138	42.846

The performance results given are taken from the NETL GTL study case that considered CO_2 capture. (37) The performance summary is provided in Exhibit 7-45.

Exhibit 7-45 Performance summary

Performance Summary	
Item	**50,000 bbl/d (kW$_e$)**
CO_2 Compressor	6,700

7.6.6 Capture Integration

For the purposes of this study, it will be assumed that an additional, 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, in order to make this case comparable to the other cases considered in this study, the cost for cooling must be included in the Greenfield breakeven price. Therefore, a stand-alone unit is included.

7.6.7 Power Source

The power consumption for this case is 6.7 MW. Power consumption estimates for the cooling water unit were scaled as described in Section 6.4. The power consumption is approximated as 7.0 MWh/hr, which includes all power required by the compression train and the cooling water system. Power will be purchased at a rate of $58.59/MWh, as discussed in Section 6. 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. The base plant power export values given in Section 7.6.1 already account for compression power requirements. Therefore, using the current NETL GTL study as an example, the 41 MWe to be exported to the grid would only need to satisfy the cooling water system power requirement.

While the specific NETL GTL case has excess power that would be able to satisfy this study’s power requirements, this scenario should be looked at on a case-by-case basis. Depending on the size and location of the facility, there could be multiple reasons to produce the required power on-site.

7.6.8 Economic Analysis Results

The O&M, TOC for Greenfield, and breakeven costs are given in this section for analysis and discussion. Owner’s costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner’s costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

Exhibit 7-46 Capital and O&M costs for GTL Greenfield site

TOTAL PLANT COST SUMMARY

Case: GTL
Plant Size: 50,000 bbl/d **Estimate Type:** Conceptual
Emissions: 1,858,766 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

Acct No.	**Item/Description**	**Bare Erected Cost $**	**Eng'g CM H.O.& Fee**	**Contingencies**	**TOTAL PLANT COST**	
					$	**$/tonnes per year**
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$3,367	$286	$731	$4,384	$3
1.1	CO_2 Compression (including intercoolers)	$16,970	$1,442	$3,682	$22,094	$14
1.2	Cooling Water Chiller Unit	$2,070	$176	$449	$2,695	$2
1.3	Balance of Plant (Instruments, Site, Buildings, etc)	$2,241	$190	$486	$2,917	$2
	TOTAL COST	**$24,648**	**$2,094**	**$5,348**	**$32,090**	**$20**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$1,497	$1
	Property Taxes and Insurance				$642	$0
	Total Fixed O&M				$2,139	$1
	Variable O&M (Maintenance Material Cost)				$3,209	$2
	Consumables (water treatment, etc)				$227	$0
	Purchased Power				$3,054	$2
	Total				**$8,629**	**$5**
	Owner's Cost				$7,745	$5
	Total				**$7,745**	**$5**
	Total Overnight Costs (TOC)				**$39,835**	**$25**
	TASC Multiplier				1.022	
	Total As-Spent Cost (TASC)				**$40,711**	**$26**

The first year breakeven cost for a Greenfield site was calculated and is shown below in Exhibit 7-47. The financial assumptions used represent the assumptions detailed in Section 5.

Exhibit 7-47 Breakeven cost for 50,000 bbl/d GTL Greenfield

BREAKEVEN COST

Case: GTL
Plant Size: 50,000 bbl/d
Emissions: 1,579,951 tonnes CO_2/yr
Capacity Factor: 0.85
Cost Base (June): 2011

Total Plant Cost	**Cost of CO_2**	
Capital Charges	3.83	\$/tonne CO_2
Fixed O&M	1.35	\$/tonne CO_2
Variable O&M	2.03	\$/tonne CO_2
Consumables	0.14	\$/tonne CO_2
Purchased Power	1.93	\$/tonne CO_2
First year Breakeven Cost	9.29	\$/tonne CO_2

7.6.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

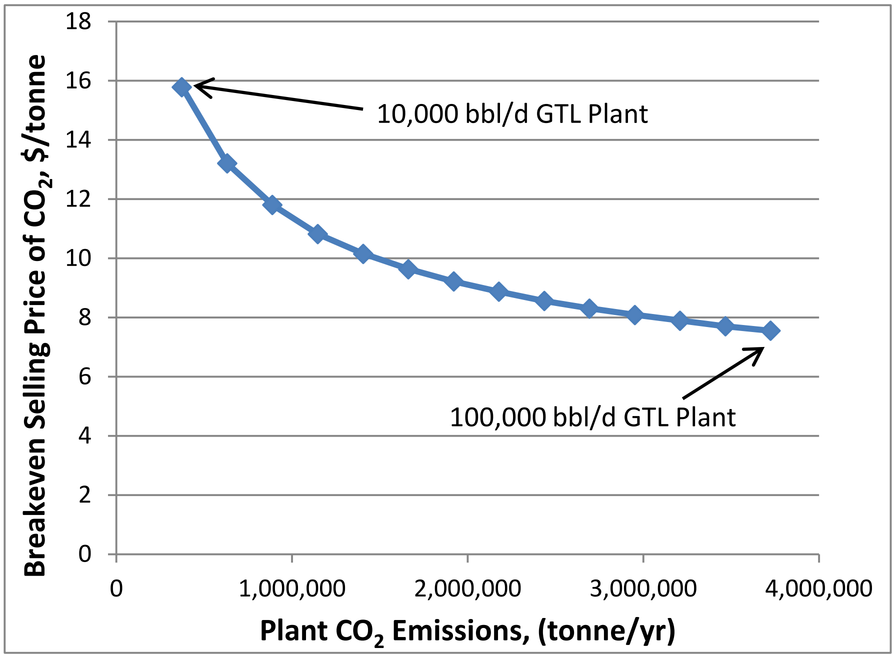
- Plant Size
- COE
- CO_2 Purity
- CCF

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore the only sensitivity discussed in this section. The other three sensitivities, COE, CO_2 purity, and CCF, are discussed and compared across the other eight cases in section 9.2.

7.6.9.1 Plant Size

Plant size was varied over the range of 10,000 bbl/d to 100,000 bbl/d. This range is an approximation, as the plant CO_2 emissions are varied, and then the plant production in bbl/d is back-calculated using a CO_2/bbl emission factor. The emission factor is obtained from the GTL reference study base case by dividing the CO_2 stream flow by the liquid production, yielding a 0.102 tonnes CO_2/bbl factor. As shown in Exhibit 7-48, as the plant size decreases over the given range, the breakeven cost increases by \$8.24/tonne CO_2.

Exhibit 7-48 Plant size sensitivity



Source: NETL

7.6.10 GTL Conclusion

The high-purity CO_2 stream produced from GTL plants makes them an attractive industrial process since the plant performs the CO_2 separation as a part of normal operations.

A compression system for a 50,000 bbl/d GTL plant was modeled and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be \$9.29/tonne CO_2 for a Greenfield site. A retrofit case was not considered, as it is assumed that any new GTL plant builds would include CO_2 capture and compression, thus not presenting an opportunity for retrofit. Sensitivities were performed on the following variables: COE, CCF, and plant size. The sensitivity tests for plant size showed the largest change in breakeven selling price, where a range of 10,000 – 100,000 bbl/d resulted in a rise in the breakeven selling price of \$8.24/tonne CO_2.

The sensitivity tests for COE and CCF also demonstrated a change in breakeven selling price. Variation of COE over a \$20-140/MWh range resulted in a \$3.96/tonne CO_2 increase in the breakeven price. Variation of CCF over a 25 percent range, 10-35 percent, resulted in a \$6.30/tonne CO_2 increase in the breakeven price.

Further GTL examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

8 Low Purity Sources

The sources discussed in this section are considered low purity sources, meaning the available CO_2 requires purification, using a MDEA AGR unit in this study, to meet EOR pipeline specifications.

8.1 Refinery Hydrogen

Refineries are one industrial source that currently deploys gas separation technology for the production of hydrogen. Air Products and Chemicals Inc. is currently capturing CO_2 from two steam methane reformer (SMR) hydrogen production plants at the Port Arthur facility in Texas for use in EOR projects in the West Hastings oil field. (38) The project concentrates CO_2 to a final purity of greater than 97 percent, and is expected to capture approximately 1 M tonnes CO_2/year, resulting in an additional 1.6 to 3.1 M barrels of domestic oil production. (38) 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 utility.

With 9 to 12 tonnes of CO_2 produced with every tonne of hydrogen (depending on feedstock) (39), at a rate of 9.1-10 million tons of hydrogen produced a year in the U.S. (40), hydrogen is a viable industrial source for CCS. NETL has also previously studied hydrogen with CO_2 capture using four cases involving steam methane reforming and coal gasification. (12)

In addition to the hydrogen plant at refineries, the fluid catalytic cracking (FCC) unit is also a potentially viable source of CO_2. This scenario has been investigated elsewhere (41), but for the purposes of this study, only the hydrogen plant at refineries is considered.

8.1.1 Size Range

Size range varies widely depending on the industry. Ninety-five percent of hydrogen produced is also consumed at the same facility. (42) Oil refinery hydrogen production is the largest consumer at 59 percent of the U.S. production. Therefore, that is the application this study will pursue. (43)

U.S. refinery hydrogen production ranged from 900 to 204,000 tonnes/year in 2003. Twenty U.S. locaitons fell within a range of 54,400 to 63,500 tonnes hydrogen/year. (42) It is reasoned that future construction capacity will be based on the largest current production to take advantage of economies of scale. Therefore, this study will assume a production of 59,000 tonnes H_2/year for the representative plant.

In addition, hydrogen produced at the oil refinery sites that are not used for processing is sold as compressed gas. This compressed, excess gas is often referred to as “merchant gas.” Merchant gas production ranged in size from 7 to 82,500 tonnes/year, with a median of 990 tonnes/year. Six locations produce between 54,400-63,500 tonnes merchant gas/year. (43)

8.1.2 CO_2 Point Sources

When producing hydrogen via steam reforming, there are three point sources in which CO_2 can be collected or vented from the plant. (39) Only one of these, the pressure swing absorption (PSA) tail gas, is a source with CO_2 concentration elevated over that of a flue gas.

In the 2010 NETL study, only a single point source, the steam reformer (SR) flue gas, was analyzed. (12)

Guido Collodi of Foster Wheeler published a study for a typical hydrogen plant, reforming natural gas, and producing 89.5 MMscfd that has the PSA tail gas characteristics shown in Exhibit 8-1. (39)

Exhibit 8-1 Hydrogen plant CO_2 characteristics

Stream	CO_2 Concentration (% mol)	CO_2 Flowrate (tonne/hr)	CO_2 Partial Pressure (psia)
PSA Tail Gas	45.10	43.5	8.70

Using the Foster Wheeler study previously mentioned with a plant capacity of 89.5 MMscfd, which equates to 78,000 tonnes H_2/year, and a CO_2 flow rate in the PSA tail gas stream of 43.5 tonne CO_2/hr, our assumed reference plant capacity of 59,000 tonnes H_2/year would have a PSA tail gas CO_2 flow rate of 32.8 tonnes CO_2/hr (36.2 tons CO_2/hr).

The Foster Wheeler study specifies the molar concentration of CO_2 in the PSA tail gas; however, the remaining composition of the stream is not given. Therefore, a study by Pierre, Dai, and Dalton of the University of Saskatchewan (44) was obtained in which a full PSA tail gas stream characterization was given. The information in this study was used to characterize the balance of the PSA tail gas stream. The final PSA tail gas stream assumptions are shown in Exhibit 8-2.

Given in the 2010 NETL hydrogen plant study (12) are different types of hydrogen plant configurations, all of which utilize PSA technology to obtain a H_2 product. The PSA tail gas temperature and pressure were specified as 100°F and 20 psia.

Combining these parameters into one table, the specification of the assumed hydrogen plant is represented in Exhibit 8-2.

Exhibit 8-2 Hydrogen plant study parameters

Hydrogen Plant Characteristics			
Capacity	59,000 tonnes H_2/year		
PSA Tail Gas Stream Characteristics			
Temperature	100 °F		
Pressure	20 psia		
PSA Tail Gas Component	Molar Flow (kgmol/hr)	Mole Fraction	Mass Flow (ton/hr)
CH_4	295.85	0.1761	5.23
C_2H_6	0.00	0.0000	0.00
C_3H_8	0.00	0.0000	0.00
CO	112.20	0.0668	3.46
CO_2	747.90	0.4452	36.28
H_2	503.85	0.2999	1.11
H_2O	0.00	0.0000	0.00
N_2	19.48	0.0116	0.60
O_2	0.00	0.0000	0.00
He	0.72	0.0004	0.00
H_2S	0.00	0.0000	0.00
SO_2	0.00	0.0000	0.00
Total	1,680	1.0000	46.69

Given the assumptions set forth in Exhibit 8-2, the amount of CO_2 that will be available for capture in the PSA tail gas stream is 273,860 tonnes CO_2/year after allowances for losses during purification.

8.1.3 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:

- An average plant will be represented by a production rate of 59,000 tonne H_2/year (42)
- Only hydrogen produced for refinery usage will be pursued
- The gas from the PSA tail gas is assumed 44.52 mol percent CO_2 and will be the one high-purity point source (44) (39)
- The CO_2 amount to be captured, as calculated above, is 273,860 tons CO_2/year (44)
- The CO_2 stream temperature is 100 °F (12)
- The CO_2 stream pressure is 20 psia (12)
- The CO_2 quality will be based on the EOR "pipeline" standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters (10)

8.1.4 CO_2 Capture System

With an assumed concentration of only 44.52 mole percent in the PSA tail gas, separation will be required to obtain QGESS EOR pipeline specifications. In some plant configurations (45) (12), the PSA tail gas is recycled to the SMR so as to recover the fuel value still present in this stream.

The stream may be sent directly to be burned in the SMR or supplemented with an air/fuel mixture, and then burned in the SMR. This requirement will be assumed here.

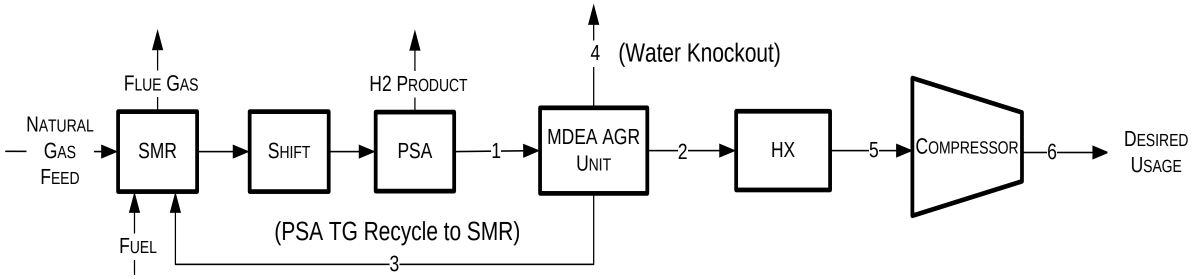
The MDEA unit discussed in Section 6.2 is employed here for the PSA tail gas purification. Auxiliary requirements are scaled based on product CO_2 flow.

The MDEA AGR unit requires low pressure steam at 71 psia to regenerate the MDEA solvent. These steam needs were met with the package boiler discussed in Section 6.3.The compression system used will be the reciprocating compression train discussed in Section 6.1.

8.1.5 Block Flow Diagram, Stream Table, and Performance Summary

Since the PSA tail gas stream CO_2 concentration is 44.52 mole percent, further purification is required. This is achieved using an MDEA process. The PSA tail gas (stream 1) is fed to the MDEA separation unit, resulting in three streams. Any inlet water is removed in a water knockout stream, H_2 and CH_4 are recycled in a PSA tail gas recycle stream that would be sent to the SMR to recover any remaining fuel value, and the third stream is the purified CO_2 stream. The CO_2 stream is then cooled to the specified CO_2 compression train inlet temperature and enters the compression train where it is compressed, resulting in a product stream of 2,215 psia and 120°F. Exhibit 8-3 shows the BFD for this process, and Exhibit 8-4 shows the stream table for this process. The SMR, Shift, and PSA blocks are shown only to represent the base plant, and to demonstrate where the PSA tail gas stream will be recycled. No stream data is given for the Natural Gas Feed, Fuel, Flue Gas, or H_2 Product streams. In a plant configuration without capture, stream 1, the PSA tail gas, would be recycled in the same way as stream 3 is currently represented.

Exhibit 8-3 CO_2 capture block flow diagram



Source: NETL

Exhibit 8-4 Refinery hydrogen stream table

	1	2	3	4	5	6
V-L Mole Fraction						
CH_4	0.1761	0.0000	0.3052	0.0000	0.0000	0.0000
C_2H_6	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
C_3H_8	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0668	0.0000	0.1158	0.0000	0.0000	0.0000
CO_2	0.4452	0.9999	0.0386	0.0000	0.9999	0.9999
H_2	0.2999	0.0000	0.5197	0.0000	0.0000	0.0000
H_2O	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.0116	0.0000	0.0201	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
He	0.0004	0.0000	0.0007	0.0000	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
AR	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	0.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	1,680	710	969	0	710	710
V-L Flowrate (kg/hr)	42,359	31,263	11,096	0	31,263	31,263
Solids Flowrate (kg/hr)	0	0	0	0	0	0
Temperature (°C)	38	49	38	---	27	49
Pressure (MPa, abs)	0.14	0.14	0.03	---	0.1	15.3
Enthalpy (kJ/kg)^A	49.49	40.19	102.62	---	21.4	-171.6
Density (kg/m^3)	1.3	2.2	0.2	---	2.1	673.4
V-L Molecular Weight	25.221	44.008	11.449	---	44.008	44.008
V-L Flowrate (lb_{mol}/hr)	3,703	1,566	2,137	0	1,566	1,566
V-L Flowrate (lb/hr)	93,386	68,923	24,463	0	68,923	68,923
Solids Flowrate (lb/hr)	0	0	0	0	0	0
Temperature (°F)	100	120	100	---	81	120
Pressure (psia)	20	19.7	5.0	---	17.4	2,215.6
Enthalpy (Btu/lb)^A	21.3	17.3	44.1	---	9.2	-73.8
Density (lb/ft^3)	0.084	0.140	0.010	---	0.133	42.037

The performance results are based on compressor quote discussed in Section 6.1. The performance summary is provided in Exhibit 8-5.

Exhibit 8-5 Performance summary

Performance Summary	
Item	**59,000 tons H_2/year (kW_e)**
CO_2 Compressor	3,600
AGR MDEA	1,200
Total	4,800

8.1.6 Capture Integration

With steam and air requirements for the PSA method of hydrogen production, integration into a facility is possible given any amount of over sizing in the air system or additional waste heat. If steam requirements for MDEA purification of the PSA tail gas can be met with waste heat, inclusion of an additional boiler to satisfy the MDEA steam requirements may be excluded. The cooling water system is stand-alone however there is a potential to integrate make-up water to feed or partially feed the cooling system thereby reducing the unit size or replacing it completely with a simple heat exchanger depending on the size of the plant. Additionally, recycling the waste stream that still contains fuel value to the SMR to recover that fuel value is considered integration since this sends a product of the capture system for use in the base plant, even though this stream exists in the base plant without capture.

8.1.7 Power Source

The power consumption for this case is 4.8 MW. Power consumption estimates for the stand-alone cooling water system were scaled as described in Section 6.4. The total power requirement was approximated to be 6.0 MWh/hr, which includes all power required by the compression train, package boiler, and the cooling water system. Power will be purchased at a rate of $58.59/MWh as discussed in Section 6. To satisfy the steam requirements for the MDEA separation process, a package boiler was included, and natural gas will be purchased at a rate of $6.13/MMBtu to fuel the boiler as discussed in Section 6.

8.1.8 Economic Analysis Results

The O&M, TOC for Greenfield and retrofit, and breakeven costs are given in this section for analysis and discussion. Owner's costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner's costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

Exhibit 8-6 Capital and O&M costs for refinery hydrogen Greenfield site

TOTAL PLANT COST SUMMARY

Case: Refinery H_2

Plant Size: 59,000 tonne H_2/yr **Estimate Type:** Conceptual

Emissions: 273,860 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)

Capacity Factor 0.85

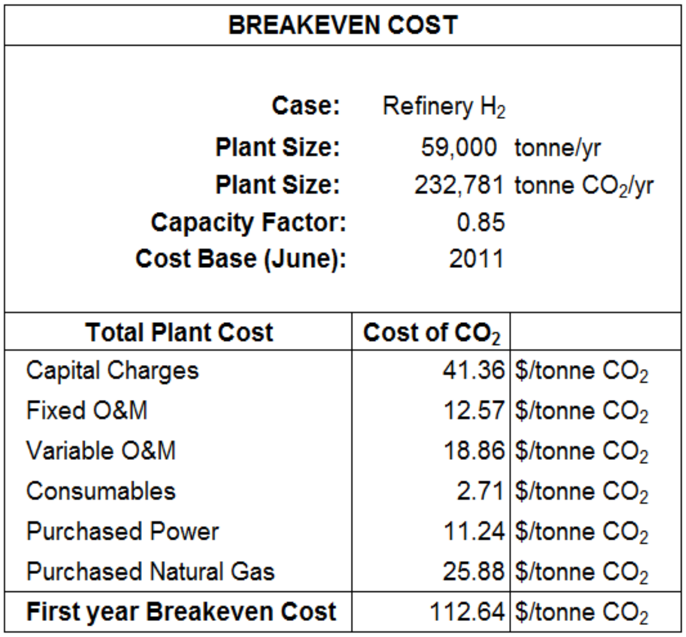
Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST	
					$	**$/tonne per year**
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$881	$75	$191	$1,147	$5
1.1	MDEA Purification Unit	$16,423	$1,396	$3,564	$21,383	$92
1.2	CO_2 Pre-cooler	$131	$11	$28	$171	$1
1.3	CO_2 Compression (including intercoolers)	$7,631	$649	$1,656	$9,936	$43
1.4	Cooling Water Unit	$3,988	$339	$865	$5,192	$22
1.5	LP Steam Boiler	$1,606	$137	$349	$2,091	$9
1.6	Balance of Plant (Instruments, Site, Buildings, etc)	$3,066	$261	$665	$3,992	$17
	TOTAL COST	**$33,726**	**$2,867**	**$7,319**	**$43,911**	**$189**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$2,049	$9
	Property Taxes and Insurance				$878	$4
	Total Fixed O&M				$2,927	$13
	Variable O&M (Maintenance Material Cost)				$4,391	$19
	Consumables (water treatment, corrosion inhibitor, MDEA solvent, Activated Carbon, Boiler Water, etc)				$630	$3
	Purchased Natural Gas				$6,025	$26
	Purchased Power				$2,618	$11
	Total				**$16,591**	**$71**
	Initial Fill MDEA Solvent				$123	$1
	Initial Fill Corrosion Inhibitor				$8	$0
	Owner's Cost				$10,667	$46
	Total				**$10,798**	**$46**
	Total Overnight Costs (TOC)				**$54,709**	**$235**
	TASC Multiplier				1.099	
	Total As-Spent Cost (TASC)				**$60,125**	**$258**

Exhibit 8-7 Total overnight costs for retrofit

		TOTAL PLANT COST		Retrofit	
Retrofit Factor	1.05	**$**	**$/tonne per year**	**$**	**$/tonne per year**
Total Overnight Costs (TOC)		**$54,709**	**$235**	**$57,445**	**$247**
TASC Multiplier		1.099			
Total As-Spent Cost (TASC)		**$60,125**	**$258**	**$63,132**	**$271**

The first year breakeven costs for a Greenfield site and a retrofit site were calculated and are shown in Exhibit 8-8 and Exhibit 8-9. The financial assumptions and retrofit factor used represent the assumptions detailed in Section 5.

Exhibit 8-8 Breakeven cost for 59,000 tonnes/yr Greenfield

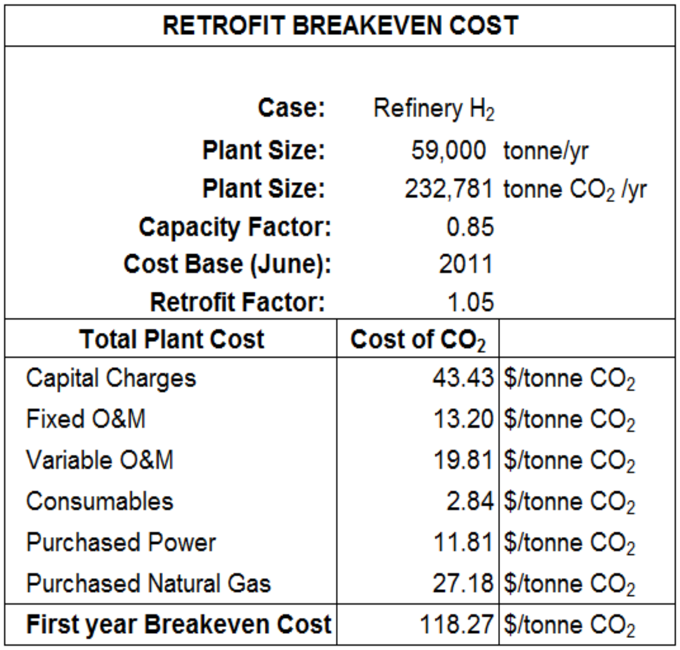


BREAKEVEN COST

Case:	Refinery H_2	
Plant Size:	59,000	tonne/yr
Plant Size:	232,781	tonne CO_2/yr
Capacity Factor:	0.85	
Cost Base (June):	2011	

Total Plant Cost	**Cost of CO_2**	
Capital Charges	41.36	\$/tonne CO_2
Fixed O&M	12.57	\$/tonne CO_2
Variable O&M	18.86	\$/tonne CO_2
Consumables	2.71	\$/tonne CO_2
Purchased Power	11.24	\$/tonne CO_2
Purchased Natural Gas	25.88	\$/tonne CO_2
First year Breakeven Cost	112.64	\$/tonne CO_2

Exhibit 8-9 Breakeven cost for 59,000 tonnes/yr retrofit



RETROFIT BREAKEVEN COST

Case:	Refinery H_2	
Plant Size:	59,000	tonne/yr
Plant Size:	232,781	tonne CO_2 /yr
Capacity Factor:	0.85	
Cost Base (June):	2011	
Retrofit Factor:	1.05	

Total Plant Cost	**Cost of CO_2**	
Capital Charges	43.43	\$/tonne CO_2
Fixed O&M	13.20	\$/tonne CO_2
Variable O&M	19.81	\$/tonne CO_2
Consumables	2.84	\$/tonne CO_2
Purchased Power	11.81	\$/tonne CO_2
Purchased Natural Gas	27.18	\$/tonne CO_2
First year Breakeven Cost	118.27	\$/tonne CO_2

As compared to the previous high purity cases, the breakeven cost for refinery hydrogen is significantly higher. This is the result of the lower CO_2 stream concentration, and the necessity for the stream to be purified before compression. The capital cost of the purification system adds a large cost increase to the breakeven cost, but also the cost of the associated purification utilities such as natural gas and power are significant contributors to the breakeven cost. The most costly utility associated with the purification process is the steam requirement. The capital and operating costs for the package boiler used to provide steam for the purification process may be excluded if the base plant has the capacity to provide the necessary steam. This would reduce all of the line items in Exhibit 8-8. It would also eliminate the Purchased Natural Gas portion of the breakeven cost, which is currently \$26/tonne CO_2.

8.1.9 Sensitivities Analysis

Sensitivity analysis was performed on the following variables:

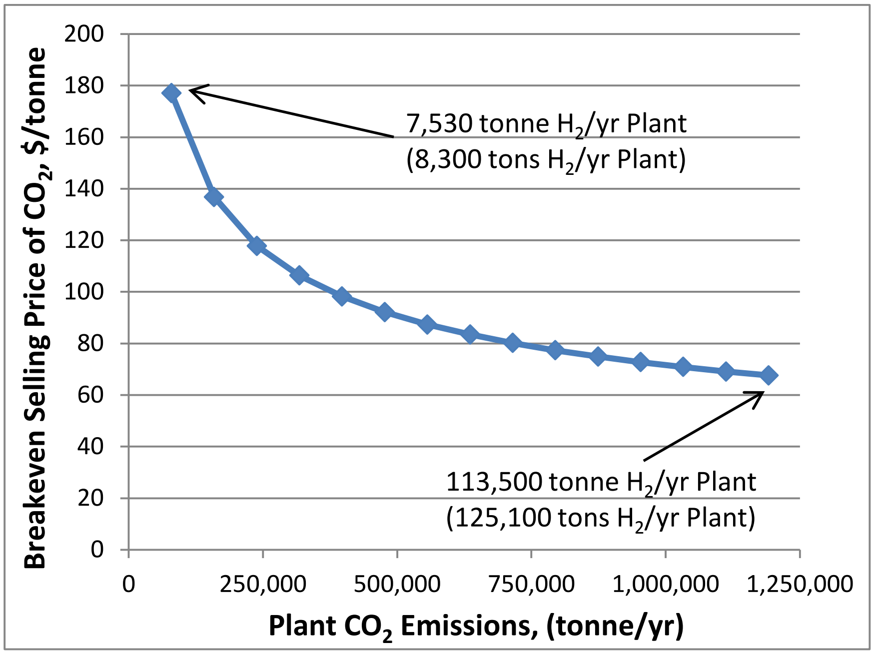
- Plant Size
- COE
- CO_2 Purity
- CCF
- Natural Gas Price

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore the only sensitivity discussed in this section. The other four sensitivities, COE, CO_2 purity, CCF, and Natural Gas price, are discussed and compared across the other cases in Section 9.2.

8.1.9.1 Plant Size

Plant size was varied over the range of 7,530 tonnes H_2/yr to 113,500 tonnes H_2/yr. These values use the same set of Hydrogen plant assumptions set forth in Section 8.1.2. As shown in Exhibit 8-10, as the plant size decreases over the given range, the breakeven cost increases by \$109.50/tonne CO_2. This variation over the plant size range is much more significant than in the high purity cases.

Exhibit 8-10 Plant size sensitivity



Source: NETL

8.1.10 Refinery Hydrogen Conclusion

As compared to the previous high purity cases, the lower purity CO_2 stream produced from the PSA tail gas found in refinery hydrogen results in a high breakeven selling price.

A purification and compression system for a 59,000 tonnes of H_2/yr capacity refinery hydrogen plant was modeled and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be $112.64/tonne CO_2 for a Greenfield site, and $118.27/ tonne CO_2 for a retrofit site. The disparity is the result of differences in equipment required for a Greenfield versus a retrofit site, assuming adequate plot plan space for the retrofit case. Sensitivities were performed on the following variables: COE, CCF, CO_2 purity, and plant size. The sensitivity tests for plant size showed the largest change in breakeven selling price; when varying over the range, a change in price of $109.50/tonne CO_2 was observed. With higher base breakeven costs, the losses in economies of scale when moving to smaller-sized plants are much more significant than in the previous cases, and, therefore, result in a large change in price for the plant size sensitivity.

The sensitivity tests for COE and CCF also demonstrated a change in breakeven selling price. Variation of COE over a $120/MWh range resulted in a $23.03/tonne CO_2 increase in the breakeven price. Variation of the CCF over a 25 percent range, 10-35 percent, resulted in a $58.76/tonne CO_2 increase in the breakeven price.

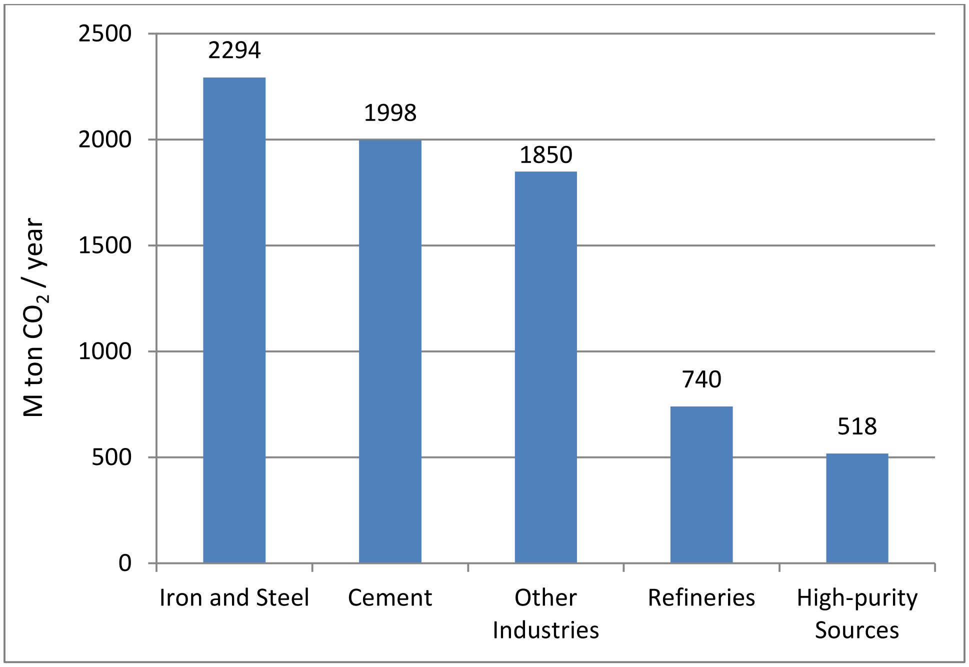
The expected trend for the CO_2 purity sensitivity was also demonstrated and is discussed in more detail in Section 9.2. As with all cases, as the CO_2 purity approaches lower concentrations, the breakeven selling price will increase dramatically.

Further refinery hydrogen examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

8.2 Steel/Iron

According to a presentation given by the International Energy Agency (IEA), (46) in 2008, industry accounted for 6.7 G tonnes (7.4 G tons) of CO_2 emissions. Exhibit 8-11 shows the distribution of these emissions by industrial processes. (46) The Iron and Steel industry accounted for 31 percent or about 2,100 M tonnes (2,300 M tons) of CO_2 emissions in 2008. Due to the large amount of emissions available for capture, the iron and steel industry is particularly attractive as a process to be included in this study.

Exhibit 8-11 Industrial CO_2 emissions



Note: Units are in short tons.

Source: NETL

Recent developments in the steel/iron sector include the agreement between the Abu Dhabi National Oil Company (ADNOC) and Masdar Carbon to move forward with a project to capture CO_2 from the Emirates Steel complex at Mussafah. The project is reported to capture 0.725 M tonnes/year (0.8 M ton/year) of CO_2 for use in EOR. (47)

8.2.1 Size Range

According to a study published by the American Iron and Steel Institute, there are 116 steel plants in the U.S., accounting for approximately 80 M tonnes of steel production in 2010. (48) Of these 116 steel plants, 99 plants produce steel using an Electric Arc Furnace (EAF) and 17 plants produce steel using the more traditional BOF. The main difference between the EAF and BOF processes has to do with the raw materials used as inputs as well as the furnace design. The resulting steel product from an EAF process contains 90 percent recycled steel on average, whereas the BOF product contains 25 percent recycled steel on average. (49) 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) (50). The combination of generally smaller EAF plants and lower concentration of EAF plant CO_2 emissions results in a high cost of capture 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 American Iron and Steel Institute for BOF plants in the U.S. in 2010, was 41.2 M tonnes. For the purposes of this study, it will be assumed that the BOF plant, located in Braddock, PA, will represent our average plant, with its 2010 capacity of 2.54 M tonnes/year (2.8 M tons/year). Applying the 2010 World Steel Association utilization factor of 73.8 percent, this study's Braddock, PA plant production comes to 1.87 M tonnes of steel/year.

8.2.2 CO_2 Point Sources

A study by Wiley et al., (50) published in 2010, assessed the opportunities for CO_2 capture in Australian iron and steel mills. They utilize stream data from an Australian BOF steel mill, with a specific configuration. For their 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 in their configuration. The CO_2 is produced in the blast furnace when iron ore is reduced to molten ore. Since the BOF process utilizes a larger amount of iron ore than the EAF process, the BOF process will produce more blast furnace CO_2. The Blast furnace gas (BFG), instead of being vented, is cleaned and used in the plant as low-grade fuel. The BFG and the coke oven gas (COG) streams are used in the plant to produce electricity and allow the plant to limit or eliminate purchasing energy from the grid. (50) The relevance to this study is that instead of having a high content CO_2 point source from the BFG, the CO_2 is distributed throughout the plant as smaller CO_2 point sources. This will increase the cost of CO_2 capture in the steel plant. The smaller CO_2 point sources available to be captured include the power plant stack (PPS), COG, blast furnace stove (BFS), sinter stack, blown oxygen steelmaking (BOS) stack, hot strip mill stack, plate mill stack, and lime kiln for the configuration given by Wiley, et al. (50) 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.

The Wiley study (50) assumed a production capacity for an integrated steel mill BOF plant of 5.5 M tons/year, which is higher than our reference plant capacity of 2.8 M tons/year. For the purposes of the present study, the CO_2 emitted and volumetric flow rate are scaled down to the assumed production capacity of 2.8 M tons/year (2.54 M tonnes/year) as given by the Braddock plant. The scaling results for the three point sources considered are shown in Exhibit 8-12.

Exhibit 8-12 BOF iron and steel plant characteristics (50)

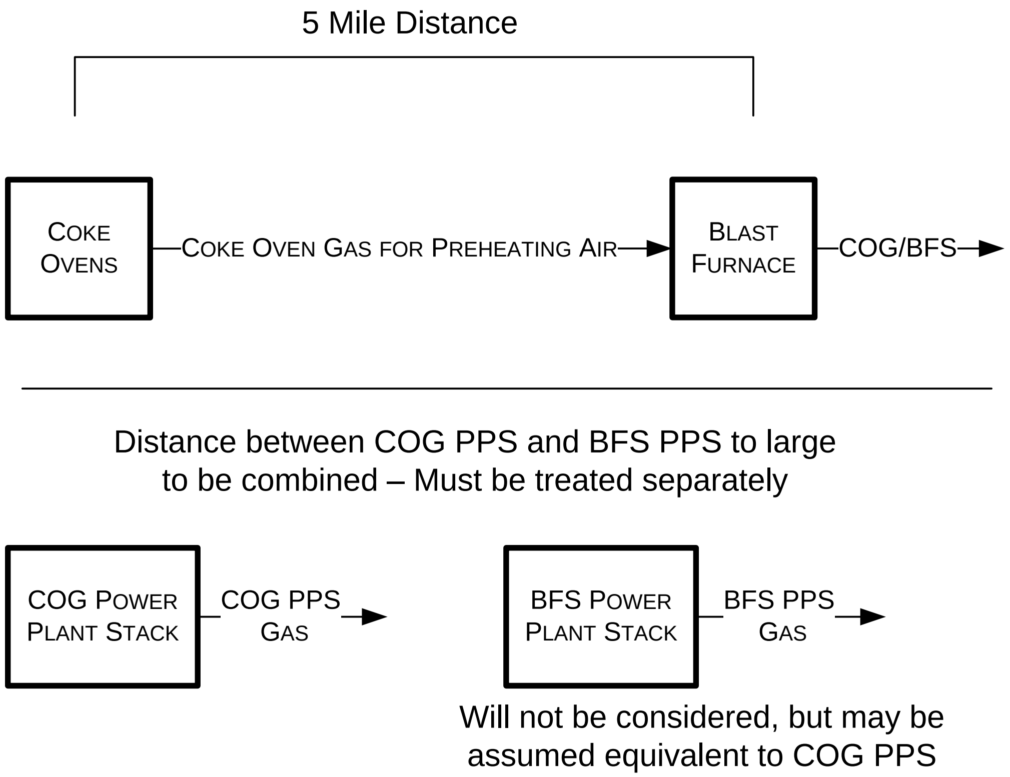
Description	Power Plant Stack (PPS)	Coke Oven Gas (COG)	Blast Furnace Stoves (BFS)
CO_2 Emitted (Mton/yr) 2.8 Mt/yr Basis	2.07	0.97	1.09
Flow Rate (Nm3/s) 2.8 Mt/yr Basis	203.64	67.20	7.13
Pressure (kPa)	101.30	101.30	101.30
Temperature (°C)	300.00	100.00	300.00
Composition (vol %)			
N_2	67.00	67.00	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
CO	-	-	-
H_2	-	-	-

The Braddock, PA, steel mill facility will be the representative base plant for this study; however, due to the lack of stream data available from the Braddock plant, the stream data given by Wiley, et al. will be paired with the Braddock plant capacity and plot plan to give a representative plant for this study. Personal communication with a former U.S. Steel employee yielded 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. It was also mentioned that a pressure difference exists between these two streams; however, this will not be considered as the existing stream data shows equal pressure between the two streams. A final point of emphasis is that there is a PPS for both the COG and BFS at the Braddock plant, but at too great a distance to be combined. Since the Wiley reference stream data considers only one PPS, it is assumed that the Wiley PPS data will represent the Braddock plant COG PPS, and the Braddock plant BFS PPS will not be considered. Section 8.2.8 will discuss the calculated breakeven costs, and will show that each of these point sources result in relatively equal breakeven costs. Therefore, while we are excluding the Braddock plant BFS PPS, it is assumed that the breakeven cost for this point source would be similar to the breakeven cost calculated for the COG PPS.

If the excluded BFS PPS were to be included, there would not be gains in economies of scale because, due to the distance between point sources, this BFS PPS could not be combined with another point source. The inclusion of this point source would require a third, separate, purification and compression scheme, eliminating any gains in economies of scale from combining point sources.

Shown below, in Exhibit 8-13, is a simplified BFD of the plot plan description to be implemented here.

Exhibit 8-13 Braddock steel mill plot plan



Source: NETL

Of the eight point sources listed by Wiley, five have CO_2 concentrations that are equal to or less than a typical CFPP flue gas stream and, therefore, will not be included. Only the three higher CO_2 concentration streams, the PPS, COG, and BFS, as given by Wiley, will be used. These stream's data are applied to the reference Braddock plant. Given the configuration of the reference plant, the COG and BFS point sources are combined to form a single stream, labeled COG/BFS.

8.2.3 Design Input and Assumptions

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

- BOF integrated steel mill is the base plant
- The base plant will be represented by the Braddock plant steel production rate of 2.54 M tonnes/year (2.8 M ton/year)
- 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; COG PPS will utilize its own separation and compression facility
- Stream data from the Wiley, et al. study will be implemented for the COG and BFS, and the PPS data as given by Wiley will represent the COG PPS, all on a 2.54 M tonnes/year (2.8 M ton/yr) production basis
- The CO_2 quality will be based on the EOR "pipeline" standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters. (10)

8.2.4 CO_2 Capture System

The stream concentrations for the three point sources are relatively low; however, they are above typical CFPP flue gas concentrations. They will require purification before compression to attain EOR pipeline standards.

The MDEA purification system discussed in Section 6.2 will be used here. Two systems will be used: one for the COG/BFS stream and one for the COG PPS stream.

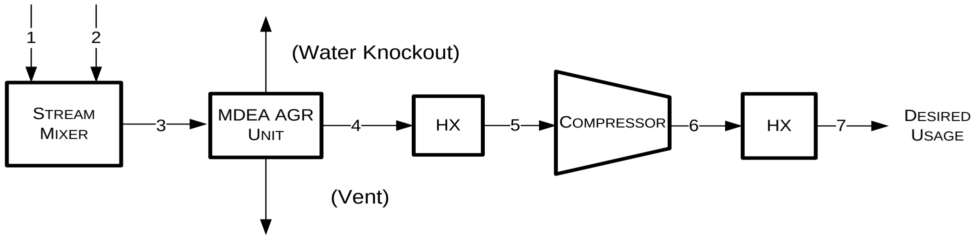
Steam for solvent regeneration will be provided by the package boiler discussed in Section 6.3.

Two compression systems, both integrally geared centrifugal compression trains, will be used for the COG/BFS stream and COG PPS stream, and will be scaled from the quote discussed in Section 6.2.

8.2.5 Block Flow Diagrams, Stream Tables, and Performance Summary

For the COG/BFS case, the COG stream and BFS stream are mixed and sent to the MDEA separation unit. Three streams result from the MDEA separation unit; a water knockout stream, a vent stream that contains mostly nitrogen, oxygen, and CO_2, and the purified CO_2 stream. The CO_2 stream is then cooled to the specified CO_2 compression train inlet temperature and enters the compression train where it is compressed, resulting in a product stream of 2,215 psia. The compression train discharge stream is then cooled to EOR pipeline specifications. Exhibit 8-14 shows the BFD for this process, and Exhibit 8-15 shows the stream table for this process.

Exhibit 8-14 CO_2 capture block flow diagram for COG/BFS



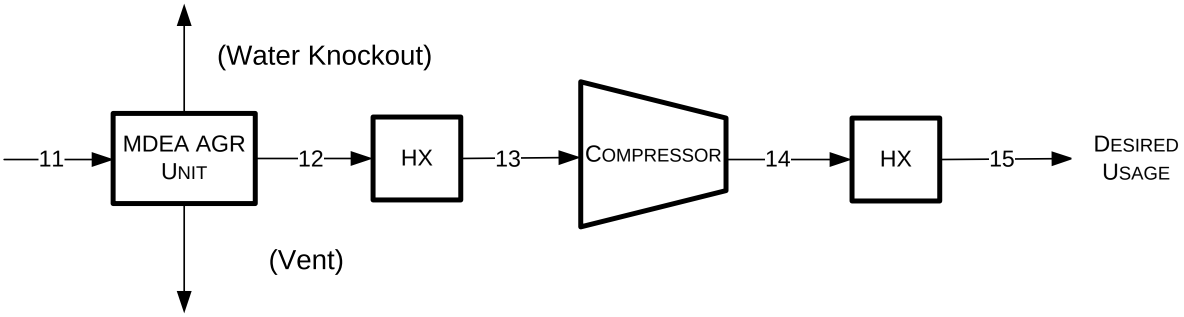
Source: NETL

Exhibit 8-15 COG/BFS stream table

	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	0.2642	0.9956	0.9956	1.0000	1.0000
COS	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	0.0548	0.0043	0.0043	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6700	0.6800	0.6710	0.0000	0.0000	0.0000	0.0000
O_2	0.0100	0.0100	0.0100	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)	10,793	1,145	11,938	3,010	3,010	2,997	2,997
V-L Flowrate (kg/hr)	344,007	34,828	378,835	132,132	132,132	131,897	131,897
Solids Flowrate (kg/hr)	0	0	0	0	0	0	0
Temperature (°C)	100	300	119	49	27	134	49
Pressure (MPa, abs)	0.10	0.1	0.1	0.1	0.1	15.3	15.3
Enthalpy (kJ/kg)^A	170.15	469.1	197.6	44.7	25.5	30.6	-171.6
Density (kg/m^3)	1.0	0.6	1.0	2.2	2.3	259.4	673.3
V-L Molecular Weight	31.872	30.413	31.732	43.896	43.896	44.009	44.009
V-L Flowrate (lb_{mol}/hr)	23,795	2,525	26,320	6,636	6,636	6,607	6,607
V-L Flowrate (lb/hr)	758,406	76,783	835,189	291,302	291,302	290,782	290,782
Solids Flowrate (lb/hr)	0	0	0	0	0	0	0
Temperature (°F)	212	572	247	120	80	274	120
Pressure (psia)	14.7	14.7	14.7	19.7	19.0	2,215.8	2,215.1
Enthalpy (Btu/lb)^A	73.1	201.7	85.0	19.2	11.0	13.1	-73.8
Density (lb/ft^3)	0.065	0.040	0.062	0.140	0.145	16.192	42.036

For the COG PPS case, the COG PPS stream is sent to an MDEA separation unit. Three streams result from the MDEA separation unit; a water knockout stream, a vent stream that contains mostly nitrogen, oxygen, and CO_2, and the purified CO_2 stream. The CO_2 stream is then cooled to the specified CO_2 compression train inlet temperature and enters the compression train where it is compressed, resulting in a product stream of 2,215 psia. The compression train discharge stream is then cooled to EOR pipeline specifications. Exhibit 8-16 shows the BFD for this process, and Exhibit 8-17 shows the stream table for this process.

Exhibit 8-16 CO_2 capture block flow diagram for COG PPS



Source: NETL

Exhibit 8-17 COG PPS stream table

	11	12	13	14	15
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	0.9927	0.9927	1.0000	1.0000
COS	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	0.0073	0.0073	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6768	0.0000	0.0000	0.0000	0.0000
O_2	0.0101	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	32,380	7,198	7,198	7,146	7,146
V-L Flowrate (kg/hr)	1,002,554	315,436	315,436	314,494	314,494
Solids Flowrate (kg/hr)	0	0	0	0	0
Temperature (°C)	300	49	27	134	49
Pressure (MPa, abs)	0.1	0.1	0.1	15.3	15.3
Enthalpy (kJ/kg)^A	434.6	47.8	28.5	30.7	-171.6
Density (kg/m^3)	0.7	2.2	2.3	259.3	673.3
V-L Molecular Weight	30.962	43.820	43.820	44.009	44.009
V-L Flowrate (lb_{mol}/hr)	71,386	15,870	15,870	15,754	15,754
V-L Flowrate (lb/hr)	2,210,254	695,418	695,418	693,340	693,340
Solids Flowrate (lb/hr)	0	0	0	0	0
Temperature (°F)	572	120	80	274	120
Pressure (psia)	14.7	19.7	19.0	2,215.8	2,215.1
Enthalpy (Btu/lb)^A	186.8	20.6	12.3	13.2	-73.8
Density (lb/ft^3)	0.041	0.140	0.145	16.186	42.035

The performance results are based on compressor quotes discussed in Section 6.1. The performance summary is provided in Exhibit 8-18.

Exhibit 8-18 Performance summary

Performance Summary	
Item	**2.54 M tonnes steel/year (kW_e)**
COG/BFS CO_2 Compressor	12,930
COG PPS CO_2 Compressor	30,930
COG/BFS AGR MDEA	5,200
COG PPS AGR MDEA	12,400
Steam Boiler Total	1,950
Total	63,410

8.2.6 Capture Integration

The BOF process integrated steel mill, as stated in the previous section, is largely self-sufficient due to the use of the BFS and COG as low grade fuel for electricity generation. Due to this set-up, integration in terms of available excess power, steam, or cooling loads between the base plant and capture system may be limited.

The cooling water system 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 size or replacing it completely with a simple heat exchanger. This would have to 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.

8.2.7 Power Source

The power consumption for compression and separation for this case is 63.41 MW. Power consumption estimates for the stand-alone cooling water unit were scaled as discussed in Section 6.4. The total power requirement was approximated to be 74.0 MWh/hr, which includes all power required by the compression trains, MDEA units, package boiler, and the cooling water system. Power will be purchased at a rate of $58.59/MWh as discussed in Section 6. For the current example, integrated BOF plants remain energy self-sufficient by recycling streams as low grade fuel. Therefore, energy generation on site to support the addition of capture and compression equipment may be justified, but it is not known if excess power is available.

8.2.8 Economic Analysis Results

The O&M, TOC for retrofit, and breakeven costs for retrofit are given in this section for analysis and discussion. Owner's costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner's costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs. Greenfield costs are not given for this case, as BOF steel plants are no longer being constructed.

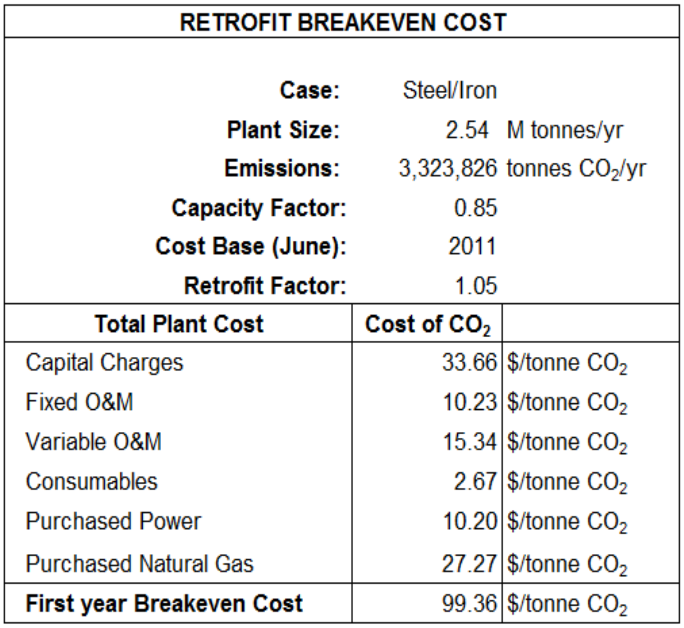
Exhibit 8-19 Capital and O&M costs for steel/iron retrofit site

TOTAL PLANT COST SUMMARY
Case: Steel/Iron Retrofit
Steel Plant Size: 2.54 M tonnes/yr **Estimate Type:** Conceptual
Emissions: 3,910,384 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

Acct No.	**Item/Description**	**Bare Erected Cost $**	**Eng'g CM H.O.& Fee**	**Contingencies**	**TOTAL PLANT COST $**	**TOTAL PLANT COST $/tonne per year**
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$5,950	$506	$1,291	$7,746	$2
1.1	MDEA Purification Unit COG + BFS	$72,751	$6,184	$15,787.04	$94,722	$28
1.2	MDEA Purification Unit COG PPS	$173,468	$14,745	$37,642.64	$225,856	$68
1.3	Pre-Cooler COG + BFS	$442	$38	$95.92	$576	$0
1.4	Pre-Cooler COG PPS	$887	$75	$192.53	$1,155	$0
1.5	CO_2 Compression COG +BFS (including intercoolers)	$9,613	$817	$2,085.97	$12,516	$4
1.6	CO_2 Compression COG PPS (including intercoolers)	$22,920	$1,948	$4,973.74	$29,842	$9
1.7	CO_2 After-cooler COG + BFS	$2,884	$245	$625.90	$3,755	$1
1.8	CO_2 After-cooler COG PPS	$5,780	$491	$1,254.31	$7,526	$2
1.9	Cooling Water Chiller Unit COG + BFS	$12,398	$1,054	$2,690.45	$16,143	$5
1.10	Cooling Water Chiller Unit COG PPS	$24,848	$2,112	$5,392.07	$32,352	$10
1.11	LP Steam Boiler COG + BFS	$7,113	$605	$1,543.46	$9,261	$3
1.12	LP Steam Boiler COG PPS	$16,960	$1,442	$3,680.23	$22,081	$7
1.13	Balance of Plant (Instruments, Site, Buildings, etc)	$35,602	$3,026	$7,725.63	$46,354	$14
	TOTAL COST	**$391,617**	**$33,287**	**$84,981**	**$509,886**	**$153**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$23,795	$7
	Property Taxes and Insurance				$10,198	$3
	Total Fixed O&M				$33,993	$10
	Variable O&M (Maintenance Material Cost)				$50,989	$15
	Consumables (water treatment, corrosion inhibitor, MDEA solvent, Activated Carbon, etc)				$8,869	$3
	Purchased Natural Gas				$90,628	$27
	Purchased Power				$33,897	$10
	Total				**$218,377**	**$76**
	Initial Fill MDEA Solvent COG + BFS				$519	$0
	Initial Fill MDEA Solvent COG PPS				$1,237	$0
	Initial Fill Corrosion Inhibitor COG +BFS				$35	$0
	Initial Fill Corrosion Inhibitor COG PPS				$82	$0
	Owner's Cost				$123,940	$37
	Total				**$125,813**	**$38**
	Total Overnight Costs (TOC)				**$635,699**	**$191**
	TASC Multiplier				1.099	
	Total As-Spent Cost (TASC)				**$698,633**	**$210**

The first-year breakeven cost for a retrofit site was calculated and is shown in Exhibit 8-20. The financial assumptions and retrofit factor used represent the assumptions detailed in Section 5.

Exhibit 8-20 Breakeven cost for 2.54 M tonnes/yr retrofit



RETROFIT BREAKEVEN COST		
Case:	Steel/Iron	
Plant Size:	2.54	M tonnes/yr
Emissions:	3,323,826	tonnes CO_2/yr
Capacity Factor:	0.85	
Cost Base (June):	2011	
Retrofit Factor:	1.05	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	33.66	\$/tonne CO_2
Fixed O&M	10.23	\$/tonne CO_2
Variable O&M	15.34	\$/tonne CO_2
Consumables	2.67	\$/tonne CO_2
Purchased Power	10.20	\$/tonne CO_2
Purchased Natural Gas	27.27	\$/tonne CO_2
First year Breakeven Cost	99.36	\$/tonne CO_2

This steel case is unlike any of the other cases considered in this study, in that, this case utilizes two distinct separation and compression facilities to address three separate point sources. However, the overall retrofit breakeven cost value calculated represents the totality of these two systems, and CO_2 flows, simplified to one breakeven cost. Therefore, the breakeven costs for the COG/BFS and COG PPS were separated to identify if either of these point sources was a major contributor to the overall plant breakeven cost. Using the same methodology (not shown), the COG/BFS retrofit breakeven cost was calculated to be \$101.31/tonne and the COG PPS retrofit breakeven cost was calculated to be \$98.84/tonne. With an overall retrofit breakeven cost of \$99.36/tonne, neither of the two point sources contributes a disproportionate amount to the overall retrofit breakeven cost.

As discussed in section 8.2.2, the BFS PPS was not considered as a point source for capture. However, given the stream data by the Wiley study, it can be concluded that the BFS PPS retrofit breakeven cost if calculated, would closely mirror the individual retrofit breakeven cost calculated for the COG PPS, a value of \$98.84/tonne. Inclusion of the BFS PPS as previously discussed, would not improve economies of scale as the plot plan given for the Braddock plant does not dictate easy combination of point sources due to distance. The only benefit of considering the BFS PPS point source would be to increase the total amount of CO_2 available to sell from the overall steel plant. In practical application of this study, if the base plant considered was closely located to existing EOR pipeline infrastructure, and the demand for additional CO_2 was sufficient, then inclusion of the BFS PPS as a point source to be captured would be warranted. However, this scenario is not considered in this study.

The breakeven cost for steel is significantly high. This is the result of the lower CO_2 stream concentration, and the necessity for the stream to be purified before compression to meet EOR pipeline specifications as given in the QGESS. (10) However, as compared to refinery hydrogen, which also requires separation and purification, the cost of steel is \$13.28/tonne less. These results are interesting, as the stream purity for the steel case is less (23-27 percent CO_2) than the refinery hydrogen case (44 percent CO_2), yet the breakeven cost for the lower purity steel case is less than the higher purity refinery hydrogen case. This is a result of the amount of CO_2 available for capture, and therefore economies of scale.

The most costly utility associated with the purification process is the steam requirement. The capital and operating costs for the package boiler used to provide steam for the purification process may be excluded if the base plant has the capacity to provide the necessary steam. This would reduce all of the line items in Exhibit 8-20, and eliminate the need to purchase natural gas, thus significantly reducing the breakeven cost.

8.2.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

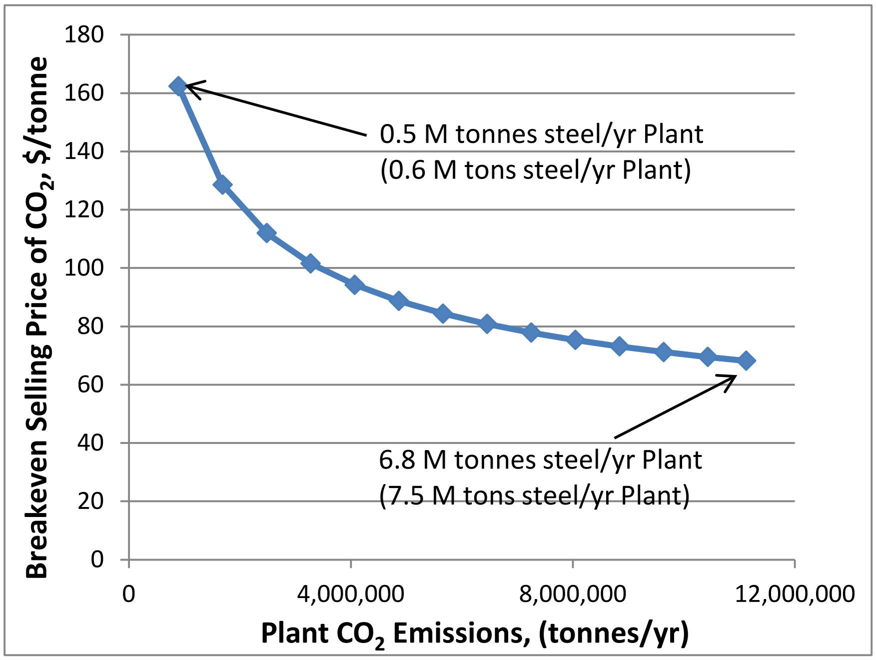
- Plant Size
- COE
- CO_2 Purity
- CCF
- Natural Gas Price

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore the only sensitivity discussed in this section. The other four sensitivities, COE, CO_2 purity, CCF, and Natural Gas Price are discussed and compared across the other cases in Section 9.2.

8.2.9.1 Plant Size

Plant size was varied over the range of 0.5 M tonnes steel/yr to 6.8 M tonnes steel/yr, representing the smallest and largest U.S. BOF steel plant capacities. These values use the same set of steel plant assumptions set forth in section 8.2.3. However, for other steel plants additional point sources may be located in a closer proximity than the 5 miles assumed here, and could possibly take advantage of economies of scale by combining streams. As shown in Exhibit 8-21, as the plant size decreases over the given range, the breakeven cost increases by $94.24/tonne CO_2. This variation over the range is much more significant than in the high purity cases, but not as impactful as in the refinery hydrogen case.

Exhibit 8-21 Plant size sensitivity



Source: NETL

8.2.10 Steel/Iron Conclusion

As compared to other low purity CO_2 cases, steel is an attractive option due to its lower breakeven selling price, and significantly higher supply of CO_2.

A purification and compression system for a 2.54 M tonnes/yr capacity steel plant was modeled and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be \$99.36/tonne CO_2 for a retrofit site. Sensitivities were performed on the following variables: COE, CCF, CO_2 purity, and plant size. The sensitivity tests for plant size showed the largest change in breakeven selling price; when varying over the range, a change in price of \$94.24/tonne CO_2 was observed. With higher base breakeven costs, the losses in economies of scale when moving to smaller-sized plants are much more significant than in higher purity cases, and, therefore, result in a large change in price for the plant size sensitivity. However, as compared to the refinery hydrogen case, the change in breakeven selling price is less (as plant size is varied for steel). This is a result of the amount of CO_2 available for capture.

The sensitivity tests for COE and CCF also demonstrated a change in breakeven selling price. Variation of COE over a \$120/MWh range resulted in a \$20.89/tonne CO_2 increase in the breakeven price. Variation of CCF over a 25 percent range, 10-35 percent, resulted in a \$47.82/tonne CO_2 increase in the breakeven price.

The expected trend for the CO_2 purity sensitivity was also demonstrated. As with all cases, as the CO_2 purity approaches lower concentrations, the breakeven selling price will increase dramatically.

Further steel/iron examination and its comparison with the other eight industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

8.3 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 2008, the U.S. cement industry produced approximately $10 billion dollars of cement product. (51) That same year, the U.S. consumed 96.8 M tonnes of Portland cement (PC), (52) while it only produced 86.3 M tonnes of PC. (53) Due to this imbalance, imports filled the PC production gap. This level of consumption was a 17 percent decrease over the previous year and was, in part, attributed to decreases in residential construction. (51) Production trends continued to fall in 2009, but rebounded in 2010 and 2011, as shown by the U.S. Geological Survey (USGS) in Exhibit 8-22. (52)

Exhibit 8-22 USGS cement production trends

Year	2007	2008	2009	2010	2011[1]
PC Production, M tonnes	95.5	86.3	63.9	66.5	67.7
Apparent PC Consumption, M tonnes	116.6	96.8	71.5	71.2	72.3
U.S. Market Satisfied by U.S Production, %	81.9	89.2	89.4	93.4	93.6
PC Price, $/tonne	104.00	103.50	99.00	92.00	91.00

[1] - Estimated

The 2008 U.S. production rate equated to an 82 percent cement plant capacity utilization rate, (51) but at this production rate the remaining demand still needed to be satisfied with imports. The use of imports can be problematic, as imports are subject to availability of foreign supply and availability of ships for transport. Also, the U.S. cement industry is regional in nature due to the high cost of shipping cement over long distances.

There are two types of PC processes: wet kiln and dry kiln. The number of the more energy-intensive wet process kilns in the U.S. has declined by 80 percent from 234, in 1974, to 46, in 2008, while the number of dry process kilns was reduced from 198 to 131 over the same period. Since 2008, approximately 85 percent of U.S. cement is produced using the dry kiln process. (51)

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 8-23is a breakdown of the fuel type consumed by kiln process for 2010 as reported by the USGS. (52) The values were originally given in various volume and mass units and were subsequently converted to heat usage, given as B Btu (Billion Btu). The assumptions for these conversions are given as footnotes below Exhibit 8-23.

Exhibit 8-23 2010 U.S. Portland cement fuel consumption by process

Process Type	Number of Plants	Conventional Fuels				Waste Fuels		
		Coal (B Btu)[1]	Pet Coke (B Btu)[2]	Oil (B Btu)[3]	Natural Gas (B Btu)[4]	Tires (B Btu)[5]	Solid (B Btu)[6]	Liquid (B Btu)[7]
Wet	15	12,808	4,845	257	991	1,720	986	9,074
Dry	83	131,425	43,215	956	9,184	8,929	5,869	25,321

Both	2	4,990	-	-	265	-	-	1,637
Total	100	149,222	48,060	1,213	10,439	10,648	6,855	36,032

[1] – Coal is described as bituminous. HHV of as-received Illinois No. 6 coal of 11,666 Btu/lb used.

[2] – Pet coke HHV of 14,850 Btu/lb used.

[3] – Oil used a Fuel Oil No. 2 HHV of 150,110 Btu/gal.

[4] – Natural gas HHV from Bituminous Baseline report of 1,032 Btu/scf used.

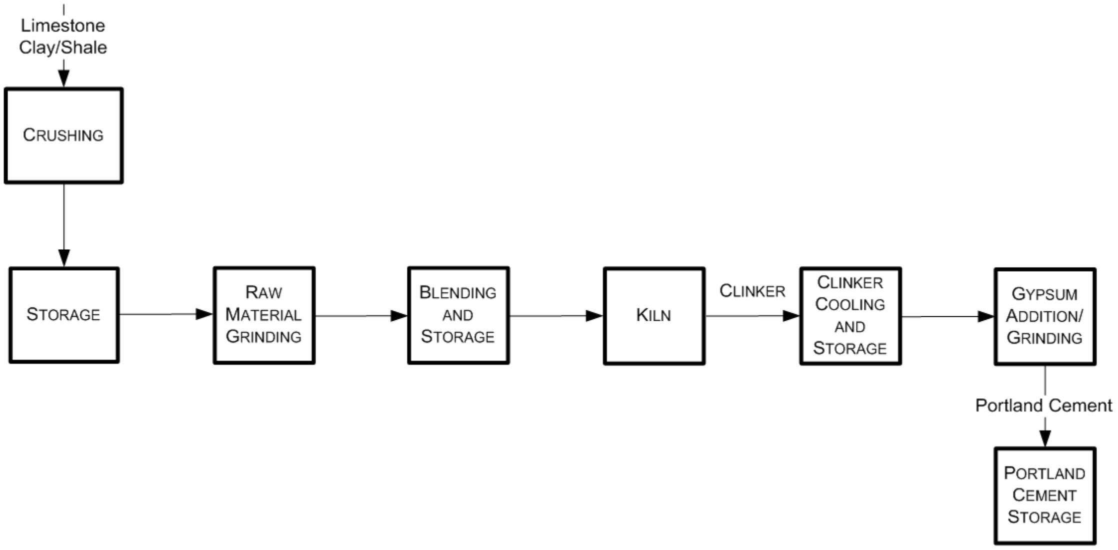
[5]– Tires used HHV of 15,000 Btu/lb.

[6] – Solid waste fuel used a HHV of 16 GJ/ton, converted to 7,584 Btu/lb.

[7] – Description of liquid waste fuel used was not specified; therefore, the HHV for Fuel Oil No. 2 of 150,110 was used as a proxy.

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. (54) Exhibit 8-24 shows the traditional PC production process, as adapted from Hassan (2005). (55)

Exhibit 8-24 Portland cement production process



Source: NETL

8.3.1 Size Range

In 2010 there were 100 U.S. cement plants, including both wet and dry processing kilns, in operation. (52) The capacity utilization rate for these 100 plants in 2010 was 67 percent, (56) giving a total capacity of 99.25 M tonnes, and an average plant capacity of 992,500 tonnes/year, which will serve as the basis for this study. For reference, 63 plants fall within the range of 0.5-

1.5 M tonnes cement/year, and 40 plants fall within the range of 0.75-1.25 M tonnes/year, which adequately brackets the assumed plant size for this study.

Cement production creates on average 1.2 tonnes CO_2 per tonne cement, (57) 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 0.43 tonnes CO_2 per tonne cement, and the average calcination emissions factor is 0.78 tonne CO_2 per tonne cement. (57) For our reference plant capacity, these emissions factors give 774,150 tonnes CO_2/year from calcinations of raw materials, and 426,775 tonnes CO_2/year from fuel burning, totaling 1.2 M tonnes CO_2/year from one point source.

These CO_2 production amounts are based on 100 percent capacity utilization. As given previously, the utilization factor in 2008 was found to be 82 percent, (51) whereas the utilization factor in 2010 was found to be 67 percent. (56) This large swing in plant capacity utilization over a two-year period reflects the potential volatility of the cement industry and the close reliance of domestic cement production to economic strength, particularly in the area of residential construction. Therefore, due to the fluctuation of plant utilization in recent years, a plant utilization of 100 percent is assumed for this study, which represents a best-case scenario.

8.3.2 CO_2 Point Sources

A techno-economic analysis of CO_2 capture from a cement plant used as a reference plant was the St. Mary's cement plant located in Ontario, Canada. Specifics given for that plant as of 2004 are shown below, in Exhibit 8-25. (55)

Exhibit 8-25 St. Mary's cement plant characteristics

St. Mary's Cement Plant Characteristics	
Clinker Production (tonnes)	637,000
Kiln Off-gas Temperature (°F)	320
Kiln Off-gas Pressure (psia)	14.7
Kiln Off-gas Mole Flow (kmol/hr)	9,851
Composition (mole %)	
H_2O	7.2
CO_2	22.4
N_2	68.1
O_2	2.3

For the current study, the main point source available for capture is the kiln off-gas, and the concentrations given for the St. Mary's cement plant will be assumed as representative. It is inherently assumed that the kiln off-gas requires only CO_2 removal and no other clean-up; however, it has been suggested that this might not be the case.

A study done by the International Energy Agency's Greenhouse Gas R&D Program (IEA GHG) in 2009 looked to estimate 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. (58) They point out that for post-combustion CO_2 capture to be implemented, there are several issues that must be addressed.

They include SO_2 concentration in the stream to be treated, which is dependent on the sulphide concentration in the raw meal; NO_2 concentration in the stream to be treated, which may result in solvent degradation; and dust present in the stream to be treated, 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 amine capture. However, a sensitivity is performed to account for these issues by adding an SCR unit to treat NO_x and an FGD to remove SO_x.

8.3.3 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:

- An average plant will be represented by a cement production capacity of 992,500 tonnes/year.
- The CO_2 amount available is 1.14 M tonnes CO_2/year at 100 percent cement plant capacity factor after accounting for AGR processing losses.
- The CO_2 stream available for capture is 22.4 mole percent CO_2
- Due to the lower purity, separation, compression, and cooling are required. Separation will be accomplished using an MDEA AGR.
- The temperature of the CO_2 available is 320 °F
- The pressure of the CO_2 available is 14.7 psia
- The CO_2 quality will be based on the EOR "pipeline" standard as mentioned in the NETL QGESS for CO_2 Impurity Design Parameters. (10)

8.3.4 CO_2 Capture System

The kiln off-gas stream concentrations is relatively low; however, it is above typical CFPP flue gas concentrations and will require purification before compression to attain EOR pipeline standards.

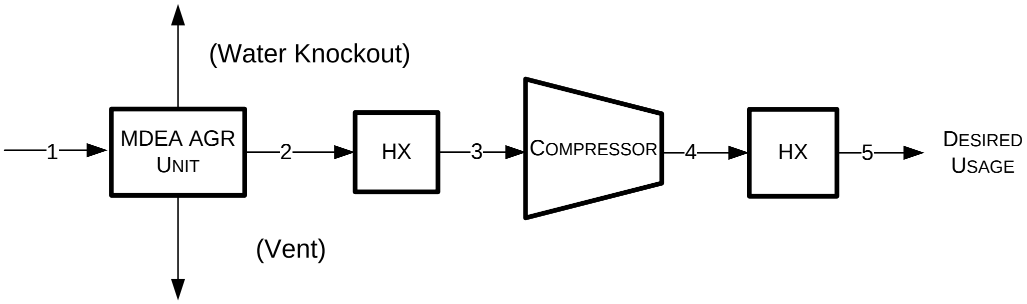
The purification system used is the MDEA system discussed in Section 6.2. Steam for solvent regeneration will be provided by the package steam boiler discussed in Section 6.1.

One integrally geared centrifugal compression train will be used, and scaled from the quote discussed in Section 6.1 based on product CO_2 flow.

8.3.5 Block Flow Diagram, Stream Table, and Performance Summary

Since the cement process releases low purity CO_2, separation along with cooling and compression is required for the CO_2 stream. As shown in Exhibit 8-26, the kiln off-gas is sent to the MDEA separation unit. Three streams result from the MDEA unit; a water knockout stream, a vent stream that contains mostly nitrogen, oxygen, and some CO_2, and the purified CO_2 stream. The CO_2 stream is then piped through a heat exchanger to reduce temperature, compressed (with inter-stage cooling), and then after-cooled before reaching the EOR pipeline Exhibit 8-26 shows the BFD for this process, and Exhibit 8-27 shows the stream table for this process.

Exhibit 8-26 Cement CO_2 capture block flow diagram



Source: NETL

Exhibit 8-27 Cement 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	0.2240	0.9932	0.9932	1.0000	1.0000
COS	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	0.0067	0.0067	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000
N_2	0.6810	0.0000	0.0000	0.0000	0.0000
O_2	0.0230	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (kg_{mol}/hr)	13,904	2,979	2,979	2,959	2,959
V-L Flowrate (kg/hr)	430,581	130,577	130,577	130,216	130,216
Solids Flowrate (kg/hr)	0	0	0	0	0
Temperature (°C)	160	49	27	134	49
Pressure (MPa, abs)	0.1	0.1	0.1	15.3	15.3
Enthalpy (kJ/kg)^A	268.7	47.2	27.9	30.7	-171.6
Density (kg/m^3)	0.9	2.2	2.3	259.3	673.3
V-L Molecular Weight	30.968	43.835	43.835	44.009	44.009
V-L Flowrate (lb_{mol}/hr)	30,653	6,567	6,567	6,523	6,523
V-L Flowrate (lb/hr)	949,268	287,873	287,873	287,078	287,078
Solids Flowrate (lb/hr)	0	0	0	0	0
Temperature (°F)	320	120	80	274	120
Pressure (psia)	14.7	19.7	19.0	2,215.8	2,215.1
Enthalpy (Btu/lb)^A	115.5	20.3	12.0	13.2	-73.8
Density (lb/ft^3)	0.054	0.140	0.145	16.186	42.035

The performance results are based on the compressor discussed in Section 6.1. The performance summary is provided in Exhibit 8-28.

Exhibit 8-28 Performance summary

Performance Summary	
Item	**992,500 tonnes/year (kW_e)**
Kiln CO_2 Compressor	12,800
Kiln AGR MDEA	5,100
Steam Boiler Total	569
Total	18,469

8.3.6 Capture Integration

The cooling water system is stand-alone however there is a potential to integrate make-up water to feed or partially feed the cooling water system thereby reducing the unit size or replacing it completely with a simple heat exchanger. This would have to 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.

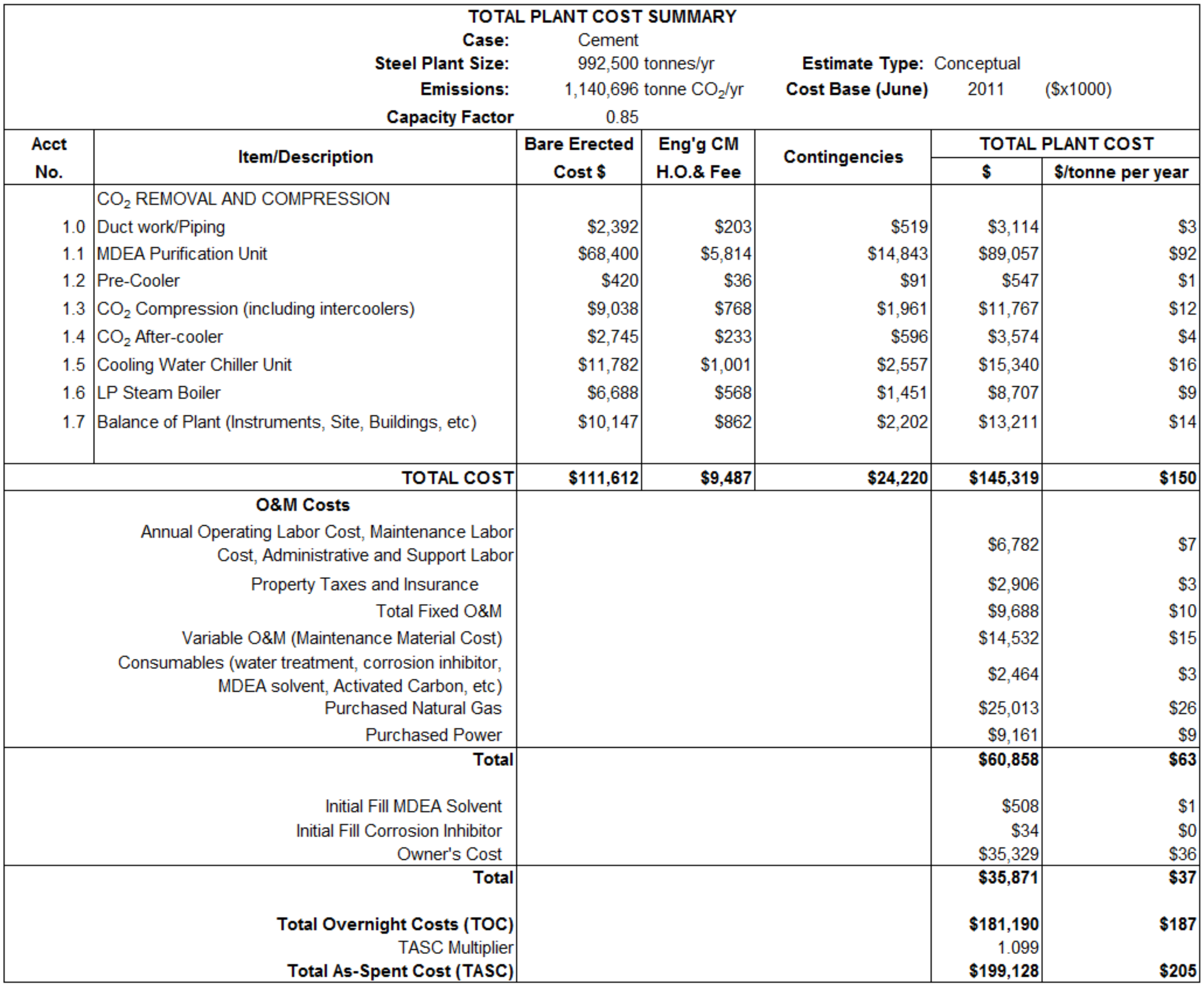
8.3.7 Power Source

The power consumption for compression and separation for this case is 18.47 MW, which was scaled from the equipment outlined in Section 6. Power consumption estimates for the stand-alone cooling water system were scaled as described in Section 6.4. The total power requirement was approximated to be 21.5 MWh/hr, which includes all power required by the compression train, MDEA unit, and the cooling water system. Power will be purchased at a rate of $58.59/MWh as discussed in Section 6.

8.3.8 Economic Analysis Results

The O&M, TOC for Greenfield and retrofit, and breakeven costs are given in this section for analysis and discussion. Owner's costs are calculated using the same methodology employed in the Bituminous Baseline Report. (1) Owner's costs account for items such as preproduction start-up costs for operation, maintenance, consumables, and other, inventory capital, financing costs, and other costs.

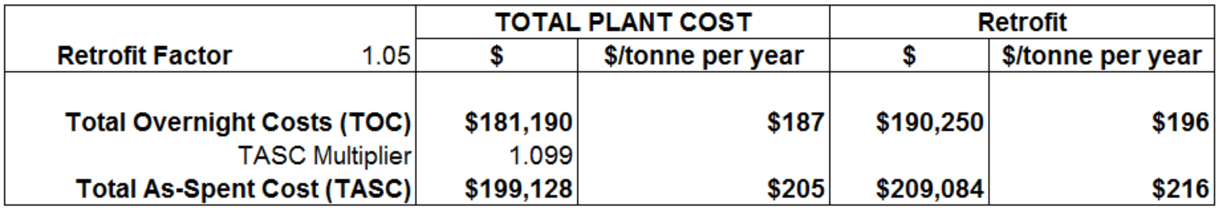
Exhibit 8-29 Capital and O&M costs for cement Greenfield site



TOTAL PLANT COST SUMMARY
Case: Cement
Steel Plant Size: 992,500 tonnes/yr **Estimate Type:** Conceptual
Emissions: 1,140,696 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST $	TOTAL PLANT COST $/tonne per year
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$2,392	$203	$519	$3,114	$3
1.1	MDEA Purification Unit	$68,400	$5,814	$14,843	$89,057	$92
1.2	Pre-Cooler	$420	$36	$91	$547	$1
1.3	CO_2 Compression (including intercoolers)	$9,038	$768	$1,961	$11,767	$12
1.4	CO_2 After-cooler	$2,745	$233	$596	$3,574	$4
1.5	Cooling Water Chiller Unit	$11,782	$1,001	$2,557	$15,340	$16
1.6	LP Steam Boiler	$6,688	$568	$1,451	$8,707	$9
1.7	Balance of Plant (Instruments, Site, Buildings, etc)	$10,147	$862	$2,202	$13,211	$14
	TOTAL COST	**$111,612**	**$9,487**	**$24,220**	**$145,319**	**$150**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$6,782	$7
	Property Taxes and Insurance				$2,906	$3
	Total Fixed O&M				$9,688	$10
	Variable O&M (Maintenance Material Cost)				$14,532	$15
	Consumables (water treatment, corrosion inhibitor, MDEA solvent, Activated Carbon, etc)				$2,464	$3
	Purchased Natural Gas				$25,013	$26
	Purchased Power				$9,161	$9
	Total				**$60,858**	**$63**
	Initial Fill MDEA Solvent				$508	$1
	Initial Fill Corrosion Inhibitor				$34	$0
	Owner's Cost				$35,329	$36
	Total				**$35,871**	**$37**
	Total Overnight Costs (TOC)				**$181,190**	**$187**
	TASC Multiplier				1.099	
	Total As-Spent Cost (TASC)				**$199,128**	**$205**

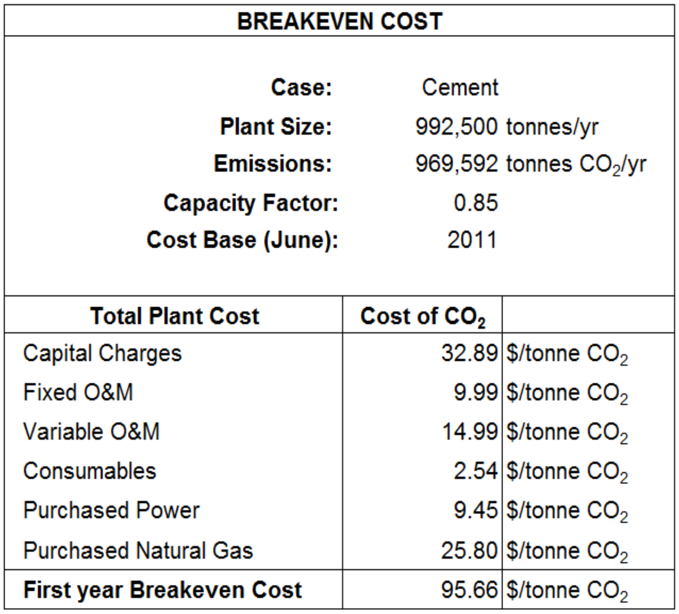
Exhibit 8-30 Total overnight costs for retrofit



		TOTAL PLANT COST		Retrofit	
Retrofit Factor	1.05	**$**	**$/tonne per year**	**$**	**$/tonne per year**
Total Overnight Costs (TOC)		**$181,190**	**$187**	**$190,250**	**$196**
TASC Multiplier		1.099			
Total As-Spent Cost (TASC)		**$199,128**	**$205**	**$209,084**	**$216**

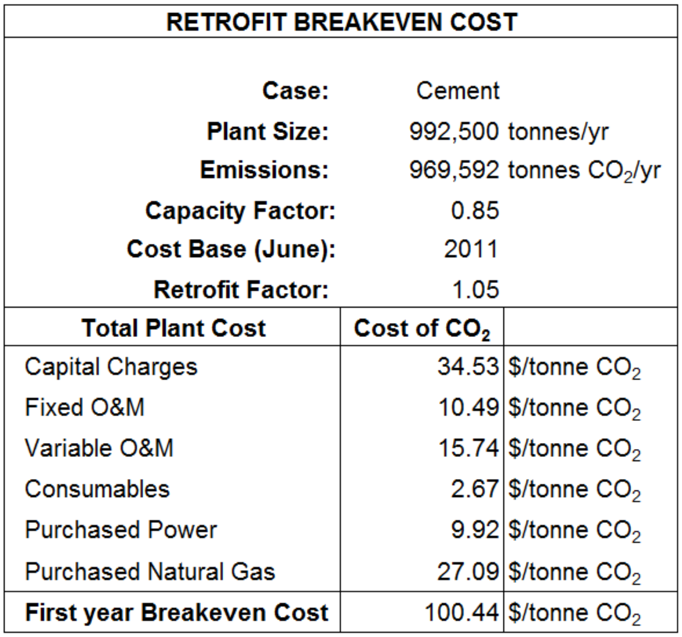
The first year breakeven cost for a Greenfield site and retrofit site were calculated and are shown in Exhibit 8-31 and Exhibit 8-32. The financial assumptions and retrofit factor used represent the assumptions detailed in Section 5.

Exhibit 8-31 Breakeven cost for 992,500 tonne/yr Greenfield



BREAKEVEN COST		
Case:	Cement	
Plant Size:	992,500 tonnes/yr	
Emissions:	969,592 tonnes CO_2/yr	
Capacity Factor:	0.85	
Cost Base (June):	2011	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	32.89	\$/tonne CO_2
Fixed O&M	9.99	\$/tonne CO_2
Variable O&M	14.99	\$/tonne CO_2
Consumables	2.54	\$/tonne CO_2
Purchased Power	9.45	\$/tonne CO_2
Purchased Natural Gas	25.80	\$/tonne CO_2
First year Breakeven Cost	95.66	\$/tonne CO_2

Exhibit 8-32 Breakeven cost for 992,500 tonne/yr retrofit



RETROFIT BREAKEVEN COST		
Case:	Cement	
Plant Size:	992,500 tonnes/yr	
Emissions:	969,592 tonnes CO_2/yr	
Capacity Factor:	0.85	
Cost Base (June):	2011	
Retrofit Factor:	1.05	
Total Plant Cost	**Cost of CO_2**	
Capital Charges	34.53	\$/tonne CO_2
Fixed O&M	10.49	\$/tonne CO_2
Variable O&M	15.74	\$/tonne CO_2
Consumables	2.67	\$/tonne CO_2
Purchased Power	9.92	\$/tonne CO_2
Purchased Natural Gas	27.09	\$/tonne CO_2
First year Breakeven Cost	100.44	\$/tonne CO_2

8.3.9 Sensitivity Analysis

Sensitivity analysis was performed on the following variables:

- Plant Size
- FGD and SCR
- COE

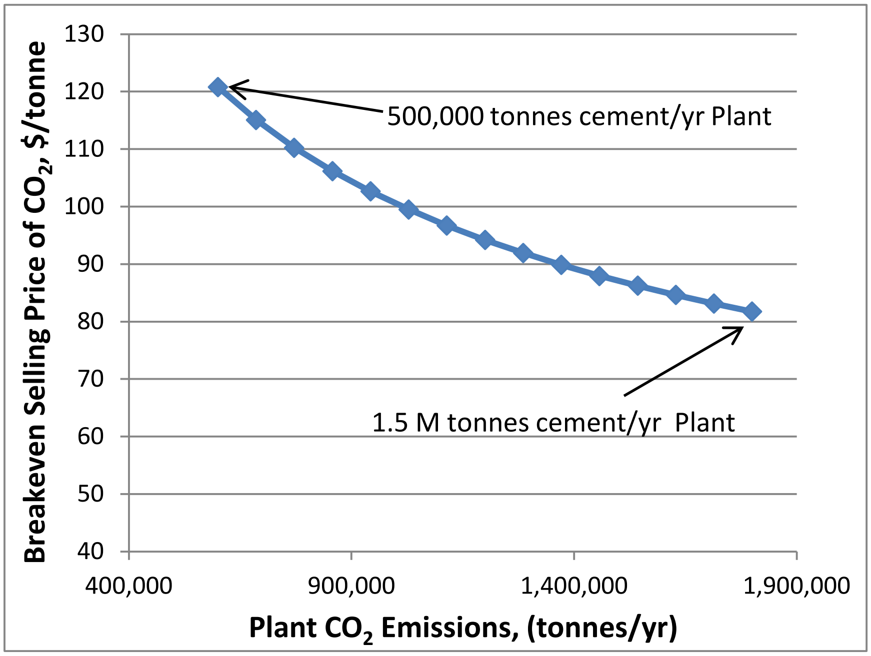
- CO_2 Purity
- CCF
- Natural Gas Price

The plant size sensitivity is process specific and difficult to compare across cases, and is therefore discussed in this section. The FGD and SCR sensitivity is also specific to the cement case, and is discussed in this section only. The other four sensitivities, COE, CO_2 purity, CCF, and Natural Gas Price, are discussed and compared across the other cases separately in Section 9.2.

8.3.9.1 Plant Size

Plant size was varied over the range of 0.5 to 1.5 M tonnes cement/yr, representing 63 of the current 100 U.S. cement plants. These values use the same set of cement plant assumptions set forth in section 8.3.3. As shown in Exhibit 8-33, as the plant size decreases over the given range, the breakeven cost increases by \$39.08/tonne CO_2.

Exhibit 8-33 Plant size sensitivity

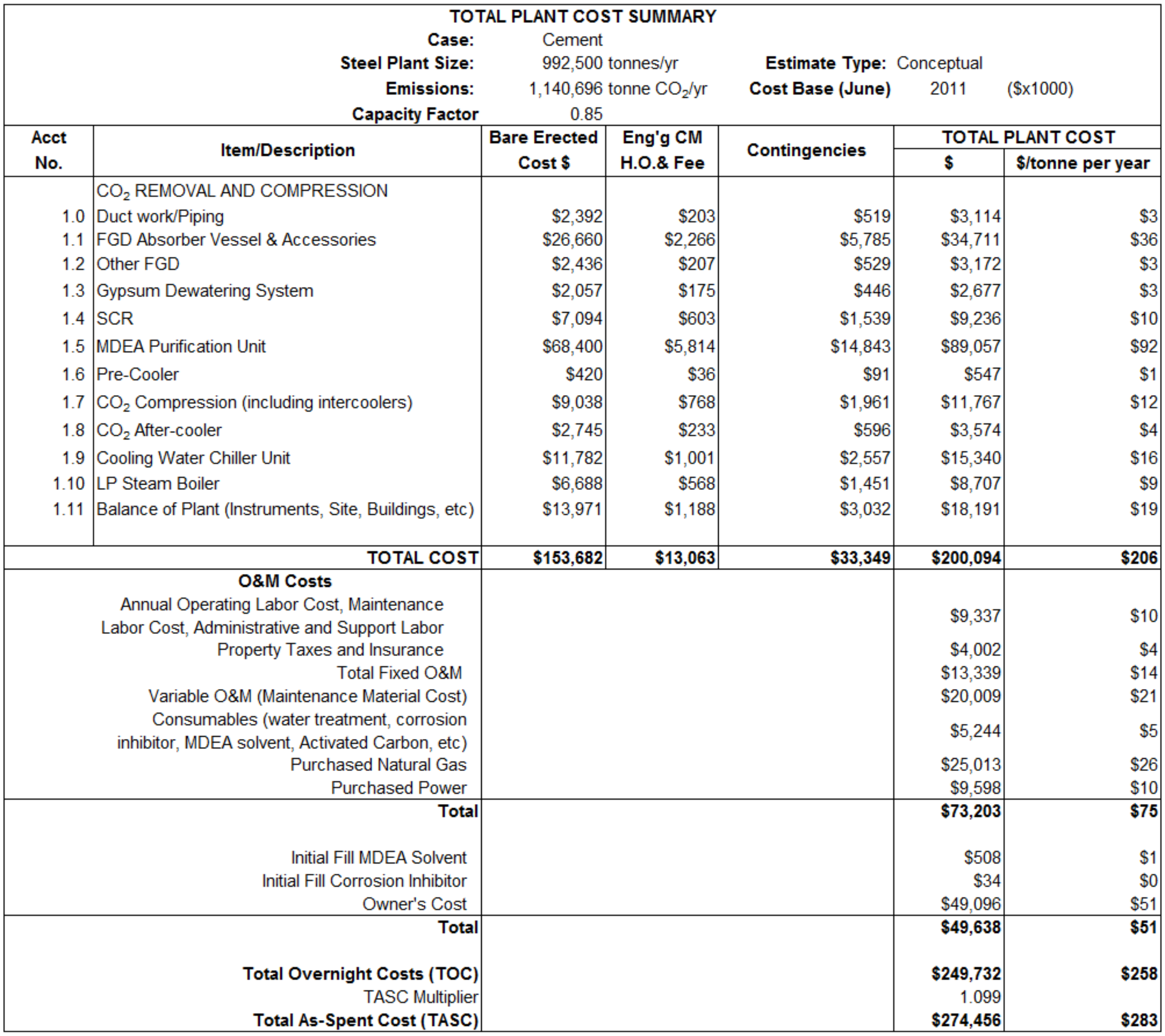


Source: NETL

8.3.9.2 FGD and SCR

As stated previously, the kiln off-gas may require additional treatment prior to CO_2 removal so as to maximize the efficiency of the MDEA removal system and prevent solvent degradation. Data for cement kiln off-gas SO_x/NO_x concentration was not available. 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 the Bituminous Baseline report, specifically PC case 12 based on quantity of gas treated. The results are shown in Exhibit 8-34 below.

Exhibit 8-34 Capital and O&M costs for cement Greenfield site with FGD and SCR



TOTAL PLANT COST SUMMARY

Case: Cement
Steel Plant Size: 992,500 tonnes/yr **Estimate Type:** Conceptual
Emissions: 1,140,696 tonne CO_2/yr **Cost Base (June)** 2011 ($x1000)
Capacity Factor 0.85

Acct No.	Item/Description	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies	TOTAL PLANT COST $	$/tonne per year
	CO_2 REMOVAL AND COMPRESSION					
1.0	Duct work/Piping	$2,392	$203	$519	$3,114	$3
1.1	FGD Absorber Vessel & Accessories	$26,660	$2,266	$5,785	$34,711	$36
1.2	Other FGD	$2,436	$207	$529	$3,172	$3
1.3	Gypsum Dewatering System	$2,057	$175	$446	$2,677	$3
1.4	SCR	$7,094	$603	$1,539	$9,236	$10
1.5	MDEA Purification Unit	$68,400	$5,814	$14,843	$89,057	$92
1.6	Pre-Cooler	$420	$36	$91	$547	$1
1.7	CO_2 Compression (including intercoolers)	$9,038	$768	$1,961	$11,767	$12
1.8	CO_2 After-cooler	$2,745	$233	$596	$3,574	$4
1.9	Cooling Water Chiller Unit	$11,782	$1,001	$2,557	$15,340	$16
1.10	LP Steam Boiler	$6,688	$568	$1,451	$8,707	$9
1.11	Balance of Plant (Instruments, Site, Buildings, etc)	$13,971	$1,188	$3,032	$18,191	$19
	TOTAL COST	**$153,682**	**$13,063**	**$33,349**	**$200,094**	**$206**
	O&M Costs					
	Annual Operating Labor Cost, Maintenance Labor Cost, Administrative and Support Labor				$9,337	$10
	Property Taxes and Insurance				$4,002	$4
	Total Fixed O&M				$13,339	$14
	Variable O&M (Maintenance Material Cost)				$20,009	$21
	Consumables (water treatment, corrosion inhibitor, MDEA solvent, Activated Carbon, etc)				$5,244	$5
	Purchased Natural Gas				$25,013	$26
	Purchased Power				$9,598	$10
	Total				**$73,203**	**$75**
	Initial Fill MDEA Solvent				$508	$1
	Initial Fill Corrosion Inhibitor				$34	$0
	Owner's Cost				$49,096	$51
	Total				**$49,638**	**$51**
	Total Overnight Costs (TOC)				**$249,732**	**$258**
	TASC Multiplier				1.099	
	Total As-Spent Cost (TASC)				**$274,456**	**$283**

The addition of SCR and FGD increases the total plant cost (TPC) over the base case Greenfield cost by approximately $54.7 M. The majority of this increase is attributed to the FGD absorber vessel, which accounts for $34.7M of the TPC. Fixed and Variable O&M costs also increase. The consumables costs also increase by $2.7M. This is due to the requirement of limestone for the FGD as well as 19 percent NH_3 for the SCR. The SCR catalyst is assumed to be included with equipment purchase.

The auxiliary requirements for the FGD and SCR are scaled from the Bituminous Baseline case 12, and total 610 kW.

The resulting breakeven selling price for the FGD/SCR sensitivity is shown in Exhibit 8-35.

Exhibit 8-35 Breakeven cost for 992,500 tonne/yr Greenfield site with FGD and SCR

FGD/SCR BREAKEVEN COST

Case:	Cement	
Plant Size:	992,500	tonnes/yr
Emissions:	969,592	tonnes CO_2/yr
Capacity Factor:	0.85	
Cost Base (June):	2011	

Total Plant Cost	**Cost of CO_2**	
Capital Charges	45.33	\$/tonne CO_2
Fixed O&M	13.76	\$/tonne CO_2
Variable O&M	20.64	\$/tonne CO_2
Consumables	5.41	\$/tonne CO_2
Purchased Power	9.90	\$/tonne CO_2
Purchased Natural Gas	25.80	\$/tonne CO_2
First year Breakeven Cost	120.83	\$/tonne CO_2

The result of this sensitivity is that the breakeven selling price increases from \$95.66/tonne CO_2 without FGD and SCR to \$120.83/tonne CO_2, an increase of \$25.17/tonne. At \$120.83/tonne, this cement sensitivity case is the highest breakeven selling price of any of the nine processes considered in this report. This breakeven selling price sensitivity is an approximation, as actual plant SO_x/NO_x 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.

8.3.10 Cement Conclusion

As compared to the previous lower purity CO_2 cases, cement is an equally attractive option to steel/iron, and more attractive than refinery hydrogen, in terms of breakeven cost. From a total industry amount of CO_2 available standpoint, referring to Exhibit ES-2, the cement industry has about 30 M tonnes/yr more CO_2 available than steel/iron, and about 10 M tonnes/yr more CO_2 available than refinery hydrogen based on this study's assumptions. However, as with the other lower purity sources, when compared to the higher purity cases, cement as expected results in a significantly higher breakeven selling price.

A compression system for a 992,500 tonne/year cement plant was modeled and techno-economic analysis was performed. The results showed the breakeven price of CO_2 to be \$95.66/tonne CO_2 for a Greenfield site, and \$100.44/tonne CO_2 for a retrofit site. The small disparity is the result of small differences in equipment required for a Greenfield versus retrofit site, assuming adequate plot plan space for the retrofit case. Sensitivities were performed on the following variables: COE, CCF, CO_2 purity, plant size, and additional stream impurities requiring treatment with an FGD and SCR. The sensitivity tests for plant size showed a large change in breakeven selling price, where a range of 0.5 – 1.5 M tonnes cement/yr, representing 63 of the 100 existing U.S. plants, resulted in a rise in the breakeven selling price of \$39.08/tonne CO_2.

The sensitivity tests for COE and CCF also demonstrated a change in breakeven selling price. Variation of COE over a \$20-140/MWh range resulted in a \$19.35/tonne CO_2 increase in the breakeven price. Variation of CCF over a 25 percent range, 10-35 percent, resulted in a \$46.72/tonne CO_2 increase in the breakeven price.

The sensitivity addressing additional impurities requiring treatment with FGD and SCR showed that the addition of these removal steps increased the Greenfield breakeven selling price from the base case of \$95.66/tonne CO_2 to \$120.83/tonne CO_2, a \$25.17/tonne CO_2 increase. It is not clear whether this sensitivity would be a common occurrence in existing U.S. cement plants, or a special, isolated case dependent on the raw materials used to make cement.

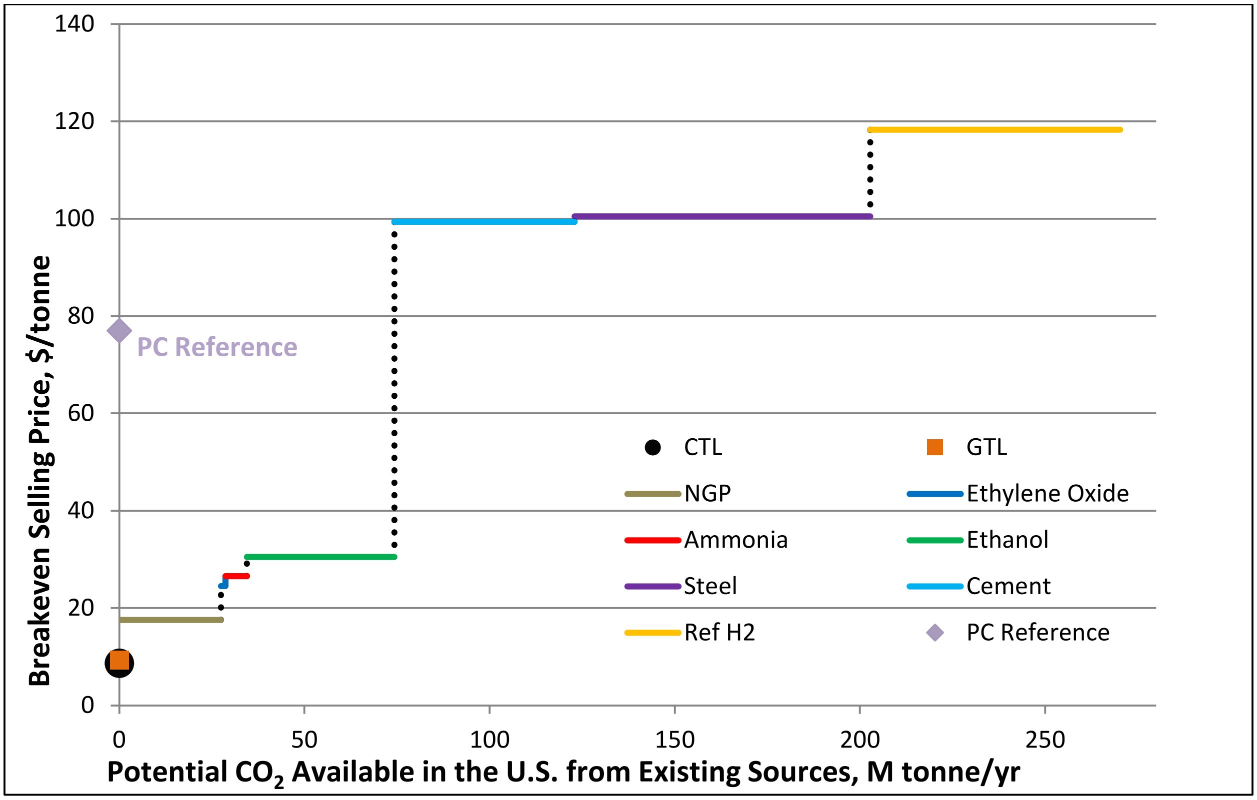
Further cement examination and its comparison with the other industrial processes can be found in Economic Analysis and Conclusions, Sections 9 and 10, respectively.

9 Economic Analysis

9.1 Economic Results

Exhibit 9-1 below shows the breakeven selling price results of each industry considered plotted against the approximated U.S. CO_2 supply each industry potentially has available for capture and use. A combination of both low breakeven selling price and large CO_2 supply may differentiate one industry from another. For example, EO and ammonia both present as low cost, but ammonia has a larger projected CO_2 supply. However, given that ethanol has only a slightly higher breakeven selling price as compared to ammonia, \$30.46/tonne versus \$26.52/tonne respectively, but almost 7 times more CO_2 available industry-wide, ethanol may be perceived as a more attractive option.

Exhibit 9-1 Breakeven selling price versus CO_2 supply



Note: The breakeven selling prices used are for a representative plant for each industry and do not account for the size variability and associated economies of scale for the actual sources in each industry.

Source: NETL

Brief descriptions of the assumptions used to project the 'Potential CO_2 Available' axis numbers follow:

Natural Gas Processing:

- Value is based on the Gas Technology Institute (GTI) Gas Resource Database for formations producing gas containing more than 2 percent CO_2, below which is generally accepted as pipeline quality gas. The potential CO_2 available from these formations was calculated by summing the potential of individual formations, which was calculated by multiplying the average CO_2 formation concentration by the annual formation production, giving 27.5 M tonnes CO_2.

Ethylene Oxide:

- The CO_2 potential from EO production (1.2 M tonnes) was calculated by applying the EO:CO_2 mass ratio (6:2) to the total U.S. EO capacity in 2007, which was determined to be 3.6 M tonnes by summing all 10 plants' reported capacity. (32)

Ammonia:

- The total 2006 U.S. ammonia production was given by the U.S. Environmental Protection Agency (EPA) as 8.2 M tonnes/yr. Applying a factor of 1.87 tonne CO_2/tonne NH_3, which represents the total plant CO_2 produced as given by Strait and Nagvekar, (4) and applying this study's assumptions of urea CO_2 consumption and CO_2 distribution over the two point sources, results in the total CO_2 available for capture of 5.8 M tonnes/yr.

Ethanol:

- Total March 2012 U.S. ethanol production was 13,894 M gal. (17) Using the equation given in the ethanol section, Section 7.1.2, for determining CO_2 produced during the fermentation process, and using the emission factor of 6.31 lb CO_2/gal ethanol, it was determined that the CO_2 available for capture totals 39.7 M tonnes/yr.

Steel/Iron

- BOF plant capacity was given by the American Iron and Steel Institute (48) as 41.2 M tonnes steel/yr, in November 2010. The literature shows a utilization ratio of 73.8 percent, an emissions factor of 2.2 tonne CO_2/tonne steel, and that 72.8 percent of the total CO_2 available results from higher purity point sources. Applying these factors results in 48.7 M tonnes/yr CO_2 available from higher purity BOF process point sources.

Cement:

- In 2010, the USGS reported 100 cement plants in the U.S. producing 66.5 M tonnes of cement. (52) Applying a factor of 1.2 tonnes CO_2/tonne cement (57) gives 79.8 M tonnes CO_2 available.

Refinery Hydrogen:

- Total U.S. hydrogen production was given as 9.1-10 M tonnes/yr (median of 9.5 M tonnes/yr used in calculation), with refineries representing 59 percent of U.S. production. The CO_2 production factor was given as 9-12 tonnes CO_2/ton H_2 (12 tonnes CO_2/tonne H_2 used), resulting in 67.5 M tonnes CO_2/yr available from refinery hydrogen production.

CTL/GTL:

- CTL and GTL are represented by single points on the y-axis of Exhibit 9-1, indicating that there are no existing U.S. CTL or GTL plants from which to calculate a total amount of CO_2 available for capture. Sections 7.5 and 7.6 use previous NETL studies that assume a reference liquids output, resulting in specific amounts of CO_2 available so as to calculate a breakeven selling price.

There is a large disparity in breakeven selling price between the higher purity, lower breakeven selling price industries (natural gas processing, EO, ammonia, and ethanol), and lower purity, higher breakeven selling price industries (steel, cement, and refinery hydrogen). However, the lower purity sources are able to provide a significantly larger supply.

A more detailed description of the methodology used to calculate the 'Potential CO_2 Available' axis for each industry can be found in Appendix A.

9.2 Sensitivity Analysis Comparison

Sensitivity analysis was performed on the following variables for all cases:

- Plant Size

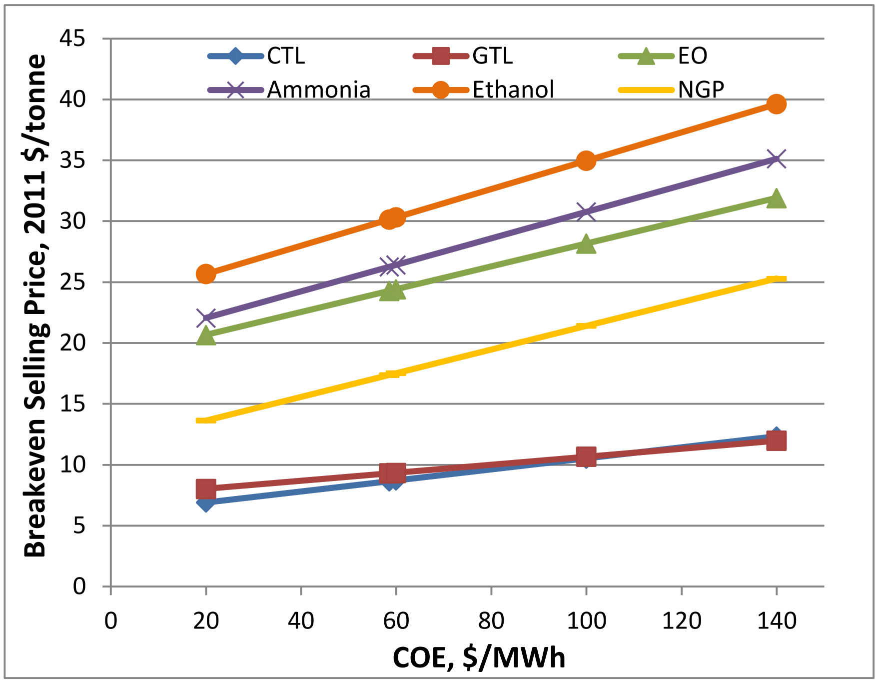
- Cost of Electricity
- CO_2 Purity
- Capital Charge Factor
- Retrofit Factor

Sensitivities pertaining to plant size are discussed in the corresponding industry's section as this sensitivity is process specific and not easily comparable across different industries. Sensitivity analysis to natural gas price was also performed for only the low purity cases that require steam generated by the natural gas boiler.

9.2.1 Cost of Electricity

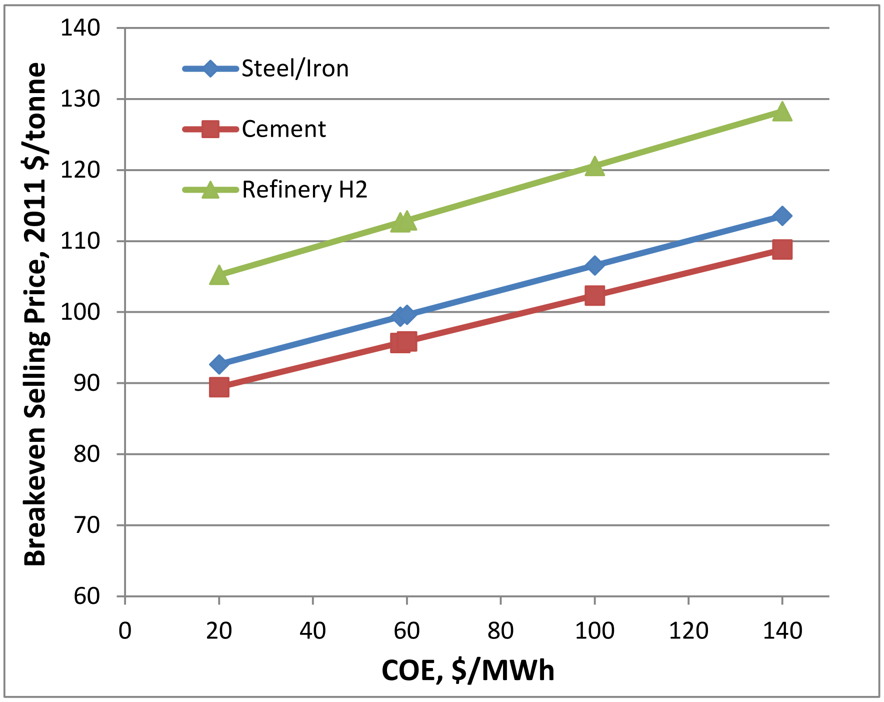
The COE purchased was varied over the range of 20-140 $/MWh for all cases. The base case used a value of $58.59/MWh, obtained from the Bituminous Baseline Report Case 13, NGCC without capture, and this value was used in all cases for the Greenfield and retrofit breakeven cost. The results of the high purity and low purity sensitivities are plotted in Exhibit 9-2 and Exhibit 9-3.

Exhibit 9-2 High purity sources cost of electricity sensitivity comparison



Source: NETL

Exhibit 9-3 Low purity sources cost of electricity sensitivity comparison



Source: NETL

For the high purity sources, the change in COE has about the same effect: a slight increase in the breakeven selling price from a \$20/MWh COE to \$140/MWh COE. The most dramatic change comes in the case of ethanol where the increase in the breakeven selling price over the COE range is the largest, giving a value of \$13.96/tonne increase for ethanol.

In the high purity sources, the slope of the CTL and GTL cases are noticeably different from the other four high purity cases. This is explained as a result of economies of scale. The CO_2 processed per MWh of auxiliary load for CTL and GTL is 18.8 and 25.8 tonnes CO_2/MWh, respectively. The remaining four cases all fall in the range of 7-9 tonnes CO_2/MWh. Since the amount of CO_2 that may be processed per MWh of electricity required is greater in the CTL and GTL cases, as the COE of purchased power increases, the change in breakeven selling price will not be as drastic, resulting in a less steep slope for the CTL and GTL cases presented in Exhibit 9-2 as compared to the other high purity cases.

For the low purity sources, the change in COE is slightly higher as compared to the high purity sources. Over the same \$20/MWh COE to \$140/MWh COE variable range, Steel/Iron shows a breakeven selling price increase of \$20.89/tonne, Cement increases by \$19.35/tonne, and Refinery Hydrogen shows the largest increase of \$23.03/tonne.

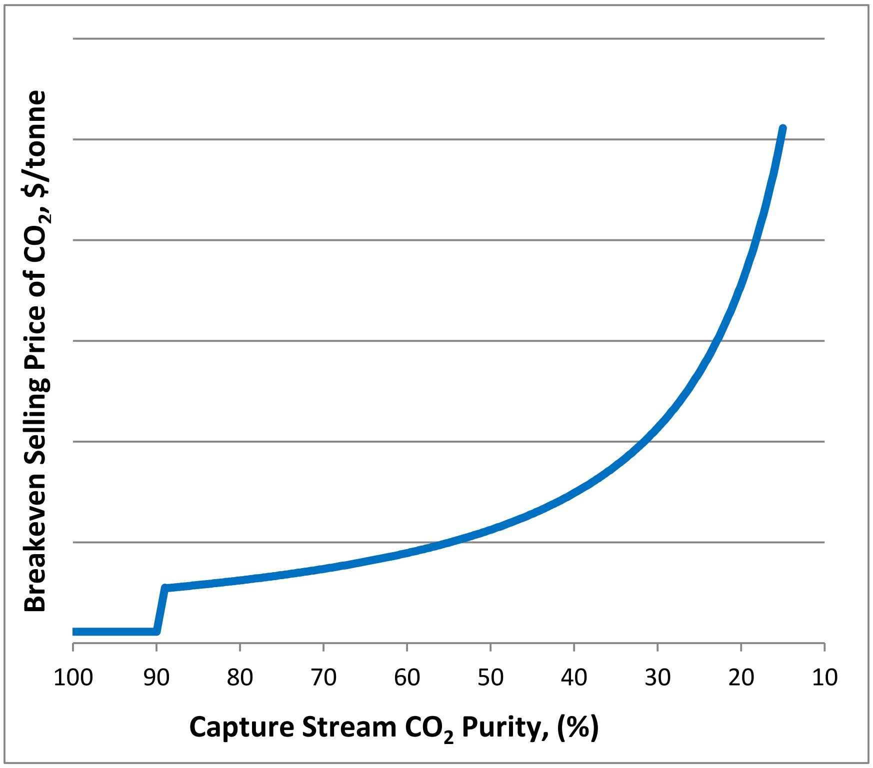
One difference between the low and high purity sources is the amount of equipment required. The only equipment that high purity sources require is compression and a cooling water unit, whereas the low purity sources require compression, a cooling water unit, as well as the MDEA AGR unit for separation. Therefore, the electrical auxiliary loads for the low purity sources are higher than the electrical auxiliary load for the high purity sources. The effect is that as the COE purchased increases, the breakeven selling price of the low purity sources will increase faster than the high purity sources, due to the larger auxiliary load required by the low purity sources.

9.2.2 CO_2 Purity

CO_2 purity was listed as a sensitivity study, but presents challenges in calculating sensitivity values. The QGESS for CO_2 Impurity Design Parameters (10) gives a literature range for EOR CO_2 purity of 90-99.8 volume percent. For a CO_2 purity within this range, purification of the EOR stream is not required, assuming that contaminants in the balance of the stream are also within required QGESS specifications. For example, given a stream of 90 volume percent CO_2, if the balance of the stream were H_2S, a contaminant, this H_2S concentration would fall outside the QGESS maximum H_2S stream concentration of 1.3 volume percent, and would require further purification. This would present as a different case with a different set of assumptions, and less as a sensitivity to the present study. Given our current set of stream conditions, purity variations are not anticipated.

For the purposes of this study, purity variations within the QGESS range will not affect the breakeven cost of CO_2, implying the assumption that contaminants are within QGESS limits as well, but variations from 90 percent CO_2 and below will affect the breakeven cost of CO_2, as this will require additional equipment to obtain acceptable EOR CO_2 purity. Given in Exhibit 9-4 is an approximation of how the breakeven selling price of CO_2 would be affected by stream purity issues for the high purity cases.

Exhibit 9-4 CO_2 purity sensitivity for high purity cases

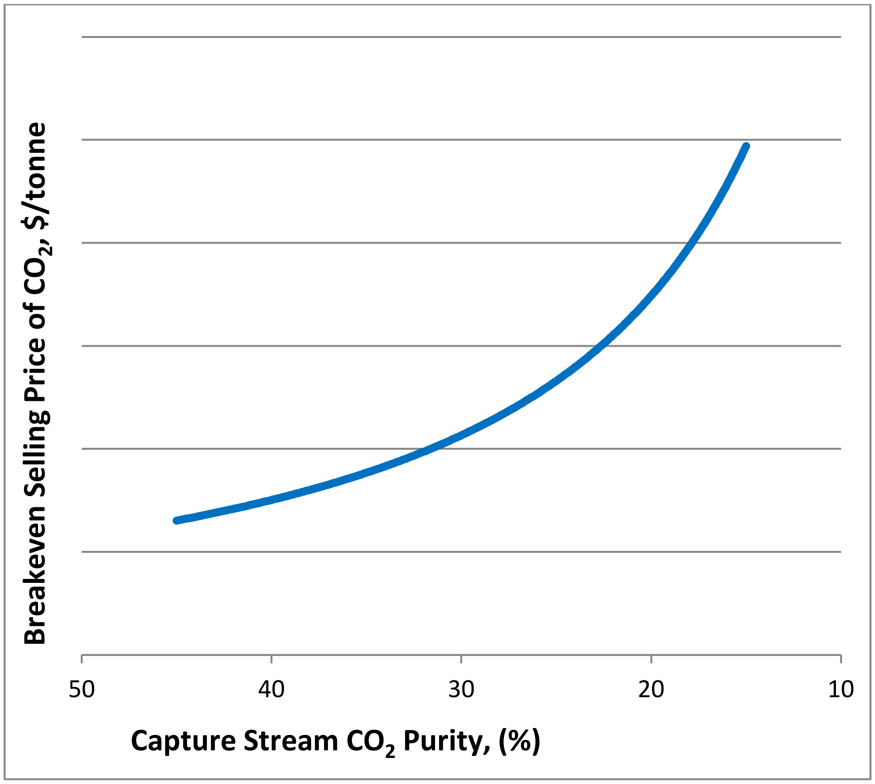


Source: NETL

The y-axis values given in Exhibit 9-4 for breakeven cost are not intended to represent real values, but rather serve as an approximation to demonstrate the trend. A CO_2 purity of 90-100 percent represents the high purity sources in this study, and a purity of approximately 15 percent would represent CFPP flue gas concentrations. There is little or no change in the breakeven cost over the range of 90-100 percent CO_2 purity, assuming reasonable impurity concentrations. As the CO_2 purity goes below 90 percent, the breakeven cost begins to increase due to the requirement for purification equipment to reach QGESS stream specifications, and as the purity approaches CFPP flue gas concentrations, the breakeven cost increases significantly.

The effects of decreasing CO_2 purity in the low purity processes are much more straightforward as compared to the high purity processes. In the low purity cases, separation and purification equipment are already present, and therefore, a decrease in the CO_2 purity does not require the addition of equipment, but rather only requires a greater degree of separation to achieve QGESS CO_2 product stream requirements. The increase in cost can be accounted for in higher operating costs stemming from a larger stream requiring treatment, more solvent make-up required for the larger stream and the inevitably higher solvent losses. Shown in Exhibit 9-5 is an approximation of the breakeven selling price trend for the low purity cases. As with the high purity plot, this low purity plot is not intended to represent real values, but only approximate the trend.

Exhibit 9-5 CO_2 purity sensitivity for low purity cases



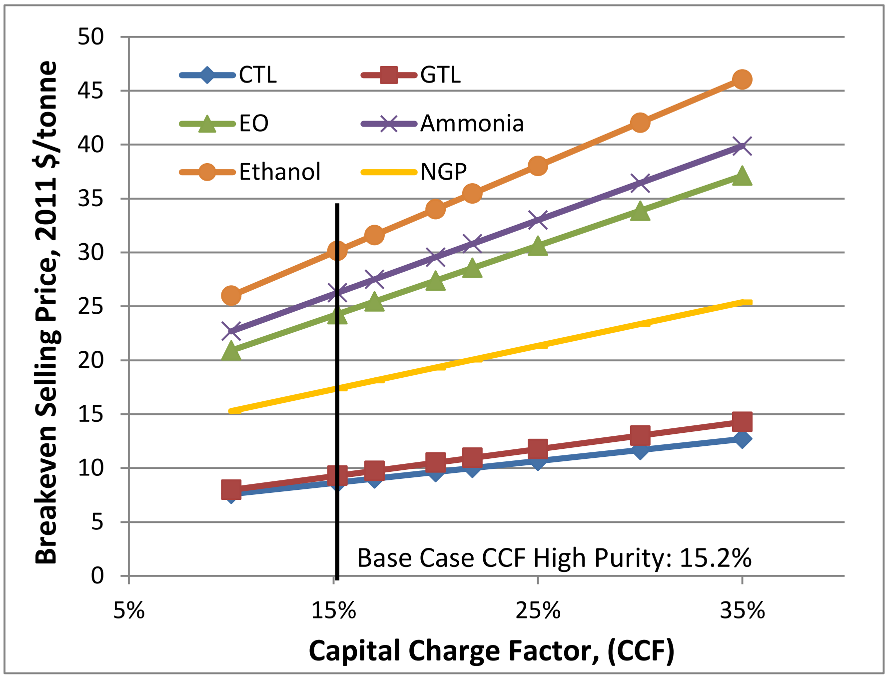
Source: NETL

In both the high purity and low purity processes, as the CO_2 purity approaches CFPP concentrations the breakeven selling price increases. The major difference is observed over the transition from 90 percent CO_2 to below 90 percent CO_2, the point at which the addition of purification equipment is required.

9.2.3 Capital Charge Factor

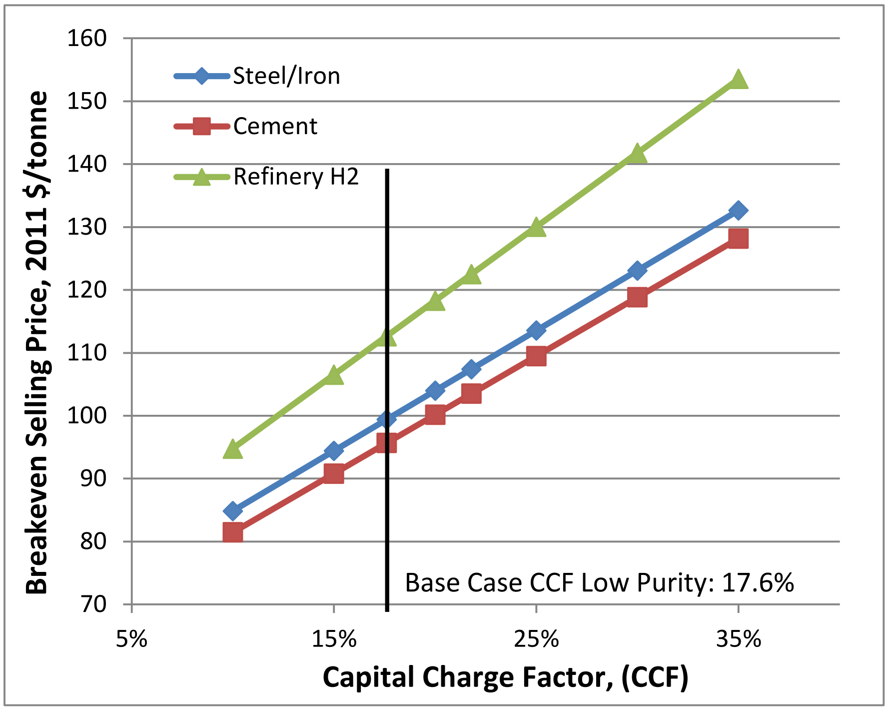
CCF is an important parameter used in this study, and should not be kept constant when comparing different processes, such as ethanol versus steel/iron. For a case where only compression and cooling is required such as ethanol, the time to install will be less than a case where compression, cooling, and capture are required, such as steel/iron. This difference in time necessitates an adjustment in the capital expenditure period assumed, and directly affects the CCF value that should be used. This was the basis presented previously in Section 5.1 for using a different CCF for the low versus high purity cases. The CCR for high and low purity cases is 15.2% and 17.6% respectively. The extension of this basis is to perform a sensitivity to the CCF to bound effects caused by changes to the capital expenditure period, or other parameters effecting the CCF. Results of the CCF sensitivity are shown in Exhibit 9-6 and Exhibit 9-7.

Exhibit 9-6 Capital charge factor sensitivity for high purity sources



Source: NETL

Exhibit 9-7 Capital charge factor sensitivity for low purity sources



Source: NETL

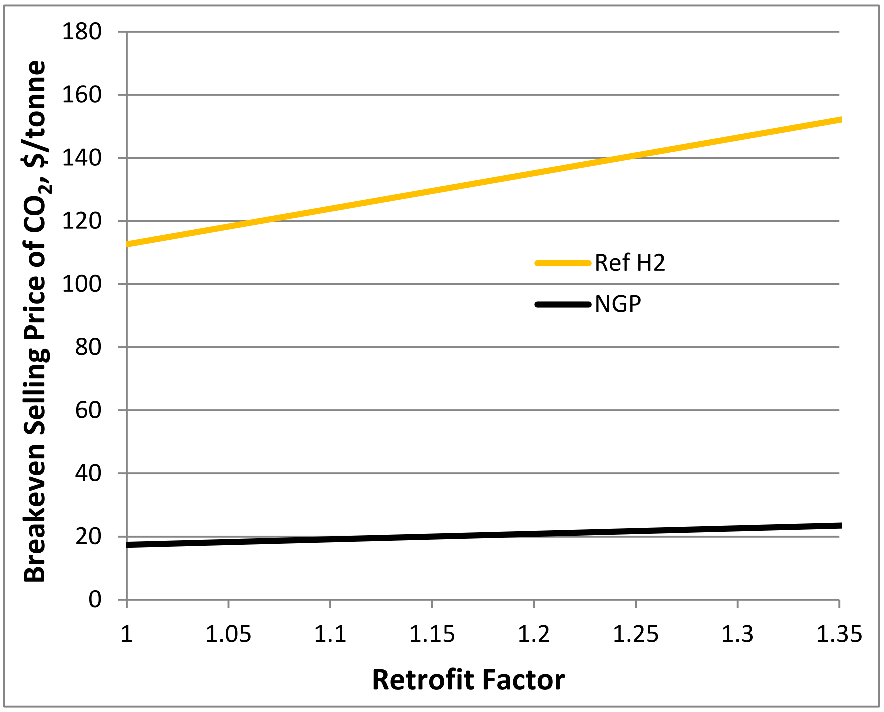
The results of this sensitivity show that change in the CCF can have a very large effect on the breakeven selling price calculated. In the high purity cases, the largest change over the range of 10-35 percent CCF is observed in the ethanol case, a change of $20.07/tonne CO_2. In the low purity cases, the effect is larger. This is expected as the low purity cases require more capital investment due to the requirement of capture equipment. The largest change in the low purity cases occurs for refinery hydrogen, where a $58.76/tonne CO_2 change in the breakeven selling price is observed. The range in CCF from 10% to 35% is equivalent to a change in the weighted average cost of capital (WACC) from around 8% to 20% assuming a 3 year capital expenditure period For a one year capital expenditure period, this range is equivalent to a change in WACC from 9% to 22%.

The CCF's used for the high purity and low purity cases, details of which have been given previously in Section 5.1, are representative of a Commercial Fuels project CCF. Other groupings include areas such as a standard CFPP, where the CCF depending on financial structure may be as low as 12.4 percent, or lower. As mentioned previously, different industries may have access to different costs of capital, and this can affect the CCF. But additionally, the maturity of a technology, specifically a capture technology like the MDEA unit employed in this study, may also affect the CCF. The CCF for the CFPP is low in part due the maturity of the system. As capture system's become more prevalent, and the learning curve is accelerated, lower CCF to those used in this study may become more reasonable, and the low end of the CCF sensitivity curve demonstrated here may become a more reasonable representation.

9.2.4 Retrofit Factor Sensitivity

As previously discussed in Section 5.2, a retrofit factor of 1.01 was applied to the TOC as a blanket retrofit cost increase for compression only cases, and a retrofit factor of 1.05 was applied to the TOC as a blanket retrofit cost increase for cases requiring compression and purification. Given the uncertainty surrounding the choice of retrofit factors, a sensitivity was performed over a retrofit factor range of 1 to 1.35. Shown in Exhibit 9-8 are the results of the sensitivity.

Exhibit 9-8 Retrofit factor sensitivity



Source: NETL

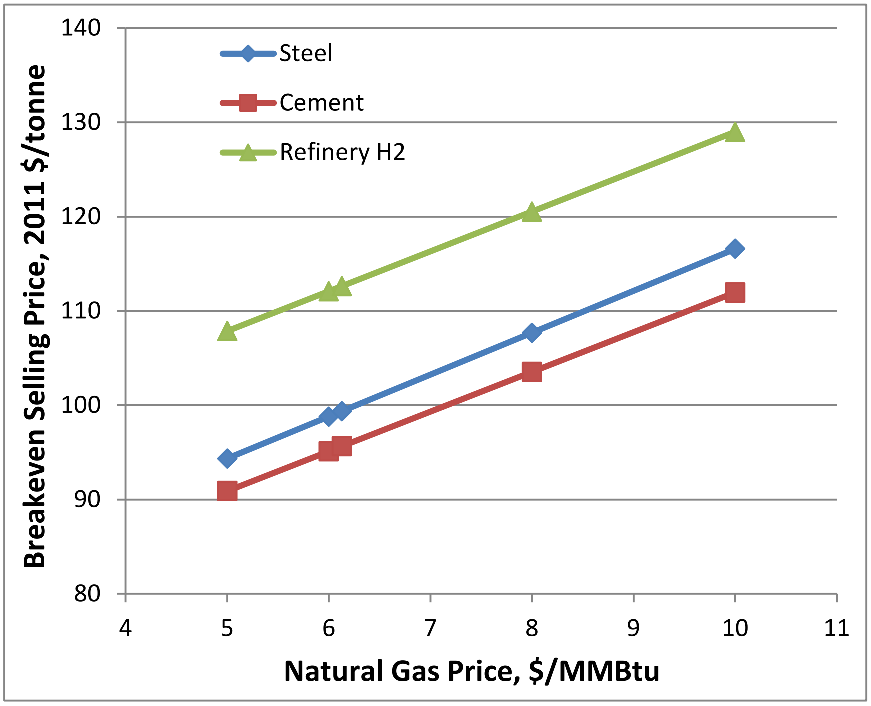
The two cases that were used for this sensitivity were refinery hydrogen, representing the highest Greenfield breakeven selling price of all cases, and natural gas processing, representing the lowest Greenfield breakeven selling price of cases with existing U.S. plants. Over the sensitivity range of 1 to 1.35, the rise in breakeven selling price is linear for both cases, with an increase for refinery hydrogen of $39.42/tonne, and an increase for natural gas processing of $6.08/tonne.

The result, as expected, shows for higher cost Greenfield cases, the retrofit factor will have a larger impact than for lower cost Greenfield cases.

9.2.5 Natural Gas Price

Natural gas is used in only the low purity cases in this study. In the low purity cases, purification of the CO_2 stream is required and accomplished using a MDEA AGR unit. The solvent is regenerated by steam, which is provided via a natural gas package boiler, as discussed in Section 6.3. Since the price of natural gas can have an effect on the overall breakeven selling price of CO_2 and a sensitivity was performed over the range of $5-10/MMBtu. The base natural gas price for each low purity case was assumed to be $6.13/MMBtu. The results of the sensitivity are shown in Exhibit 9-9.

Exhibit 9-9 Natural gas price sensitivity for low purity cases



Source: NETL

As expected, as the price of natural gas increases, the breakeven selling price increases. Over the $5-10/MMBtu range, the largest change is observed with steel at $22.24/tonne CO_2, followed by refinery hydrogen at $21.11/tonne CO_2, and finally cement with a change of $21.04/tonne CO_2.

10 Conclusion

Nine different industrial sources were examined: ethanol, ammonia, natural gas processing, ethylene oxide, CTL, GTL, refinery hydrogen, steel/iron, and, cement. Plant sizes were chosen based on different factors. These factors included the next representative plant size expected to be built in the industry (ammonia, refinery hydrogen), plant sizes representative of the majority of production for the industry (ethanol, steel/iron, ethylene oxide, cement), or plant sizes that would justify the addition of capture equipment (natural gas processing). Plant sizes for CTL and GTL were driven by previous NETL studies.

Both Greenfield and retrofit costs were determined. The retrofit costs were derived by application of a retrofit factor to the applicable capital equipment cost accounts. Engineering judgment was used to determine a representative retrofit factor.

The results of this study show that CTL gives the smallest required Greenfield breakeven selling price for the CO_2 product, a value of \$8.66/tonne. This is justifiable as CTL presents with a highly pure CO_2 source, as well as the most available CO_2 for capture. This combination of high availability coupled with high purity results in the lowest price, and is a function of CO_2 capture being required as part of CTL (and GTL) processes. The most costly option was refinery hydrogen, with a Greenfield cost of \$112.64/tonne. For refinery hydrogen, the low purity, although higher than CFPP flue gas concentrations, requires purification equipment to attain EOR pipeline standards. Along with purification issues, the amount of CO_2 available for capture is low, and these two factors coupled together result in the highest Greenfield breakeven price required for all of the nine cases examined here. The remaining cases fall in between the maximum and minimum cases as follows: GTL at \$9.29/tonne, natural gas processing at \$17.38/tonne, EO at \$24.28/tonne, ammonia at \$26.26/tonne, ethanol at \$30.15/tonne, cement at \$95.66/tonne, and finally steel/iron at \$99.36/tonne. The assumed CO_2 concentrations for GTL, natural gas processing, EO, ammonia, and ethanol were very pure, the same purity as the CTL case. The reason for the increasing prices given similar purity is the amount of CO_2 available for capture. In general, among the high purity cases, as the amount of CO_2 available for capture increases, the breakeven selling price decreases. The exception to this is EO, because EO does not require pre-cooling or post-cooling to meet EOR pipeline requirements, whereas some of the other high purity cases do require pre- and post-cooling.

Sensitivity analyses of CCF and COE show minimal change in the breakeven selling price 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, as the amount of CO_2 available for capture increases, the breakeven selling price decreases. The largest effect is observed with refinery hydrogen, where a change of \$109.50/tonne CO_2 was observed when plant size was varied over the range of 7,530 tonnes of H_2 production to 113,500 tonnes of H_2 production. The base case production was 59,000 tonnes of H_2.

This analysis shows that the ideal plant for this application has two specific characteristics; (1) high CO_2 purity so that further purification is not required; and (2) large amounts of CO_2 available. CO_2 purity, as expected, plays a large role in the calculation of a breakeven selling price; however, the amount of CO_2 and, therefore, the varying economies of scale from one industrial process to another, also play a large role.

11 Future Work

Future work in this area targeted at expanding the existing analysis to other industries should look to plants with the aforementioned characteristics of high purity and large supply. Potential recommendations include plants where CO_2 removal is crucial to the base plant process. A perfect example of this is ammonia and urea, 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.

11.1 In-depth Process Analysis

There are several opportunities where this study 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 requires specific assumptions as to how the base ammonia plant allocates CO_2. However, 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.

11.2 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 “break-even” cost for CO_2 capture and make some processes more feasible when combined with others.

11.3 Additional Processes

Methanol and a variety of other commodity chemical manufacturing facilities could be potential processes assuming appropriate feedstock to justify capture. Additionally, as mentioned in Section 8.1, the FCC unit at refineries is another viable point source for CO_2 capture. This may be investigated separately, or it could be included in the Multiple Process Scenario, where the FCC unit and the refinery hydrogen unit are combined to take advantages of economies of scale.

11.4 Technical/Economic Analysis of CO_2 Distribution to EOR Fields

As stated previously in Section 6.1, 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 breakeven selling price. A reduction in pressure would result in a lower compressor capital cost, as well as reduced power consumption and, therefore, 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 CO_2 breakeven cost.

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Appendix A

Given in this appendix is a more detailed description of how the CO_2 supply values were calculated for the exhibits found in the executive summary and economic results section.

In order to calculate the 'CO_2 Available' axis, significant and broad assumptions are necessary. For some cases, such as ammonia where only 24 plants are operational in the U.S., or ethylene oxide where only 10 plants are operational, CO_2 supply values may be more accurate. Whereas, for natural gas processing, the more than 490 processing plants, and widely differing CO_2 concentrations of the raw natural gas to be processed make this calculated supply value less reliable. Therefore, the assumptions used to project these numbers are detailed as follows:

Natural Gas Processing:

- The Gas Technology Institute (GTI) Gas Resource Database was used for this calculation. This database characterizes gas producing formations with annual production as well as average CO_2 composition. Several sources have stated that pipeline CO_2 concentrations are restricted to 2 percent CO_2 or less. (33) (25) Therefore, only formations producing gas above 2 percent CO_2 are considered. The average CO_2 fraction for formations above 2 percent CO_2 was multiplied by the annual formation production, giving 27.5 M tonnes CO_2. This value is used in Exhibit ES-2 above, and represents total potential CO_2 available. However, for low CO_2 content gas that does not meet pipeline specifications (2-5 percent CO_2 content), in certain situations, this gas is blended such that the total product meets pipeline specifications, and is, therefore, not processed in this way, reducing the amount of CO_2 potentially available in this calculation. When considering only gas at or above 5 percent CO_2 content, the total potential CO_2 supply is calculated as 20.7 M tonnes/year. This value is not considered in this study, but represents alternate assumptions about the natural gas industry. Also not considered is gas requiring processing due to higher than acceptable N_2 or H_2S fractions. However, processing to remove these species will not significantly impact the amount of CO_2 available.

Ethylene Oxide:

- The total United States (U.S.) ethylene oxide (EO) capacity in 2007 was determined to be 3.6 M tonnes spread among 10 plants. (32) Literature shows that CO_2 is produced during the EO process in a ratio of 6:2 EO:CO_2, giving a total amount of CO_2 produced of 1.2 M tonnes.

Ammonia:

- The total 2006 U.S. production was given by the U.S. Environmental Protection Agency (EPA) as 8.2 M tonnes/yr. Applying a factor of 1.87 tonne CO_2/tonne NH_3, which represents the total CO_2 produced from both point sources of an ammonia plant as given by Strait and Nagvekar, (4) the total CO_2 produced was determined to be 15.33 M tonnes/yr. When considering total U.S. consumption of ammonia (approximated by the U.S. Geological Survey (USGS) to be 14 M tonnes/yr in 2006), the CO_2 produced from that consumption is determined to be 26.2 M tonnes/yr. This would represent the scenario where U.S. imports of ammonia are stopped; instead all U.S. consumption is supported by U.S. production, and CO_2 available for capture from this industry is maximized. This scenario was not employed here. The study assumptions are that 66 percent of the total plant CO_2 produced comes from the CO_2 stripping vent, the only high purity source, and the balance is emitted as coal-fired power plant (CFPP) flue gas type concentrations. This gives a total CO_2 available in high purity form as 10.12 M tonnes CO_2/yr. For plants that also produce urea, (18 of 24 U.S. plants), 28 percent of the total CO_2 available is assumed to be used for urea production, and captured from the stripping vent. Applying this amount as unavailable, the total CO_2 available for capture was determined to be 5.8 M tonnes/yr.

Ethanol:

- Total March 2012 U.S. ethanol production was given as 13,894 M gal. (17) Using the equation given in the ethanol section 7.1.2 for determining CO_2 produced during the fermentation process, it was determined that the CO_2 available for capture totals 39.7 M tonnes/yr. This equation uses an emission factor of 6.31 lb CO_2/gal ethanol as given by the Illinois State Geological Survey, (19) which assumes corn is the main feedstock of the ethanol plant.

Steel/Iron

- Blown oxygen furnace (BOF) plant capacity was given by the American Iron and Steel Institute (48) as 41.2 M tonnes/yr in November 2010, with a utilization ratio of 73.8 percent, giving an actual 2010 steel production of 30.4 M tonnes from BOF plants. The literature shows that for this study's assumptions, 72.8 percent of the total CO_2 available from the steel plant comes from the assumed higher purity point sources. The emissions factor was given as 2.2 tonne CO_2/tonne steel for BOF processes, and applying these factors results in 48.7 M tonnes/yr CO_2 available for capture, from BOF processes only, from the higher purity CO_2 point sources.

Cement:

- In 2010 there were 100 cement plants in the U.S. utilizing both dry and wet processing kilns, and according to the USGS, these 100 plants produced 66.5 M tonnes of cement. (52) Cement production creates 1.2 tonnes CO_2/tonne cement (57), and applying this factor to the 2010 production gives 79.8 M tonnes CO_2 available. As will be discussed in section 8.3, cement production has seen a sharp decline in recent years due in large part to decreases in construction. The reported 2010 cement production is 29 M tonnes less than cement production in 2007. Projections show that cement production is expected to increase, and as it does, so too will the amount of CO_2 available for capture and reuse from this industry.

Refinery Hydrogen:

- Total U.S. hydrogen production was given as 9.1-10 M tonnes/yr (10-11 M tons/yr), (9.5 M tonnes/yr used), with refineries consuming 59 percent of U.S. production, giving 5.6 M tonnes H_2/yr used/produced by refineries. The CO_2 production factor was given as 9-12 tonnes CO_2/ton H_2, and the high end of 12 tonnes CO_2/tonne H_2 was used. This results in 67.5 M tonnes CO_2/yr available from refinery hydrogen production.

Coal-to-Liquids (CTL)/ Gas-to-Liquids (GTL):

- CTL and GTL are represented by single points on the y-axis, indicating that there are no existing U.S. CTL or GTL plants from which to draw a total amount of CO_2 available for capture. Sections 7.5 and 7.6 use previous National Energy Technology Laboratory (NETL) studies that assume a reference liquids output, resulting in specific amounts of CO_2 available so as to calculate a breakeven selling price, but no actual CTL or GTL plants are currently in operation in the U.S.

It should be noted that the individual values read on the x-axis do not indicate the amount available, but the total range for each industry on the x-axis represents the amount available. For example, the starting point for ammonia is 28.7 M tonnes CO_2/yr, but this does not represent the amount of CO_2 available. Rather, the ammonia range is 28.7 – 34.5, representing 5.8 M tonnes CO_2/yr available. It should also be noted that the breakeven selling price reported does not include transportation costs.

Appendix B

The carbon balance for the Ethanol case is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		**kg/hr(lb/hr)**
Fermentation Stream	4,455 (9,823)	CO_2 Captured Stream	4,455 (9,823)
		Convergence Tolerance	0 (0)
Total	**4,455 (9,823)**	**Total**	**4,455 (9,823)**

The carbon balance for the Ammonia case is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		**kg/hr(lb/hr)**
Stripping Vent	14,281 (31,485)	CO_2 Captured Stream	14,281 (31,485)
		Convergence Tolerance	0 (0)
Total	**14,281 (31,485)**	**Total**	**14,281 (31,485)**

The carbon balance for the Natural Gas Processing case is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		**kg/hr(lb/hr)**
Stripping Vent	20,266 (44,590)	CO_2 Captured Stream	20,226 (44,590)
		Convergence Tolerance	0 (0)
Total	**26,266 (44,590)**	**Total**	**26,226 (44,590)**

The carbon balance for the Ethylene Oxide case is shown below.

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)
		Convergence Tolerance	0 (0)
Total	**3,785 (8,345)**	**Total**	**3,785 (8,345)**

The carbon balance for the Coal-to-Liquids case is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		**kg/hr(lb/hr)**
Gasification AGR Unit	263,744 (581,461)	CO_2 Captured Stream	998,088 (2,200,423)
FT AGR Unit	734,344 (1,618,962)		
		Convergence Tolerance	0 (0)
Total	**998,088 (2,200,423)**	**Total**	**998,088 (2,200,423)**

The carbon balance for the Gas-to-Liquids case is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		**kg/hr(lb/hr)**
Stripping Vent	212,188 (467,794)	CO_2 Captured Stream	212,188 (467,794)
		Convergence Tolerance	0 (0)
Total	**212,188 (467,794)**	**Total**	**212,188 (467,794)**

The carbon balance for the Refinery Hydrogen case is shown below.

Carbon In	Carbon Out

	kg/hr(lb/hr)		kg/hr(lb/hr)
PSA Tail Gas Stream	8,981 (19,800)	CO_2 Captured Stream	8,532 (18,810)
		Tail Gas Recycle	449 (990)
		Convergence Tolerance	0 (0)
Total	**8,981 (19,800)**	**Total**	**8,981 (19,800)**

The carbon balance for the Steel case coke oven gas (COG)/blast furnace stove (BFS) stream is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		kg/hr(lb/hr)
COG Stream	35,002 (77,167)	COG/BFS CO_2 Captured Stream	35,996 (79,358)
BFS Stream	2,888 (6,368)	Recycle	1,895 (4,177)
		Convergence Tolerance	0 (0)
Total	**37,891 (83,535)**	**Total**	**37,891 (83,535)**

The carbon balance for the Steel case COG power plant stack (PPS) stream is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		kg/hr(lb/hr)
COG PPS Stream	90,346 (199,179)	COG PPS CO_2 Captured Stream	85,829 (189,220)
		Recycle	4,517 (9,959)
		Convergence Tolerance	0 (0)
Total	**90,346 (199,179)**	**Total**	**90,346 (199,179)**

The carbon balance for the Cement case is shown below.

Carbon In		Carbon Out	
	kg/hr(lb/hr)		kg/hr(lb/hr)
Kiln Off-gas Stream	37,408 (82,470)	Kiln EOR Stream	35,537 (78,347)
		Clean Flue Gas	1,870 (4,124)
		Convergence Tolerance	0 (0)
Total	**37,408 (82,470)**	**Total**	**37,408 (82,470)**

Appendix C

Shown in this appendix is a similar table of literature search results that appears in Section 3. Given the volume of sources used in this study, in text citation of the sources used to develop this table was not reasonable. Therefore, this appendix shows the literature search results table with full citations.

Process	Average Cost Curve ($/tonne CO_2 Avoided)	Average CO_2 Produced per Plant (kTonne/Yr)	Previous NETL Study	Total Estimated U.S. CO_2 Supply Potential (M tonne/year)
Ammonia	$18.20[1,2,3,4]	710[1,4,5,6]	No	15.3[24,25]
Coal to Liquids	$9.70[7]	1,445[7]	Yes[23]	*No U.S. Plants*
Ethanol	$57.90[8]	410[3,19,20,21]	No	69.8[26,27]
Ethylene Oxide	*N/A*	180[1]	No	8.8[1,28]
Gas to Liquids	*N/A*	*N/A*	Current	*No U.S. Plants*
Hydrogen Plant	$44.50[1,9]	600[9]	Yes[22]	100.0[9,18]
Iron and Steel	$158.90[2,3,4,8,12,13,14,15]	7,150[4,13,14,15,16,17]	No	105.1[13,29]
Natural Gas Processing	$20.90[3,4]	1,000[3,4]	No	*N/A*
Refinery Hydrogen	$74.80[1,2,8,18]	22,108[33]	No	28.6[9,18]
Methanol	$17.50[32]	*N/A*	No	*Feedstock Dependent*
Cement	$65.40[1,2,3,4,10,11]	355[4,10,11]	No	79.8[30,31]

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